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Probabilistic modeling of innovative clean coal technologies: Implications for technology evaluation and research planning. (Volumes I and II)

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PROBABILISTIC MODELING OF INNOVATIVE CLEAN COAL TECHNOLOGIES:

Implications for Technology Evaluation and Research Planning

VOLUME 1

A DISSERTATION

SUBMITTED IN PARTIAL FULFILLMENT OF THE REQUIREMENTS

for the degree

DOCTOR OF PHILOSOPHY

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by

Henry Christopher Frey

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THESIS

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Abstract

Computer modeling of innovative technologies in early phases of development can provide insights into the feasibility, optimal design, uncertainties, applications, and risks of the new process. Mass and energy balances characterizing the performance of a new process can be developed and refined based on on-going experimental work, providing the basis for iterative development of engineering models for technology assessment. Uncertainties in the performance of a new technology, based on limited test results, can be explicitly characterized using probabilistic modeling techniques such as Monte Carlo simulation or variants.

A probabilistic approach is described which allows the explicit and quantitative representation of the uncertainties inherent in innovative technologies. The method is applied to analyses of selected clean coal technologies. Probabilistic analyses provide insights into the uncertainties in process performance and cost not possible with conventional deterministic or sensitivity analysis. Applications of the probabilistic modeling framework are illustrated via analyses of the performance and cost of the fluidized bed copper oxide process, an advanced technology for the control of SO_2 and NO_x emissions from coal-fired power plants, and three integrated gasification combined cycle (IGCC) systems. An engineering model of a conceptual commercial-scale system for each technology provides the basis for the analysis. The models capture key interactions between process areas, as well as between performance and cost.

For each technology evaluated, uncertainties in performance and cost parameters of the engineering models were explicitly characterized using probability distributions. Estimates of uncertainty were based on literature review, data analysis, and elicitation of the expert judgment of process engineers involved in technology development. Typically, 20 to 50 model parameters were treated probabilistically in an analysis.

The engineering models were exercised in probabilistic modeling environments to characterize the uncertainties in key measures of process performance and cost. Probabilistic simulation considers the simultaneous interaction among all uncertain input variables. The resulting uncertainties in performance and cost provide an explicit, quantitative measure of the risk of either poor performance or high cost associated with innovative process technologies. Furthermore, using statistical or "probabilistic sensitivity" techniques, the key input uncertainties that drive uncertainty in performance and cost can be identified and prioritized. Thus, probabilistic analysis has direct implications for cost estimating, risk assessment, and research planning.

Competing technologies can be compared probabilistically to obtain explicit quantification of the probability that an advanced technology will have higher performance and lower cost than conventional technology. Additional research is assumed to reduce the uncertainty in key input parameters. Therefore, the expected pay-off from additional research can be evaluated using alternative assumptions regarding uncertainties. Engineering model results are used as inputs to decision models, to gain further insights regarding technology selection and research strategies.

Technology-specific case studies illustrate in detail how probabilistic modeling is used to characterize uncertainties, identify and prioritize key uncertainties, evaluate design trade-offs under uncertainty, and identify strategies for further research. In addition, for one IGCC system, the judgments of more than one expert were elicited for two major process areas. The judgments were used as inputs to separate case studies, and the results compared to identify the implications of the alternative judgments.

For most of the analyses considered here, the probabilistic approach is found to yield higher estimates of cost and lower estimates of plant performance than obtained from traditional deterministic approaches to technology evaluation. A key benefit from probabilistic analysis is the explicit characterization of skewed uncertainties in innovative technologies, which are a key source of cost growth often overlooked.

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In memory of my grandfather, Lawrence W. Sagle.

A self-taught man of arts and letters who valued education.

He remains an inspiration to his family and those who knew him.

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NOMENCLATURE

English Letter Symbols

A	=	Heat transfer surface area (ft^2)
A_L	=	Area of land required (acres)
AF	=	Allowance for funds during construction (fraction)
ALR	=	Average labor rate ($/hour)
C_{elec}	=	Cost of electricity produced by plant (mills/kWh)
c_f	=	Plant annual capacity factor (fraction)
C_i	=	Cost of item i ($)
$C_{PC,i}$	=	Process contingency cost for plant section i ($)
$C_{PJ,i}$	=	Project contingency cost for plant section i ($)
$CAT_{i,j}$	=	Initial catalyst requirement for chemical i in plant section j (mass units)
$CHEM_{i,j}$	=	Initial chemical requirement for chemical i in plant section j (mass units)
CTP	=	Coal throughput per gasifier (lb/hr/gasifier)
d	=	Diameter (ft)
d_{min}	=	Minimum diameter (ft)
D_b	=	Sorbent bulk density (lb/ft^3)
DC_j	=	Direct capital cost of plant section j ($1,000 January 1989)
e_a	=	annual escalation rate for plant equipment (fraction)
E(x)	=	Expected (mean) value of variable x.
f_{ash}	=	Fraction of ash in coal (fraction)
f_{att}	=	Zinc ferrite attrition rate per 80 cycles (fraction)
f_{CaCO3}	=	Purity of limestone (fraction)
f_{cr}	=	Capital recovery factor (fraction)
f_{EHO}	=	Engineering and home office cost factor (fraction)
f_{HS}	=	Volume fraction of hydrogen sulfide in syngas (fraction)
f_{ICC}	=	Indirect capital cost factor (fraction)
$f_{M,i}$	=	Maintenance cost factor for plant section i (fraction)
$f_{PC,i}$	=	Process contingency for plant section i (fraction)
f_{PJ}	=	Project contingency (fraction)
f_s	=	Fraction of sulfur in coal (fraction)
f_{SO2}	=	Fraction of off-gas volume flow rate that is sulfur dioxide
f_{vclf}	=	Variable cost levelization factor (ratio)
FOC	=	Fixed operating cost ($/year)
i	=	Interest cost for spent funds (fraction)
I_{PCI}	=	*Chemical Engineering* plant cost index
IC	=	Inventory capital ($1,000)
IDC_i	=	Indirect capital cost of plant section i ($1,000)
INT(x)	=	Nearest integer value of x

H = Height (ft)
HV_{coal} = Heating value of coal (Btu/lb)
IC = Inventory capital ($1,000)
L_S = Sulfur loading in sorbent (weight fraction)
$m_{i,j,k}$ = Mass flow rate of species i at the plant section j inlet or outlet k, (lb/hr in all cases except for coal, where units are tons/day)
$M_{i,j,k}$ = Molar flow rate of species i at the plant section j inlet or outlet k (lbmole/hr)
MW_j = Electrical output of plant section j (megawatts)
n = Number of data points used in a regression analysis (integer)
N = Construction period (years)
$N_{o,j}$ = Number of operating trains of plant section j (integer)
$N_{s,j}$ = Number of spare trains of plant section j (integer)
$N_{T,j}$ = Number of total trains of plant section j (integer)
OC_i = Operating cost for category i ($/year)
$P_{j,k}$ = Pressure at the plant section j inlet or outlet k (psia)
PP_i = Preproduction cost of category i ($1,000)
PPC = Total preproduction cost ($1,000)
q = heat flux (Btu/hr)
Q_{coal} = Energy flow of coal (MMBtu/hr)
$R_{Ca/S}$ = Calcium-to-sulfur molar ratio (lbmole Ca/lbmole S)
R^2 = Coefficient of Determination (decimal)
r_{tax} = sales tax (fraction)
$R_{Ca/S}$ = Molar ratio of calcium to sulfur (lbmole Ca/lbmole S)
s = sample standard deviation estimated from a data set
S_c = Fresh sorbent charge (lb)
S_s = Spent sorbent charge (lb)
SF = Shift factor (labor shifts/day)
t_a = Absorber cycle time (hours)
$T_{j,k}$ = Temperature at the plant section j inlet or outlet k (^oF)
TCR = Total capital requirement ($1,000)
TDC = Total direct cost ($1,000)
TIC = Total indirect cost ($1,000)
TPC = Total process capital ($1,000)
TPI = Total plant investment ($1,000)
U = Univeral heat transfer coefficient (Btu/(ft^2-hr-^oF)
UC_i = Unit cost of item i ($/mass unit)
V_c = Volume of sorbent charge (ft^3)
$V_{i,j,k}$ = Volume flow rate of species i at plant section j inlet or outlet k (acfm)
V_s = Superficial velocity (ft/s)
Var(x) = Variance of variable x
VOC = Total variable operating cost ($/year)
$W_{e,j}$ = Electricity requirement for plant section j (kW)

$W_{s,j}$ = Shaft work for plant section j (kW)

Greek Letter Symbols

η_{HS} = Hydrogen sulfide removal efficiency (fraction)

η_{Ox} = Fraction of oxidant that is oxygen, molar basis (fraction)

ΔT_{LM} = Log mean temperature difference (^oF)

σ = population standard deviation

Subscripts

a = ambient
AS = Administrative and support labor
EHO = engineering and home office costs
EP = environmental permitting
f = fresh
FC = fixed operating cost
Fuel = fuel
i = inlet
IC&C = initial catalyst and chemicals
ICC = indirect capital cost
M = maintenance
ML = maintenance labor
MM = maintenance materials
OC = variable operating cost
PC = process contingency
PJ = project contingency
o = outlet
s = spent

Species

ads = adsorbent
ash = ash
bent = bentonite
byp = byproduct (either sulfur or sulfuric acid)
cat = catalyst
cf = coal feed
chem = chemical
ci = corrosion inhibitor
cm = coal feed, moisture and ash free basis
coke = coke
cons = consumables

Cl = chlorine
COS = carbonyl sulfide
CW = cooling water
fines = fines
fo = fuel oil
hps = high pressure steam
hy = hydrazine
HS = hydrogen sulfide
lime = lime
L = limestone
LPG = liquified petroleum gas
mo = morpholine
O = oxygen
Ox = oxidant
pw = polished water
rw = raw water
s = sulfur
sa = sulfuric acid
sbd = scrubber blowdown
sh = sodium hydroxide
so = soda ash
sp = sodium phosphate
ss = Selexol solvent
ssy = saturated syngas
su = surfactant
syn = syngas
ww = waste water
zf = zinc ferrite

Equipment/Plant Sections

BF = Boiler Feed Water System
BS = Beavon-Stretford Tailgas Treating System
C = Claus Plant Sulfur Recovery Plant
CH = Coal Handling
CM = Air Boost Compressor
CW = Cooling Water
CY = High Temperature, High Pressure Cyclones
G = Gasification, High Temperature Gas Cooling, Ash Removal, and Particulate Scrubbing
GF = General Facilities
GT = Gas Turbine
HR = Heat Recovery Steam Generator
L = Limestone Handling
LT = Low Temperature Gas Cooling and Fuel Gas Saturation
OF = Oxidant Feed

PC	=	Process Condensate Treatment System
PR	=	Air Boost Compressor Precooler
S	=	Selexol Acid Gas Removal
SA	=	Sulfuric Acid Plant
SC	=	SCOT Tail Gas Treating System
SF	=	Sulfation
SS	=	Sub-Systems for Lurgi Gasifier
ST	=	Steam Turbine
SU	=	Sulfation
ZF	=	Zinc Ferrite

1.0 INTRODUCTION

The purpose of research, development, and demonstration (RD&D) is to provide and improve information regarding the feasibility, promising applications, optimal designs, uncertainties, and risks associated with a new process technology. The information generated from research can be used by process developers to refine the technology and ultimately by potential process adopters to make a decision about whether, and under what circumstances, to use the new technology. Of concern to a process developer is the selection of appropriate technologies for research and the prioritization of research needs.

Particularly in process engineering fields, proper research planning is often hindered by the unreliability of performance and cost estimates prepared during early stages of technology development. According to Hess and Myers (1989):

> Ultimately, all advanced technology R&D programs directed at the marketplace, be they public or private efforts, must be guided in very substantial part by the cost of the product of advanced technology relative to the current technology. Unfortunately, accurate assessment of the costs of advanced technologies has always been one of the most difficult and uncertain tasks facing an R&D planner. [p.1]

Preliminary performance and cost estimates for a new technology are inherently uncertain because of the lack of large scale experience required to verify expectations. In spite of this, these estimates are often presented as deterministic point-values without regard to their degree of confidence. Poorly informed decisions regarding research planning and technology adoption may result, at considerable cost in terms of wasted resources devoted to projects that, given a more complete characterization of known information, might not have been pursued. An important aspect of any RD&D program should be a systematic method for identifying and prioritizing research activities, allocating funds to RD&D, and maximizing the probability of success for an RD&D program.

This research addresses issues related to research planning for innovative technologies which are in an early stage of research or development. Explicit characterization of uncertainty in process performance and cost is postulated as a key feature of a robust research planning method. A number of questions motivate such an approach to research planning, including:

- What is the expected commercial performance of the new technology based on what is known from small scale tests and mass and energy balances?
- How reliable are these performance and estimates for a mature, commercial plant?
- How do variations in design affect cost?

- What are the key factors driving uncertainty in process performance and cost?
- What are the risks and pay-offs of the new technology vis-a-vis conventional technology?
- What are the potential market niches for the new technology?
- What are the expected results from further process RD&D?
- How much does RD&D cost?
- Is it worth it?

In this research, a systematic quantitative method is developed and applied to help answer these questions. The key features of the research planning method are:

- Selection of candidate technologies for evaluation;
- Development of engineering performance and cost models of the technologies to be evaluated, based on available performance and cost information;
- A probabilistic modeling capability to incorporate uncertainties about performance and cost parameters;
- Elicitation or development of technical judgments regarding performance and cost parameter uncertainties,
- Exercising of the models to answer these questions:
    - What are the key process design trade-offs?
    - What are the uncertainties that most affect overall costs?
    - What are the potential pay-offs and risks vis-a-vis conventional technology?
    - What is the likely effect and value of additional research?
    - What are the likely real costs of a first-of-a-kind demonstration of the new technology?
- Decision analysis regarding:
    - How policy-based objectives and decision maker's preferences influence selection of the optimal technology for further RD&D; and
    - How policy-based objectives and decision maker's preference influence planning RD&D strategies for a given technology.

The research planning method is applied to case studies of several innovative clean coal technologies for electric power generation. These case studies are intended to demonstrate the approach and to yield technology-specific conclusions regarding research strategies and potential application niches.

1.1 Modeling Innovative Technologies

Shortcomings in traditional approaches to predicting the performance and cost of innovative process technologies are a key motivation for this research. These shortcomings revolve around the incomplete characterization of the limitations of data and assumptions in performance and cost parameters. In addition, an incomplete scope of modeling of new technologies may fail to reveal important process interactions that affect technical and economic feasibility.

1.1.1 Decisions During RD&D

An innovative technology is a concept which departs in some fundamental way from existing technology and which holds the promise of a significant improvement in performance and/or cost over conventional technology. The transformation of an innovative concept into a commercialized technology involves many decisions at various stages of development, as shown in Figure 1. Figure 1 is based on a discussion in Merrow et al (1981). Typically, a new concept may be evaluated theoretically and tested at a small (e.g., bench-top) scale. If promising technical results are obtained, a preliminary cost estimate of a commercial scale design may be made. If the cost estimates are high, the project may be dropped or research may continue to identify less costly variants of the technology. If the costs are promising, research is likely to continue to a larger scale test and to a development phase. Several pilot plants, of varying size and design, may be built during the research and development phase. As confidence in the technology improves, a more definitive cost estimate may be commissioned from an outside group as a final screening prior to designing a full-scale demonstration plant. At this stage, there may still be significant uncertainties in cost and performance that only a full-scale demonstration plant would help resolve.

The types of decisions made during RD&D include whether the new technology should be rejected as infeasible, identification and prioritization of technical uncertainties for focused research, and identification of improvements that can be made to optimize the process. In practice, decisions about research prioritization are often made based on incomplete consideration of potential interactions between the new technology and its operating environment. For example, in identifying research priorities for a new emission control technology, research planners may fail to consider potential interactions between the new technology and the power plant.

1.1.2 Modeling Performance and Cost

Predictions of the performance of commercial implementations of an innovative concept may involve several challenges. In the very early stages of process development, predictions may be based on limited experimental work and rely heavily on simple mass and energy balances. These estimates may tend to assume ideal conditions and to overlook potential limitations, such as reaction kinetics or energy losses in the system. As a concept proceeds to a small scale testing or pilot plant phase, laboratory data may become available to help identify more realistic values for key process parameters. However, uncertainties in interpretation of test data stem from: (1) statistical errors in the data; (2) differences in

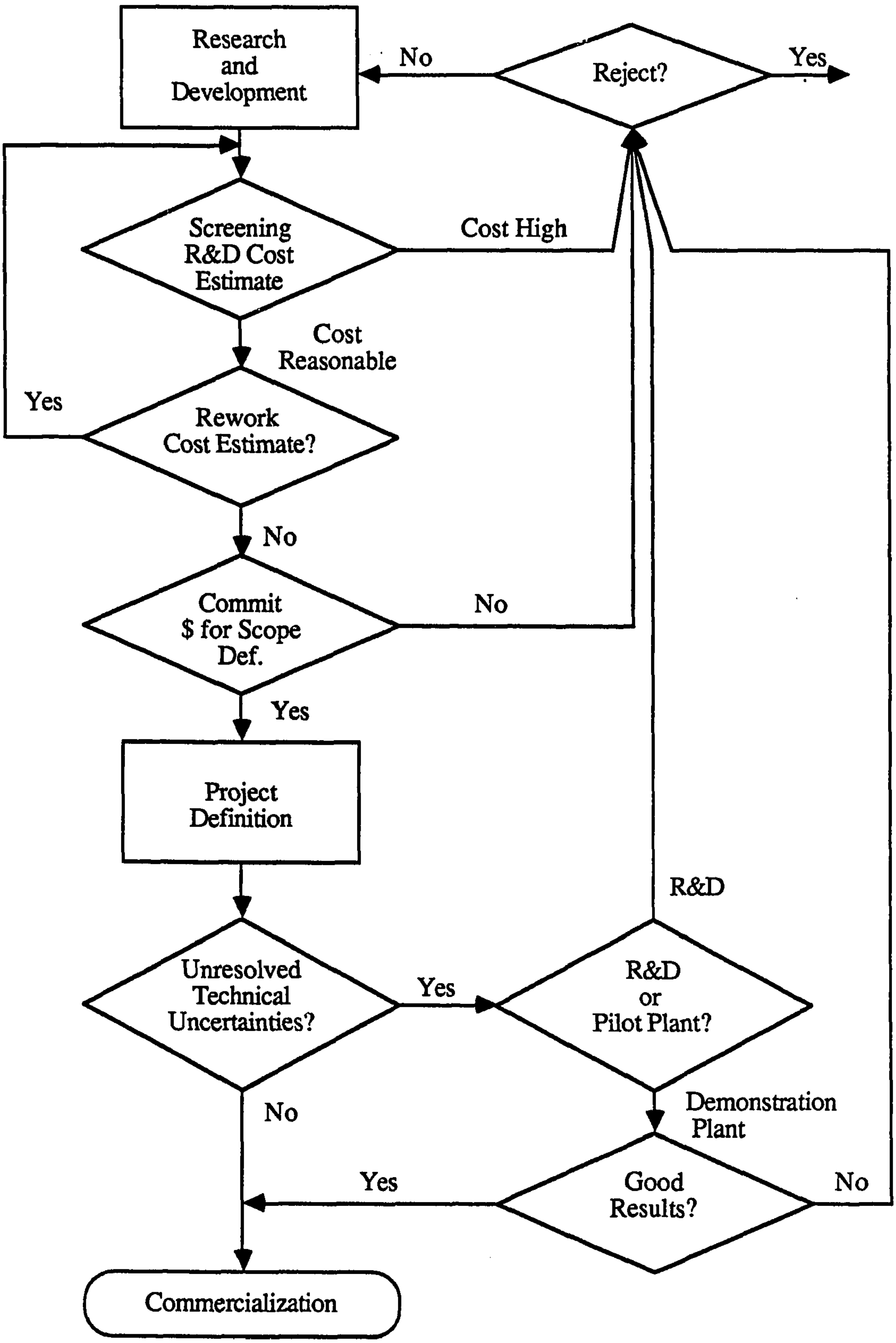

Figure 1. Decisions in the Research, Development, and Demonstration Process

configuration between the PDU and a commercial-scale plant, and (3) potential problems in scale-up from the PDU to commercial-size equipment.

Uncertainties in key performance parameters of an innovative process typically result in uncertainty in predicting key performance or environmental characteristics of the technology, such as plant efficiency and emissions of pollutants. Uncertainties in system performance (e.g, flowrates, pressures) lead to uncertainties in the required size of process equipment and the consumption of materials (e.g., sorbent) and parasitic power. This results in uncertainties in capital and operating costs, which, along with key performance characteristics, are the ultimate measures of interest for comparative analysis. Furthermore, even if process performance were known with certainty, uncertainties regarding the costs of equipment (particularly equipment not previously used in commercial scale service) and reagents remain. For example, preliminary cost estimates may not capture all of the costs that would be revealed by a final estimate based on more detailed engineering analysis. Therefore, the process area costs developed in conceptual design studies may tend to be underestimated. Potential problems that could be encountered in a commercial-scale plant, such as corrosion or fouling, also may not be anticipated. Hence, performance and cost estimates developed in early stages of technology development could prove incorrect.

The Rand Corporation has performed a number of studies regarding problems with estimating the performance and cost of first-of-a-kind innovative process plants. These studies include: evaluation of cost growth and performance shortfalls for the first-of-a-kind (demonstration) plant (Merrow, Phillips, and Myers, 1981); evaluation of the potential benefits of building a first-of-a-kind plant (Hess, 1985); an assessment of problems and R&D requirements for technology that involves processing of solids (E.W. Merrow, 1986); evaluation of factors involved in construction schedule slippage and increased startup costs for first-of-a-kind plant (Myers and Shangraw, 1986); an evaluation of industry's approach to developing contingency factors for cost estimates (Milanese, 1987); and an evaluation of cost estimating methods used for evaluating coal-to-substitute natural gas (SNG) systems (Hess and Myers, 1989).

Typical of the findings of the Rand are: (1) bias and uncertainty in performance and cost estimates results from low levels of process and project understanding, particularly for new technologies; (2) cost-underestimation of new technologies is widespread and systematically related to low levels of project definition and the amount of unproven technology employed; and (3) performance over-estimation is associated with unproven technology in a process concept (Merrow et al, 1981).

1.1.3 Handling Uncertainties

Nearly all analyses of energy and environmental control technologies that are still in the research phase involve uncertainties. In developing performance and cost estimates of technologies that are in early stages of development, the most common approach is for engineers to assume a "best guess" point-value judgment for key parameters. These judgments may be intended to represent neither undue optimism or pessimism regarding the technology, or they may be intended to incorporate a degree of conservatism. However, the basis for many assumptions, and the scope of thought that went into them, are often not explicitly documented in conceptual design studies. Thus, the degree of confidence that a decision-maker should place in the performance and cost estimate is often not rigorously considered.

The most common approach to handling uncertainties is either to ignore them or to use simple "sensitivity" analysis. In sensitivity analysis, the value of one or a few model input parameters are varied, usually from "low" to "high" values, and the effect on a model output parameter is observed. Meanwhile, all other model parameters are held at their "nominal" values. In practical problems with many input variables which may be uncertain, the combinatorial explosion of possible sensitivity scenarios (e.g., one variable "high", another "low," and so on) becomes unmanageable. Furthermore, sensitivity analysis provides no insight into the *likelihood* of obtaining any particular result. Thus, while they may indicate that a range of possible values may be obtained, sensitivity results do not provide any explicit indication of how a decision-maker should weigh each possible outcome.

A specific approach to handling uncertainty in capital cost estimates, whether for a new or existing technology or for a preliminary or detailed cost estimate, employs "contingency factors." The contingency often is the single largest expense in the cost estimate, and yet it is also the least documented. A contingency is used to represent additional costs that are likely to occur, but that are not included explicitly in the cost estimate (Milanese, 1987).

Generally, all capital cost methods involve estimating one or more contingency factors. In the electric power industry, perhaps the most widely used cost estimating method for research planning is that of the Electric Power Research Institute (EPRI). EPRI uses two types of contingency factors: project and process contingency (EPRI, 1986). The project contingency is intended to cover the costs of additional equipment or other costs that would result from a more detailed design of a definitive project at a specific site. This implies that as costing proceeds from a preliminary to a detailed final estimate, the project contingency factor should be reduced. The process contingency is intended to quantify the uncertainty in the technical performance and commercial scale cost of a new technology. Uncertainties in performance are implicitly assumed only to affect equipment design and not to affect the overall performance characteristics of the technology. This contingency factor is reduced as a technology proceeds from bench scale to full commercial use. Both of these contingency factors are deterministic estimates of additional costs that are expected to occur. In the EPRI Technical Assessment Guide, there is little substantive discussion of how these factors should be derived; suggestions for selecting values of both the project and process contingency values appear to be merely "rule-of-thumb" recommendations.

The Gas Research Institute (GRI) also sponsors studies of clean coal technologies. GRI requires an estimate of "process development allowances" (PDA) for all major plant sections (GRI, 1983). The PDA is intended to account for increases in cost as the design definition of a new technology is increased, revealing additional equipment required for a commercial-scale plant, and as a technology proceeds from early stages of development through commercialization. The PDA is an average of assessments, on a percentage of direct cost basis, of expected cost increases based on the state of the technology, the availability of experimental data, assumptions in the performance and cost estimate that have not been tested, and expected difficulty of control and operation. For each of these categories, the GRI Guidelines suggest a number of areas to consider when developing the assessments. The PDA is similar to the EPRI definition of process contingency.

Contingency factors are only applied to capital cost estimates. Analogous factors are not used for annual (fixed and variable operating cost) estimates, nor are they used explicitly in developing performance estimates.

The contingency factor approaches used by EPRI and others have not been validated by actual data. The Rand Corporation conducted a survey of 18 companies in the chemical and petroleum industries to determine the actual methods used to develop contingency factors (Milanese, 1987). The study indicates that contingency factors are badly under-estimated, which may be leading to bad decisions about certain projects. Factors such as project definition, owner characteristics, nature of the company (oil vs. non-oil), state of the technology (innovation or proven), project characteristics, management characteristics, and who actually estimated the contingency (e.g., project engineer or management) where considered in the study. The results were that what little theory exits regarding contingency factors is not applied, and that some of the factors which seem intuitively to be important are not captured in the contingency factor. Rand recommends the greater and more formalized use of experience, the use of a "delphi" technique to get multiple expert inputs, and the inclusion of costs associated with risks and innovation.

Although some conceptual design studies prepared for the Electric Power Research Institute (EPRI) have included a "risk analysis" involving probabilistic simulation, such as a recent study of an innovative clean coal technology by Heager and Heavan (1990), the specification of uncertainties has been only on cost-related parameters. Furthermore, the analysis of uncertainty has been confined only to capital costs, and most analyses are insufficiently documented to allow critical evaluation of the modeling results.

1.2 Innovative Coal-Based Power Generation Technology

Because of current environmental, economic, security, and political concerns, the U.S. government and others are becoming extensively involved in research, development, and demonstration (RD&D) of so-called "clean coal technologies." These are coal-based energy conversion technologies in which emissions of potentially harmful pollutants (gas, liquid, and solids) are reduced compared to commercially available technology. Improvements in plant efficiency and reductions in plant cost are also being sought. A clean coal technology would be termed "innovative" if, compared to conventional technology, it held the promise of significant improvements in several of the following ways: (1) reduced in environmental discharges (air, liquid, or solid); (2) improved plant efficiency; (3) reduced plant costs; and (4) improvement flexibility in terms of construction and plant operation.

The largest consumer of coal in the United States is the electric utility sector. In 1985, the utility sector used 15.5 quads (quadrillion Btu) of the total of 18.2 quads of coal used in the U.S. (EIA, 1987). Emissions of sulfur dioxide from coal-burning power plants were estimated to be 15.6 million tons in 1985, and the emissions of nitrogen oxides were approximately 6 million tons. Total emissions of these pollutants in the U.S. are approximately 23 and 20 million tons for sulfur dioxide and nitrogen oxides, respectively (DOE, 1987). Both of these pollutants are chemical precursors to acid rain, which has been postulated to cause a variety of impacts in in many areas of the U.S. and Canada. These impacts include: (1) acidic and low pH lakes and streams, with resultant stress on aquatic life; (2) possible effects on forests; (3) contribution to physical damage of cultural and construction materials; (4) reduced visibility due to sulfate and other aerosol species;

(5) effects on human health; and (6) economic effects resulting from the previously listed impacts.

Although the environmental impacts of coal combustion pose significant challenges to emission control design, coal is the most abundant of the U.S. domestic fossil fuel resources. The demonstrated coal reserve base could supply current needs for 260 years (EIA, 1987).

1.2.1 Need for New Power Plants

While demand for electric power in the U.S. has been increasing at an average rate of about three percent annually in recent years, new capacity has been brought on line at the rate of only about one percent per year. Thus, in the coming years large capacity additions will be required to meet load growth and to replace retiring plants. In the 1990's, it is expected that 100,000 to 300,000 MW of new capacity will be required. Whiie some of this capacity will be supplied by independent power producers who sell excess electricity to utilities, electric utilities will have to build most of the new capacity themselves. In the short-term, natural gas is expected to be available at attractive prices; thus, a significant portion of utility capacity addition may take the form of natural gas-fired gas turbine installations. General Electric reportedly predicts that 80 percent of new capacity through the next decade will be based on gas turbines. However, 90 percent of new capacity additions already on order, representing between 10 to 30 percent of the capacity needed for the next decade, are coal-fired steam turbine-generator systems (Smock, 1990).

While increasing concern over the environmental impact of power plants will make siting and permitting of these facilities more difficult, the use of indigenous and abundant coal may be economically less risky than natural gas as a long term utility (or independent power producer) fuel. Therefore, it is important to identify and promote development of coal-fueled power plant technologies that are both environmentally acceptable and economically attractive.

1.2.2 Current Environmental Regulations

Current U.S. Environmental Protection Agency (EPA) new source performance standards (NSPS) applicable to coal-fired power plants require up to 90 percent sulfur dioxide (SO_2) removal, over 99 percent particulate matter (PM) removal, and moderate (about 50 percent) reduction of nitrogen oxides (NO_x) emissions. A conventional emission control system for a new pulverized coal (PC) power plant typically consists of a wet limestone flue gas desulfurization (FGD) system for SO_2 control, an electrostatic precipitator (ESP) for PM removal, and combustion controls for NO_x reduction. These systems are all commercially available and well-demonstrated. However, recent commercial experience in Japan and Germany with selective catalytic reduction (SCR) indicates that 80 to 90 percent NO_x removal may be feasible, although SCR has not yet been applied with U.S. coals (Cichanowicz and Offen, 1987; Damon and Giovanni, 1987). For some types of coal-fueled power plant systems which are not yet commercialized, such as integrated gasification combined cycle (IGCC) systems, an explicit NSPS does not yet exist (Simbeck et al, 1983).

Specific emissions sources may also be subject to permitting under Prevention of Significant Deterioration (PSD) rules, which are intended to preserve air quality based on ambient atmospheric concentration standards. Such permits are issued on a case-by-case basis and are often more stringent than NSPS. In extreme cases, a new source may be required to obtain emission offsets from other facilities before it can be permitted.

Conventional electric power plants are required to comply with EPA Effluent Guidelines and Standards for liquid discharges, including wastewater, cooling tower blowdown, boiler blowdown, ash transport water, process condensate, and purge water (Bechtel, 1988a). However, in many cases, a more stringent permit may be issued on a case-by-case basis under the National Pollutant Discharge Elimination System (NPDES) (Simbeck et al, 1983). Typically, water treatment systems are required in order to comply with either of these standards.

Solid wastes from coal-fueled power plants are regulated under the Resource Conservation and Recovery Act (RCRA). Power plant solid wastes are usually nonhazardous, and as such are disposed of in accordance with the nonhazardous material guidelines of RCRA.

1.2.3 Changes in Environmental Regulations

The most recent major revision in environmental control strategy in the U.S. is the Clean Air Act Amendment (CAAA) signed by the President on November 15, 1990. Prior to 1995, 110 of the largest SO_2-emitting stations are targeted for specific emission reductions. By 2000, the CAAA requires a reduction in national SO_2 emissions by 10 million tons/year compared to 1980 levels. After 2000, a nationwide SO_2 emission cap of 8.9 million tons/year will be in effect. In addition, emissions from virtually all power plants larger than 75-MW will be required not to exceed 1.2 lb SO_2 per million BTU (lb/MMBtu) of coal consumed. In its full implementation, the CAAA is a market-based approach to emission control, unlike the "command and control" NSPS regulations now in effect. Under the market-based system, each emitter must possess an emission allowance for each ton of SO_2 emitted annually. In principle, emitters are free to buy, sell, and bank emission credits to meet their needs and to comply with the national emission cap at lowest cost (Leone, 1990).

Under the CAAA, each emitter faces economic incentive to reduce emissions to the point where the marginal cost of pollution control equals the cost of an emission credit. Thus, technologies which can economically achieve high removal efficiencies can provide a direct financial benefit to the utility. Thus, the CAAA may promote more rapid innovation in clean coal technology.

The CAAA also calls for a 2 million ton/year reduction in national NO_x emissions by 2000 (Lee, 1991). EPA is required to set command-and-control NO_x standards for tangentially-fired and dry-bottom wall-fired boilers, as well as for other boilers specified in the amendment. EPA must also set new NSPS for other fossil-fueled units. Trading between SO_2 and NO_x is not part of the current CAAA, but may be studied later for possible inclusion in a future amendment (Leone, 1990).

1.2.4 Emerging Coal-Based Power Generation Technologies

With the prospect of increasingly stringent emission control has evolved the concept of integrated environmental control. This concept is illustrated in Figure 2. The concept has several dimensions. One is to consider interactions among control methods for air, water, and solid waste emissions, so that reductions in one type of discharge do not unduly increase others. Another is the integrated use of pre-combustion, combustion, and post-combustion control methods (as distinct from one approach alone). A third dimension is the development of new processes for combined pollutant removal in lieu of separate processes for individual pollutants. Other process innovations not directly related to emission control may also affect emissions. Thus, integrated environmental control represents good design practice and provides opportunities to minimize costs for a given set of emission reduction requirements (Carr, 1986).

Key objectives of emissions control research, embodied in the notion of integrated environmental control, have been system simplification and cost reduction. Examples of integrated concepts for pulverized coal-fired power plants include combining flue gas SO_2 and NO_x removal in a single reactor vessel, and coupling the designs of the power plant and emission control systems.

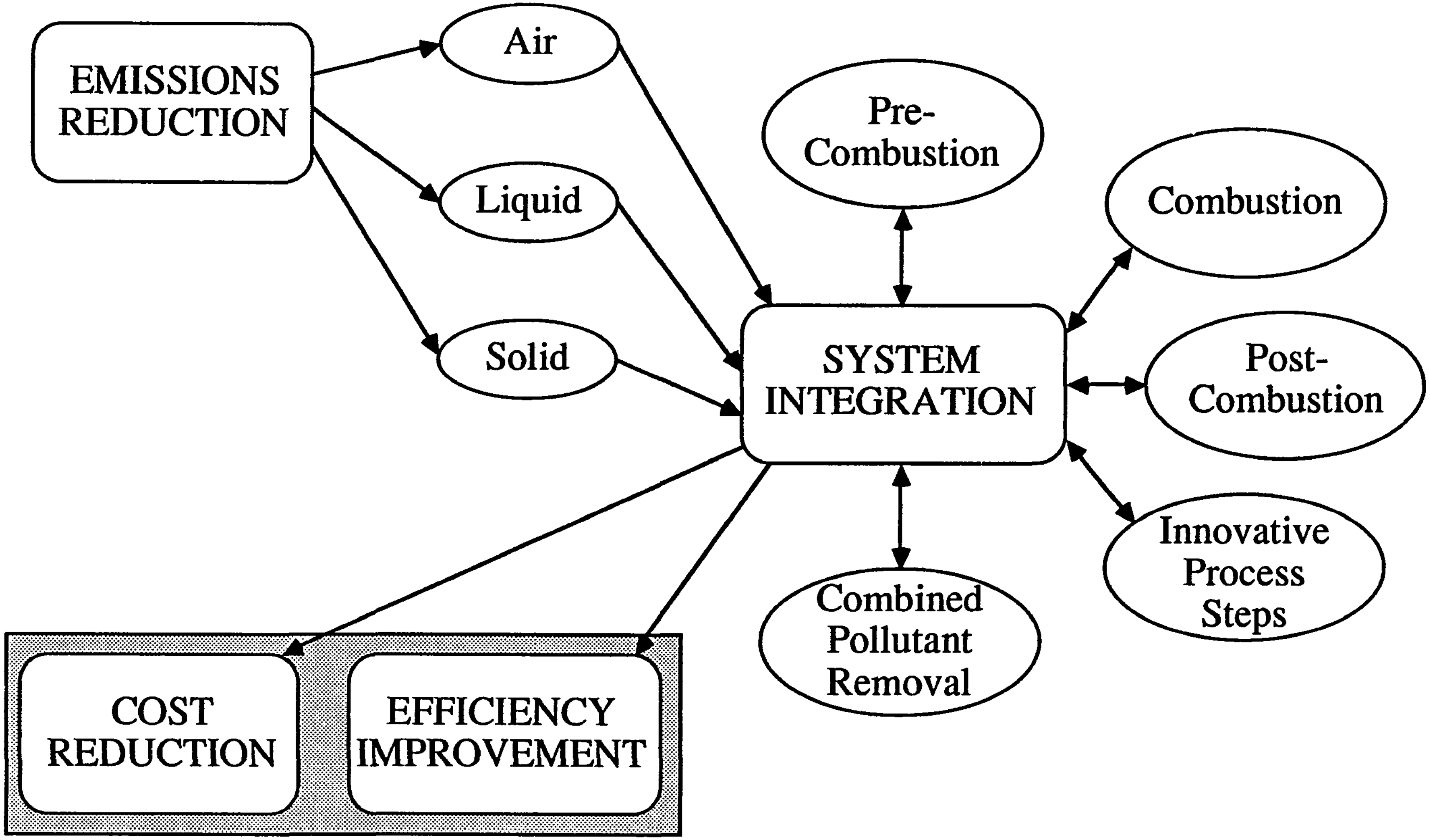

Figure 2. The Concept of Integrated Environmental Control

DOE and others have supported development of more advanced alternatives for control of SO_2 and NO_x emissions from coal-fueled power plants. One alternative, integrated gasification combined cycle (IGCC), represents a new approach for the clean and efficient use of coal in electric power generation. As emission control requirements have increased, so has the cost of conventional PC power plants, while their thermal efficiency has decreased, due to power requirements for emission control systems (DOE, 1987). Natural gas- and oil-fired systems based on gas turbine combined cycle technology have high efficiencies, but consume expensive premium fuels. In a combined cycle plant, fuel is burned in a gas turbine, and the hot exhaust gas is used to generate steam for a steam cycle. Electric generators on both the gas turbine and steam turbine generate electricity. By substituting synthetic fuel gas derived from coal for natural gas or oil, a coal-fueled gas turbine combined cycle power plant results. By integrating the steam cycle to generate steam from the high temperature coal gasification process, the overall thermal efficiency can be optimized. Advantages of IGCC over PC plants include higher thermal efficiency, a capability for high (over 95 percent) sulfur removal efficiency, lower NO_x emissions, low particulate emissions, reduced solid waste due to byproduct recovery of elemental sulfur, reduced cooling water requirements (because gas turbines, rather than boiler/furnaces, generate a large portion of the power), reduced land requirements and a capability to burn coal, oil, or natural gas (SFA, 1983).

Unlike PC plants, IGCC systems are characterized by a modular design which allows phased construction of the system, flexibility in fuel use, and flexibility in design. A phased approach to IGCC construction might be based on an initial natural gas- or oil-fired simple cycle gas turbine. In later phases a combined cycle and a coal gasification system are added, gradually increasing the electrical output of the facility and resulting ultimately in an IGCC power plant. The advantage of phased construction is that a utility can add new capacity incrementally, reducing the amount of capital that is at risk at any given time to uncertain electric load growth forecasts. The lead time required for simple cycle and combined cycle power plants is significantly less than that for PC power plants. Furthermore, the utility is not committed to natural gas or oil as a long-term fuel in a phased IGCC project (Fluor, 1986). Because IGCC plants are modular, they can be built in a wide range of sizes and with a variety of options for specific equipment, such as gasifiers and gas turbines.

Compared to PC power plants, the notion of integrated environmental control is extended further in IGCC processes. In IGCC systems, environmental control is required not just to meet environmental regulations, but also for proper plant operation. For example, pollutants such as sulfur species and ash particles have deleterious effects on key components of IGCC systems, such as the gas turbine, and therefore must be controlled. In addition, the environmental control systems significantly affect the thermal cycle and, hence, plant efficiency.

1.2.5 U.S. Clean Coal Technology Program

From 1975 to 1986, it is estimated that electric utilities have spent over $60 billion for SO_2 control. To utilize the nation's strategic coal resource in cost-effective, efficient, and environmentally acceptable manner, the U.S. government has embarked on a major program to reduce the technical risk and promote the commercial adoption of new clean coal technologies through demonstration projects. In 1985, the U.S. Congress appropriated $400 million for a first round of projects as part of the Clean Coal Technology (CCT) program. As part of the program, an equal or greater amount of funding for each demonstration project must be contributed from other sources (e.g., industry, state governments). In 1987, partly in response to Canadian concerns about transboundary acid rain resulting from emissions in the U.S., the administration requested that federal funding of the CCT program be increased to $2.5 billion. The U.S. Congress has since appropriated a total of $2.75 billion for the CCT program. Of this amount, $1.55 billion has already been committed to three rounds of solicitations from which 39 demonstration projects have been selected. The remaining $1.2 billion in the CCT program is to be distributed in fourth and fifth rounds of solicitations. The program is administered by DOE. Because electric utilities are the major consumer of coal in the U.S., the program is geared toward coal-fueled power plant or emission control system demonstration projects (DOE, 1987; GAO, 1989; 1990).

DOE has also sponsored and conducted research and development of a number of clean coal technologies as part of its regular mission. In particular, the DOE Pittsburgh Energy Technology Center (PETC) has been the focal point for DOE in-house and contract research on coal combustion based systems. The DOE Morgantown Energy Technology Center (METC) has conducted research on alternative coal conversion processes such as coal gasification and direct coal-fueled internal combustion engines and gas turbines. Private organizations in the U.S. that have been extensively involved in clean coal technology projects include the Electric Power Research Institute, the Gas Research Institute, and others.

DOE's involvement in the development of clean coal technologies suggests that its capability to evaluate new technologies and make decisions about research planning be critically evaluated. According to Merrow et al (1981), DOE is in a poor position to interpret performance and cost estimates. For example, DOE lacks a database of projects that it has already undertaken to provide insight into historical sources of difficulties in performance and cost estimating. DOE does not have a corporate experience or memory to analyze the problems of cost growth and performance shortfalls. Further, Merrow et al assert that DOE is subject to a "political environment in which it is very tempting to attribute problems to inflation and regulation" [p. 89]

The participation of the federal government in environmental regulation and clean coal technology development makes such development a matter of public policy. Furthermore, different branches of the government can have either synergistic or interfering effects of the development of government sponsored technology. For example, regulations developed by the EPA are a criteria by which new technology must be judged. However, the development of new technology with reduced emissions will also spur changes in environmental regulations. The new CAAA provides explicit new economic incentives for development of innovative cost-effective emission control systems.

1.2.6 Greenhouse Gas Emissions

One additional concern related to coal utilization which is receiving increasing attention is "global warming." Global climate change is postulated by many to result from changes in the chemical composition of the atmosphere that lead to changes in the earth's atmospheric energy balance. Certain gases transmit short wavelength incoming solar energy, but absorb the infrared energy radiated from the earth's surface. These radiatively important gases (RIG's) are the so-called greenhouse gases. An increase in the concentration of RIG's is expected to lead to re-radiation of infrared energy back to the earth, leading to an increase in atmospheric and surface temperatures. The most significant RIG is believed to be CO_2, because of its spectral absorption range, and the fact that its atmospheric concentration is increasing significantly. Predictions of the "greenhouse effect" a fraught with uncertainties resulting from simplifications or lack of understanding about how the atmosphere and carbon cycle work (e.g., UNEP, no date).

Because fossil fuel power plants are a major emission source of CO_2, they are likely to be affected by any policy which aims to reduce greenhouse gas emissions. Thus, even though the threat of "global warming" is uncertain, it may be prudent to promote the development of more efficient clean coal technologies, which reduce the emission of CO_2 associated with a unit of generated electricity.

1.3 Objectives

The objective of this work is develop and apply a method for research planning for innovative process technologies. While current approaches to technology modeling and decision making may be appropriate for well-established, commercial technology, they are inadequate as a basis for research planning. Deterministic performance and cost estimates based on "best-guess" assumptions are not likely to provide insight into interactions among uncertainties which are sources of performance shortfall or cost growth. They are not likely to provide explicit insight into the specific process parameters which may contribute to technical or cost risk, nor are they likely to provide an explicit quantitative measure of the likelihood that a new technology will fail compared to conventional technology.

The Rand studies have indicated that a systematic approach to incorporating expert judgments about potential sources of performance shortfalls and cost growth is needed. In any type of modeling effort, the limitations of data and of knowledge about the system should be reflected in the model results. Clearly, uncertainties abound in the early stages of technology development, and they must be considered as an integral part of research planning.

To satisfy requirements for research planning, it is necessary to: (1) identify robust solutions to process design questions in the face of uncertainty to eliminate inferior design options; (2) identify key problem areas in a technology that should be the focus of further research to reduce the risk of technology failure; (3) compare competing technologies on a consistent basis to determine the risks associated with adopting a new technology; and (4) evaluate the effects that additional research might have on comparisons with conventional technology.

An important class of process technologies are electric power plants. In particular, innovative clean coal technologies, as discussed in Section 1.2, are expected to play a key role in the energy and environmental future of the U.S., as well as in other countries. Research planning for innovative clean coal technology development is an important part of energy and environmental policy. Thus, the research planning method developed here is applied to case studies involving several clean coal technologies. The purpose of the case studies is both to demonstrate the research planning method and to obtain technology-specific conclusions regarding research strategies.

1.4 Overview of Dissertation

A research planning method is developed and applied to analyze research strategies for several selected advanced coal-fueled power plant technologies. The generalizable features of the methodology are discussed in Chapter 2. One baseline and four advanced coal-fueled power plant technologies were selected to demonstrate application of the research planning method. Chapter 3 describes the selection and modeling of these systems.

Technical judgments about uncertainties in performance, cost, and economic parameters in the engineering and economic models are required as part of the quantitative approach to RD&D planning. The basis for the uncertainties assigned to model parameters for each of the five systems investigated in this work are documented in Chapter 4.

In Chapter 5, the models discussed in Chapters 3 are applied using the judgments about uncertainty described in Chapter 4 and the methodology for uncertainty and decision analysis discussed in Chapter 2. A variety of case studies are given in Chapter 5 which illustrate ways in which probabilistic analysis of sufficiently integrated models can provide insights not readily obtained from traditional deterministic or sensitivity analysis.

Several implications of the probabilistic modeling approach and the case study results are discussed in Chapter 7. Conclusions regarding both the generalizable features of the methodology employed here and the specific results of the case studies for each technology are given in Chapter 8.

In this work, a considerable effort was devoted to model development and the elicitation of judgments about uncertainties. The complete documentation of these aspects of the research was not appropriate for the main body of the dissertation. For the interested reader, however, an extensive set of appendices is included which present in detail the basis for the engineering models (Appendix A) and approach used to estimate and elicit uncertainties (Appendix B).

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2.0 METHODOLOGY

In Section 1.3, objectives for a research planning method are discussed. In this Chapter, a method which satisfies these objectives is presented. The method is shown schematically as a flow diagram in Figure 3. The purpose of the method is to provide a more robust and useful tool than traditional approaches for evaluation of innovative process technologies. The method also provides a quantitative means for research planning. Each of the major features of the method are discussed in the following sections. These include: (1) selection of candidate technologies for detailed evaluation; (2) development of appropriate engineering models of the selected technologies; (3) elicitation of expert technical judgments about uncertainties; (4) a modeling environment for performing probabilistic analysis; and (5) applications of the probabilistic models to address concerns in research planning. These applications include: characterizing uncertainty in key measures of plant performance, emissions, and cost; identifying robust design trade-off decisions in the face of uncertainty; identifying the key uncertainties in model input parameters that drive uncertainty in model output variables for the purpose of research prioritization; and comparison of alternative technologies when faced with uncertainty.

Technologies which continue to look promising after the detailed evaluation are then considered for further research. Probabilistic analysis of a technology provides a quantitative basis for focusing research expenditures on the specific aspects of the process which most significantly contribute to the risk of technology failure. The probabilistic analysis also provides a basis for bounding the expenditures that should be committed to further research, by providing a quantitative measure of the expected pay-off from further research.

2.1 Identification of Case Studies

Prior to performing an analysis, one must decide on criteria by which alternatives are to be judged and select a set of alternatives that are to be evaluated.

2.1.1 Identifying Decision Criteria

In the context of government-sponsored research, there are public policy concerns such as emissions, consumption of key natural resources, and cost which motivate the selection of decision criteria for evaluating and selecting innovative process technologies. In the context of corporate-sponsored research, concerns may be similar, but approached

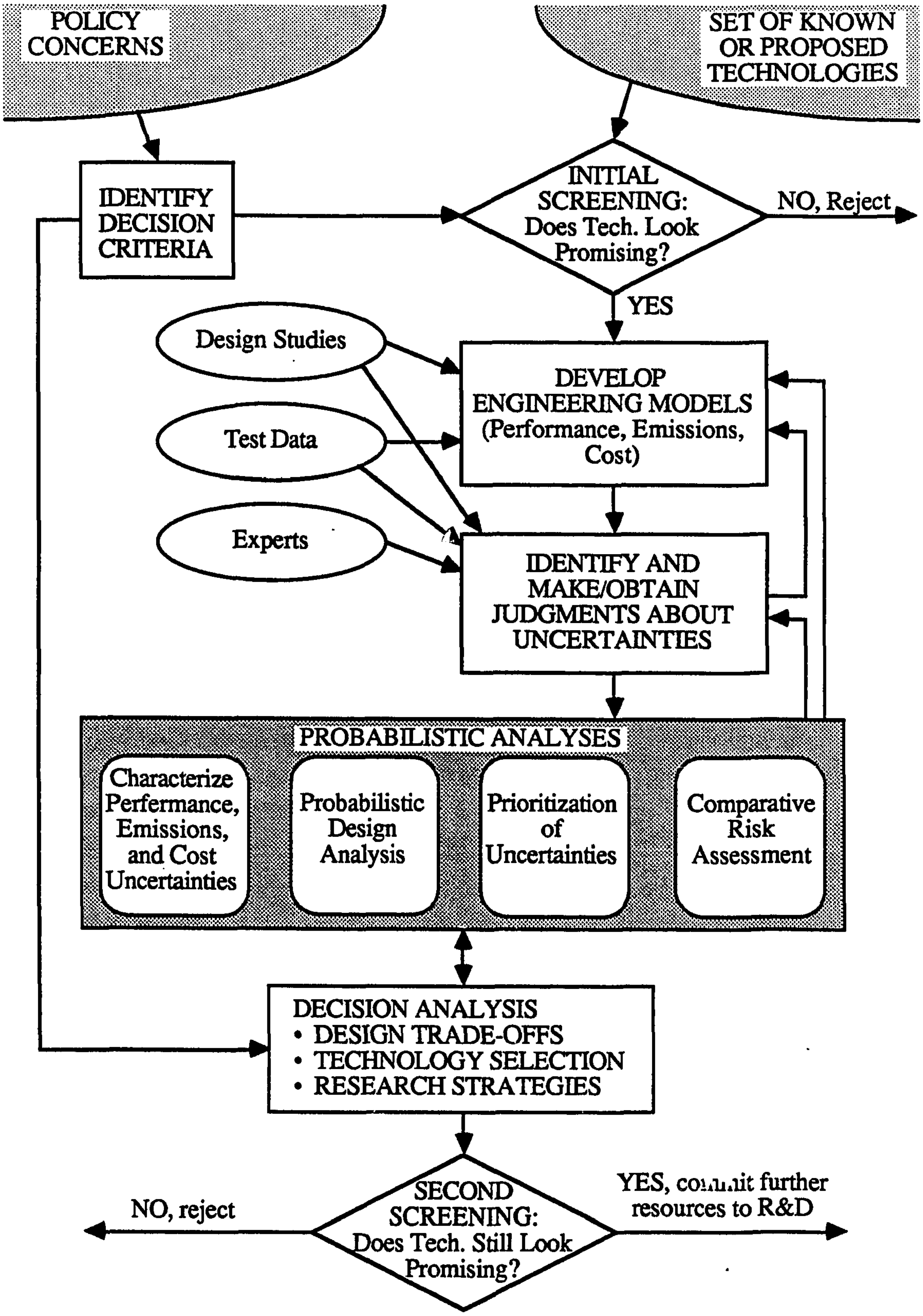

Figure 3. Method for Evaluation of Innovative Process Technologies

from a different perspective. For example, cost-related concerns may be more important, and environmental concerns only considered to the extent that they are imposed by existing environmental standards. In contrast, publicly-sponsored research may be more concerned with identifying long-term technology options capable of achieving substantial improvements or reductions in areas such as resource consumption and emissions.

Form the electric utility perspective, decisions regarding the adoption of a new electric power generating technology for commercial use are made on the basis of a number of attributes, such as familiarity with the process, previous experience with similar systems, perceptions about the risk that the new system will have either poor performance or high cost compared to conventional technology, and ability of the new technology to comply with environmental regulations.

Analysis of case studies in this work will focus on consideration of objective measures of plant performance, cost, and environmental discharges. Other considerations can easily be included in a decision analysis framework, if desired. Various measures of the attractiveness of a new technology can be combined into a single multi-variate "utility" function which weights various attributes according to the preferences of a decision-maker. Decision analysis techniques are discussed further in Section 2.5.

The engineering models developed in this work will be exercised to estimate key measures of plant performance and cost. These include material requirements, plant efficiency, emissions, total capital cost, annual operating costs, and the levelized cost of electricity. Examples of parameters that could be compared across process flowsheets are:

- Plant thermal efficiency (percent of input chemical heating value converted to electricity delivered at plant fenceline)
- Normalized emission rates (e.g., lb SO_2/MMBtu, lb CO_2/kWh)
- Normalized consumption rates (e.g., raw water or coal flow per kWh)
- Normalized waste flows (e.g., lb ash/kWh)
- Capital Cost ($/kW)
- Fixed Operating Cost ($/kW-yr)
- Incremental Variable Operating Cost (mills/kWh)
- Byproduct Credit (mills/kWh)
- Fuel Cost (mills/kWh)
- Cost of Electricity (mills/kWh)
- Construction Time (years)
- Modularity (increments of capacity that can be added at a time, MW)

A technology would be favored if it had higher efficiency, reduced material consumption, reduced emissions and waste discharges, reduced cost, reduced construction time, and increased modularity compared to conventional technology. A technology would be rejected if it could not comply with environmental standards, regardless of the other attributes.

The cost of electricity (COE) usually provides the single most important measure of process feasibility, if all other minimum requirements are met. Because the cost of electricity is based on levelized capital and annual costs normalized to the plant energy output, the effects of plant efficiency, costs of consumables, and partial costs of some environmental discharges (e.g., ash disposal, pollution control) are already included. Other costs, such as externalities arising from air emissions, are not included. However, the CAAA will result in an equivalent price for SO_2 emissions which can be included in a decision analysis.

Decisions about innovative technologies can be viewed from either an economic or decision analysis perspective. In an economic perspective, a power plant produces technically interdependent joint products of electricity and emissions. Electricity is priced in an economic market, while emissions are currently an externality that are not priced due to market failure. In the context of a cost evaluation of a technology, electricity has a positive price, while emissions have a negative price. In addition, other joint products are possible, such as salable byproducts with positive prices. Command and control regulations do impose costs associated with fixed emission reductions, but do not provide an economic incentive to control the quantity of emissions to optimize total production cost. However, the CAAA will create a market for selected pollutants--at least for SO_2 in the short term and perhaps in five or ten years NO_x also. Thus, the cost of production will be a function of the electricity and byproducts sold and the pollutants emitted. The power producer thus would seek to optimize the revenue from the joint products less the costs of production, subject to any constraints on maximum emission rates.

An alternative perspective is that of decision analysis. In a decision analysis framework, a research planner would seek to select alternative research strategies which maximize a policy-based utility function. The utility function can consider multiple attributes and timing of outcomes. For example, the cost of electricity would be one attribute by which technologies could be selected. As previously discussed, this attribute takes into account a variety of interactions among performance and cost in the power plant, including plant efficiency. However, the environmental insult from the technology may need to be considered apart from the cost of electricity. Thus, a multi-attribute utility function might include cost of electricity, and emissions of pollutants such as SO_2, NO_x, and CO_2 as attributes.

2.1.2 Selection of Alternatives

When evaluating innovative clean coal technology, the objective is to identify the most promising candidates which hold promise of reduced emissions, improved plant efficiency, and reduced costs compared to conventional technology. Initially, there may be a large set of potential candidate technologies that could be selected for more detailed evaluation. An initial screening of such technologies may be required to identify a manageable subset for further study. The screening would be based on the decision criteria identified based on policy or other concerns.

The selection of process flowsheets for analysis should be based on expectations regarding system configurations that are likely to adopted into commercial use, if key uncertainties can be resolved with positive outcomes during research and development. Usually, candidates for RD&D are identified based on early estimates which indicate promising performance and cost. More detailed modeling and evaluation should then be completed to help focus and bound expenditures for further research.

2.2 Performance and Cost Modeling

A key step in evaluating a process technology is the development of an appropriately detailed engineering model. The scope of a complete engineering model includes mass and energy balances for major process areas (plus additional technical detail as warranted), characterization of emissions of key pollutants, and characterization of capital, annual, and levelized plant costs. The selection of chemical species to include in the model for analysis of environmental discharges is motivated by the same set of policy concerns that influences decision criteria. For example, if acid rain is of concern, decision criteria will include SO_2 and NO_x emissions, which must then also be included in the engineering models.

2.2.1 Purpose of Modeling

The purpose of engineering models of innovative technologies that are in early stages of research is to reasonably and completely characterize the existing state of knowledge about the new technology. A second objective is to try to predict the performance and cost of a commercialized system based on the existing knowledge. However, existing knowledge about an innovative concept is often incomplete. The model can then be used to identify the key weaknesses in understanding that require further investigation, if uncertainties in knowledge can be explicitly represented in the model.

An engineering model should be sensitive to key parameters that are known to affect either performance or cost. An engineering model should be sufficiently detailed to capture: (1) performance interactions among process areas, include feedbacks or recycle streams; and (2) interactions between specific performance and design parameters and cost. However, while it may often be possible to build complex models involving hundreds of variables, usually only a handful of variables are found to be important determinants of performance and cost. Thus, there is usually a decreasing return in investment for building larger and more complex models than are needed.

The modeling philosophy used here is to build models of sufficient complexity to capture all expected important interactions or all known sources of uncertainty, but to keep the models within a manageable size to facilitate running the models and interpreting results. Thus, the models are developed to characterize generic features of the process technologies. They are not intended to include all of the detail that would be needed for equipment sizing for every piece of equipment in the plant or for estimating the costs of a site-specific construction project.

2.2.2 Modeling Performance

A number of fundamental constraints on process performance exist which can form the basis for modeling any process technology in the earliest phases of research. Because many aspects of the technology may be poorly understood initially, the most robust approach to developing performance models is to begin with relatively simple mass and energy balances. In a simple model, key performance areas which are poorly understood or which require additional empirical data can often be parameterized. For example, conversion rates of chemical reactions may be specified as model inputs rather than calculated based on reaction kinetics, which may be poorly understood. In this manner, it is possible to include structural features in the model that represent important process interactions, even when faced with limited information. Furthermore, it is possible to represent uncertainty in those features as uncertainties in specific model parameters. Thus, if the chemistry of a particular reaction is poorly understood, the reaction conversion rate may be treated as uncertain.

In addition to estimating major performance measures, such as the consumption of key reagents or plant efficiency, a performance model should be sufficiently detailed to track the key environmental species of concern. This may require mass balances for dozens of chemical species, depending on the system being modeled.

As more information is obtained from early research, the models can be refined to include additional information regarding reaction kinetics or other details as appropriate. Additional constraints on the mass and energy balances, or new features of the technology, may be added at varying stages of technology development based on theoretical expectations, experience with analogous systems, or preliminary experimental results with small-scale versions of the innovative system. Also, as the quality of information about the technology improves, the values assumed for performance and cost parameters in the model can also be refined. Thus, model development should be considered as an iterative process.

2.2.3 Modeling Cost

There are a variety of approaches to developing cost estimates for process plants. These approaches differ in the level of detail with which costs are disaggregated into separate line items, as well as in the simplicity or complexity of analytic relationships used to estimate line item costs. The level of detail appropriate for the cost estimate depends on: (1) the state of technology development for the process of interest; and (2) the intended use of the cost estimates. The models developed here are intended to estimate the costs of innovative coal-to-electricity systems for the purpose of evaluating the comparative economics of alternative process configurations. The models are intended to be used only for preliminary or "study grade" estimates using representative (generic) plant designs and parameters.

In the electric utility and chemical process industries, there are generally accepted guidelines regarding the approach to developing cost estimates. EPRI (1986) has defined four types of cost estimates: simplified, preliminary, detailed, and finalized. The cost estimates developed here are best described as "preliminary." The differences between different types of cost estimates are briefly described below.

A *simplified* cost estimate is based on information about major stream flow rates and design parameters from a simple process flow diagram. The cost information used in a simplified estimate typically includes published cost curves or scaling relationships for generic process areas or for the plant as a whole. A simplified cost estimate may also be based on adjusting costs from similar published or in-house work on the basis of a single performance parameter. A simplified estimate is thus sensitive to only one (or a few) major performance parameter(s), such as the coal feed rate or the plant electrical output.

A *preliminary* cost estimate is based on a more disaggregated consideration of the costs of specific process areas and specific equipment items. A preliminary estimate also includes the use of ratio or scaling relationships to adjust costs for a variety of operating conditions. The preliminary estimate is sensitive to a larger number of performance parameters (perhaps a few dozen) than the simplified estimate.

Detailed and *finalized* cost estimates are generally developed only for site-specific projects that are intended for construction. For a large process plant, these types of estimates may cost millions of dollars to prepare. They are based on vendor quotations for specific equipment costs in response to specifications developed by an architect/engineering firm.

For the purposes of evaluating alternative technologies, and for research planning, preliminary cost estimates are the most appropriate. Preliminary cost estimates are sensitive to the performance and design parameters that are most influential in affecting costs. Thus, the goal of this study is to develop preliminary cost estimates for the systems under study.

A major constraint on cost model development is the availability of data from which to develop cost versus performance relationships for specific process areas or for major equipment items. Data from published studies can be used to develop cost models for specific process areas using regression analysis. Regression analysis is used extensively for cost model development in this study (and elsewhere). An overview of the key concepts of regression analysis, and the philosophy of this study in applying regression analysis, is given briefly in the next section. Alternatively, cost models for process areas consisting of only one major equipment item can be based on published equipment cost curves, either in place of or as a supplement to regression analysis.

2.2.4 Role of Regression Analysis in Model Development

Regression analysis is used to help understand the interrelationships among a given set of variables. The use of regression analysis here is oriented toward developing useful and reasonable relationships primarily between process area costs and key performance parameters. In a few cases, it is also used to develop useful relationships between performance variables. The emphasis here is not on the use of extensive formal statistical tests but rather on the practical application of regression analysis for cost model development. Thus, some statistical tests, along with engineering judgments and the availability of data, are used to guide the selection of parameters, the representation of relationships in the regression models, and validation of the models. The "goodness" of the regression models are indicated with common summary statistics, graphical comparison of the model predictions with the actual data, and evaluation of the appropriateness of the model relationships with *a priori* engineering expectations.

The approach used in the development of regression models in this research is described in detail in Appendix A.2.4. The issues related to developing the regression models include developing a data set for analysis, selecting parameters for inclusion in the model, and validating the model. One measure of how well a regression model fits the data upon which it is based is the standard error of the model. The standard error is one type of uncertainty that can be explicitly considered as part of probabilistic modeling of the clean coal technologies. It represents the variability in cost (or other quantity predicted by the model) that is not explained by the regression model.

2.3 Characterizing Uncertainties

As discussed in Section 1.1.3, analyses of technologies that are still in the research phase involve uncertainties, which are often ignored or treated in a limited way using sensitivity analysis. However, sensitivity analysis suffers from shortcomings resulting from the difficulty in evaluating the effect of *simultaneous* variations in several parameters and the lack of insight into the *likelihood* of obtaining any particular result.

A more robust approach is to represent uncertainties in model parameters using probability distributions. Using probabilistic simulation techniques, simultaneous uncertainties in any number of model input parameters can be propagated through a model to determine their combined effect on model outputs. The result of a probabilistic simulation includes both the possible range of values for model output parameters and information about the likelihood of obtaining various results. This provides insights into the risks or potential pay-offs of a new technology. Statistical analysis on the input and output data can be used to identify trends (e.g., key input uncertainties affecting output uncertainties), without need to re-run the analysis. Thus, probabilistic analysis can be used as a research planning tool to identify the uncertainties in a process that matter the most, thereby focusing research efforts where they are most needed. Probabilistic analysis may be referred to elsewhere as "range estimating" or "risk assessment."

There are three general areas of uncertainty that should be explicitly reflected in engineering models. These are uncertainties in: (1) process performance parameters (e.g., flowrates), (2) process area capital costs, and (3) process operating costs. For example, in calculating the cost of a gasifier, there may be uncertainty (because of the lack of commercial experience with the design) in the coal flow rate required to achieve a given electrical output. This leads to uncertainty in the size (hence, cost) of the gasifier for a particular system. However, for a given gasifier size and type, there is also a probability that the equipment cost could be higher or lower than the nominal estimate (e.g., due to expected improvements in equipment design and cost, or to potential problems with fouling and corrosion, requiring more expensive materials, design modifications, or additional maintenance). The same type of uncertainties may apply to operating and maintenance cost factors. The uncertainties associated with advanced systems or subsystems will typically be much larger than for conventional technology. A probabilistic engineering modeling framework is required to evaluate the overall uncertainty in process cost as a result of performance and cost uncertainties in specific process areas to determine the overall technical and cost risks and to identify research priorities.

The development of ranges and probability distributions for model input parameters can be based on information available in published studies, statistical data analysis and/or the judgments of process engineers with relevant expertise. The approaches to developing probability distributions for model parameters are similar in may ways to the approach one might take to pick a single "best guess" number for deterministic (point-estimate) analysis or to select a range of values to use in sensitivity analysis. However, the development of estimates of uncertain usually requires more detailed thinking about possible outcomes and their relative likelihoods.

2.3.1 Philosophy of Uncertainty Analysis

The "classical" approach in probability theory requires that estimates for probability distributions must be based on empirical data. However, in many practical cases, the available data may not be relevant to the problem at hand. For example, test results from a process development unit (PDU) under a given set of conditions may not be directly applicable to estimating the performance of a fifth-of-a-kind commercial scale plant under a different set of operating conditions. Thus, statistical manipulation of data may be an insufficient basis for estimating uncertainty in a real system of interest. Engineering analysis or judgments about the data may be required.

An alternative approach differs in how probability distributions are interpreted. In the so-called "Bayesian" view, the assessment of the probability of an outcome is based on a "degree of belief" that the outcome will occur, based on all of the relevant information an analyst currently has about the system. Thus, the probability distribution may be based on empirical data and/or other considerations, such as technically-informed judgments or predictions. People with different information may estimate different distributions for the same variable (Morgan and Henrion, 1990). The assessment of uncertainties requires one to think about all possible outcomes and their likelihoods, not just the "most likely" outcome. This is an advantage for the analyst, because by thinking systematically and critically about uncertainties, one is more likely to anticipate otherwise overlooked problems, or to identify otherwise overlooked potential pay-offs of a system.

2.3.2 Types of Uncertain Quantities

There are a number of types of uncertainty that one might consider when developing a probability distribution for a variable. Some of these are summarized briefly here.

Statistical error is associated with imperfections in measurement techniques. Statistical analysis of test data is thus one method for developing a representation of uncertainty in a variable.

Empirical measurements also involve *systematic error*. The mean value of a quantity may not converge to the "true" mean value because of biases in measurement and procedures. Such biases may arise from imprecise calibration, faulty reading of meters, and inaccuracies in the assumptions used to infer the actual quantity of interest from the observed readings of other quantities. Estimating the possible magnitude of systematic error may involve an element of engineering judgment. For example, data on sorbent attrition in a PDU may be used to estimate the sorbent attrition in a fifth-of-a-kind commercial-scale system. The conditions in the PDU differ from that in the commercial scale unit; therefore, there may be a systematic error involved in using the PDU data for design purposes.

Variability can be represented as a probability distribution. Some quantities are variable over time. For example, the composition of a coal (or perhaps a sorbent) may vary over time.

Uncertainty may also arise due to lack of actual experience with a process. This type of uncertainty often cannot be treated statistically, because it requires predictions about something that has yet to be built or tested. This type of uncertainty can be represented using technical estimates about the range and likelihood of possible outcomes. These judgments may be based on a theoretical foundation or experience with analogous systems.

2.3.3 Encoding Uncertainties as Probability Distributions

As indicated in the previous sections, there are two fundamental approaches for encoding uncertainty in terms of probability distributions. These include statistical estimation techniques and engineering judgments. A combination of both methods may be appropriate in many practical situations. For example, a statistical analysis of measured test data may be a starting point for thinking about uncertainties in a hypothetical commercial scale system. One must then consider the effect that systematic errors, variability, or uncertainties about scaling-up the process might have on interpreting test results for commercial scale design applications.

Statistical Techniques

Statistical estimation techniques involve estimating probability distributions from available data. The fit of data to a particular probability distribution function can be evaluated using various statistical tests. For example, the cumulative probability distribution of a set of data may be plotted on "probability" paper. If the data plot as a straight line, then the distribution is normal. Procedures for fitting probability distribution functions are discussed in many standard texts on probability and are not reviewed here. Rather, the focus of this discussion is on the situations where statistical analysis alone may be insufficient, because engineering insights may be required to interpret whatever limited data are available.

Judgments about Uncertainties

In making judgments about a probability distribution for a quantity, there are a number of approaches (heuristics) that people use which psychologists have observed. Some of these can lead to biases in the probability estimate. Three of the most common are briefly summarized.[1]

1) *Availability.* The probability that experts assign to a particular possible outcome may be linked to the ease (availability) with which they can recall past instances of the outcome. For example, if tests have yielded high sorbent durability, it may be easier to imagine obtaining a high sorbent durability in the future than obtaining lower durabilities.

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[1] The discussion here is based on Morgan and Henrion, *Uncertainty: A Guide to Dealing with Uncertainty in Quantitative Risk and Policy Analysis,* Cambridge University Press, 1990.

Thus, one tends to expect experts to be biased toward outcomes they have recently observed or can easily imagine, as opposed to other possible outcomes that have not been observed in tests.

2) *Representativeness* has also been termed the "law of small numbers." People may tend to assume that the behavior they observe in a small set of data must be representative of the behavior of the system, which may not be completely characterized until substantially more data are collected. Thus, one should be cautious in inferring patterns from data with a small number of samples.

3) *Anchoring and adjustment* involves using a natural starting point as the basis for making adjustments. For example, an expert might choose to start with a "best guess" value, which represents perhaps a median or most likely (modal) value, and then make adjustments to the best guess to achieve "worst" and "best" outcomes as bounds. The "worst" and "best" outcomes may be intended to represent a 90 percent probability range for the variable. However, the adjustment from the central "best guess" value to the extreme values is often insufficient, with the result that the probability distribution is too tight and biased toward the central value. This phenomena is *overconfidence*, because the expert's judgment reflects less uncertainty in the variable than it should. The "anchor" can be any value, not just a central value. For example, if an expert begins with a "worst" case value, the entire distribution may be biased toward that value.

Judgments also may be biased for other reasons. One common concern is *motivational bias*. This bias may occur for reasons such as: a) a person may want to influence a decision to go a certain way; b) the person may perceive that they will be evaluated based on the outcome and might tend to be conservative in their estimates; c) the person may want to suppress uncertainty that they actually believe is present in order to appear knowledgeable or authoritative; and d) the expert has taken a strong stand in the past and does not want to appear to contradict themself by producing a distribution that lends credence to alternative views.

Designing an Elicitation Protocol

From studies of how well calibrated judgments about uncertainty are, it appears that the most frequent problem encountered is overconfidence (Morgan and Henrion, 1990). Knowledge about how most people make judgments about probability distributions can be used to design a procedure for eliciting these judgments. The appropriate procedure depends on the background of the expert and the quantity for which the judgment is being elicited. For example, if an expert has some prior knowledge about the shape of the distribution for the quantity, then it may be appropriate to ask him/her to think about extreme values of the distribution and then to draw the distribution. On the other hand, if a technical expert has little statistical background, it may be more appropriate to ask him/her a series of questions. For example, the expert might be asked the probability of obtaining a value less than or equal to some value x, and then the question would repeated for a few other values of x. The judgment can then be graphed by an elicitor, who would review the results of the elicitation with the expert to see if he/she is comfortable with the answers.

To overcome the typical problem of overconfidence, it is usual to begin by thinking about extreme high or low values before asking about central values of the distribution. In general, experts' judgments about uncertainties tend to improve when: (1) the expert is forced to consider how things could turn out differently than expected (e.g., high and low extremes); and (2) the expert is asked to list reasons for obtaining various outcomes.

While the development of expert judgments may be flawed in some respects, it does permit a more robust analysis of uncertainties in a process when limited data are available. Furthermore, in many ways, the assessment of probability distributions is qualitatively no different than selecting single "best guess" values for use in a deterministic estimate. For example, a "best guess" value often represents a judgment about the single most likely value that one expects to obtain. The "best guess" value may be selected after considering several possible values. The types of heuristics and biases discussed above may play a similar role in selecting the value. Thus, even when only a single "best guess" number is used in an analysis, a seasoned engineer usually has at least a "sense" for "how good that number really is." This may be why engineers are often able to make judgments about uncertainties easily, because they implicitly make these types of judgments routinely.

2.3.4 Some Types of Probability Distributions

Examples of several types of probability distributions are shown in Figure 4 as both probability density functions (pdf's) and cumulative distribution functions (cdf's). The pdf is a graphical means of representing the relative likelihood or frequency with which values of a variable may be obtained. The pdf also clearly illustrates whether a probability distribution is symmetric or skewed. In a symmetric unimodal distribution, the mean (average), median (50th percentile), and mode (peak) coincide. In a positively skewed distribution (e.g., lognormal), the mean is greater than the median, and both are greater than the mode.

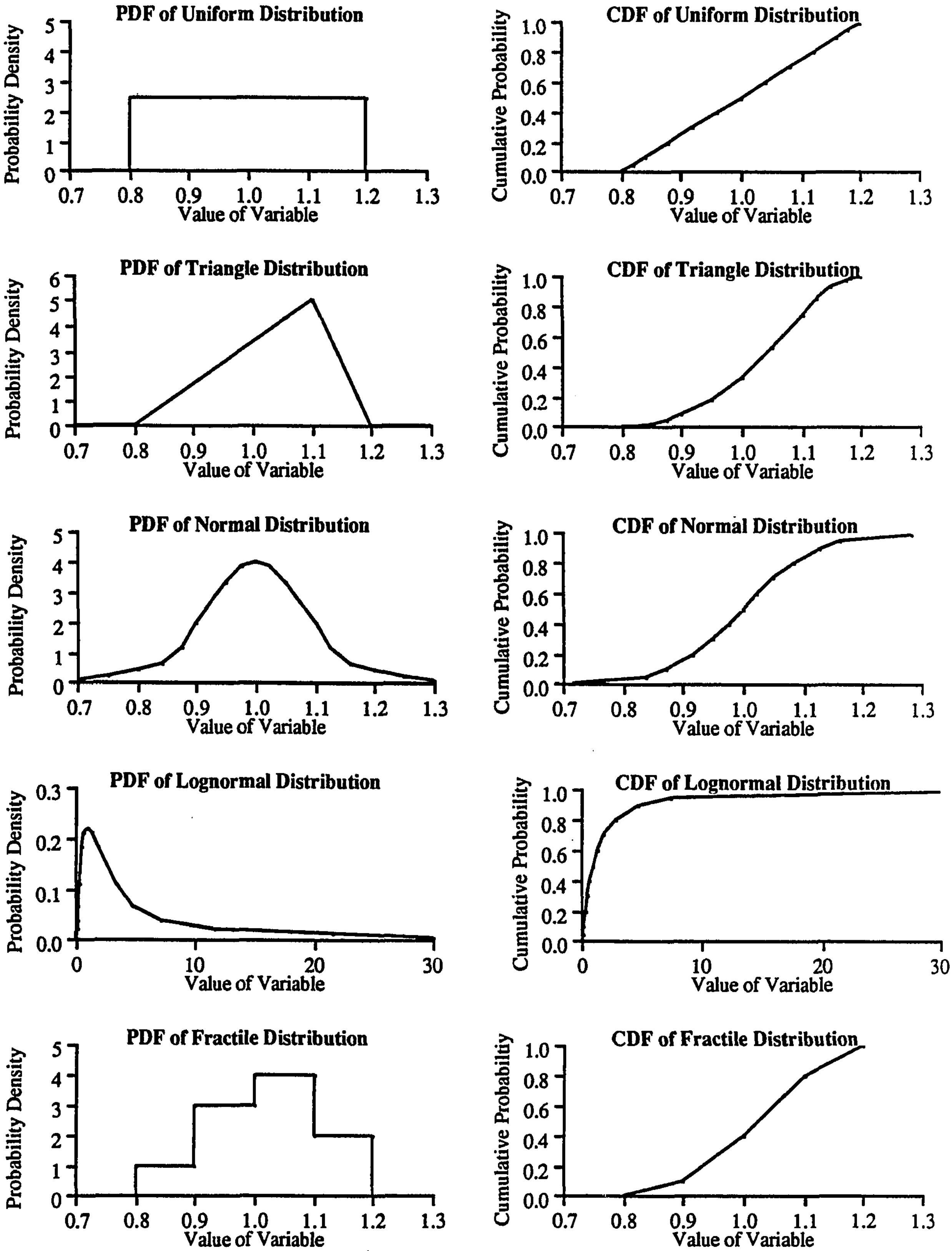


Figure 4. Some Types of Probability Distributions.

An alternative way to represent a probability distribution is the cdf. The cdf shows probability fractiles on the y-axis and the value of the distribution associated with each fractile on the x-axis. The cdf is a way to represent any probability distribution when there is information about various fractiles of the distribution (e.g., the values of the 5th, 50th and 95th percentiles).

A brief description of several types of probability distributions and their applications is given here:

- Uniform: Uniform probability of obtaining a value between upper and lower limits. Useful when an expert is willing to specify a finite range of possible values, but is unable to decide which values in the range are more likely to occur than others. The use of the uniform distribution is also a signal that the details about uncertainty in the variable are not known. Useful for screening studies.
- Triangle: Similar to uniform except a mode is also specified. Use when an expert is willing to specify both a finite range of possible values and a "most likely" (mode) value. The triangle distribution may be symmetric or skewed (as in Figure 4). Like the uniform, this distribution indicates that additional details about uncertainty are not yet known. The triangle distribution is excellent for screening studies and easy to obtain judgments for.
- Normal: A symmetric distribution with mean, mode, and median at the same point. Often assumed in statistical analysis as the basis for unbiased measurement errors. The normal distribution has infinite tails; however, over 99 percent of all values of the normal distribution lie within plus or minus three standard deviations of the mean. Thus, when used to represent uncertainty in physical quantities which much be greater than zero, the standard deviation should not be more than about 20 or 30 percent of the mean, or else the distribution must be truncated.
- Lognormal: A positively skewed distribution (it has a long tail to the right). This distribution is usually used to represent uncertainty in physical quantities which must be non-negative and are positively skewed, such as the size of an oil spill or the concentration of a pollutant. This distribution may be used when uncertainties are expressed on a multiplicative order-of-magnitude basis (e.g., factor of 2) or when there is a probability of obtaining extreme large values.
- Loguniform: A uniform distribution in log space (each decade has equal probability, not shown in Figure 4).
- Fractile: The finite range of possible values is divided into subintervals. Within each subinterval, the values are sampled uniformly according to a specified frequency for each subinterval. This distribution looks like a histogram and can be used to represent any arbitrary data or judgment about uncertainties in a parameter, when the parameter is continuous. Explicitly shows detail of the judgments about uncertainties.
- Chance: This is like the fractile distribution, except that it applies to discrete, rather than continuous, variables. An example of a discrete variable is the number of trains of equipment, which must be an integer (e.g., 30% chance of one train, 70% chance of two).

2.4 Probabilistic Modeling

In order to analyze uncertainties in innovative process technologies, a probabilistic modeling environment is required. A typical approach is the use of Monte Carlo simulation, as described by Ang and Tang (1984) and others. In Monte Carlo simulation, a model is run repeatedly, using different values for each of the uncertain input parameters each time. The values of each of the uncertain input parameters are generated based on the probability distribution for the parameter. If there are two or more uncertain input parameters, one value from each is sampled simultaneously in each repetition in the simulation. Over the course of a simulation, perhaps 20, 50, 100, or even more repetitions may be made. The result, then, is a set of sample values for each of the model output variables, which can be treated statistically as if they were an experimentally or empirical observed set of data.

Although the generation of sample values for model input parameters is probabilistic, the execution of the model for a given set of samples in a repetition is deterministic. The advantage of Monte Carlo methods, however, is that these deterministic simulations are repeated in a manner that yields important insights into the sensitivity of the model to variations in the input parameters, as well as into the likelihood of obtaining any particular outcome.

Monte Carlo methods also allow the modeler to use any type of probability distribution for which values can be generated on a computer, rather than to be restricted to forms which are analytically tractable.

In random Monte Carlo simulation, a random number generator is used to generate uniformly distributed numbers between 0 and 1 for each uncertain variable. Note from Figure 4 that all cdf's have an ordinate axis ranging from zero to one. Thus, uniformly distributed random numbers are used to represent the fractile of the random variable for which a sample is to be generated. The sample values for the random variables are calculated using the inverse cdf functions based on the randomly generated fractiles.

Using Monte Carlo techniques, it is therefore possible to represent uncertainty in a model of a process technology by generating sample values for uncertain variables, and running the model repetitively. Instead of obtaining a single number for model outputs as in deterministic simulation, a set of samples is obtained. These can be represented as cdf's and summarized using typical statistics such as mean and variance.

An alternative to random Monte Carlo simulation is Latin Hypercube Sampling (LHS). In LHS methods, the fractiles that are used as inputs to the inverse cdf are not randomly generated. Instead, the probability distribution for the random variable of interest is first divided into ranges of equal probability, and one sample is taken from each equal probability range. However, the ranking (order) of the samples is random over the course of the simulation, and the pairing of samples between two or more random input variables is usually treated as independent. In median LHS, one sample is taken from the median of each equal-probability interval, while in random LHS one sample is taken from random within each interval (Morgan and Henrion, 1990).

LHS methods guarantee that values from the entire range of the distribution will be sampled proportional to the probability density of the distribution. Because the distributions are sampled over the entire range of probable values in LHS, the number of samples required to adequately represent a distribution is less for LHS than for random Monte Carlo sampling. LHS is the technique employed in this study.

2.4.1 Modeling Environments

Two probabilistic modeling environments are used in this study. One is an non-procedural interactive environment developed by Henrion (Henrion, 1982; Henrion and Wishbow, 1987). Key uncertainties in process parameters can be characterized using a variety of probability distributions. The resulting uncertainty distributions for model outputs are then estimated using median LHS.

The other modeling environment is ASPEN, a chemical process simulator. ASPEN is described further in Section 3.2.2. As part of DOE-supported research at Carnegie-Mellon University, a probabilistic modeling capability has been added to the publicly available version of ASPEN (Diwekar, Rubin, and Frey, 1989). An initial step in this effort was the identification of suitable software for sampling probability distributions and performing output analysis. A Fortran program developed by Iman and Shortencarier (1984) using Latin hypercube sampling (LHS) was identified as the best publicly available program for assigning probability distributions to model parameters and generating samples from those distributions. A Fortran program developed by Iman, Shortencarier, and Johnson (1985) was identified for analysis of model output. This program uses partial correlation coefficients and standardized regression coefficients for measuring linear correlations and partial rank correlation coefficients and standardized rank regression coefficients for measuring nonlinear correlations.

The LHS sampling and output analysis programs have been implemented into ASPEN through a new unit operation block, which is documented in a new user's manual (Diwekar, 1989) and technical reference manual (Diwekar and Rubin, 1989). The stochastic block assigns user-specified distributions to the key input parameters selected by the user, uses the LHS program to generate samples from the distributions, and passes the sampled values of each uncertain parameter to the flowsheet. After a flowsheet simulation is run, the output variables of interest are collected. The simulation is then repeated for a new set of samples selected from the probabilistic input distributions. A new Fortran block is used to control the cycling of the stochastic block, and another Fortran block is used to access and assign samples to model parameters. The probabilistic modeling capability in ASPEN has both Monte Carlo and random LHS options. There is also a capability to specify rank order correlations in input variables.

2.4.2 Selecting Sample Size

The sample size corresponds to the number of repetitions used in the probabilistic simulation. The selection of sample size is usually constrained at the upper end by the limitations of computer software, hardware, and time, and at the lower end by the acceptable confidence interval for model results. In cases where the analyst is most interested in the central tendency of distributions for output variables, the sample size can often be relatively small. However, in cases were the analyst is interested in low probability outcomes at the tails of output variable distributions, large sample sizes may be needed. As sample size is increased, computer runtime, memory use, and disk use may become excessive. Therefore, it may be important to use no more samples than are actually needed for a particular application.

One approach to selecting sample size is to decide on an acceptable confidence interval for whatever fractile level is of most concern in the investigation (Morgan and Henrion, 1990). For example, we may wish to obtain a given confidence that the value of the p^{th} fractile will be bounded by the i^{th} and k^{th} fractiles. In a Monte Carlo simulation, we can use the following relations to estimate the required sample size:

$$i = mp - c\sqrt{mp(1 - p)} \qquad (1)$$

$$k = mp + c\sqrt{mp(1 - p)} \qquad (2)$$

The relations in Equations (1) and (2) yield a confidence interval for the p^{th} fractile if the sample size is known, where c is the standard deviation of the standard normal distribution associated with the confidence level of interest. To calculate the number of samples required, the expressions above can be rearranged to calculate the confidence interval $(Y_{p-\Delta p}, Y_{p+\Delta p})$ as follows:

$$m = p(1 - p)\left(\frac{c}{\Delta p}\right)^2 \tag{3}$$

For example, if we wish to be 90 percent confident that the value of the 90th percentile will be enclosed by the values of the 85th and 95th fractiles, then c would be 1.65 and m would be 98.

However, another factor to consider in selecting sample size is whether a high degree of simulation accuracy is really needed. In screening studies based on a first-pass set of expert judgments, it may be unnecessary to obtain a high degree of confidence in specific fractiles of the output distribution, because initial estimates of uncertainty may be subject to considerable empirical uncertainty themselves.

In the work described here, computational limitations, particularly with respect to time, are significant factors in limiting sample size. For most studies, sample sizes of 100 or 150 have been found to be adequate to reasonably characterize the cumulative distribution functions for output variables. The relations in Equations (1) and (2) can be used to develop an explicit confidence interval for an entire CDF. In general, the magnitudes of the confidence intervals for the central fractiles of the distribution are "tighter" than for the tails. The confidence interval on the CDF is an indicator of how "good" the simulation is at estimating the "true" value of the CDF assuming that all model input assumptions are "true," but the true value of the CDF may remain uncertain based on limitations of the input assumptions.

The approach to selecting sample size described above is appropriate for use with the Monte Carlo simulation technique. In this work, LHS is employed as discussed previously. The approach to estimating the precision of modeling results based on confidence intervals will typically overestimate the required sample size needed with LHS.

2.4.3 Analyzing Results

Sample correlation coefficients are a simple but useful tool for identifying the linear correlations between uncertain variables. Other techniques are available as well in the newly implemented probabilistic capability for the ASPEN process simulator, as developed by Iman, Shortencarier, and Johnson (1985). These output analysis techniques are described here briefly.

A partial correlation coefficient (PCC) analysis is used to identify the degree to which correlations between output and input random variables may be linear, and it is estimated in conjunction with multi-variate linear regression analysis. In PCC analysis, the input variable most highly correlated the output variable of interest is assumed as the starting pointing for construction of a linear regression model. In the regression model, the output variable is treated as the dependent variable and the most highly correlated input variable is treated as a predictive variable. The partial correlation technique then searches for another input variable which is most highly correlated with the *residuals* of the regression model already containing the first input variable. The residual is the difference between the actual sample value of the dependent variable and the estimated sample values, using the linear regression model containing the first input variable. The process is repeated to add more variables in the analysis. The partial correlation coefficient is a measure of the unique linear relationship between the input and dependent variables that cannot be explained by variables already included in the regression model.

Standardized regression coefficients (SRC) can be used to measure the relative contribution of the uncertainty in the input variables on the uncertainty of the output variables. This analysis involves standardization of all the sample values for the model input variables and a multi-variate regression of an output variate based on the inputs. The regression coefficients for each input variate then indicate the relative importance of that variate as a factor determining the output. SRCs measure the shared contribution of the input to the output, because all of the simulation input uncertainties are included in the regression analysis simultaneously. The SRCs are the partial derivatives of the output variable with respect to each input variable. Because PCCs are a measure of the unique contribution of each parameter, and SRCs measure the shared contribution, they do not always lead to the same conclusions.

PCC and SRC analysis is limited to cases where the relationship between input and output variables is linear; however, by basing the regression analysis on the ranks of the samples for each variable, rather than on the values of the samples, the PCC and SRC techniques can be extended to non-linear cases. These techniques are known as partial rank correlation coefficients (PRCC) and standardized rank regression coefficients (SRRC) (Iman, Shortencarier, and Johnson, 1985).

While regression analysis of input and output sample vectors is an important tool for prioritizing input uncertainties that are most "sensitive," it is important to understand the limitations of partial correlation coefficients when using a given sample size. Edwards

(1984) provides a clear discussion of tests of significance for correlation coefficients. When using partial correlation coefficients for output analysis, we are interested in testing the null hypothesis that the coefficient is equal to zero. For independent random variables, the t-test can be used and the value of t is calculated as follows:

$$t = \frac{r}{\sqrt{1 - r^2}} \sqrt{m - n} \tag{4}$$

The degrees of freedom m-n is given by the number of samples m and the number of input variables n used in the regression analysis. The t statistic calculated in Equation (4) can then be compared to values in a table of the t-distribution for a given significance level and degrees of freedom. If the statistic calculated above is greater than the value from the table, the null hypothesis is regarded as sufficiently improbable that it can be rejected. As an example, for 100 samples, 50 independent variables used in a regression analysis, and a significance level of 0.01 for a one-sided test, an obtained value for r of greater than 0.322 or less than -0.322 would be grounds for rejection of the null hypothesis. Treatment of partial rank correlation coefficients is similar.

2.5 Making Decisions

Data obtained from probabilistic analysis of process technologies can be used to answer several questions, such as:

- Is one technology preferred over another?
- Is additional research merited?
- What should be the research strategy?
- How much is additional research worth?
- Under what conditions does the decision strategy change? (How robust is the decision strategy?)

These questions can be answered using decision analysis as an analytical tool for evaluating alternative technology options and research strategies. Decision analysis techniques are discussed in many texts (e.g., Clemen, 1988; Dawes, 1988; Watson and Buede, 1987). First, we will consider decisions based on a single attribute, such as expected cost savings, and then briefly consider a more detailed decision model incorporating the risk attitudes of a decision maker and the time value of research outcomes. The approach can be extended to consider other decision attributes.

There are two general types of decisions that are considered in technology evaluation. The first is a choice between an advanced process based on current knowledge of the process and a conventional technology. The second type of decision is that regarding specific options for further research and development of the advanced process. It is assumed here that additional research would not eliminate uncertainty in the advanced process. Instead, research is expected to reduce the range of uncertainties in one or more uncertain model input variables. This would affect the uncertainty in the performance or cost of the technology. Thus, additional research would change the probability distribution for the difference in cost between the advanced and conventional technologies.

As discussed previously, the levelized cost of electricity is perhaps the most important attribute that a decision maker would consider when comparing two competing technologies. The technology with the highest probability of lower cost would usually be preferred by a decision maker. However, there may be a possibility that the advanced technology could result in higher costs, even though it may be *likely* to have lower costs. A decision-maker may be adverse to the possibility of a bad outcome. A sufficiently risk averse decision maker may prefer to stay with the conventional technology, rather than take a risk of higher cost from the new technology.

The decision model can easily be refined to consider the risk attitude of a particular decision maker using expected utility, rather than expected cost savings, as the basis for decision making. Utility is a measure of the personal value a decision maker places on a specific outcome, and it may differ from the monetary value of the outcome (Dawes, 1988). Furthermore, because the results of research may not be obtained for 5 to 15 years, the time value of the outcomes can be modeled using discounting. One possible utility function for such a decision model is thus:

$$u(x) = \left\{ \frac{x(i,n) - x_l(i,n)}{x_h(i,n) - x_l(i,n)} \right\}^b \tag{5}$$

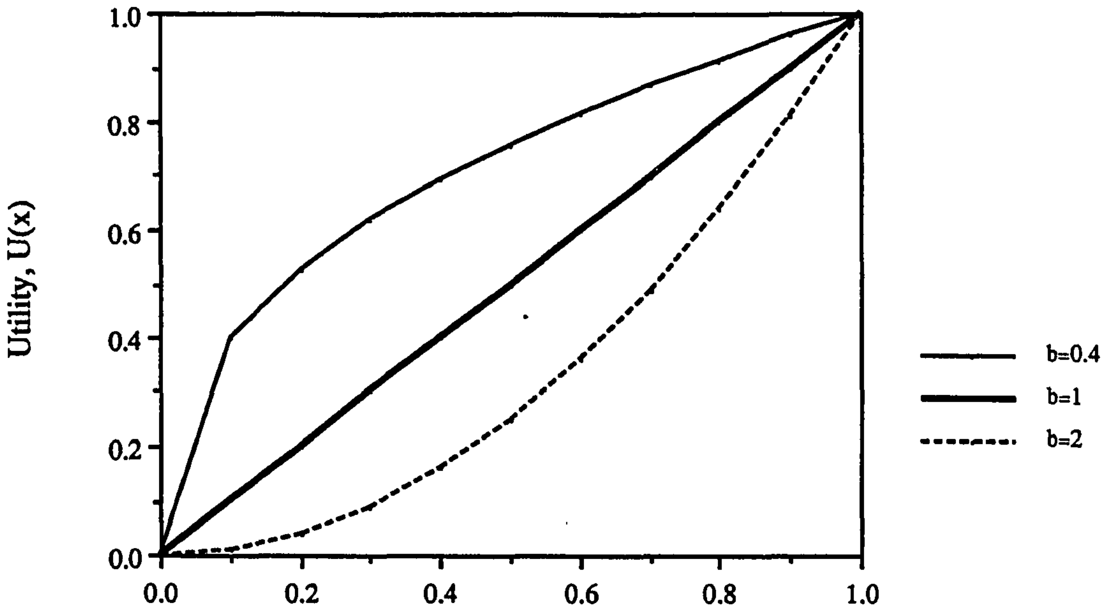
where,

$$x(i,n) = \frac{x}{(1+i)^n}$$

and,

x	=	discounted outcome of a given alternative
i	=	discount rate
n	=	time period (years)
x_l	=	lower limit of x for all alternatives
x_h	=	upper limit of x for all alternatives
b	=	risk attitude exponent

For a risk neutral decision-maker, b=1. A risk averse decision maker prefers a sure outcome over an alternative with a slightly higher expected value and a risk of a loss.



Normalized Value of Outcome, x

Figure 5. Power Utility Function

Thus, a risk averse decision maker tends to be "conservative." For a risk averse decision maker, b<1. Conversely, a risk seeking decision maker is willing to forfeit an increase in expected value to play a riskier game, and in this case b>1. The utility function is plotted for normalized values of x and selected values of b in Figure 5. A nominal value of b=0.6 (risk averse) is considered a reasonable assumption in expected utility analysis.

When comparing two technologies, the timing of outcomes may be important. The conventional technology assumed as a baseline in a comparison may be available immediately. The advanced technology may be available in the near term, but with considerable uncertainty in performance and cost. Alternatively, additional research may be conducted on the new technology to reduce the uncertainties. However, the pay-offs from such research may take 5 to 20 years. Therefore, in making comparisons, the decision-maker may choose between the conventional technology, the advanced technology based on current information, or the advanced technology based on reductions in uncertainty that would be obtained in 5 to 20 years. The decision maker may prefer an outcome today to the same outcome in 10 years. Thus, the time preference of the decision maker may be represented using discounting, as indicated in Equation (5).

Decisions regarding technology selection and research planning may be sensitive not only to the comparisons of performance and cost between the two technologies, but also to the assumptions regarding risk attitude, timing of outcomes, and the discount rate.

In cases where more than one attribute is important to a decision maker, a multi-attribute utility function can be used. For example, if differences in emission rates for key pollutants are criteria for selecting one technology over another, in addition to cost criteria, an additive utility function might be used which weighs each criteria according the decision maker's preference (e.g., Clemen, 1988).

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3.0 ENGINEERING MODELS OF SELECTED CLEAN COAL TECHNOLOGIES

This chapter provides a description of the five coal-based power generation and integrated environmental control systems selected for analysis in this research. Systems were selected on the basis of promising costs, plant performance, and emission reductions. These technologies include two options for pulverized coal fired power plants and three gasification-based systems. The engineering performance, emissions, and cost models of each system are also described.

Two different types of coal-based power generation technologies are evaluated here because of the changing nature of the technology. As discussed in Chapter 1, changes in nature of environmental regulations are responsible, at least in part, for spurring the development of new technologies. The technologies most commonly employed today for coal-based power generation are pulverized coal-fired systems. However, integrated gasification combined cycle (IGCC) systems are emerging as a promising alternative that is receiving increasing attention in the utility sector, both in the U.S. and abroad (Epstein, 1990).

These two types of systems are discussed in Sections 3.1 and 3.2, respectively. Representative technologies for each of these categories have been selected for detailed analysis in this research, involving application of the probabilistic analysis method for technology evaluation and research planning.

3.1 Pulverized Coal Power Plant Technologies

As discussed in Section 1.2.2, the current commercial state-of-the-art in power plant technology capable of stringent emission control of SO_2, NO_x, and PM is a system with wet limestone FGD for SO_2 control, low-NO_x burners, and an ESP for PM control. SCR, a technology already commercialized in Japan and Germany, can also be employed, in conjunction with low NO_x burners, to achieve an overall NO_x reduction of 90 percent compared to uncontrolled emissions. Thus, a PC power plant with FGD and SCR is the most readily available option for coal-based power generation capable of reducing both SO_2 and NO_x emissions by 90 percent. For this reason, a coal power plant with FGD/SCR is selected as the baseline technology for this study.

To meet possible future emission regulations, DOE and others have sponsored development of advanced alternative emission control systems for coal-fired power plants.

In particular, the DOE Pittsburgh Energy Technology Center (PETC) has conducted research on a number of technologies that combine SO_2 and NO_x removal into a single reactor, and that reduce the solid waste produced by air pollution control systems. Research goals for the development of advanced technologies at PETC have been system simplification, reduction in cost, and capability of 90 percent or more removal of both SO_2 and NO_x. One of these technologies, which is used as a case study in this research, is the fluidized bed copper oxide process. Key features of the copper oxide process are that, unlike a wet FGD/SCR system, (1) it combines SO_2 and NO_x removal in a single reactor vessel, and (2) it is regenerative (i.e. the reagent is reused rather than disposed of) which produces a saleable sulfur or sulfuric acid byproduct. The solid waste from a copper oxide system consists only of the fly ash collected in a conventional fabric filter.

Alternatives to conventional pulverized coal (PC) combustion, such as integrated coal gasification combined cycle (IGCC) systems, are capable of NO_x emissions comparable to those of PC plants equipped with SCR, as well as high (over 95 percent) levels of SO_2 control (Cool Water, 1986). Furthermore, political concern over acid rain (for which SO_2 and NO_x are precursors) may accelerate the time table for more stringent emission regulations of conventional PC plants. Therefore, there is incentive to develop technology options to reliably achieve stringent emission reductions at minimum cost in a timely fashion.

The modeling environment used here to evaluated the pulverized-coal based power plant technologies, the technologies themselves, and the engineering models of the technologies are described further in the following sections.

3.1.1 IECM Modeling Environment

The copper oxide process is a technology in an early phase of development, for which limited test data and no commercial design or operating experience are available. Even in the case of the conventional emission control system assumed here, SCR has not yet been applied with U.S. coals, and uncertainties remain regarding catalyst performance. Uncertainties in system performance at the commercial scale lead to uncertainties in the required size of process equipment and the consumption of materials (e.g., sorbent) and parasitic power. These uncertainties result in uncertainties in capital and operating costs, which are the ultimate measures of interest for comparative analysis. Furthermore, even if process performance were known with certainty, uncertainties regarding the costs of equipment (particularly equipment not previously used in commercial scale service) and reagents remain. To explicitly characterize these uncertainties, and to evaluate the overall

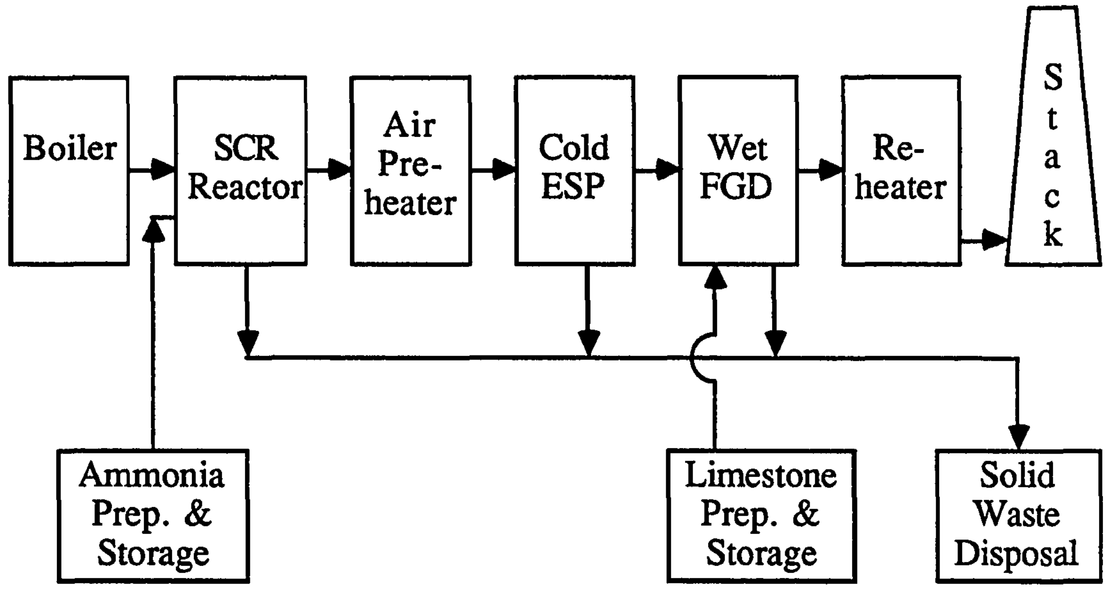

Figure 6. Simple Schematic of Pulverized Coal-Fired Power Plant with Flue Gas Desulfurization and Selective Catalytic Reduction.

uncertainty in process costs, a probabilistic engineering modeling framework has been developed.

Analytic models of the performance and cost of a variety of emission control systems for pulverized coal-fired power plants have been developed by Rubin et al (1986) as part of the Integrated Environmental Control Model (IECM). The IECM was developed under contract to PETC for the purpose of evaluating advanced technology options for emission control of coal-fired power plants, including pre-combustion, combustion, and post-combustion approaches to integrated environmental control.

A unique feature of the IECM is the ability to characterize uncertainty in model input parameters using probability distributions. This feature is obtained by implementing the analytic models within a probabilistic modeling environment developed by Henrion (Henrion, 1982; Henrion and Wishbow, 1987). The resulting uncertainty distributions for model outputs are calculated using Latin hypercube sampling, a technique described in Section 2.4.

3.1.2 Baseline Power Plant Technology

A simple schematic of a PC power plant with FGD and SCR is shown in Figure 6. In this system, coal is combusted in a pulverized coal furnace. Low-NO_x burners are used to reduce NO_x emissions by about 50 percent, compared to uncontrolled emissions using more conventional burners. The heat released from combustion is used to generate steam in a series of boilers, and the steam is used in the plant steam cycle to generate power. Before exiting the boiler, the flue gas passes through a heat exchanger called an economizer, which is used to preheat boiler feedwater going to the steam cycle. The hot flue gas exiting the boiler then passes through a number of devices, most of which are flue gas treatment devices for environmental control.

In the baseline technology assumed here, a "hot" side SCR system is employed. In the SCR process area, ammonia is injected into the flue gas upstream of a catalyst bed. NO_x in the flue gas reacts with the ammonia, in the presence of the catalyst, to form nitrogen and water vapor. The flue gas passes through the air preheater, which is used to cool the flue gas and preheat the combustion air entering the furnace. The flue gas then passes through a "cold" side (downstream of the air preheater) ESP for particulate control. The low dust flue gas exiting the ESP then passes through reactor vessels in the FGD system, where SO_2 in the flue gas reacts with a wet slurry containing limestone sorbent. Because the FGD system results in additional cooling of the flue gas, reheat is required to provide sufficient buoyancy for the flue gas to rise up the stack and disperse into the atmosphere. Particulate matter which drops out of the flue gas in the SCR reactor and which is captured in the ESP system must be collected for landfill disposal. Also, the spent sorbent from the FGD system is a "throw-away" waste stream.

Engineering performance, emissions, and cost models of the baseline power plant technology are available as components of the IECM (Rubin et al; 1986). These models characterize the key mass and energy balance for each process and major component within each process. The models track environmental species, including SO_2, NO_x, and PM. The economics models calculate capital, operating, and levelized costs in accordance with typical utility industry practice.

3.1.3 Fluidized Bed Copper Oxide Process

The fluidized bed copper oxide process is representative of dry, regenerable, combined SO_2/NO_x removal processes for coal-fired power plants. The technology is described in the next section. Then, the performance and cost models of the technology are described.

3.1.3.1 Process Overview

The copper oxide process is designed to achieve 90 percent removal of both SO_2 and NO_x from power plant flue gases. The copper oxide process combines SO_2 and NO_x removal in a single reactor vessel. The process is regenerative, producing a marketable sulfur or sulfuric acid byproduct in lieu of a solid waste containing spent sorbent. A simple

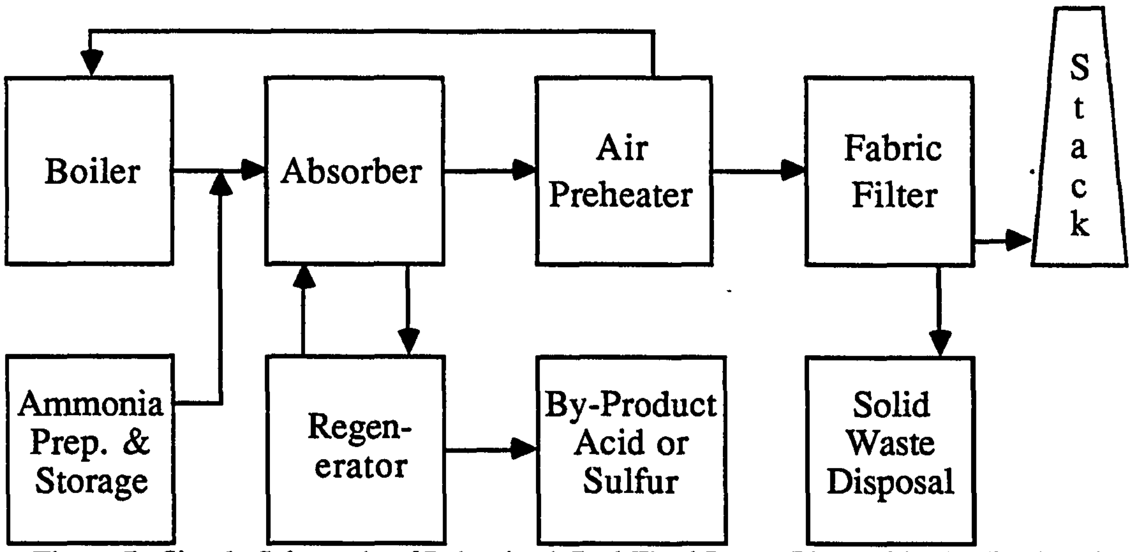

Figure 7. Simple Schematic of Pulverized Coal-Fired Power Plant with Fluidized Bed Copper Oxide Process for Simultaneous SO_2/NO_x Removal.

schematic of a power plant with the copper oxide process is shown in Figure 7. A more detailed schematic of the copper oxide process is given in Figure 8.

In a commercial-scale process, a bed of copper-impregnated sorbent, consisting of small diameter (e.g., 1/8 inch) alumina spheres, is fluidized by the power plant flue gas. Removal of SO_2 and SO_3 in the flue gas occurs by reaction with copper oxide in the sorbent:

$$CuO + SO_2 + 1/2\ O_2 \rightarrow CuSO_4 \quad (6)$$

$$CuO + SO_3 \rightarrow CuSO_4 \quad (7)$$

NO_x is removed by reaction with ammonia injected into the flue gas upstream of the absorber. The reaction is catalyzed by copper sulfate and promoted by the mixing within the fluidized bed (Drummond et al, 1985):

$$4\ NO + 4\ NH_3 + O_2 \rightarrow 4\ N_2 + 6\ H_2O \quad (8)$$

$$2\ NO_2 + 4\ NH_3 + O_2 \rightarrow 3\ N_2 + 6\ H_2O \quad (9)$$

The absorber reactions are exothermic. For a high sulfur coal, the increase in flue gas temperature may be as much as 100 ^oF for 90 percent sulfur removal (Frey, 1987). This incremental thermal energy can be recovered in the power plant air preheater, resulting in an energy credit.

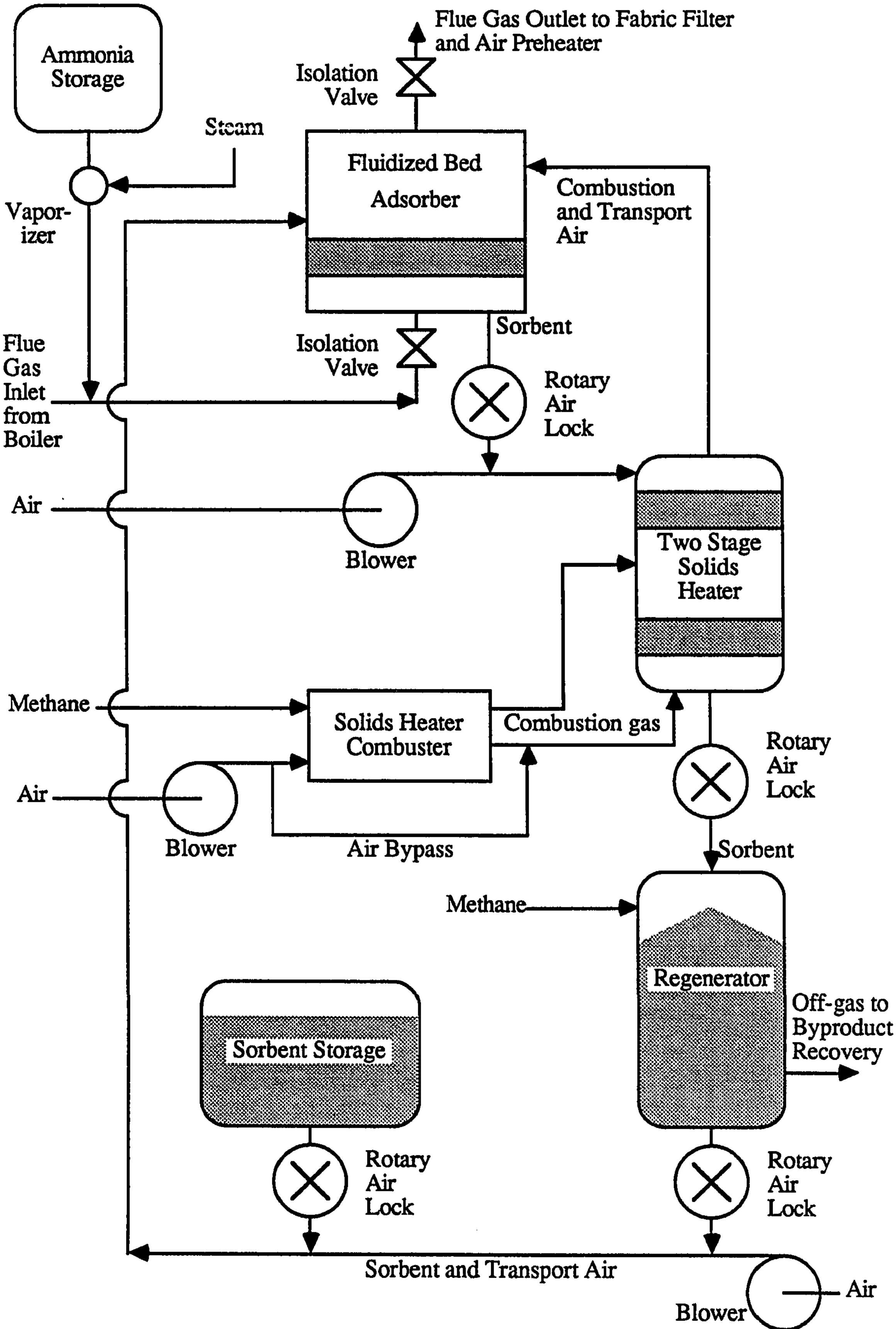

Figure 8. Schematic of the Fluidized Bed Copper Oxide Process.

The sulfated sorbent is transported from the fluidized bed absorber to a solids heater and then to a regenerator. Regeneration of the sorbent occurs by reaction with methane, converting the copper sulfate and unreacted copper oxide to elemental copper, and producing an off-gas containing sulfur dioxide:

$$CuSO_4 + {}^1/2\ CH_4 \rightarrow Cu + SO_2 + {}^1/2\ CO_2 + H_2O \tag{10}$$

$$4\ CuO + CH_4 \rightarrow 4\ Cu + CO_2 + 2\ H_2O \tag{11}$$

The off-gas is further processed to recover either sulfuric acid or elemental sulfur. The sorbent is transported back to the absorber. When the sorbent contacts the transport air an exothermic chemical reaction occurs:

$$Cu + {}^1/2\ O_2 \rightarrow CuO \tag{12}$$

Additional details of process development may be found in Radian (1984), Demski et al (1982), Yeh et al (1984), Plantz et al (1986) and Williamson et al (1987).

3.1.3.2 Process Performance Models

The copper oxide process performance model was developed by Frey (1987) as part of the IECM. It includes the fluidized bed absorber, sorbent heater, regenerator, solids transport system, and ammonia injection system. The model also characterizes the performance of an integrated sulfuric acid or sulfur plant, and the power plant air preheater. Material and energy balances are developed for 12 chemical species that are traced throughout the system. These species include components of the flue gas, sorbent, transport air, regenerator off-gas, and a number of other process streams. The mass balances account for the stoichiometry and conversion rate of reactions (6) through (12). An energy balance for each process area is calculated based on the mass balance, heat of chemical reactions, and enthalpy data for each chemical species. Details of the performance model are available elsewhere (Frey, 1987). However, the copper oxide process model was modified as part of the current work. These modifications include replacement of a regression model for estimating the copper-to-sulfur molar ratio with a kinetics-based sulfation reaction model, and the addition of a performance and cost model of an elemental sulfur recovery plant. These modifications are discussed here briefly and in Appendix A.1 in more detail.

The sorbent requirement for sulfur absorption determines the required sorbent flow rate. The sorbent flow rate is, in turn, a key parameter that affects the mass and energy balances, sizing, and cost of most components of the copper oxide process. The total sorbent mass flow rate, including copper oxide, unregenerated copper sulfate, and the alumina substrate, is given by:

$$m_s = \left(\frac{W_{CuO}}{X_{Cu}}\right)[M_{CuO} + M_{CuSO_4}(1 + 1.260X_{Cu})] \tag{13}$$

where

$$M_{CuO} = \frac{1}{r}(M_{SO_2} + M_{SO_3}) \tag{14}$$

and

$$M_{CuSO_4} = \left(\frac{\eta_s}{\eta_r}\right)(1 - \eta_r)(M_{SO_2} + M_{SO_3}) \tag{15}$$

The conversion rates of reactions (6) through (9) in the fluidized bed absorber are based on the emission control requirements. The available copper-to-sulfur (Cu/S) molar ratio required to achieve a specified SO_2 reduction requirement is estimated based on a first-order sulfation reaction kinetics model developed by PETC (Yeh and Drummond, 1986). This model assumes ideal flow in a plug flow reactor and may be written as:

$$\eta_s = \frac{\exp[B(1\text{-}r)] - 1}{\exp[B(1\text{-}r)] - r} \tag{16}$$

where,

$$B = \frac{kDAGZC_o}{W_{Cu}V_o}$$

and

$$r = \frac{M_{SO_x}}{M_{CuO}}$$

The PETC model accounts for the molar ratio of sulfur oxides to inlet available copper, r, and.the amount of available copper oxide initially resident in the fluidized bed per unit molar gas flow, B. Since incomplete regeneration reduces the available copper relative to fresh sorbent, the PETC model must be modified to explicitly include the effect of regeneration efficiency on the Cu/S ratio requirement. An explicit representation for C_o, the weight ratio of copper oxide to alumina in the substrate, can be given as a function of the regeneration efficiency:

$$C_o = \left(\frac{X_{Cu}}{R - X_{Cu}}\right)\frac{1}{\left[1 + \left(\frac{\eta_s}{\eta_r}\right)(1 - \eta_r)\,r\right]} \tag{17}$$

This expression for C_o is used in Equation (16). For the case study, the SO_2 removal efficiency is a model parameter, and Equation (16), with the modified value of C_o, is used to solve for the required copper-to-sulfur molar ratio. The modified equation is solved numerically.

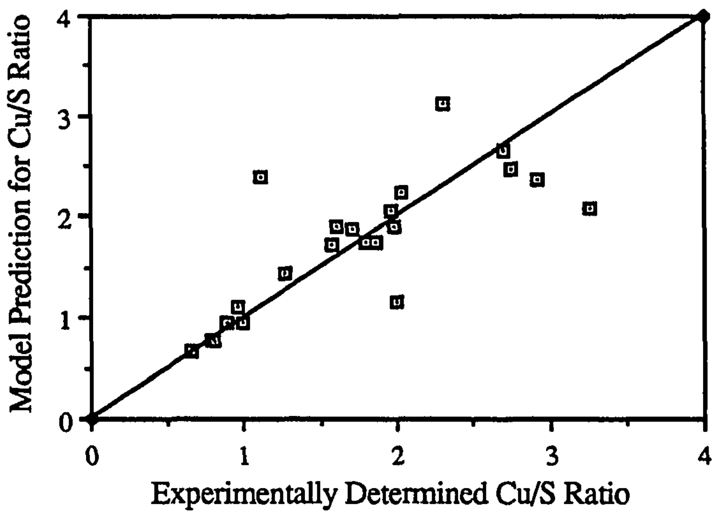

Figure 9. Comparison of Experimental and Predicted Copper-to-Sulfur Molar Ratio.

Figure 9 shows model predictions for the copper-to-sulfur molar ratio based on experimental data reported by Yeh and Drummond (1986). The model is a function of eight variables, each of which is subject to measurement error. Thus, scatter in the plotted data is expected. The scatter in the predicted Cu/S ratio can be represented statistically by a standard error, which is an indicator of the variance in the observed Cu/S ratio that is not explained by the analytical model.

Additional detail regarding the copper oxide sulfation model is presented in Appendix A.1.

For byproduct recovery, the elemental sulfur process is capable of 95 percent sulfur conversion, in contrast to 99.5 percent recovery for sulfuric acid plant. Thus, to achieve an overall 90 percent SO_2 removal, the copper oxide process must be operated at 90.5 percent removal efficiency with sulfuric acid recovery or 94.7 percent removal efficiency with sulfur recovery. Other possible designs, such as recycling tail gas from the sulfur plant to the power plant flue gas, or sulfur plant tail gas treating, were not considered in this study. A more detailed discussion of the newly developed Claus plant model is given in Appendix A.1.

The molar ratio of ammonia-to-nitrogen oxides required to achieve a given level of NO_x removal is estimated using a regression model developed from PETC test data (Frey, 1987). The model is sensitive to the fluidized bed height and the required NO_x removal efficiency:

$$r_{an} = \exp\left(\frac{\eta_n - 0.131\ H - 0.782}{0.367}\right) \tag{18}$$

The model in Equation (18) does not explicitly account for the catalytic effect of copper sulfate that is believed to promote NO_x removal (Drummond et al, 1985). However, the catalytic effect has not been well-characterized in the available literature.

3.1.3.3 Economic Model

Direct capital costs, indirect capital costs, and fixed plus variable operating costs have been modeled for each component of the copper oxide system as part of the IECM (Frey, 1987). The direct capital costs include the absorber, solids heater, regenerator, solids transport, ammonia injection, flue gas handling, incremental air preheater sizing, and byproduct recovery plants. These costs are estimated using "capacity-exponent" scaling relationships based on key process parameters.

Indirect costs, including engineering, design, supervision, contractor, construction expense, and contingencies, are based on the total direct costs. Other items included in the total capital requirement are interest during construction, royalties, pre-production costs, inventory capital, initial chemicals, and land. Fixed operating costs include operating and maintenance labor, maintenance materials, and administrative labor. Variable operating costs include makeup sorbent, ammonia, methane, electricity, and credits for byproduct sales and recovered energy in the air preheater, which reduces coal consumption.

Since the performance and cost of major components of environmental control systems are inter-related, cost comparisons between the copper oxide process and the conventional FGD/SCR technology are based on the total pollution control cost for the power plant. Thus, the total levelized costs presented here include the copper oxide process, a fabric filter particulate collector, solid waste disposal, and any incremental changes to the power plant associated with air preheater modifications and energy credits. For integrated designs employing coal cleaning, the incremental cost of the cleaned coal also is included. All costs of pollution control are expressed as an incremental cost of electricity generation, and constant dollars, which exclude inflation effects, are used throughout the analysis. The capital recovery and variable cost levelization factors are calculated using standard methods described by the Electric Power Research Institute (1986). Economic models for other pollution control systems (used for comparative analyses) are described in Rubin et al (1986).

3.2 IGCC Systems

This section discusses the basis for selecting candidate IGCC systems for detailed evaluation, and describes the engineering performance and cost models of each technology. A total of three IGCC systems were selected for case studies. While each system has unique features, they also have many process areas in common. The technical description of each system is provided in the next section. In Section 3.2.2, the performance models for the three IGCC technologies are described. In Section 3.2.3, the approach used to develop cost models for each technology is discussed. In addition, the interested reader will find complete documentation of the IGCC capital, annual, and levelized cost models in Appendix A.

3.2.1 Selection of Candidate IGCC Systems

A number of variations of IGCC power plant designs exist, based primarily on differences in the coal gasifier technology. Both oxygen and steam are necessary reactants in the coal gasification process, which produces a syngas containing carbon monoxide and hydrogen. Alternate gasifier designs may use either oxygen or air as the oxidant. The primary difference in gasifier design is the type of reactor bed in which the coal is gasified. The three generic types of gasifiers are moving-bed, fluidized-bed, and entrained-flow. In a moving bed gasifier, coal flows downward counter-current with the steam and oxidant, and the highest temperatures are reached toward the bottom of the reactor. A prominent example of this type of gasifier is the Lurgi design. In a fluidized bed reactor, the coal is well-mixed with steam and oxidant, leading to a more uniform temperature distribution in the gasifier. The Kellogg-Rust-Westinghouse (KRW) gasifier is an example of a fluidized bed design. In an entrained flow gasifier, the coal is gasified in a plug flow reactor in which the coal and reactants move co-currently through the reactor. The Texaco gasifier is the most common entrained-flow design. The gasifier design affects the temperature of the fuel gas, composition of the fuel gas (e.g., methane content, presence of tars and oils), ability to handle certain coals (e.g., caking coals), ability to handle fines, and oxidant and steam requirements, among other factors (Simbeck et al, 1983).

The Electric Power Research Institute (EPRI), a privately funded research consortium in the electric utility industry, has sponsored a number of performance and cost evaluations of IGCC technologies. These include technologies based on the Texaco, KRW, Shell, British Gas Corporation/Lurgi (BGC/L), and Dow gasifier technologies (e.g., Fluor, 1983a, 1983b; Fluor, 1984; Fluor, 1985; Parsons, 1985; Fluor Daniel, 1989). Most of the studies sponsored by EPRI have focused on the entrained-flow Texaco gasifier technology. This is partly because the Texaco gasifier has more operating experience on the demonstration plant level than other technologies such as Shell, KRW, and British Gas/Lurgi (Simbeck et al, 1983). EPRI, in cooperation with others, has co-sponsored the Cool Water gasification program, the first IGCC demonstration plant, based on the Texaco technology. The Cool Water IGCC plant was in service for five years. Emissions of SO_2 and NO_x were well below both NSPS and the more stringent local emission permit limits for Cool Water (Cool Water, 1988).

An oxygen-blown (oxygen used as the oxidant) KRW-based system may offer some advantages over a comparable Texaco-based system, including reduced oxygen consumption, a lower temperature and pressure gasifier, a syngas with a higher heating value, and fewer parasitic loads (due largely to reduced oxygen consumption). Initial comparisons of Texaco and KRW based systems indicate that the heat rates (efficiency), capital costs, and levelized costs for these systems are nearly identical, even though a larger capital cost contingency factor was used for the KRW cost estimate (Fluor, 1985). A schematic of an oxygen-blown KRW-based IGCC system is shown in Figure 10.

Lurgi gasification technology is the oldest of the technologies most commonly considered for IGCC systems. The Lurgi dry ash coal gasification process was developed in the 1930s in Germany, and over 150 gasifiers have been installed internationally since, most notably nearly 90 gasifiers in South Africa. Because the Lurgi gasifier is a moving-bed design in which the coal flows countercurrently with the steam and oxidant, the temperature varies throughout the reactor. The exiting gas temperature is lower than for other gasifier designs, and the "cold gas efficiency" (percent of chemical energy in the coal contained in the syngas) is higher than for other gasifier designs (SFA, 1983). However, the syngas typically contains oils and tars, which must be removed from the syngas in conventional IGCC designs to avoid deposition on downstream equipment. The removal of tars and oils reduces the heating value of the syngas, and requires additional scrubbing equipment, increasing capital costs (Parsons, 1985; Bechtel et al, 1988c). Alternatives to conventional gas cleanup systems may eliminate the requirement for syngas cooling in Lurgi-based systems, therefore preventing the condensation of tars and oils and eliminating the associated gas scrubbing equipment, resulting in significant cost savings (Corman, 1986).

The U.S. Department of Energy's Morgantown Energy Technology Center (DOE/METC) has sponsored, and continues to sponsor, a number of system analysis studies to identify potentially promising advanced IGCC process configurations and to

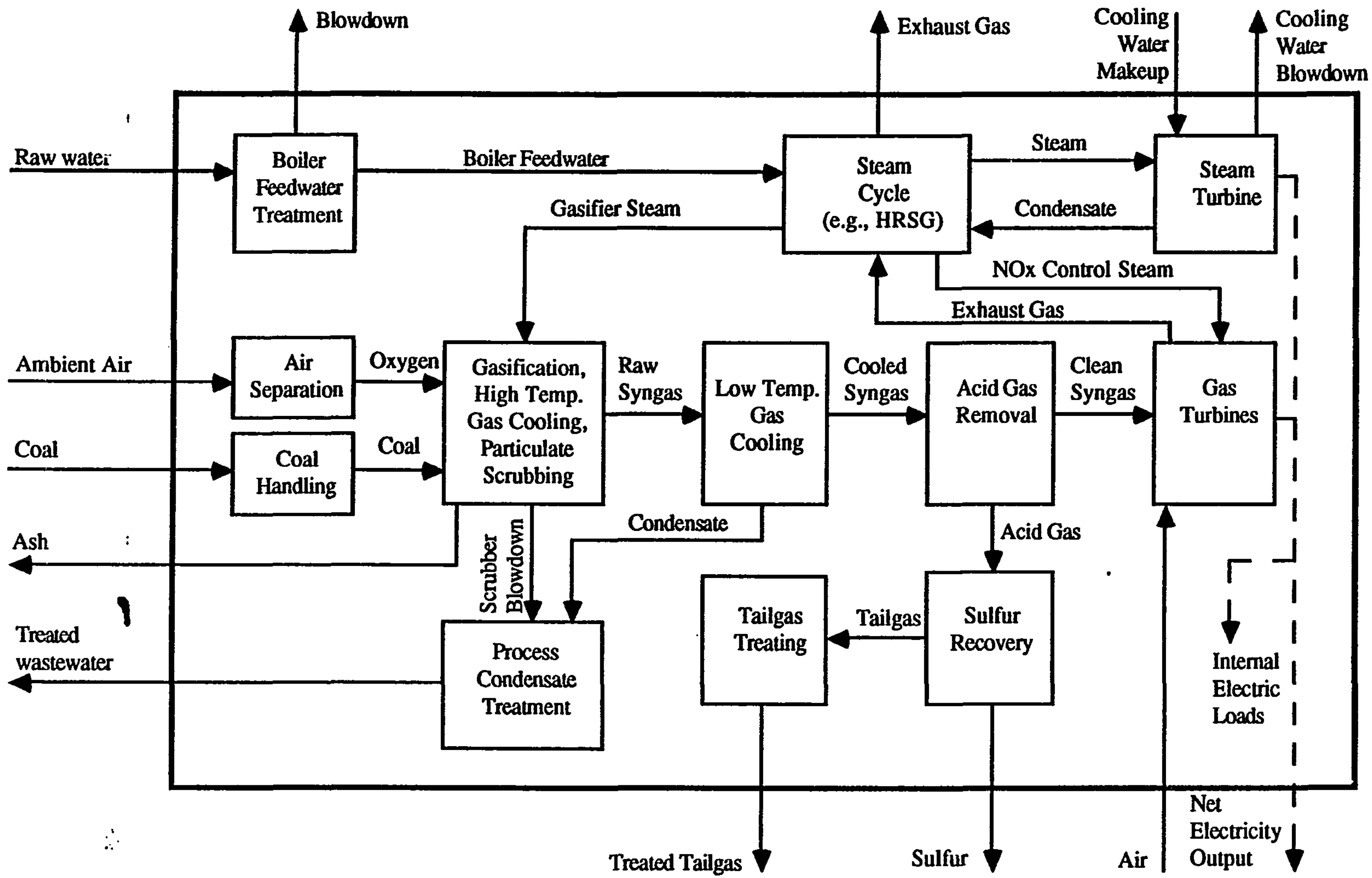

Figure 10. Schematic of Oxygen-Blown KRW-based IGCC System with Cold Gas Cleanup

provide performance and, in some cases, cost data for these. These include conceptual design studies of new and repowered Texaco-based IGCC plants (UTC, 1983), a phased Westinghouse (now KRW)-based plant (WEC, 1983), oxygen-blown Lurgi-based systems for power generation and synthetic natural gas production (Cincotta, 1984; Zahnstecher, 1984), simplified air-blown (air used as the oxidant) systems using Lurgi gasifiers and "hot" gas cleanup (Corman, 1986), and the performance and cost of "hot" gas cleanup and sulfur recovery systems (Klett et al., 1986; O'Hara, Chow, and Findley, 1987), as opposed to the lower temperature "cold" gas sulfur removal systems assumed in EPRI studies. A METC-sponsored study, prepared by Southern Company Services, of air-blown KRW-based IGCC systems featuring "hot" gas cleanup will soon be published, as of Spring 1991.

Conventional IGCC designs, such as that of the Cool Water demonstration project, are based on "cold" gas cleanup, in which the fuel gas from the gasifier is cooled to a sufficiently low temperature (e.g., 100 °F) that the Selexol or similar sulfur removal process can be used to separate H_2S from the fuel gas. A focus of research at the METC is the development of "hot" gas cleanup systems, in which sulfur compounds may be removed from the gasifier or the fuel gas at high temperature (e.g., 1,000 °F). Hot gas cleanup eliminates the capital cost associated with heat exchangers needed to the cool the fuel gas and process condensate treatment systems needed to handle condensate resulting from fuel gas cooling. Hot gas cleanup also reduces the thermal efficiency penalty associated with gas cooling, allowing the sensible heat of the high temperature fuel gas to be supplied directly to the gas turbine.

One of the most promising hot gas cleanup configurations is an air-blown Kellogg-Rust-Westinghouse (KRW) IGCC system. A schematic of this technology is shown in Figure 11. The hot gas cleanup system system features in-bed desulfurization in the fluidized bed gasifier with limestone or dolomite, subsequent sulfur removal from the fuel gas with a zinc ferrite sorbent, and high efficiency cyclones and ceramic filters for particulate removal. The off-gas from the zinc ferrite reactor, which contains sulfur compounds, is recycled to the gasifier. The advantage of such a system compared to a base case oxygen-blown system with cold gas cleanup is that, (1) it does not require an expensive and energy consuming oxygen plant, (2) it eliminates the capital costs associated with sulfur recovery (all sulfur is disposed with the spent limestone or dolomite), and (3) it reduces the amount of fuel gas cooling required prior to combustion in the gas turbine, thereby improving the plant thermal efficiency.

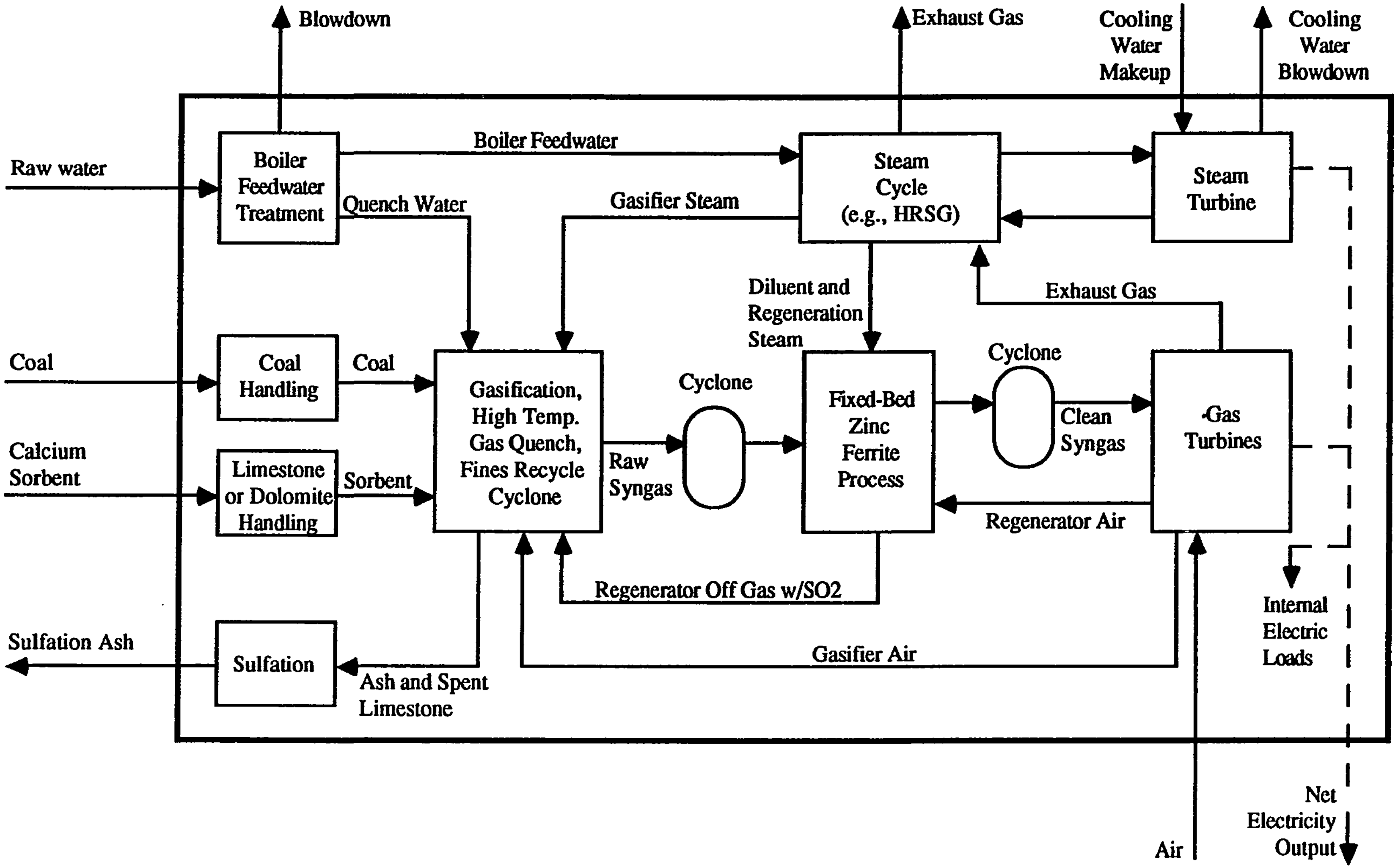

Figure 11. Simplified Schematic of Air-Blown KRW IGCC System with Hot Gas Cleanup

Testing of an air-blown KRW-based system with hot gas cleanup at the process development unit (PDU) level has been conducted (KRW, 1988). M.W. Kellogg has presented some results of a performance and cost analysis of such a system, although no detail was provided on costs (Banchik and Cover, 1988). M.W. Kellogg and Bechtel, under a cooperative agreement with DOE as part of the clean coal technology program, began design of a 63.5 MW demonstration plant of a KRW gasifier IGCC system with hot gas cleanup using the fixed-bed zinc ferrite process (KRW, 1988). However, the project was cancelled due to problems finding a site and obtaining financing (Gallaspy, 1990).

In the fixed-bed zinc ferrite process, sulfur is removed from the syngas by reaction with a sorbent consisting of zinc ferrite pellets. Absorption occurs until just before "breakthrough", at which point the sorbent is saturated. The absorber is then taken off-line, and the syngas is diverted to another zinc ferrite reactor vessel containing regenerated sorbent. Sulfided sorbent is regenerated using air as a reactant and steam as a diluent, to prevent the heat released in the exothermic regeneration reactions from sintering the sorbent. The regeneration off-gas containing sulfur dioxide is then recycled to the gasifier, in KRW-based designs. Because a significant amount of U.S. government resources have been committed to developing the air-blown KRW-IGCC in-situ desulfurization hot gas cleanup concept, it is appropriate to consider this as an important case to include in model development and technology evaluation.

METC has also sponsored development of IGCC systems with hot gas cleanup based on the Lurgi gasification technology. A schematic of this technology is shown in Figure 12. The higher cold gas efficiency of Lurgi gasifiers compared to other gasifiers assumed for IGCC design can result in a higher plant efficiency, because a larger portion of the energy input enters the combined cycle system through the fuel gas rather than only through the steam cycle. The conversion efficiency of energy entering the gas turbine is much higher than that of energy entering the steam cycle. The exit temperature of syngas from a Lurgi or similar gasifier provides a more direct match with the temperature window of hot gas cleanup systems, thereby eliminating any requirement for syngas cooling. Lurgi-based IGCC systems with hot gas cleanup therefore offer the potential for simplified plant designs. General Electric, under contract to METC, has been involved in analysis, testing, and development of hot gas cleanup systems for simplified Lurgi-based IGCC plants. These efforts include conceptual cost and design studies (Cincotta, 1984; Corman, 1986), proof-of-concept system design studies (e.g., Smith, 1987), and construction of a proof-of-concept system for a moving bed gasifier with hot gas cleanup (e.g., Cook, 1989).

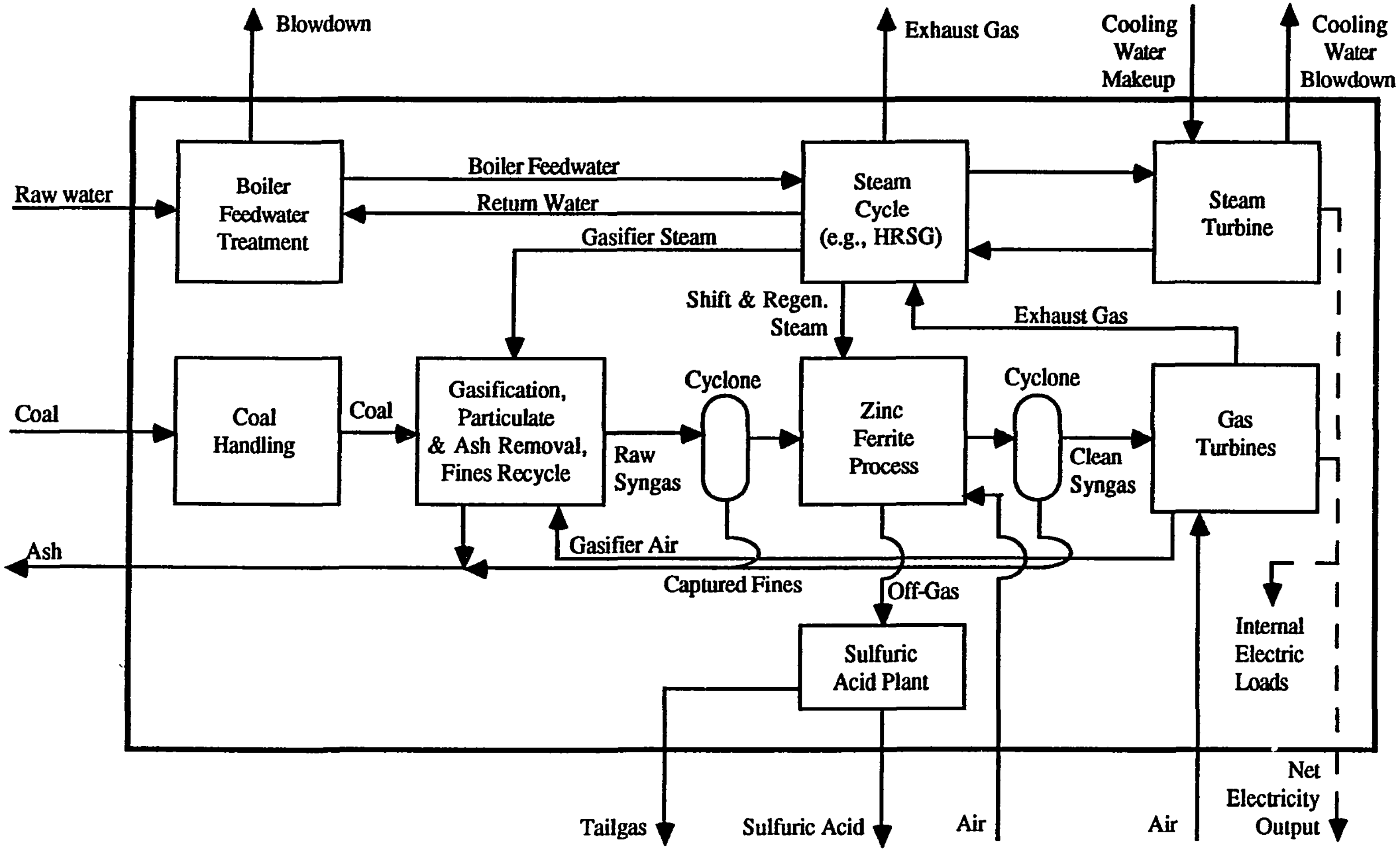

Figure 12. Schematic of Air-Blown Dry-Ash Lurgi Gasifier IGCC System with Hot Gas Cleanup

The focus of the General Electric research program is testing of a moving-bed zinc ferrite desulfurization system, in which sorbent circulates continuously between an absorber and regenerator vessel, as opposed to the fixed-bed system in which the sorbent remains in one vessel which is cycled between absorption and regeneration duty. The moving bed design offers advantages in terms of a steady flow of regeneration off-gases and the elimination of a requirement for steam as a diluent (Smith, 1987). However, at this time, only limited design data and no detailed cost data are publicly available for this proprietary system.

Based on a review of published design studies and research efforts, three IGCC technologies were selected for detailed evaluation. These include one system featuring cold gas cleanup up, which is intended to be representative of conventional IGCC technology, and two advanced alternatives featuring hot gas cleanup, representing innovative process technologies. The advanced systems differ in the approach used for gasification and fuel gas desulfurization. These systems are:

- Oxygen-blown KRW-based IGCC with cold gas cleanup (Figure 10)
- Air-blown KRW-based IGCC with hot gas cleanup, featuring gasifier in-bed bulk desulfurization and external fuel gas polishing desulfurization using the fixed bed zinc ferrite process (Figure 11)
- Air-blown Lurgi-based IGCC with hot gas cleanup, featuring external fuel gas bulk desulfurization using the fixed bed zinc ferrite process (Figure 12)

Additional details regarding the performance of these technologies is included in Appendix B. In particular, detailed technical reviews of information about the Lurgi gasifier, KRW gasifier, zinc ferrite desulfurization system, and gas turbine process areas are given in Appendices B.3, B.4, B.5, and B.6, respectively.

3.2.2 IGCC System Performance Models

Performance models developed by the U.S. Department of Energy were available and obtained as a starting point for modeling the IGCC systems studied in this research. However, several modifications to the performance models were required. Furthermore, no cost models for these systems were available, and these had to be developed. The development of cost models for all three systems is described in Section 3.2.3.

3.2.2.1 ASPEN Modeling Environment

METC has developed a number of performance simulations of IGCC systems in the ASPEN (Advanced System for Process ENgineering) modeling environment. ASPEN is a Fortran-based deterministic steady-state chemical process simulator developed by the Massachusetts Institute of Technology (MIT) for DOE to evaluate synthetic fuel technologies (MIT, 1987). The ASPEN framework includes a number of generalized unit operation "blocks", which are models of specific process operations or equipment (e.g., chemical reactions, pumps). By specifying configurations of unit operations and the flow of material, heat, and work streams, it is possible to represent a process plant in ASPEN. In addition to a varied set of unit operation blocks, it is possible to include Fortran programs within the simulation models as "Fortran blocks," or to call external Fortran subroutines. ASPEN contains an extensive physical property database, which allows a modeler to include a wide range of chemical species in the model. ASPEN also includes convergence algorithms for calculating results in closed loop systems, which are modeled using "design specification." A design specification causes part or all of the simulation model to be executed iteratively, varying the value of a selected flowsheet input variable to achieve a specified value of a key design variable within a given tolerance. These combinations of features make ASPEN a powerful tool for process simulation.

The METC IGCC performance models are used by DOE to calculate mass and energy balances for IGCC systems, to conduct sensitivity analyses of performance parameters, and to evaluate the effect of design modifications on plant performance. The IGCC designs that have been modeled to date include oxygen-blown systems with cold gas cleanup based on the Texaco, BGC/L, and KRW gasifier technologies, an air-blown KRW-based system with in-bed limestone desulfurization and hot gas zinc ferrite sulfur removal, and an air-blown Lurgi-based system with hot gas zinc ferrite sulfur removal. In some cases, such as the air-blown KRW and Lurgi based systems, documentation is not available for these models, other than the ASPEN input file. These models typically consist of about 80 unit operation blocks and 4 to 8 major flowsheet sections. While the bulk of the models are comprised of generalized unit operation blocks, there are a number of Fortran blocks and design specifications which are specific to IGCC systems or to a flowsheet. There also are user models to handle coal properties, and there is a Fortran block used as a summary report writer to concisely present plant performance results. The flowsheets have been developed in a modular approach to allow sections to be "borrowed" from other flowsheets, substantially reducing development time of new IGCC simulation models (Stone, 1985).

As noted in Section 2.4, a newly developed probabilistic modeling capability for ASPEN is available for evaluating process technologies in the face of uncertainty. This capability is utilized here for the evaluation of uncertainties in advanced IGCC system concepts.

The performance models adopted from METC for each of the three systems evaluated as part of this study are described in the following sections. However, several modifications were required. These are summarized briefly here, and in more detail in Appendix A.2.

- A new design specification was added to all flowsheets to set the gas turbine compressor inlet air flow rate based on choked conditions at the turbine inlet nozzle (see Appendix B.6 for a discussion of choked flow at the turbine inlet). This feature is required so that the size of the gas turbines in the performance model properly corresponds with the basis for the gas turbine cost model. The flowsheet input variables were adjusted to represent plant performance based on three gas turbines.

- The gas turbine model was modified by adding more a detailed representation of the cooling air circuit, which is required to accurately estimate gas turbine efficiency.

- NO and NO_2 were added to the component list of all the flowsheets, and chemical reactions representing both thermal and fuel NO_x generation were added to the gas turbine combustor unit operation block. This modification was needed to characterize the environmental performance of the IGCC systems more completely than in the original DOE versions of the models. With the modification, it is now possible to specify the fraction of fuel bound nitrogen (i.e. ammonia) converted to NO_x in the gas turbine combustor, the proportion of NO to NO_2, and the formation of thermal NO_x. See Appendix B.6 for more discussion of NO_x emissions from the gas turbine.

- A new Fortran block was added to provide one location for initializing the key design parameters that were previously initialized in individual unit operation blocks, Fortran blocks, and design specifications. This facilitates specification of assumptions for both deterministic and probabilistic case studies.

- A new stochastic flowsheet section was added for flowsheet control, variable assignment, and output analysis for the purpose of probabilistic simulation, using the newly developed stochastic block for the ASPEN simulator.

- A new Fortran block was added to the flowsheets for the two systems with hot gas cleanup to specify the ammonia yield from the gasifier based on the inlet coal nitrogen content. The assumptions for gasifier ammonia yield are discussed in Appendices B.3 and B.4.

- A set of unit operation and Fortran blocks were added for the two systems with hot gas cleanup to represent the consumption of fuel gas for reductive regeneration of the zinc ferrite sorbent used for fuel gas desulfurization.

- The flowsheet of the Lurgi-based system originally included a performance model of a moving-bed zinc ferrite desulfurization system. This was replaced by a model of the fixed-bed zinc ferrite system "borrowed" from the flowsheet of the KRW-based system with hot gas cleanup.

- A set of unit operation models was added to the air-blown KRW-based system to represent the sulfation unit, including waste heat generated from oxidation of calcium sulfide in the spent limestone sorbent to calcium sulfate and combustion of residual carbon in the gasifier ash. The oxidation of calcium sulfide is required prior to waste disposal, as discussed in Appendix B.4.4.7. The waste heat (less boiler losses) is provided to the plant steam cycle for energy recovery.

- Information regarding environmental discharges from the IGCC systems are summarized using Fortran blocks. These include plant SO_2, NO_x, CO, and CO_2 emissions, as well as solid waste and wastewater discharges, plant water consumption, and byproduct production.
- Detail regarding plant auxiliary power requirements is modeled as part of the newly developed cost models. The auxiliary power requirements affect the net plant thermal efficiency. These models are discussed in Appendix A.7.4.

3.2.2.2 Oxygen-Blown KRW IGCC With Cold Gas Cleanup

The ASPEN simulation of the oxygen-blown fluidized-bed KRW IGCC system with cold gas cleanup was originally based on a conceptual design of a 570 MW plant (Bechtel and WE, 1983c), and is documented in a METC report (Stone, 1985). An updated version is based on a more recent study (Fluor, 1985), but is not documented. The only information about the recent flowsheet is the input file itself (Stone and Craig, 1988). The main difference between the earlier and recent version of the flowsheet is in the pressure levels assumed in the steam cycle. Figure 13 shows a diagram of the flowsheet sections for the ASPEN flowsheet. The components or characteristics included in each flowsheet section, the number of unit operation blocks used in the ASPEN simulation, and a brief description of the function of each section are listed in Table 1. The material stream flows that link each flowsheet section are shown in Figure 13. Not shown in detail are "information" streams that show the flow of heat from various flowsheet sections to the steam cycle. As discussed above, several modifications have been included in the performance model regarding the characterization of the emissions from and performance of the gas turbine.

3.2.2.3 Air-Blown KRW IGCC With Hot Gas Cleanup

An ASPEN flowsheet of a KRW-IGCC with in-bed limestone sorbent, zinc ferrite hot gas cleanup, an advanced gas turbine (GE MS7000F) and a reheat steam cycle has recently been developed at METC. The only current documentation of this model, however, is the ASPEN computer code (Craig, 1988). The original computer model included 73 unit operation blocks in four major plant sections. From examination of the ASPEN input file, a simple conceptual diagram of the flowsheet was developed and is given in Figure 14. This flowsheet is considerably simpler than the system with cold gas cleanup, reflecting the simpler design of hot gas cleanup configurations of IGCC systems. Some features of the model are listed in Table 2. The features include the modifications to the performance model implemented for this study, which were discussed previously.

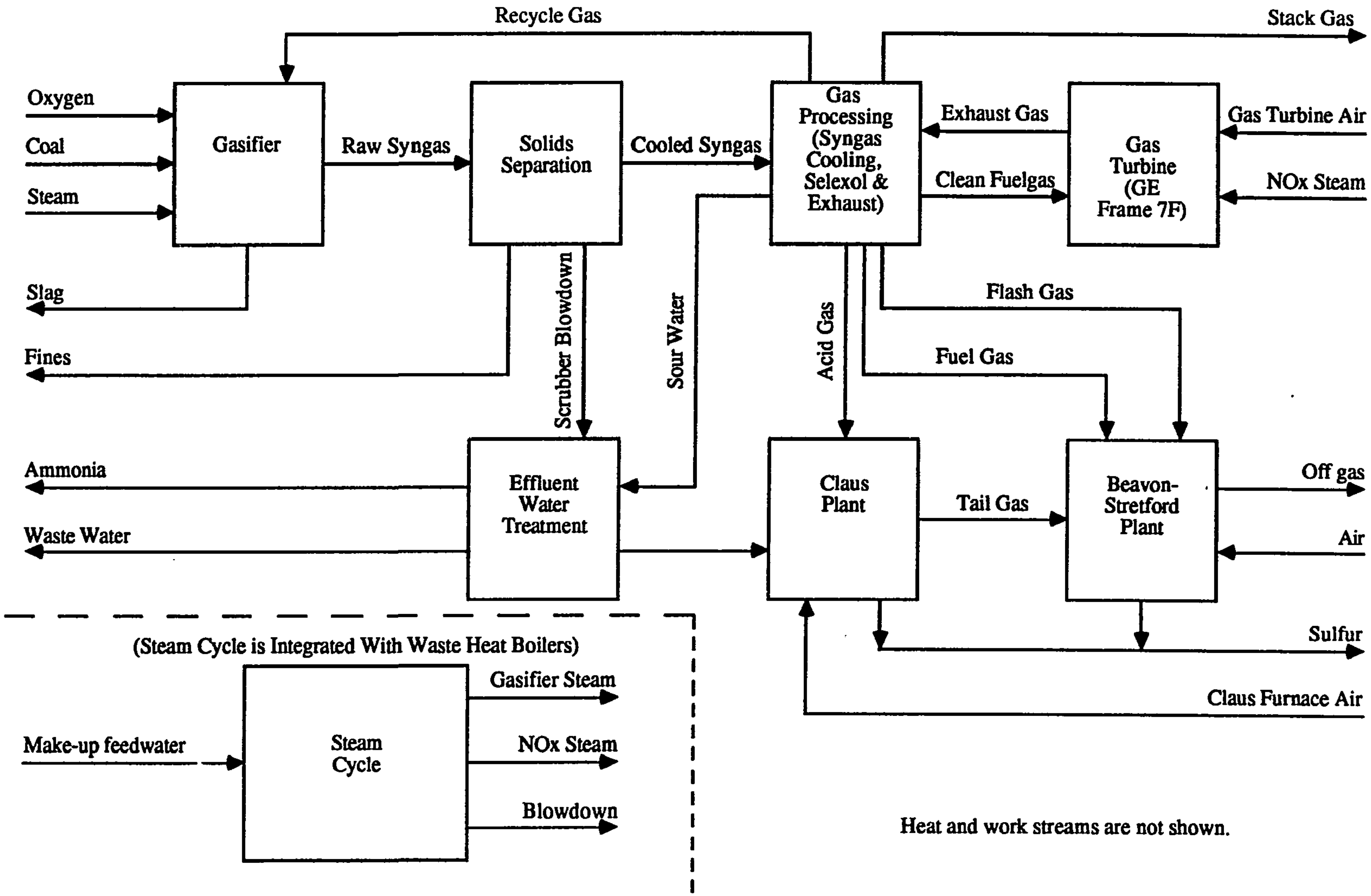

Figure 13. Conceptual Diagram of ASPEN Flowsheet for Oxygen-Blown KRW IGCC With Cold Gas Cleanup

Table 1. METC ASPEN Model of KRW-Based IGCC System: Plant Sections and Unit Operations

Flowsheet Section[a]	Components/Processes/Comments
Gasifier (6)	Coal decomposed based on ultimate analysis; Stoichiometric reactor model for slag and fines; Mix decomposed coal, steam, oxygen, recycle; Equilibrium reactor model for gasification; Ash removal after cyclones.
Solids Separation (4)	Gas cooling Particulate scrubbing to separate liquids and solids from gas, and split effluent to fines and purge water; Cooling (to reheat clean fuel gas); Water vapor knock-out, recycled to scrubber.
Gas Processing (8) (Cleanup and cooling)	Ammonia separation; Fuel gas recycle (to gasifier); ammonia/H_2S to effluent water treatment system; Selexol: acid gas removal to Claus plant, flash gases to Beavon-Stretford unit; Fuel split: fuel gas to Stretford unit; Reheat fuel gas (from Solid Sep. cooling); Cooling of gas turbine exhaust in high-pressure and low pressure HRSG.
Gas Turbine (18)	Reheated fuel gas mixed with steam and compressed air; Mixture burned in stoichiometric reactor; Multi-stage compression of air, bleeds for turbine blade cooling (added more detail); Multi-stage expansion of combustion gases; Heat loss and pressure drop considered. New design specification for estimating compressor inlet air mass flow rate. NO_x Emissions (newly added)
Effluent Water Treatment (4)	Purge water (scrubber) stream class change; Mix purge water with ammonia/H_2S stream; Cool mixture, separate wastewater, H_2S (to Claus), and ammonia.

(Continued on next page)

Table 1 (Continued). METC ASPEN Model of KRW-Based IGCC System: Plant Sections and Unit Operations

Flowsheet Section[a]	Components/Processes/Comments
Claus Plant (7)	Compress air, mix with acid gas and H_2S from water treatment section; Three stoichiometric reactor models to convert 95 percent of H_2S to elemental sulfur; Waste heat boiler; Separate sulfur and tail gas.
Beavon-Stretford Unit (7)	Compress air, compress Claus tail gas; Mix air, tail gas, fuel gas, and flash gas from the Selexol unit; Three stoichiometric reactor models to eliminate remaining H_2S; Separate sulfur and off-gas.
Steam Cycle (33)	Linked to other sections by heat streams; Condenser, makeup water, pump, deaerator;
HRSG (9)	Steam sides of HRSG and associated pumps;
Auxiliaries (11)	Heat recovery from Claus, Stretford, effluent water treatment;
Steam Turbine (7)	Four stage, no reheat; Steam from or split to Selexol, water treatment, and ammonia removal.

[a] Numbers in parenthesis are number of unit operation blocks in each ASPEN flowsheet section

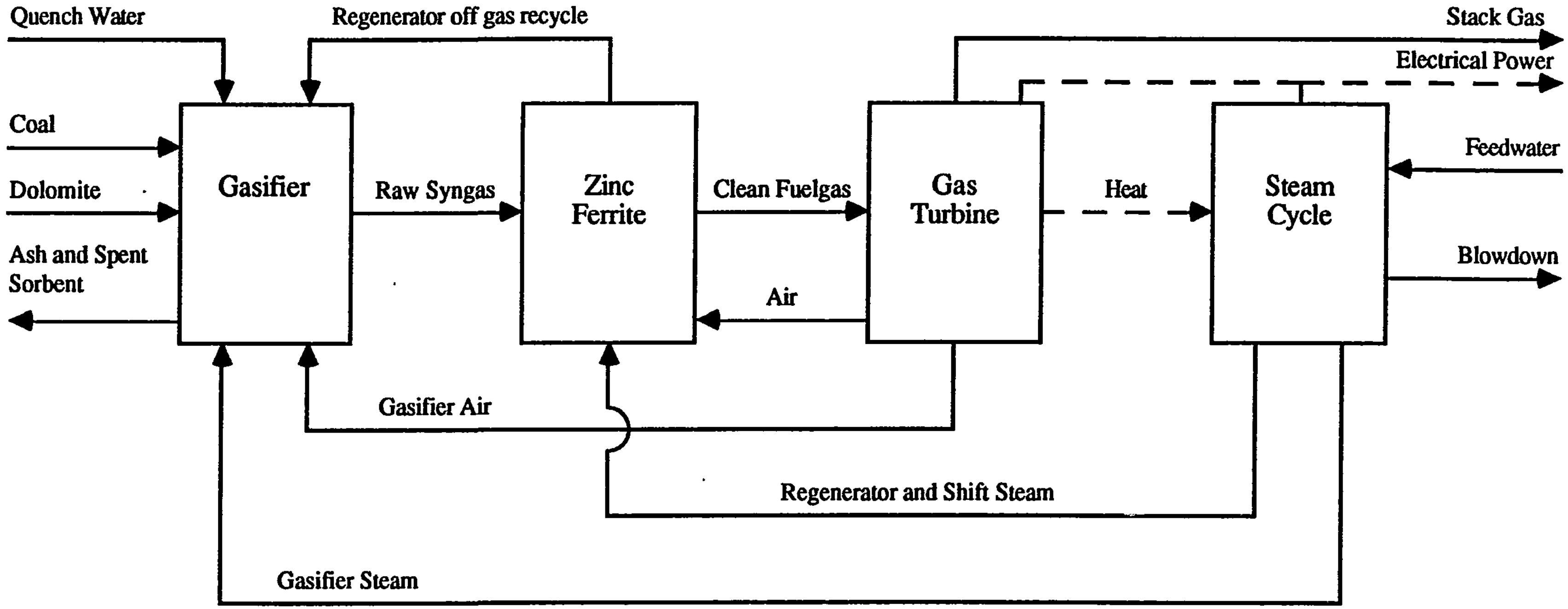

Figure 14. Conceptual Diagram of ASPEN Flowsheet of Air-Blown KRW IGCC System with In-Bed Desulfurization and Zinc Ferrite Process

Table 2. METC ASPEN Model of an Air-Blown KRW IGCC System With Hot Gas Cleanup: Plant Sections and Unit Operations

Flowsheet Section[a]	Components/Processes/Comments
Gasifier (10)	Air-blown KRW gasifier with in-bed desulfurization; characterization of fines, slag, and dolomite based on chemical analysis, in-bed desulfurization reaction based on specified conversion rate, gasification reactions based on equilibrium reaction model, recycle gas, raw gas quench cooling; newly added Fortran block for gasifier NH_3 yield.
Sulfation (4)	Oxidation of a user specified fraction of calcium sulfide in spent sorbent from gasifier to $CaSO_4$; conversion of a user-specified fraction of carbon retained in gasifier bottom ash; generation of waste heat, with a specification for boiler efficiency, for the plant steam cycle.
Zinc Ferrite (18)	Two-vessel zinc ferrite desulfurization system;
Stream Class Change (6)	ASPEN feature for handling gases and solids;
Absorption (5)	Mixing of steam and raw syngas for shift equilibrium to protect sorbent , sulfidation reactions;
Regeneration (7)	Mixing of diluent steam, regeneration air, and spent sorbent, preheating of inlet gas streams against outlet gas streams, regeneration reactions; newly added blocks for reductive regeneration using fuel gas.
Gas Turbine (26)	General Electric model MS7000F Gas Turbine with air extraction; boost compressor for gasifier and zinc ferrite regeneration air; and gas-side heat exchange in HRSG;
Compressors (6)	Three stage compression, pressure ratio of 12, extraction for turbine vane cooling; New design specification for estimating compressor inlet air mass flow rate.
Combustion (5)	Stoichiometric reactor model for combustion reactions, heat loss, temperature control for gas turbine firing temperature (e.g., 2,300 ^{o}F);
Turbines (7)	Two stage expansion, inlet vane cooling air; NO_x Emissions (newly added)
Booster (4)	Booster precoolers/compressor for gasification air
HRSG (4)	Exhaust gas cooling.
Steam Cycle (22)	Steam cycle, including steam-side HRSG, steam turbine, condenser, and deaerator.
HRSG (6)	1545 psia and 570 psia steam generation, 1465 psia superheated steam at 975 ^{o}F, 510 psia reheated steam;
Steam Turbine (8)	Reheat steam turbine;
Miscellaneous (12)	Deaerator, condenser, pumps, splitting of steam flows to gasifier, zinc ferrite regenerator, and zinc ferrite absorber shift reaction.

[a] Numbers in parenthesis are number of unit operation blocks in each ASPEN flowsheet section

A notable modification to the structure of the flowsheet is the addition of unit operations representing the sulfation unit, which is required to treat the spent limestone prior to waste disposal. The sulfation unit also is used to recover energy both from sorbent oxidation and from combustion of unconverted carbon in the gasifier ash. This energy is recovered as waste heat for the plant steam cycle.

The gasifier section of the flowsheet models processes associated with coal gasification including the gasifier reactor vessel, ash removal from the gasifier, recycle of syngas, and high temperature gas cooling. The gasification section includes models to represent coal in terms of its chemical components, fines, slag, and ash, and a model to decompose dolomite into components of CaO, MgO, and CO_2. Inputs to the gasifier include coal, limestone, steam, and air. An equilibrium reaction model is used to represent the reactions occurring in the gasifier. A separate model is used to represent the in-bed sulfur adsorption, in which the sulfur removal efficiency is specified as an input. All solids are assumed to be removed from the syngas prior to the desulfurization system. A newly added Fortran block and slight modification to the gasification reactor model allow the yield of ammonia in the fuel gas to be specified by the user.

The zinc ferrite section of the flowsheet includes models of sulfur absorption and sorbent regeneration based on a fixed-bed design. Prior to absorption, the raw syngas is mixed with steam to promote the water-gas shift reaction for improved catalyst performance. The shift reaction is modeled using an equilibrium reaction model. The raw syngas and steam mixture is then mixed with the zinc ferrite sorbent, and the desulfurization reactions are modeled using a stoichiometric reactor model, in which the reaction conversion rates are specified. The clean fuel gas is an output of this flowsheet section.

The regeneration of the sorbent occurs in two steps. The first step, oxidative regeneration, involves exothermic chemical reactions, requiring air as a reactant and steam as a diluent to prevent sintering of the sorbent. A stoichiometric reactor model, in which conversion rates are specified, is used to represent the regeneration reaction. The off-gas from the regenerator is cooled against the incoming diluent and recycled to the gasifier. Oxidative regeneration usually does not completely remove sulfur from the sorbent, and may leave residual amounts of zinc and iron sulfates.

The second step, reductive regeneration, involves reaction with the sulfates. This oxidized species must be reduced prior to the next absorption cycle. The sulfates may be converted to sulfides by reaction with hydrogen or carbon monoxide. Therefore, a portion of the fuel gas is used for reductive regeneration. Several new unit operation and Fortran blocks have been added to the original performance model to characterize reductive regeneration and the effect it has on fuel gas consumption.

The flowsheet section for the MS7000F gas turbine includes three stages for the compressor (including air extraction for turbine vane cooling), combustion of the syngas (including heat losses and combustor pressure drop), two stages of turbines with a 2,300 ^{o}F turbine inlet temperature, and a gas-side heat recovery steam generator model. As part of recent modifications, NO_x emissions have been characterized and additional detail has been added to the representation of cooling air circuits. The steam cycle model includes pumps, economizers, boilers, a superheater, a reheater, a deaerator, condensate pumps, condenser, and a reheat steam turbine with high pressure, intermediate pressure, low pressure, and very low pressure stages.

3.2.2.4 Air-Blown Lurgi IGCC with Hot Gas Cleanup

An ASPEN model of a fixed bed gasifier-based IGCC system with hot gas cleanup has been developed at METC. Like the KRW-based hot gas cleanup IGCC model, the only model documentation is the ASPEN input file (Klara, Rastogi, and Craig, 1988). The flowsheet includes a gasifier section, hot gas desulfurization using a fixed bed zinc ferrite process, a model of General Electric MS7000F gas turbine, and a model of the steam cycle. The model assumes that the off-gas from the zinc ferrite process is converted to sulfuric acid. There is no in-bed desulfurization in this model. A simple diagram of the ASPEN flowsheet is shown in Figure 15. Table 3 lists the features of the model. Modifications to this model are similar to the previous two. They include: a new Fortran block for specifying gasifier ammonia yield, modifications to the gas turbine model for improved characterization of performance, addition of NO_x species to the flowsheet to characterize emissions, and addition of a representation of reductive regeneration for the zinc ferrite process.

3.2.3 IGCC System Cost Models

Cost models for each of the three IGCC systems were developed based on a review of approximately 30 comprehensive conceptual design studies prepared for DOE, EPRI, and the Gas Research Institute (GRI), as well as other studies which focused on specific process components. The models provide "preliminary" estimates of process capital and operating costs based on the standard method developed by EPRI (1986). To link process

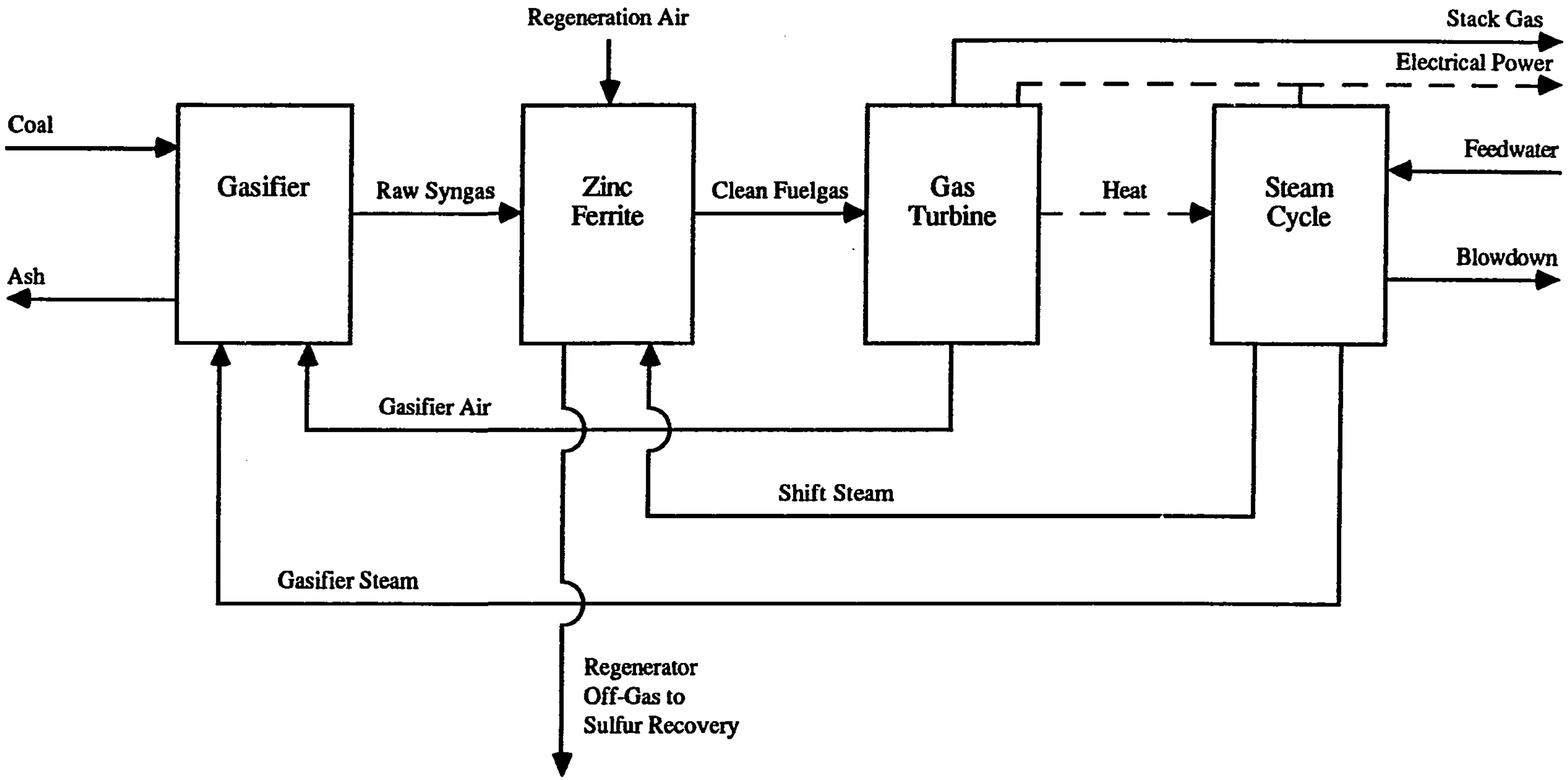

Figure 15. Conceptual Diagram of ASPEN Flowsheet of Air-Blown Fixed-Bed Gasifier IGCC System with Zinc Ferrite Process

Table 3. METC ASPEN Model of an Air-Blown Lurgi IGCC System With Hot Gas Cleanup: Plant Sections and Components

Flowsheet Section[a]	Components/Processes/Comments
Gasifier (11)	Air-blown generic fixed-bed dry-ash gasifier; characterization of coal, fines based on chemical analysis, gasification reactions based on equilibrium reaction model, allowance for specification of carbon conversion and fines carryover, provisions for in-bed desulfurization but not used in this input file; newly added Fortran block for gasifier NH_3 yield.
Zinc Ferrite (18)	Two-vessel zinc ferrite desulfurization system;
Stream Class Change (6)	ASPEN feature for handling gases and solids;
Absorption (5)	Mixing of steam and raw syngas for shift equilibrium to protect sorbent , sulfidation reactions;
Regeneration (7)	Mixing of diluent steam, regeneration air, and spent sorbent, preheating of inlet gas streams against outlet gas streams, regeneration reactions; newly added blocks for reductive regeneration using fuel gas.
Gas Turbine (26)	General Electric model MS7000F Gas Turbine with air extraction; boost compressor for gasifier and zinc ferrite regeneration air; and gas-side heat exchange in HRSG;
Compressors (6)	Three stage compression, pressure ratio of 12, extraction for turbine vane cooling; New design specification for estimating compressor inlet air mass flow rate.
Combustion (5)	Stoichiometric reactor model for combustion reactions, heat loss, temperature control for gas turbine firing temperature (e.g., 2,300 ^{o}F);
Turbines (7)	Two stage expansion, inlet vane cooling air; NO_x Emissions (newly added)
Booster (4)	Booster compressor for gasification air, compressor precoolers;
HRSG (4)	Exhaust gas cooling.
Steam Cycle (22)	Steam cycle, including steam-side HRSG, steam turbine, condenser, and deaerator.
HRSG (6)	1545 psia and 570 psia steam generation, 1465 psia superheated steam at 975 oF, 510 psia reheated steam;
Steam Turbine (8)	Reheat steam turbine;
Miscellaneous (12)	Deaerator, condenser, pumps, splitting of steam flows to gasifier, and zinc ferrite absorber shift reaction.

[a] Numbers in parenthesis are number of unit operation blocks in each ASPEN flowsheet section

flowsheet parameters with economic cost models, the methodological approach was to model all costs at the level of major plant sections for each IGCC technology. Table 4 shows a list of the sections used for the three IGCC systems. For each system, there are approximately a dozen major process sections. The cost models characterize direct and total capital costs, fixed operating costs, variable operating costs, and the annualized cost of electricity. The cost models provide a consistent basis for comparative cost analysis between baseline and advanced technology options, as well as a basis for evaluation of each technology individually. The cost models are summarized here; however, additional detail regarding the cost models is provided in Appendix A.

Cost modeling philosophy, cost data available, the approach used to implement the cost models with the ASPEN performance models, and the role and use of regression analysis for cost model development, are discussed in Appendix A.2. The direct capital cost models for the oxygen-blown KRW-, air-blown KRW-, and air-blown Lurgi-based IGCC systems are documented in Appendices A.3, A.4, and A.5, respectively. The total capital cost model, which is common to all three technologies, is documented in Appendix A.6. The fixed and variable operating cost models are documented in Appendix A.7. The annualized cost of electricity model is discussed in A.8.

3.2.3.1 Capital Cost Models

The capital cost models consist of two parts: a series of models for the direct cost of major plant sections; and a generic framework for estimating indirect and other capital costs.

The direct cost models are based on key plant performance parameters. There are 10 to 12 major plant sections per IGCC technology (excluding general facilities), as shown in Table 4. The direct capital cost of each process section was estimated separately, based on analytic relationships between direct cost and key performance parameters. These relationships were developed from published data, typically based on regression analysis (as discussed in Appendix A.2.4).

The performance parameters used in the direct cost models for the three selected IGCC technologies are summarized in Table 5. Typically, several performance variables, in addition to design variables such as the number of spare and operating trains of equipment, are included in the direct cost models. By summing the individual section costs, the total direct cost of each IGCC system is sensitive to approximately two dozen performance parameters, in addition to process design parameters.

Table 4. Major Plant Sections in the IGCC Cost Models

Area No.	Oxygen-Blown KRW with Cold Gas Cleanup	Air-Blown KRW with Hot Gas Cleanup	Air-Blown Lurgi with Hot Gas Cleanup
10	Oxygen Plant	Air Boost Compression	Air Boost Compression
20	Coal Handling	Coal Handling	Coal Handling
25	--	Limestone Handling	--
30	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasification, Coal Pressurization, Ash Depressurization
31	--	--	Coke Handling, Fines Agglomeration, Ash Removal
32	--	--	High Temperature Cyclones
35	--	Sulfation	--
40	Low Temperature Gas Cooling, Fuel Gas Saturation	--	--
50	Selexol Sulfur Removal	Zinc Ferrite Desulfurization	Zinc Ferrite Desulfurization
60	Claus Sulfur Recovery	--	Sulfuric Acid Plant
70	Tail Gas Treating	--	--
80	Steam, Condensate, Boiler Feed Water	Steam, Condensate, Boiler Feed Water	Steam, Condensate, Boiler Feed Water
85	Process Condensate Treatment	--	--
90	Combined Cycle	Combined Cycle	Combined Cycle
100	General Facilities	General Facilities	General Facilities

Table 5. Summary of Performance Parameters in the Direct Capital Cost Models

Description	Area No.	Direct Cost Model Performance Parameter(s)[a,b]: KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Oxidant Feed	10	Ambient temperature Oxygen flow rate Oxygen purity	Compressor work Precooler heat transfer area	Compressor work Precooler heat transfer area
Coal Handling	20	Gasifier coal feed	Gasifier coal feed	AR coal feed
Limestone Handling	25	--	Limestone feed rate	--
Gasification	30	Oxidant feed Coal feed (MAF) Ash removal rate	Coal feed (MAF)	Number of units (based on coal throughput)
Coke, Fines, and Ash	31	--	--	Gasifier coal feed
Cyclones	32	--	--	Syngas pressure Syngas volume flow rate
Sulfation	35	--	Ash and fines removal rate	--
Gas Cooling	40	Syngas outlet temp. Syngas outlet pres. Syngas flow rate	--	--
Selexol	50	Syngas flow rate H_2S removal eff.	--	--
Zinc Ferrite	50	--	Sorbent loading Sulfur flow rate Sorbent bulk density Syngas volume flow Design pressure	Sorbent loading Sulfur flow rate Sorbent bulk density Syngas volume flow Design pressure
Claus Plant	60	Recovered sulfur flow rate	--	--
Sulfuric Acid	60	--	--	Off-gas flow rate Concentration of sulfur dioxide Off-gas temp

(Continued on next page)

Table 5 (Continued). Summary of Parameters in the Direct Capital Cost Models

		Direct Cost Model Parameter(s)[a,b]		
Description	Area No.	KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Tail Gas	70			
SCOT		Recovered sulfur flow rate from Claus	--	--
Beavon-Stretford		Recovered sulfur flow rate from B-S	--	--
Boiler Feed Water	80	Raw water feed rate Polished water flow rate	Raw water feed rate Polished water flow rate	Raw water feed rate Polished water flow rate
Process Cond. Treatment	85	Particulate scrubber blowdown	--	--
Comb. Cycle	90			
Gas Turbine	91	Number of units (power output)	Number of units (power output)	Number of units (power output)
HRSG	92	Pressure of high pressure steam High pressure steam flow rate	Pressure of high pressure steam High pressure steam flow rate	Pressure of high pressure steam High pressure steam flow rate
Steam Turb.	93	Power output	Power output	Power output
General Fac.	100	Fraction of other direct costs	Fraction of other direct costs	Fraction of other direct costs

[a] Abbreviations are: temp. = temperature; pres. = pressure; eff. = efficiency; AR = as-received; and B-S = Beavon-Stretford; MAR = moisture- and ash-free.

[b] The direct cost models are presented in detail for these three systems in Appendices A.3, A.4, and A.5, respectively.

The direct cost models therefore represent an intermediate level of detail, in contrast to either a detailed estimate for an actual plant (involving hundreds of separate equipment items and an equivalent number of performance and design parameters), or a simplified estimate in which the total direct capital cost is estimated based on scaling of a single parameter such as plant size. The direct cost models provide sufficient detail to evaluate the effect of changes in key performance and design parameters on the total capital cost. The direct cost models can also be modified as new or more detailed information on the cost of key plant sections becomes available.

The total capital cost for an IGCC plant is estimated based on the direct costs for each plant section and a number of other capital cost items. A summary of the key parameters in the total capital cost model is given in Table 6. In addition to the parameters required for the direct cost model, the total capital cost model requires specification of approximately 60 parameters, including contingency factors, and the various factors associated with the preproduction and initial catalyst and chemical costs. These include maintenance cost factors for each plant section, number of operating personnel for each plant section, and unit costs for labor, fuel, consumables, ash disposal, and byproduct credits.

In a probabilistic analysis, process and project contingency factors are replaced with explicit representations of uncertainty in capital cost using probability distributions. See Appendix B.7.2 for more discussion of the treatment of capital cost-related uncertainties.

3.2.3.2 Operating Cost Models

The operating costs are estimated based on approximately 40 to 50 cost parameters (depending on the type of plant), which also influence the preproduction and initial catalyst and chemicals costs, as previously indicated. In addition, the material requirements for catalyst and chemicals, fuel, consumables, ash disposal, and byproduct sales are estimated based on key plant performance parameters. A summary of these parameters is given in Table 7. Therefore, the combined capital and operating cost models are based on approximately 100 performance, design, and cost parameters for each IGCC technology.

3.2.3.3 Annualized Cost Models

The total capital and operating costs are used in conjunction with an estimate of the net electric power production of the plant to estimate the cost of electricity produced by the plant. To accurately determine the net plant electrical output, estimates of the auxiliary power requirements of key plant sections were developed. The cost of electricity therefore depends on the parameters summarized in Tables 5, 6, and 7, plus an additional set of

Table 6. Summary of Key Parameters in the Total Capital Cost Model

Indirect capital cost factor (default = 0.25)
Sales tax (default = 0.06)
Engineering and home office cost factor (default = 0.15)
Cost of environmental permits (default = $1 million)
Process contingency factors (See Appendix A.6.2, Table A-7)
Project contingency factor (default = 0.20)
Interest for funds spent during construction (default = 0.10)
Inflation rate (default = 0, constant dollar basis)
Preproduction costs
    Depend on all parameters that affect operating costs
Initial catalyst and chemicals
    Depend on material requirements and unit costs for each material

Table 7. Summary of Key Parameters in the Operating Cost Model

Average labor rate (default = $19.70/hour)
Shift factor (default = 4.75)
Number of operators per shift for each process area (See Appendix A.7.1, Table A-12)
Maintenance cost factor for each process area (See Appendix A.7.1, Table A-13)
Unit costs of fuel, consumables, ash disposal, and byproduct credits (See Appendix A.7.2, Table A-14)
Mass requirements for fuel, consumables, ash disposal, and byproducts

parameters, as discussed in Appendix A.8. These additional parameters include the economic assumptions in the capital recovery factor and the performance parameters used to estimate the auxiliary power requirements. The auxiliary power requirements are discussed in Section A.7.

3.2.3.4 Cost Model Implementation

The cost models developed here are specifically intended to be coupled with an existing set of IGCC performance models developed by METC in the ASPEN simulation environment. The cost models for the IGCC systems have been coded into Fortran and implemented as subroutines along with the corresponding ASPEN performance models. The performance models determine the key material flow rates and process parameters required by the cost model to calculate capital and annual costs. Each IGCC cost model subroutine consists of a main program in which capital and annual costs are estimated based on the equations developed in the previous chapters. The cost model subroutine calls additional subroutines as needed for: (a) estimating the number of trains for selected process areas, (b) range-checking the predictive variables for the direct cost regression models, (c) estimating the fixed charge factor, and (d) estimating the variable cost levelization factor. The last two subroutines were adapted to Fortran from BASIC programs listed in the EPRI Technical Assessment Guide (1986), and they are required to estimate the total revenue requirement and the cost of electricity. Each subroutine contains extensive comments to document the code.

To implement the cost models with the ASPEN performance models, ASPEN Fortran blocks were developed to access the flowsheet performance variables required as inputs to the cost models. The Fortran blocks call the cost model subroutine and transfer the values for the performance variables via common blocks. The Fortran blocks also contain the initializations for the input parameters of the cost models, to facilitate the specification of values for both deterministic and probabilistic case studies. Key cost model results are returned to the Fortran block from the cost model subroutines. These results are collected for statistical analysis when running probabilistic simulations. The cost model subroutines record the values of the input variables and cost results to the simulation report file in a detailed summary format. Examples of the cost model outputs for deterministic analyses of each of the three IGCC systems are given in Appendix C.

3.2.3.5 Cost Model Applicability and Limitations

The cost models developed here are intended to estimate the cost of the specific IGCC systems. *These models are not intended for application to any other type of system.* For example, these models should not be used to estimate costs for coal gasification fuel cell systems, coal-to-SNG systems, or general coal gasification refinery systems. The cost models for specific process areas should not be used to estimate costs for a process environment different from the specific IGCC systems described in this study. For example, the Claus plant cost model should not be used to estimate sulfur recovery costs for petroleum refining. The design basis for each IGCC system and each specific process area should be carefully reviewed when considering application of these models to estimating costs for anything other than the specific systems described here.

Because the available data used to develop the cost models are limited in the range of coal types and stream composition, use of these models for other than the base case designs using Illinois No. 6 coal must be carefully considered. While the data used to develop the cost model for the KRW-based IGCC system with cold gas cleanup includes a variety of coals (Illinois No. 6, Pittsburgh No. 8, Wyodak subbituminous, North Dakota lignite, and Texas lignite), the data used to develop the other two IGCC system cost models are based only on eastern coals (e.g., Illinois No. 6 or Pittsburgh No. 8). Thus, the range of applicability of the gasification section cost models for the two systems with hot gas cleanup should be limited to eastern coals.

The ASPEN performance models and most of the performance and cost studies are based on the GE Frame 7F gas turbine. As described in Appendices A.3 and B.6, this is a large (150 MW) high firing temperature gas turbine. The cost of the HRSG process area, discussed in Section A.3.11, is based on large gas turbines. Therefore, these models are limited to a combined cycle configuration that includes a large gas turbine such as the GE Frame 7F.

As discussed in previously here and throughout Appendices A.3, A.4, and A.5, many of the direct cost models have been developed using regression analysis. *In general, the regression models should not be extrapolated beyond the range of the data used to develop the models.* The cases where limited extrapolation may be reasonable are noted. The limits on the range of values for the predictive variables have implications for estimating the number of trains of process equipment in each process area. As noted throughout the text, the preferred alternative to extrapolating the direct cost models is to adjust the number of trains, where possible, so that the mass flow rate per train of the key predictive variable is within the bounds of the regression model. This approach to estimating the number of trains is used in the computer models.

The computer models are designed to print warning messages when extrapolation of a direct cost model occurs. When extrapolation occurs for a direct cost model that constitutes a small portion of the total direct cost, there generally will not be a significant effect on the accuracy of the total result. However, if a model is extrapolated significantly beyond its range of validity (say, 50 percent or more above the upper limit or below the lower limit), the results may be incorrect. In such cases, the user may wish to override the assumptions regarding the number of trains in the particular process area, or take other appropriate corrective action.

In the process of developing the direct cost models, the predicted values for direct cost have been compared with published values for the purpose of validating each process area model. The models developed to estimate the consumption rate of catalysts and chemicals have been similarly compared to published estimates. The statistical measures of R^2 and standard error given in the report for each model indicate the degree to which the models replicate the data of published studies. In general, there is excellent statistical agreement, leading to good replication of published results for identical input assumptions. The use of statistical measures of goodness-of-fit for regression models is discussed further in Appendix A.2.4.

4.0 CHARACTERIZING UNCERTAINTIES IN PROCESS TECHNOLOGIES

The innovative clean coal technologies discussed in Chapter 3 represent promising new approaches for the clean and efficient use of coal for power generation, offering low levels of SO_2 and NO_x emissions. However, making predictions regarding the mature commercial-scale performance and cost of these technologies is inherently uncertain, as discussed in Chapter 1. A conceptual estimate may be intended to represent the cost of a mature fifth-of-a-kind commercial plant for the process of interest. However, such a plant may not be built for another 10 or 20 years, and currently available information for making predictions may be based on only small scale tests or theoretical models.

Yet decisions must be made today regarding which technologies to select for further research, and how to focus research on specific aspects of the technology. Historically, predictions about innovative process technologies have been biased toward optimistic outcomes, leading to potentially costly mistakes in decision-making. A feature of traditional approaches to handling uncertainty, as discussed in Section 1.1.3, is the use of simple multipliers, called "contingency factors," to represent expected cost increases associated with either process or project-related uncertainties. However, the application of contingency factors is often poorly documented. In contrast, the tendency of costs for innovative process plants to be underestimated when using contingency factors has been well-documented by RAND (Merrow et al, 1981).

Predictions about the performance and cost of innovative technologies should reflect the degree of confidence that engineers have in the input assumptions used to generate the predictions. In this research, the approach taken is to explicitly quantify both the range and likelihood of values for parameters used as inputs to the engineering models. Using probabilistic simulation techniques previously discussed, the simultaneous effect of input parameter uncertainties can be propagated through the model to yield an explicit indication of the uncertainty in output values. The uncertainty in the output variables, such as total capital cost or plant efficiency, represents uncertainty in the analyst's ability to predict performance or cost based on the limited nature of current information about the technology. Suppressing this uncertainty, as is done routinely in deterministic cost estimates, may give a misleading sense of confidence. In fact, the prediction may be uncertain by a significant range, whether the analyst chooses to acknowledge this or not. The range of uncertainty may be important information to a decision maker, and it should be explicitly considered.

The approach used to develop the estimates of uncertainty in specific model parameters in this study is discussed in the next section. Then, the parameter uncertainty estimates used for each process technology are summarized in the following section. Considerable detail on the basis for developing estimates of uncertainty in specific parameters is given in Appendix B.

4.1 Estimating Uncertainty

As discussed in Section 1.1.3 and Section 2.3.2, there are a number of types of uncertainty that an analyst faces in trying to predict the commercial scale performance and cost of an innovative process technology. The categorization of these uncertainties involves two dimensions. The first dimension is the type of uncertainty, such as statistical error, systematic error, variability, and lack of any empirical basis at all. The latter is true of concepts for which no testing has been done. The other dimension is the aspect of the evaluation that is subject to uncertainty. These aspects include process performance variables, equipment sizing parameters, process area capital costs, requirements for initial catalysts and chemicals, indirect capital costs, process area maintenance costs, requirements for consumables during plant operation, and the unit cost of consumables, byproducts, wastes, and fuel, to indicate a representative set. Model parameters in any one of these areas may be uncertain, depending on the state of development of the technology, the level of detail of the performance and cost estimate, future market conditions for new chemicals, catalysts, byproducts, and wastes, and so on.

As indicated in Section 2.3, it may not always be possible to develop estimates of uncertainty based on classical statistical analysis, nor would such an approach be appropriate in many cases. Particularly for innovative process technologies, data may be lacking regarding the sources of uncertainty a process engineer or analyst knows to exist. Thus, data analysis alone would be an insufficient basis for estimating uncertainty in a variable. When data are lacking, estimates of uncertainty must rely on the informed judgments of technical experts. Engineers are often said to have a "horse sense" about the quality of data they use in evaluations. The development of judgments about uncertainties merely requires the analyst or expert to quantify their "horse sense." As discussed in Section 2.3.3, judgments regarding uncertainties can be encoded as probability distributions.

Developing estimates of uncertainty in specific process parameters involves several steps. These include:

- Review the technical basis for uncertainty in the process
- Identify specific parameters that should be treated as uncertain
- Identify the source of information regarding uncertainty for each parameter
- Depending on the availability of information, develop estimates of uncertainty based on:
    - Published judgments in the literature (rarely available)
    - Published information, both quantitative and qualitative, that can be used to infer a judgment about uncertainty
    - Statistical analysis of data
    - Elicitation of judgments from technical experts.

Both as part of model building and uncertainty analysis, the analyst must develop an understanding of the process technology being evaluated. Published conceptual design studies or test results from small scale testing are often a valuable source of information regarding uncertainties. For example, understanding of mechanisms by which key chemical reactions occur in a process vessel, such as a gasifier, may be incompletely understood. This may result in uncertainty in predicting the amount of reagents needed to achieve a given conversion rate. These types of concerns may be discussed, if only briefly, in published studies. Estimates used in conceptual design studies may be based on extrapolation of results from small scale tests. There may be some key assumptions in such an extrapolation which are subject to uncertainty. Further, the results of testing may be subject to considerable uncertainty that could be characterized as statistic error, systematic error, or variability.

A review of published information can provide insights into the aspects of the process technology which are uncertain. In addition, technical experts can be asked which aspects of a process they think are uncertain. By developing an understanding of the underlying mechanisms that contribute to uncertainty in a process, it then becomes possible to identify the specific model parameters that should be treated as uncertain. The identification of these is specific to each process area.

In some cases, information about uncertainty can be taken directly from the literature. In rare cases, there may be explicit statements in published studies regarding high, low, and most likely values of a parameter that could be used to develop a probabilistic representation of uncertainty. More often, there may be differences in assumptions used across design studies, that reflect different judgments by process evaluators. These differing judgments can be used as information by the analyst to develop an initial representation of uncertainty for a screening study. For example, the analyst may be able to make a preliminary judgment regarding the lowest, most likely, and highest values that would be obtained for a given parameter based on a review of assumptions in design studies and the reasons given for the assumptions. Preliminary judgments regarding uncertainties in model parameters can be used in a probabilistic screening study as inputs to the engineering models. The models can then be run to identify which of the input uncertainties were most important in driving uncertainty in key output variables. Then the analyst can prioritize the input parameters for which more detailed information about uncertainty is warranted. If a particular preliminary judgment about uncertainty is found to have an influential effect in the model, then it should become a candidate for more detailed evaluation.

In an initial screening analysis of uncertainties, it is important not to prematurely eliminate potentially uncertain parameters from probabilistic treatment. For example, literature or a technical expert may indicate that a particular uncertainty is not believed to be an important determinant of uncertainty in process performance or cost. However, unless there has been a quantitative analysis to support such a conclusion, and unless the analysis was done with a sufficiently integrated performance and cost model and a sufficient variance on the uncertain parameters, there may be no reason to accept the conclusion. A screening analysis serves the purpose of identifying key uncertainties in a rigorous quantitative manner. Sometimes, results contrary to "conventional wisdom" are obtained, because conventional wisdom is often based on incomplete consideration of process interactions and potential ranges of uncertainty.

In cases were data are available to support a statistical analysis, the development of estimates of uncertainty may be straightforward. However, it is important to recognize the potential differences in the system from which the data were obtained and the system for which predictions are sought. As discussed in Section 2.3.2, there may be uncertainty in predicting information about a commercial scale system based on test results from a small scale experiment, due to statistical and systematic errors. Thus, a degree of expert judgment may be required to interpret the results of a statistical analysis for application to process evaluation.

A common approach to data analysis and model development is regression analysis. Regression models are often used by engineers to develop predictions of the mean value of a dependent (output) variable based on a set of independent (predictive) variables. However, a more appropriate perspective is that regression models are used to explain the variance in observed values of the dependent variable based on corresponding observed values of the independent variables. There is almost always a portion of the variance in the dependent variable that cannot be explained by the regression model, resulting in a "standard error" term (see Appendix A.2.4 for a detailed discussion of regression analysis and its application to process evaluation). In an uncertainty analysis, it is possible to explicitly represent the standard error of the estimate for the dependent variable with a probability distribution.

Another approach to developing estimates of uncertainties is to elicit technically-informed judgments from process experts. The approach to eliciting judgments about uncertainties is discussed in Section 2.3.3.

4.2 Estimates of Uncertainty for Pulverized Coal Plant Emission Control Technologies

For the analyses of the two emission control systems for pulverized coal fired power plants, the selection of parameters for probabilistic representation was based on a review of data, design studies, statistical analysis, and expert judgments by process developers and the author. A summary of the key parameter assumptions used in case studies is presented in Tables 8, 9, 10, and 11. Table 8 summarizes the key parameters, including emission constraints, base plant design, and financial parameters, assumed for case studies of both the conventional wet FGD/SCR system and advanced copper oxide process. The emission contraints represent the design targets assumed by process developers. Table 9 summarizes the different coals considered, including both unwashed and cleaned (30 percent sulfur reduction on an energy basis) coals. Table 10 summarizes key input values and uncertainty distributions for the conventional wet FGD/SCR emission control system. The key inputs and distributions for the copper oxide emission control system, including a fabric filter and solid waste disposal system, are summarized in Table 11. To illustrate the approaches to characterizing uncertainty in process parameters, a few examples are discussed. These examples focus on the copper oxide process. Examples are selected that illustrate a judgment made by an analyst based on published data, uncertainty estimated from statistical analysis, and a judgments elicited from an expert.

Table 8. Selected Input Parameter Assumptions for Case Studies

Model Parameter	Deterministic (Nominal) Value	Probability Distribution	Values (or σ as % of mean)
Emission Constraints			
Nitrogen Oxides	90% Reduction		
Sulfur Oxides	90% Reduction		
Particulates	0.03 lb/MMBtu		
Power Plant Parameters			
Gross Capacity	522 MW		
Gross Heat Rate	9500 Btu/kWh	-1/2 Normal[a]	(1.8 %)
Capacity Factor	65 %	Normal	(7 %)
Excess Air (boiler/total)	20 %/39 %	Normal	(2.5 %)
Ash to Flue Gas	80 %		
Sulfur to Flue Gas	97.5 %		
Economizer Outlet Temp	700 °F		
Preheater Outlet Temp	300 °F		
Financial Parameters			
Inflation Rate	0 %		
Debt Fraction	50 %		
Common Stock Fraction	35 %		
Preferred Stock Fraction	15 %		
Real Return on Debt	4.6 %	Normal	(10 %)
Real Return on Com. Stock	8.7 %	Normal	(10 %)
Real Return on Pref. Stock	5.2 %	Normal	(10 %)
Federal Tax Rate	36.7 %		
State Tax Rate	2.0 %		
Ad Valorem Rate	2.0 %		
Investment Tax Credit	0 %		
Book Life	30 years		
Real Fuel Escalation	0 %	1/2 Normal[a]	σ = 0.06 %

[a] A -1/2 normal is a negatively skewed truncated normal distribution, with the mode at the extreme upper end of the distribution. Conversely, a 1/2 normal is positively skewed with mode at the extreme lower end.

Table 9. Selected Properties of Coals Used for Case Studies (As-Fired Basis)

	Illinois No. 6 Coal		Pittsburgh Coal	
Coal Property	Run-of-Mine	Washed[a]	Run-of-Mine	Washed[a]
Heating Value, Btu/lb	10,190	10,330	13,400	12,900
Sulfur, wt %	4.36	3.09	2.15	1.66
Carbon, wt %	57.0	57.7	74.8	72.1
Hydrogen, wt %	3.7	4.0	4.6	4.5
Oxygen, wt %	7.2	8.4	5.3	5.4
Nitrogen, wt %	1.1	1.1	1.4	1.3
Moisture, wt %	12.3	17.5	2.7	7.9
$/ton (at mine)	26.10	30.68	33.40	34.99
$/ton (transport)	7.90	7.90	7.90	7.90

[a] Model results for a 30 % sulfur reduction on a lb/MMBtu basis using conventional coal cleaning (Level 3 plant design)

Table 10. Nominal Parameter Values and Uncertainties for the Conventional Environmental Control System

Model Parameter	Deterministic (Nominal) Value	Probability Distribution	Values (or σ as % of mean)[a]
Wet FGD System			
Molar Stoichiometry	(calc)	Normal	(5 %)
No. Operating Trains	4	Chance	10 % @ 1; 20 % @ 2; 40 % @ 3; 30 % @ 4
No. Spare Trains	1	Chance	75 % @ 0; 25 % @ 1
Reheat Energy	(calc)	Chance	75 % @ 0; 25 % @ x
Total Energy Use	(calc)	Normal	(10 %)
Limestone Cost	$15/ton	Uniform	$10-15/ton
Direct Capital Costs	(calc)	Normal	(10 %)
Operating Costs	(calc)	Normal	(10 %)
Selective Catalytic Reduction			
Space Velocity	2,850/hr	Normal	(10 %)
NH_3 Stoichiometry	(calc)	Normal	(10 %)
Catalyst Life	15,000 hrs	Chance	5 % @ 1,275 hrs; 30 % @ 5,700 hrs; 50 % @ 11,400 hrs; 14 % @ 17,100 hrs; 1 % @ 28,500 hrs
Energy Requirement	(calc)	Normal	(10 %)
Ammonia Cost	$150/ton	Uniform	$150-225/ton
Catalyst Cost	$460/ft^3	Normal	(7.5 %)
Direct Capital Cost	(calc)	Triangular	0.8x, x, 2x
Operating Cost (excl. Cat.)	(calc)	Normal	(10 %)
Cold-Side Electrostatic Precipitator			
Specific Collection Area	(calc)	Normal	(5 %)
Energy Requirement	(calc)	Normal	(10 %)
Total Capital Cost	(calc)	Normal	(10 %)
Operating Cost	(calc)	Normal	(10 %)
Solid Waste Disposal			
Land Cost	$6,500/acre	Normal	(10 %)
Direct Cost	(calc)	Normal	(10 %)
Operating Cost	(calc)	Normal	(10 %)

[a] For uniform distributions actual values are shown. For triangular distributions, end-points and median are shown. For chance distributions, the probabilities of obtaining specific values are shown.

Table 11. Nominal Parameter Values and Uncertainties for the Advanced Environmental Control System

Model Parameter	Deterministic (Nominal) Value	Probability Distribution[a]	Values (or σ as % of mean)[b]
Copper Oxide Process[c]			
Fluidized Bed Height	48 inches		
Sorbent Copper Loading	7 wt-%		
Regeneration Efficiency	99.2 %	-1/2 Normal	(20 %)
Fluidized Sorbent Density	400 kg/m^3	Normal	(10 %)
Standard Error, Cu/S Ratio	0	Normal	$\sigma = 0.39$
Sorbent Attrition	0.06 %	Normal	(41 %)
Ammonia Stoichiometry	(calc)	Normal	(6.25 %)
Regeneration Temp	900 ^oF	Normal	(2 %)
No. Operating Trains	4	Chance	10 % @ 1; 20 % @ 2; 40 % @ 3; 30 % @ 4
No. Spare Trains	1	Chance	50 % @ 0; 50 % @ 1
Sorbent Cost	$5.00/lb	-1/2 Normal	(25 %)
Methane Cost	$4.50/mscf	1/2 Normal	(25 %)
Ammonia Cost	$150/ton	Uniform	$150-225/ton
Sulfuric Acid Cost	$40/ton	-1/2 Normal	(30 %)
Sulfur Cost	$125/ton	-1/2 Normal	(30 %)
Absorber Direct Cap. Cost	(calc)	Uniform	1.0x - 1.5x
Solids Heater DCC	(calc)	Uniform	1.0x - 1.5x
Regenerator DCC	(calc)	Uniform	1.0x - 1.5x
Solids Transport DCC	(calc)	Uniform	1.0x - 2.0x
Sulfur Recovery DCC	(calc)	Uniform	1.0x - 1.2x
Total Capital Cost	(calc)	1/2 Normal	(10 %)
Fabric Filter			
Air-to-Cloth Ratio	2.0 $acfm/ft^2$	-1/2 Normal	(10 %)
Bag Life	(calc)	Normal	(25 %)
Energy Requirement	(calc)	Normal	(10 %)
Bag Cost	$0.80/ft^2	Normal	(5 %)
Operating Cost	(calc)	Normal	(15 %)
Total Capital Cost	(calc)	Normal	(15 %)
Solid Waste Disposal			
Land Cost	$6,500/acre	Normal	(10 %)
Direct Cost	(calc)	Normal	(10 %)
Operating Cost	(calc)	Normal	(10 %)

a A -1/2 normal is a negatively skewed truncated normal distribution, with the mode at the extreme upper end of the distribution. Conversely, a 1/2 normal is positively skewed with mode at the extreme lower end.

b For uniform distributions actual values are shown. For triangular distributions, end-points and median are shown. For chance (discrete) distributions, the probabilities of obtaining specific values are shown.

c As part of integration of the copper oxide process with the base power plant, the plant air preheater is resized to maintain an exit flue gas temperature of 300 ^oF.

A key determinant of process capital and operating costs for the copper oxide process is the sorbent circulation rate. The sorbent circulation rate is proportional to the required molar flow rate of unreacted copper oxide that must enter the absorber to achieve a given sulfur dioxide removal efficiency. As discussed in Section 3.1.2.2, the molar flow rate of unreacted copper oxide depends on the available Cu/S molar ratio. An important factor that affects the Cu/S ratio is the regeneration efficiency. Test results for the Cu/S ratio were based on low (e.g., 30 to 50 percent) regeneration efficiencies (Yeh et al,1984). SMC estimated that a properly designed and sized regenerator, coupled with appropriate heating of the sorbent to reaction temperature, can result in regeneration efficiencies of over 99 percent at a 30 minute residence time (SMC, 1984). To characterize uncertainty in the regeneration efficiency, a negatively skewed probability distribution was therefore assumed with a maximum value of 99.2 percent, representative of nominal expectations, with a small probability that the value could go below 50 percent, representative of actual experience to date. The negatively skewed distribution is qualitatively consistent with the notion of performance shortfalls that are characteristic of innovative chemical process plants (Merrow et al, 1981).

Another uncertainty regarding the Cu/S ratio results from regression analysis of experimental data. As shown in Figure 9, there is residual error in the regression model predictions of the Cu/S ratio compared to experimentally observed values for the same set of model dependent variables. This residual error is represented in the regression model as a normal probability distribution with a mean of zero, which is added to the predicted estimate of the Cu/S ratio. The standard error represents the observed uncertainty in the Cu/S ratio that is not explained by the regression model.

In experiments on a life cycle test unit, the sorbent attrition rate (another key parameter) was reduced to 0.13 weight percent of the sorbent circulation rate after modifications were made to the solids transport system (Williamson et al, 1987). The test results indicated that solids transport was the primary source of sorbent attrition. However, significant improvements in the attrition rate were expected for a commercial process. The judgment of one process developer, elicited for this study, was that the attrition would nominally be 0.06 percent, but could have a 90 percent chance of being between 0.02 and 0.10 percent. The expert indicated that the uncertainty is normally distributed. This judgment formed the basis for the distribution in Table 11.

The preceding three examples have focused on sources of uncertainty in process performance. An additional source of uncertainty is in process costs. On a commercial scale, measures such as equipment additions or redesigns, which often occur between the time of a conceptual design study and a mature commercial plant, could lead to significantly higher costs. Results from a Rand study regarding cost growth in pioneer process plants indicate that solids handling systems pose the greatest difficulties in process design and operation (Milanese, 1987). In deterministic analyses, contingency factors normally are used to represent process risks. However, in probabilistic analyses, process contingency factors are supplanted by directly specifying uncertainties as probability distributions in model parameters affecting cost. To represent the uncertainty in current estimates of capital cost, the capital costs for each major equipment area were assigned uniform probability distributions, with the current estimates at the low end of the range. The high end of the ranges represented probabilities that the actual capital cost might increase by up to 50 percent for most process areas and up to 100 percent for the solids transport system. These uncertainties were intended to be representative of the costs likely to be found in a commercial system (e.g., based on a fifth-of-a-kind plant).

Additional details regarding the basis for characterizing uncertainties in both the FGD/SCR and copper oxide systems are given in Appendix B. The uncertainties specific to the FGD/SCR system are discussed in Appendix B.1, and those specific to the copper oxide process are discussed further in Appendix B.2.

4.3 Estimates of Uncertainty for IGCC Technologies

The characterization of uncertainties in the three IGCC systems was a major effort in this research. The complete documentation of the effort is given in Appendix B, sections B.3 through B.8. In this section, the uncertainties assumed for base case analysis of the three IGCC systems are summarized.

Because the IGCC systems have several process areas common to two or all three of the systems, the approach taken was to characterize uncertainties in the performance in each process area with consideration of its application in different systems. For example, the gas turbine process area is common to all three IGCC systems. The technical review of information about uncertain considers all three types of process environments. The characterization of uncertainties also explicitly considers the differences in the set of uncertain parameters, and magnitude (variance) of uncertainties, depending on which IGCC system is assumed. Thus, uncertainties between competing technologies are characterized on a common basis that permits comparative analysis.

The characterization of process performance uncertainties focused on four major process areas. These are:

- Lurgi gasification, including the fines recycle cyclone.
- KRW gasification, including designs featuring oxygen-blown gasification and air-blown gasification with in-bed desulfurization using a calcium-based sorbent. In addition, uncertainties in the sulfation process area associated with the air-blown KRW process was also evaluated.
- Zinc ferrite desulfurization, which is common to both the air-blown KRW and Lurgi IGCC systems.
- Gas turbine, which is common to all three IGCC systems. However, there are significant differences between the IGCC process environments related to fuel gas heating value, ammonia concentration in the fuel gas, environmental performance, and cost.

In addition, uncertainties in other IGCC performance and cost model parameters were characterized. These include:

- Cost model parameters common to all three IGCC systems. These include indirect capital costs, operating cost parameters, and financial assumptions
- Direct capital costs for each process area
- Maintenance costs for each process area
- Unit costs of consumables, byproducts, and wastes associated with variable costs
- Error terms for regression models of direct capital cost and plant auxiliary power requirements.

First, the approach used to develop estimates of uncertainties for the IGCC systems will be described, with particular focus on the elicitation of judgments from technical experts. Then, the base case assumptions regarding the values and distributions for key model parameters for each of the three IGCC systems will be summarized. For several cases, there are alternative assumptions regarding uncertainties. For example, there are four sets of assessments of uncertainty for the zinc ferrite process area. These alternative cases are not summarized here in the main body of the report. However, they are documented in detail in Appendix B.

4.3.1 Obtaining Judgments From Technical Experts

A key focus of the IGCC case studies was the development of a practical yet detailed approach to the characterization of uncertainties in the cases where expert judgments are required. One challenge to obtaining judgments is distance, which may make face-to-face interviews unpractical. Another is availability. Many experts are busy people. Thus, it is often difficult to schedule visits, particularly when several experts are to be approached. A solution to the difficulty in obtaining access to specific experts was the development of an elicitation briefing packet and the use of follow-up phone conversations to clarify expert responses to the briefing packets. This eliminated the need for complex travel arrangements, but allowed for interaction with each of the experts. The following sections describe how experts were selected and how their judgments were elicited.

4.3.1.1 Identifying Experts

The primary source for expert judgments regarding process area uncertainties was DOE/METC. METC both conducts in-house and funds externally contracted research on process technologies that are components of IGCC and other coal gasification-based systems. METC process engineers have extensive practical experience obtained from work on internal research projects and their project management work on externally funded contracts. This work includes experimental and modeling studies. Many METC process engineers also have previous employment experience with companies that have been involved in research on the process areas of interest in this study. Therefore, technical experts at METC were approached for their judgments regarding uncertainties. Because their expertise is strongly performance-oriented, and less cost-oriented, the focus of the uncertainty elicitations was on performance uncertainties.

Originally, four key process were identified for which expert judgments regarding uncertainties were desired. Briefing packets for each of these, as discussed in the following section, were developed. For each process area, specific experts at METC were identified in cooperation with METC management. However, because of personnel and time constraints within METC, only three process areas could be addressed. These are: Lurgi gasification; zinc ferrite desulfurization; and gas turbine.

Initially, three experts were selected for each of the three process areas. METC management distributed the briefing packets provided by the author to the experts. The responses were collected and returned. For the zinc ferrite process area, all three experts responded. For the other two process areas, two of the three experts responded. For bookkeeping purposes, these experts are assigned arbitrary designations. The three zinc ferrite experts are referred to as Experts ZF-1, ZF-2, and ZF-3. The two Lurgi gasification experts are referred to as Experts LG-1 and LG-2. Similarly, the two gas turbine experts are referred to as Experts GT-1 and GT-2. Expert ZF-1 is the same person as Expert LG-1.

In addition to formal elicitations of uncertainty from METC process engineers, other engineers in industry were approached for information regarding uncertainties in key process areas. However, a major obstacle to obtaining information about uncertainties for some process areas involves concern about proprietary information and the competitive position of specific companies. Several of the process areas evaluated in this study involve technologies that are considered proprietary by the companies involved in developing them, even in cases where a portion of the development is government-funded. Therefore, experts within those companies may often be reticent about providing detailed information regarding the technologies, other than that which has already been published. Thus, for those cases where companies were unable or unwilling to provide detailed information regarding their technologies, the author relied on published information to make preliminary characterizations of uncertainty.

4.3.1.2 Briefing Packets

The major source of expert information regarding uncertainties involved responses to detailed briefing packets for three major process areas. A total of seven questionnaire responses were obtained from six engineers at DOE/METC. One engineer provided responses for two process areas. The briefing packets included three parts. These were: (1) Part 1, introduction to uncertainty analysis; (2) Part 2, technical background for uncertainties in the process area of interest; and (3) Part 3, a questionnaire regarding uncertainties in specific process area performance parameters, and in a few cases cost parameters also.

Part 1 was common to all of the process areas for which judgments were sought. It was written as an informal nine page paper. It included a discussion of the philosophy of uncertainty analysis, types of uncertain quantities, and methods for characterizing uncertainties. This information is contained in Section 2.3.

Part 2 was specific to each process area. Packets for four major process areas were developed, including Lurgi gasification, KRW gasification, zinc ferrite desulfurization, and gas turbine. These technical background papers included a description of the process area, a description of the IGCC process environments to which the process area is applied, a review of key design and performance assumptions, and a detailed review of the specific aspects of the process areas which may contribute to uncertainty in either performance or cost. The reviews were based primarily on information in published literature. The information that was included in each technical background paper is presented in Appendix B for each of the four process areas in Sections B.3 through B.6.

The uncertainty questionnaire, Part 3, was also specific to each process area. While questionnaires were developed for the four major process areas, only three could be distributed to experts within DOE/METC. The questionnaires for the three process areas are given in Appendix B.8.

The questionnaires were designed to provide the expert with some flexibility in responding. For example, the expert was asked to review the key design assumptions for each process area, and to provide alternative suggestions if not happy with them. The expert was also asked to explain the basis for any changes. Similarly, the expert was given a list of parameters which the elicitor believed should be treated as uncertain. The expert was asked to examine the list, and suggest any additions or deletions. For the parameters included in the list for which the expert was able to provide judgments of uncertainty, the expert was asked to consider "worst" and "best" outcomes, before considering median or modal values. An important part of the questionnaire was to request also a *basis* for the judgments. The questionnaire was intended to encourage the expert to think systematically about the range and likelihood of possible outcomes for each uncertain variable, and to explain the mechanisms by which such outcomes would be obtained. In particular, the design of the questions was sensitive to some of the concerns discussed in Section 2.3, such as overconfidence and the tendency to obtain better judgments when explanations are required.

4.3.1.3 Expert Reaction to the Briefing Packets

The reaction of the experts to the briefing materials was varied. Generally, the experts indicated that the information contained in Part 1 was more than they needed. According to one, "Part 1 was more information than I needed or wanted." Another said, "the uncertainty analysis discussion [Part 1] seemed like a justification for the approach and more than necessary to elicit answers to the questionnaire."

With respect to the technology-specific background papers, responses were generally favorable. One zinc ferrite process area expert stated in the response to the questionnaire that "the summary in Part 2 was useful, and perhaps essential to this exercise." Another zinc ferrite expert wrote, "the briefing information was needed and about the right amount of depth to stimulate thought without being too cumbersome." A Lurgi gasifier expert characterized the summary information as "useful," while the other said "Part 2 was well done." A gas turbine expert wrote, "Part 2 was very well done," and characterized the summary as "objective and unbiased." The other gas turbine expert did not respond directly to the issue of the briefing material, but indicated that the technical background paper provided a thorough summary of published information.

With respect to the questionnaires, responses were mixed. For the Lurgi gasifier and zinc ferrite process areas, the experts responded to as many questions as the scope of their expertise allowed, and they were quite candid in pointing out questions for which they could not provide a response. For the gas turbine process area, perhaps due partly to propriety concerns that limit the information available to DOE engineers, and perhaps due to time constraints, the experts reacted less favorably. The gas turbine questionnaire also was the most difficult of the three that were distributed, because it contained more questions and more detail within each question. This is because the gas turbine is common to all three IGCC systems, and there are significant differences between the three applications that require explicit characterization.

One Lurgi gasifier expert indicated that the questionnaire was "not too difficult" to respond to, and that his experience with METC's 42-inch diameter fixed bed gasifier helped him to provide judgments about the Lurgi gasifier assumed in the study. The other Lurgi gasifier expert indicated that he did not have operating experience with fixed-bed gasifiers, and that "it was difficult to develop [estimates of] the range of values for various variables due to lack of abundant actual operating data."

The zinc ferrite experts responded similarly. One stated, "considerable thinking was required in order to provide "good" judgments." Another, who provided an alternative set of judgments in addition to the case requested, stated: "it was fairly easy to make the judgments of uncertainty; however, much thought was required to arrive at what seemed like meaningful inputs." The third zinc ferrite expert did not comment on the briefing materials.

With respect to the gas turbine questionnaire, one expert indicated that the information he would require to answer the questions was not available to him. He stated:

> To answer these questions, I would need detailed and extensive statistical data in these areas. This data is not available to DOE/METC personnel. Gas turbine manufacturers would be the prime source of this data, *if it exists*. Usually, these data would be proprietary ...

The expert indicated that gas turbine manufacturers would be reluctant to release information of the type requested because they spend millions of dollars of internal research funds to develop new systems. Release of the results of such work could impair their competitive position. Furthermore, while manufacturers provide emission guarantees based on current regulations and market demands, they must be careful about releasing information that would subject them to "ever-racheting lower levels of emissions." Thus, the expert was unable to respond to questions about uncertainty because of the limited information available to him. However, he was able to comment on the design assumptions proposed in the questionnaire.

The other responding gas turbine expert was critical of the questionnaire itself:

> In part 3 you are asking the respondents to devote an unreasonable amount of time to answering 12 compound questions in considerable detail ... the format is not suited to the general reviewer, but only those personnel whose time is to be dedicated to lengthy speculative exercises.

The expert indicated that the questions should be converted to a set of multiple choice questions. He also indicated that the surveyor should develop the type of probability distribution that best fits the response. However, the difficulty with a multiple-choice format would be obtaining justifications and explanations regarding the basis for the judgments.

In general, the briefing packets appeared to be successful in communicating technical background information regarding the specific process areas. In most cases, particularly when the expert had sufficient time and a detailed knowledge base, the responses to the questionnaires were detailed and complete. In the case of the gas turbine process area, the questionnaire was perhaps too ambitious. An initial questionnaire focusing on just one case, such as the air-blown Lurgi system, might have been more appropriate than the lengthy questionnaire that tried to cover all three IGCC systems. However, limitations in propriety information regarding the gas turbine would still limit the responses.

4.3.1.4 Follow-Up Phone Interviews

For all seven of the responses received, a follow-up phone call was made to the expert to clarify ambiguities, to obtain elaboration on the basis for specific judgments, and to obtain judgments for parameters that were either not included on the questionnaire or for which responses were not obtained. These phone conversations typically lasted about half an hour. In some cases, detailed supporting information was obtained regarding mechanisms by which various outcomes could be obtained. In one case, an expert indicated that he would be willing to provide judgments about two parameters in the zinc ferrite process area that were not included in the questionnaire. These judgments were obtained during the follow-up phone conversation.

4.3.2 Other Approaches to Characterizing Uncertainties

In addition to the elicitation of expert judgments regarding uncertainties, two other approaches were used. One of these was the use of published information as the basis for making preliminary characterizations of uncertainty. This approach was necessary for the

Table 12. Summary of the Financial Assumptions for the IGCC Case Studies.

Description	Units	Value
Cost Year		January 1989
Chemical Engineering Plant Cost Index	(index)	351.5
Chemicals Cost Index	(index)	411.3
Construction Interest	%/yr	10
Construction Years	years	4
Booklife	years	30
Inflation Rate	%/yr	0.0
Sales Tax	%	5
Real Return on Debt	%/yr	4.6
Real Return on Preferred Stock	%/yr	5.2
Real Return on Equity	%/yr	8.7
Debt Ratio	fraction	0.50
Preferred Stock Ratio	fraction	0.15
Federal and State Tax Rate	fraction	0.38
Investment Tax Credit	fraction	0.0
Property Taxes and insurance	%/yr	2.0

KRW gasification and sulfation process areas, for which expert judgments could not be obtained. In addition, this approach was used for many of the cost-related uncertainties. The other approach was the use of regression analysis. Regression analysis was used extensively to develop the cost models for the three IGCC systems, as detailed in Appendix A. Regression models were developed for process area direct capital costs, consumable material requirements for plant operation, and auxiliary power requirements. The standard error from the regression model can be explicitly represented in a probabilistic simulation.

4.3.3 Base Case Estimates of Uncertainty for the Oxygen-blown KRW-based IGCC System

The financial assumptions made for all of the IGCC systems are given in Table 12. These include the cost year in which all cost estimates are reported and financial parameters used to calculate the fixed charge factor and variable cost levelization factor. These factors are calculated using the standard approach described by EPRI (1986). Based on the assumptions in Table 12, the fixed charge factor is 10.34%/yr and the variable cost levelization factor is 1. The fixed charge factor is also known as the capital carrying charge or capital recovery factor, and it is the levelized annual cost for repaying the plant total capital cost as a percentage of the capital cost. Here, the fixed charge factor is calculated using an inflation rate of zero. This results in an estimate using "constant" dollars. For all of the IGCC systems, an unwashed Illinois No. 6 coal is assumed. The characteristics of the assumed coal are given in Table 13.

Table 13. Characteristics of the Coal Assumed for IGCC System Studies

Proximate Analysis, wt-%, run-of-mine basis	
Moisture	12.0
Fixed Carbon	47.8
Volatile Matter	31.4
Ash	8.8
Ultimate Analysis, wt-%, dry basis	
Carbon	69.53
Hydrogen	5.33
Nitrogen	1.25
Chlorine	0.00
Sulfur	3.86
Oxygen	10.03
Ash	10.00
Ash Fusion Temperature, °F	2,300

The key assumptions regarding parameters for the oxygen-blown KRW-based IGCC system with cold gas cleanup are summarized in Table 14. A total of 41 parameters are treated as uncertain. These include assumptions regarding the performance of the gasifier and gas turbine process areas, capital cost parameters, direct capital cost, maintenance costs, labor rate, unit costs, and regression model error terms for direct capital and auxiliary power models. The deterministic values used here are based on assumptions used in published design studies.

For this IGCC system, the characterizations of uncertainty were developed by the author. Estimates of uncertainty in the gasification and gas turbine process areas are based on detailed reviews of technical information. The technical background information and the basis for each of the uncertainties for these two process areas are discussed in detail in Appendices B.4 and B.6, respectively.

The estimates of uncertainty in the capital cost parameters, including engineering and home office fees, indirect construction factor, and project uncertainty, are based on typical ranges of values for these parameters suggested by EPRI (1986). The basis for these estimates is discussed further in Appendix B.7.1.

For the direct costs, the deterministic values reported in Table 14 represent the process contingency factors assumed in published design studies (e.g., Dawkins et al, 1985). As a preliminary characterization of uncertainty in capital cost, it was assumed that the process contingency factors were intended to represent the mid-point of a symmetric uncertainty distribution for process area direct cost. The relative magnitudes of the contingency factor was assumed to suggest the relative magnitude of the variances that should be used between process areas. As a best case assumption, it was assumed that the capital cost would be no lower than the unadjusted estimate obtained from the direct cost models. As a worst case assumption, it was assumed that the cost could be equivalent to that obtained with double the nominal contingency factor. The one exception is the low temperature gas cooling process area, which is represented in the literature as being commercial available without any technical risk. For this process area, a chance of a small decrease in cost compared to the direct cost model estimate is assumed. In most cases, a uniform distribution between the best and worst values was assumed, while in a few other cases, a triangular distribution was assumed. The effect of a triangular distribution, compared to a uniform distribution, is to place more "weight" on the outcomes near the published contingency factor then on the extreme high or low outcomes. The triangular distribution was used in cases where the author felt that the published contingency factors were carefully developed.

An exception to the approach described above is the estimate of uncertainty in the gas turbine process area. Design studies tend to assume very low contingency factors (e.g., five percent or less) for this process area, in spite of the fact that the gas turbine requires modifications to the fuel valve to handle medium-BTU coal gas. Gas turbine capital cost uncertainty is discussed further in Appendix B.6.7.4.

Similar to the estimates of uncertainty in direct cost, estimates of uncertainty in maintenance cost factors use published deterministic values as a starting point. However, in many cases it is believed that it is more likely that maintenance costs would increase than decrease compared to the deterministic values. The underlying reason for this belief is that IGCC systems must handle material streams containing various contaminants derived from coal conversion. These contaminants are likely to cause deposition, erosion, and corrosion problems in various parts of the system, or to cause deactivation of solvents or catalysts. A response to such problems would be increased maintenance. The basis for these judgments is discussed also in Appendix B.7.3.

Other operating cost parameters include the operating labor rate, unit costs for ash disposal and byproduct sales, and a factor to account for byproduct marketing costs. These uncertainties are discussed in Appendix B.7.4.

Table 14. Summary of the Base Case Parameters Values and Uncertainties for the Oxygen-Blown KRW-based IGCC System with Cold Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]			
GASIFIER PROCESS AREA							
Gasifier Pressure	psia	465					
Gasifier Temperature	^oF	1,850					
Overall Carbon Conversion	wt-% of feed coal carbon	95	Triangular	75	to	95	(95)
Oxygen/Carbon Ratio	lbmole O_2/C	0.34	Uniform	0.33	to	0.35	
Steam/Oxygen Ratio	lbmole H_2O/O_2	1.35	Uniform	1.1	to	1.6	
Sulfur Retention in Bottom Ash	mol-% of inlet sulfur	15	Triangular	10	to	20	(15)
GAS TURBINE PROCESS AREA							
Pressure Ratio	ratio	13.5					
Turbine Inlet Temp	oF	2,300					
Exhaust Flow	lb/sec	938					
Thermal NO_x	fraction of air nitrogen fixated	5.0×10^{-5}	Uniform	2.5×10^{-5}	to	7.5×10^{-5}	
Unconverted CO	wt-% of CO in fuel gas	0.99985	Uniform	0.9998	to	0.9999	
CAPITAL COST PARAMETERS							
Engineering and Home Office Fee	fraction	0.10	Triangular	0.07	to	0.13	(0.10)
Indirect Construction Cost Factor	fraction	0.20	Triangular	0.15	to	0.25	(0.20)
Project Uncertainty	fraction	0.175	Uniform	0.10	to	0.25	
General Facilities	fraction	0.20					
DIRECT COSTS[b]							
Coal Handling	% of DC	5	Uniform	0	to	10	
Oxidant Feed	% of DC	5	Uniform	0	to	10	
Gasification	% of DC	20	Triangular	0	to	40	(20)
Selexol	% of DC	10	Triangular	0	to	20	(10)
Low Temperature Gas Cooling	% of DC	0	Triangular	-5	to	5	(0)
Claus Plant	% of DC	5	Triangular	0	to	10	(5)
Beavon-Stretford	% of DC	10	Triangular	0	to	20	(10)
Boiler Feed Water	% of DC	0					
Process Condensate Treatment	% of DC	30	Triangular	0	to	50	(30)

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Table 14 (Continued). Summary of the Base Case Parameters Values and Uncertainties for the Oxygen-Blown KRW-based IGCC System with Cold Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]			
Gas Turbine	% of DC	12.5	Uniform	0	to	25	
HRSG	% of DC	2.5	Uniform	0	to	5	
Steam Turbine	% of DC	2.5	Uniform	0	to	5	
General Facilities	% of DC	5	Uniform	0	to	10	(5)
MAINTENANCE COSTS[c]							
Coal Handling	% of TC	3					
Oxidant Feed	% of TC	2					
Gasification	% of TC	4.5	Triangular	3	to	6	(4.5)
Selexol	% of TC	2	Triangular	1.5	to	4	(2)
Low Temperature Gas Cooling	% of TC	3	Triangular	2	to	4	(3)
Claus Plant	% of TC	2	Triangular	1.5	to	2.5	(2)
Beavon-Stretford	% of TC	2					
Boiler Feed Water	% of TC	1.5					
Process Condensate Treatment	% of TC	2	Triangular	1.5	to	4	(2)
Gas Turbine	% of TC	1.5	Triangular	1.5	to	2.5	(1.5)
HRSG	% of TC	1.5					
Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					
OTHER FIXED OPERATING COST PARAMETERS							
Labor Rate	$/hr	19.70	Normal	17.70	to	21.70	
VARIABLE OPERATING COST PARAMETERS							
Ash Disposal	$/ton	10	Triangular	10	to	25	(10)
Sulfur Byproduct	$/ton	125	Triangular	60	to	125	(125)
Byproduct Marketing	fraction	0.10	Triangular	0.05	to	0.15	(0.10)
DIRECT COST REGRESSION MODEL ERROR TERMS[d]							
KRW Coal Handling	$ Million	0	Normal	- 10	to	10	
Oxygen Plant	multiplier	1.012	Lognormal	0.78	to	1.29	
KRW Gasification	$ Million	0	Normal	- 20.5	to	20.5	
Low Temperature Gas Cooling	$ Million	0	N/A				
Selexol	$ Million	0	Normal	- 5.1	to	5.1	
Claus	$ Million	0	N/A				
Beavon-Stretford	$ Million	0	N/A				

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Table 14 (Continued). Summary of the Base Case Parameters Values and Uncertainties for the Oxygen-Blown KRW-based IGCC System with Cold Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]		
Boiler Feedwater	$ Million	0	N/A			
Process Condensate	$ Million	0	N/A			
HRSG	$ Million	0	Normal	- 17.3	to	17.3
Steam Turbine	$ Million	0	Normal	- 15.8	to	15.8
AUXILIARY POWER REGRESSION MODEL ERROR TERMS[d]						
KRW Coal Handling	MW	0	Normal	- 1.6	to	1.6
Oxygen Plant	MW	0	Normal	- 6.6	to	6.6
KRW Gasification	MW	0	Normal	- 0.52	to	0.52
Low Temperature Gas Cooling	MW	0	N/A			
Selexol	MW	0	Normal	- 0.55	to	0.55
Claus Plant	MW	0	N/A			
Beavon-Stretford	MW	0	N/A			
Boiler Feedwater	MW	0	N/A			
Process Condensate	MW	0	N/A			
General Facilities	MW	0	N/A			

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range. For normal and lognormal distributions, the 99.8 percent probability range is given.

[b] For direct costs, the deterministic values represent "contingency factors" as defined by EPRI (1986) and others. For probabilistic studies, uncertainty in capital cost is represented by an uncertainty factor, which is described by a probability distribution. DC = process area direct cost.

[c] TC = process area total cost, including indirects and contingency

[d] Negligibly small error terms were not included in the simulation.

The last category of uncertainties include regression model error terms for both direct cost and auxiliary power requirement models. These models were developed as part of the IGCC cost models. The error terms are derived from statistical analysis of the models. See Appendix A for details regarding these models and Appendix B.7.5 for a brief summary of the regression model error terms.

4.3.4 Base Case Estimates of Uncertainty for the Air-blown KRW-based IGCC System

The base case assumptions regarding key parameter values for the model of the air-blown KRW-based IGCC system with hot gas cleanup are summarized in Table 15. A total of 46 parameters are treated probabilistically. Estimates of uncertainty in performance parameters include the gasification, sulfation, zinc ferrite desulfurization, and gas turbine process areas. The deterministic values are based on typical assumptions in the literature. The uncertainty estimates for the gasification and sulfation process areas are based on a review of published information, and these estimates are discussed further in Appendix B.4.5. The uncertainty estimates for the zinc ferrite process area were elicited from Expert ZF-1. The basis given by Expert ZF-1 for the judgments regarding uncertainties is summarized in Appendix B.5.3.1. The uncertainty estimates for the gas turbine process area were estimated by the author based on a review of published information, discussions with process engineers at DOE, and discussion with engineers at gas turbine manufacturers, as discussed in Appendix B.6.7.

The capital cost parameter, direct cost, maintenance cost, labor rate, unit cost, and regression model uncertainties were estimated in the same manner as for the oxygen-blown KRW system. See Appendix B.7 for more discussion of these.

4.3.5 Base Case Estimates of Uncertainty for the Air-blown Lurgi-based IGCC System

Deterministic and uncertainty assumptions for the air-blown Lurgi-based IGCC system are summarized in Table 16. A total of 47 performance and cost parameters are treated probabilistically. Performance uncertainties in three major process areas are characterized. These process areas are Lurgi gasification, zinc ferrite desulfurization, and gas turbine. For the gasification process area, the judgments of Expert LG-1 are taken as the base case assumptions. Expert LG-1 provided uncertainty characterizations for nine process performance variables. Notably, Expert LG-1 provided estimates of gasifier coal throughput conditioned on the gasifier pressure, and estimates of

Table 15. Summary of the Base Case Parameters Values and Uncertainties for the Air-Blown KRW-based IGCC System with Hot Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]
GASIFIER PROCESS AREA				
Gasifier Pressure	psia	465		
Gasifier Temperature	^oF	1,900	Triangular	1,900 to 1,950 (1,900)
Overall Carbon Conversion	wt-% of feed coal carbon	95	Triangular	90 to 97 (95)
Oxygen/Carbon Ratio	lbmole O_2/C	0.46	Triangular	0.45 to 0.47 (0.46)
Steam/Oxygen Ratio	lbmole H_2O/O_2	0.45	Uniform	0.4 to 0.5
Sulfur Retention in Bottom LASH	mol-% of inlet sulfur	90	Triangular	85 to 95 (90)
Limestone Calcium-to-Sulfur Ratio	lbmole Ca/S	2.6	Triangular	2 to 2.8 (2.6)
Gasifier Ammonia Yield,	Equiv. fraction of coal N to NH_3	0.10	Triangular	0.005 to 0.10 (0.10)
SULFATION PROCESS AREA				
SO_2 Emissions	lb /MMBtu	0.01	Triangular	0.01 to 0.05 (0.01)
NO_x Emissions	lb /MMBtu	0.15	Triangular	0.10 to 0.20 (0.15)
Conversion of CaS to $CaSO_4$	%	60	Uniform	30 to 90
Carbon Conversion	%	95	Triangular	90 to 98 (95)
ZINC FERRITE DESULFURIZATION PROCESS AREA				
Residual Sulfate After Oxidative Regen.	mol-% of captured S	7.5	Triangular	3 to 11 (7.5)
Residual Sulfide After Reductive Regen.	mol-% of S in sulfate	85	Triangular	50 to 90 (85)
Sorbent Sulfur Loading	wt-% S in sorbent	17	Normal	2.16 to 31.84 (17)
Sorbent Attrition Rate	wt-% sorbent loss/cycle	1.0	Fractile	5%: 0.17 to 0.34 20%: 0.34 to 0.5 25%: 0.5 to 1 25%: 1 to 1.5 20%: 1.5 to 5 5%: 5 to 25
Absorber Pressure Drop	psi/ft bed height	0.4	Triangular	0.29 to 0.53 (0.4)
Absorption Cycle Time	hours	30		
Max. Vessel Diameter	ft	12.5		
Max. Vessel Height	ft	37.5		

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Table 15 (Continued). Summary of the Base Case Parameters Values and Uncertainties for the Air-Blown KRW-based IGCC System with Hot Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]			
GAS TURBINE PROCESS AREA							
Pressure Ratio	ratio	13.5					
Turbine Inlet Temp	oF	2,300					
Exhaust Flow	lb/sec	938					
Thermal NO_x	fraction of air nitrogen fixated	4.25×10^{-5}	Uniform	1.0×10^{-5}	to		7.5×10^{-5}
Fuel NO_x	% conversion of NH_3 to NO_x	90	Triangular	50	to	100	(90)
Unconverted CO	wt-% of CO in fuel gas	0.9885	Uniform	0.9772	to		0.9999
CAPITAL COST PARAMETERS							
Engineering and Home Office Fee	fraction	0.10	Triangular	0.07	to	0.13	(0.10)
Indirect Construction Cost Factor	fraction	0.20	Triangular	0.15	to	0.25	(0.20)
Project Uncertainty	fraction	0.175	Uniform	0.10	to	0.25	
General Facilities	fraction	0.20					
DIRECT COSTS[b]							
Coal Handling	% of DC	5	Uniform	0	to	10	
Limestone Handling	% of DC	5	Uniform	0	to	10	
Oxidant Feed	% of DC	10	Uniform	0	to	20	(10)
Gasification	% of DC	20	Triangular	0	to	40	(20)
Sulfation	% of DC	40	Triangular	20	to	60	(40)
Zinc Ferrite	% of DC	40	Uniform	0	to	80	
Sulfuric Acid Plant	% of DC	10	Uniform	0	to	20	
Boiler Feed Water	% of DC	0					
Gas Turbine	% of DC	25	Uniform	0	to	50	
HRSG	% of DC	2.5	Uniform	0	to	5	
Steam Turbine	% of DC	2.5	Uniform	0	to	5	
General Facilities	% of DC	5	Uniform	0	to	10	
MAINTENANCE COSTS[c]							
Coal Handling	% of TC	3					
Limestone Handling	% of TC	3					
Oxidant Feed	% of TC	2	Triangular	1	to	3	(2)
Gasification	% of TC	4.5	Triangular	3	to	6	(4.5)

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Table 15 (Continued). Summary of the Base Case Parameters Values and Uncertainties for the Air-Blown KRW-based IGCC System with Hot Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]			
Sulfation	% of TC	4	Triangular	3	to	6	(4)
Zinc Ferrite	% of TC	3	Triangular	3	to	6	(3)
Boiler Feed Water	% of TC	0					
Gas Turbine	% of TC	2	Triangular	1.5	to	6	(2)
HRSG	% of TC	1.5					
Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					
OTHER FIXED OPERATING COST PARAMETERS							
Labor Rate	$/hr	19.70	Normal	17.70	to	21.70	
VARIABLE OPERATING COST PARAMETERS							
Limestone Sorbent	$/ton	18	Triangular	18	to	25	(18)
Zinc Ferrite Sorbent	$/lb	3.00	Triangular	0.75	to	5.00	(3.00)
Ash Disposal	$/ton	10	Triangular	10	to	25	(10)
DIRECT COST REGRESSION MODEL ERROR TERMS							
Boost Air Compressor	$ Million	0	Normal	-0.66	to	0.66	
Boiler Feedwater	$ Million	0	Normal	-0.78	to	0.78	
HRSG	$ Million	0	Normal	- 17.3	to	17.3	
Steam Turbine	$ Million	0	Normal	- 15.8	to	15.8	

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range. For normal and lognormal distributions, the 99.8 percent probability range is given.

[b] For direct costs, the deterministic values represent "contingency factors" as defined by EPRI (1986) and others. For probabilistic studies, uncertainty in capital cost is represented by an uncertainty factor, which is described by a probability distribution. DC = process area direct cost.

[c] TC = process area total cost, including indirects and contingency

Table 16. Summary of the Base Case Parameter Values and Uncertainties for the Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]
GASIFIER PROCESS AREA				
Gasifier Pressure	psia	(calculated)		
Gasifier Temperature	°F	1,100		
Fines Carryover from Gasifier	wt-% of Coal Feed	5.0	Fractile	5%: 0 to 1
				20%: 1 to 3.5
				25%: 3.5 to 5
				25%: 5 to 8
				15%: 8 to 15
				5%: 15 to 20
				5%: 20 to 30
Fines Capture in Recycle Cyclone	% of Carryover	95	Fractile	25%: 50 to 90
				25%: 90 to 95
				25%: 95 to 97
				25%: 97 to 98
Fines Carbon Content	wt-% of fines	79	Fractile	5%: 65 to 70
				20%: 70 to 75
				25%: 75 to 79
				25%: 79 to 84
				25%: 84 to 87
Carbon Retention in Bottom Ash	wt-% of coal feed carbon	2.5	Triangular	0.75 to 10 (2.5)
Sulfur Retention in Bottom Ash	wt-% of coal feed sulfur	3.0	Triangular	1.5 to 6 (3)
Gasifier Coal Throughput				
250 psia	lb DAF/(hr-ft^2)	266	Triangular	133 to 333 (266)
300 psia	lb DAF/(hr-ft^2)	305	Triangular	152 to 381 (305)
350 psia	lb DAF/(hr-ft^2)	341	Triangular	170 to 426 (341)
Gasifier Ammonia Yield	Equiv. fraction of coal N to NH_3	0.9	Triangular	0.5 to 1.0 (0.9)
Gasifier Air/Coal Ratio	lb air/lb DAF	3.1	Triangular	2.7 to 3.4 (3.1)
Gasifier Steam Requirement				
Air/coal = 2.7	lb H_2O/lb DAF	0.81	Uniform	0.54 to 1.08
Air/coal = 3.1	lb H_2O/lb DAF	1.55	Uniform	1.24 to 1.86
Air/coal = 3.4	lb H_2O/lb DAF	2.38	Uniform	2.04 to 2.72
ZINC FERRITE DESULFURIZATION PROCESS AREA				
Residual Sulfate After Oxidative Regen.	mol-% of captured S	7.5	Triangular	3 to 11 (7.5)
Residual Sulfide After Reductive Regen.	mol-% of S in sulfate	85	Triangular	50 to 90 (85)
Sorbent Sulfur Loading	wt-% S in sorbent	17	Normal	2.16 to 31.84 (17)

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Table 16 (Continued). Summary of the Base Case Parameter Values and Uncertainties for the Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]			
Sorbent Attrition Rate	wt-% sorbent loss/cycle	1.0	Fractile	5%: 0.17	to	0.34	
				20%: 0.34	to	0.5	
				25%: 0.5	to	1	
				25%: 1	to	1.5	
				20%: 1.5	to	5	
				5%: 5	to	25	
Absorber Pressure Drop	psi/ft bed height	0.4	Triangular	0.29	to	0.53	(0.4)
Absorption Cycle Time	hours	30					
Max. Vessel Diameter	ft	12.5					
Max. Vessel Height	ft	37.5					
<u>GAS TURBINE PROCESS AREA</u>							
Pressure Ratio	ratio	13.5					
Turbine Inlet Temp	oF	2,300					
Exhaust Flow	lb/sec	938					
Thermal NO_x	fraction of air nitrogen fixated	$4.25x10^{-5}$	Uniform	$1.0x10^{-5}$	to	$7.5x10^{-5}$	
Fuel NO_x	% conversion of NH_3 to NO_x	90	Triangular	50	to	100	(90)
Unconverted CO	wt-% of CO in fuel gas	0.9885	Uniform	0.9772	to	0.9999	
<u>CAPITAL COST PARAMETERS</u>							
Engineering and Home Office Fee	fraction	0.10	Triangular	0.07	to	0.13	(0.10)
Indirect Construction Cost Factor	fraction	0.20	Triangular	0.15	to	0.25	(0.20)
Project Uncertainty	fraction	0.175	Uniform	0.10	to	0.25	
General Facilities	fraction	0.20					
<u>DIRECT COSTS</u>[b]							
Coal Handling	% of DC	5	Uniform	0	to	10	
Oxidant Feed	% of DC	10	Uniform	0	to	20	
Gasification	% of DC	20	Uniform	10	to	30	
Cyclones	% of DC	5	Uniform	0	to	10	
Zinc Ferrite	% of DC	40	Uniform	0	to	80	
Sulfuric Acid Plant	% of DC	10	Uniform	0	to	20	
Boiler Feed Water	% of DC	0					
Gas Turbine	% of DC	25	Uniform	0	to	50	
HRSG	% of DC	2.5	Uniform	0	to	5	
Steam Turbine	% of DC	2.5	Uniform	0	to	5	
General Facilities	% of DC	5	Uniform	0	to	10	

(Continued on next page)

Table 16 (Continued). Summary of the Base Case Parameter Values and Uncertainties for the Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup.

Description	Units	Deterministic Value	Distribution	Parameters[a]			
MAINTENANCE COSTS[c]							
Coal Handling	% of TC	3					
Oxidant Feed	% of TC	2	Triangular	1	to	3	(2)
Gasification	% of TC	3	Triangular	2	to	12	(3)
Cyclones	% of TC	3	Triangular	1.5	to	4.5	(3)
Zinc Ferrite	% of TC	3	Triangular	3	to	6	(3)
Sulfuric Acid Plant	% of TC	2					
Boiler Feed Water	% of TC	1.5					
Gas Turbine	% of TC	2	Triangular	1.5	to	6	(2)
HRSG	% of TC	1.5					
Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					
OTHER FIXED OPERATING COST PARAMETERS							
Labor Rate	$/hr	19.70	Normal	17.70	to	21.70	
VARIABLE OPERATING COST PARAMETERS							
Zinc Ferrite Sorbent	$/lb	3.00	Triangular	0.75	to	5.00	(3.00)
Ash Disposal	$/ton	10	Triangular	10	to	25	(10)
Sulfuric Acid Byproduct	$/ton	40	Triangular	0	to	60	(40)
Byproduct Marketing	fraction	0.10	Triangular	0.05	to	0.15	(0.10)
DIRECT COST REGRESSION MODEL ERROR TERMS							
Boiler Feedwater	multiplier	0	Normal	-0.78	to	0.78	
HRSG	$ Million	0	Normal	- 17.3	to	17.3	
Steam Turbine	$ Million	0	Normal	- 15.8	to	15.8	
Boost Air Compressor	$ Million	0	Normal	- 0.66	to	0.66	
Lurgi Coal Handling	$ Million	0	Normal	- 14.4	to	14.4	
Sulfuric Acid Plant	$ Million	0	Normal	- 4.0	to	4.0	
AUXILIARY POWER REGRESSION MODEL ERROR TERMS							
Lurgi Coal Handling	MW	0	Normal	- 0.35	to	0.35	

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range. For normal and lognormal distributions, the 99.8 percent probability range is given.

[b] For direct costs, the deterministic values represent "contingency factors" as defined by EPRI (1986) and others. For probabilistic studies, uncertainty in capital cost is represented by an uncertainty factor, which is described by a probability distribution. DC = process area direct cost.

[c] TC = process area total cost, including indirects and contingency

the gasifier steam requirement conditioned on the gasifier oxidant requirement. The latter is a case in which the expert explicitly considered the correlation structure between two uncertainties. In later comparative case studies, the judgment of Expert LG-2 is also used.

For the zinc ferrite process areas, the judgments of Expert ZF-1 are taken as a base case. The judgment sets LG-1 and ZF-1 are the only pair of separate process areas for which judgments were obtained from the same individual. Expert ZF-1 provided estimates of uncertainties for five parameters in the zinc ferrite process area, as well as suggestions for other key deterministic design assumptions, which are included in Table 16. The judgments obtained from Experts ZF-2 and ZF-3 are also considered in later comparative analysis.

Complete details regarding the technical judgments of the gasification and zinc ferrite desulfurization experts are given in Appendix B. The basis for the estimates by Expert LG-1 are discussed in detail in Appendix B.3.4.1. The judgments of Expert ZF-1 are discussed in Appendix B.5.3.1. The judgments of the other gasifier and zinc ferrite experts are also described in Appendices B.3.4 and B.5.3, respectively.

As for the KRW systems, the estimates of uncertainty in the gas turbine process area were developed by the author based on published data and discussions with process engineers, as discussed in Appendix B.6.7. The capital cost parameter, direct cost, maintenance cost, labor rate, unit cost, and regression model uncertainties were estimated in the same manner as for the KRW systems. See Appendix B.7 for more discussion of these.

4.4 Correlation Structures

While the judgment set of Expert LG-1 provides an example of judgment regarding correlated uncertainties (i.e. gasifier oxidant and steam requirements), the other judgments regarding uncertainties were elicited or developed assuming no correlations. For the base case assumptions, the uncertain parameters are assumed to be statistically independent. However, the results of a probabilistic could be influenced by correlation structures among uncertain variables, depending on the strength of the correlations. While correlations were not elicited from the technical experts, an effort was made to identify mechanisms which would tend to cause simultaneous effects in two or more parameters. For example, according to Expert LG-1, the mechanism by which organically-bound sulfur is released from the coal is associated with the carbon conversion rate. Therefore, high carbon conversions tend to result in high sulfur release.

The information obtained from experts regarding possible relationships among performance variables was used to construct illustrative correlation structures for the gasification and zinc ferrite process areas. These assumptions are described in Appendix B. The correlation structures are considered as probabilistic "sensitivity" cases in comparison to the uncorrelated base case assumptions. The implications of correlation structures are considered in later sections.

4.5 The "Best Guess"

The point values used here for performance and cost parameters in deterministic case studies correspond to a central value of the uncertainty. Depending on the type of distribution, the deterministic value may be the same as the mode, median, or mean of the probability distribution. The experts whose judgments were obtained in this study indicated that the "best guess" was the same as the median or mode. For a symmetric probability distribution, the mean, mode, and median coincide. However, for skewed distributions, these three measures of the central tendency may be different. For example, in a unimodal positively skewed distribution, the mean is greater than the median, which is greater than the mode. Thus, the use of the mode as a best guess leads to an underestimate of the average of the distribution.

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5.0 MODELING APPLICATIONS AND RESULTS: PULVERIZED COAL-FIRED SYSTEMS

Applications of the probabilistic engineering models for evaluation of the conventional FGD/SCR and advanced fluidized bed copper oxide process emission control systems are described here. The focus here is on developing probabilistic comparisons of the cost of the copper oxide process compared to the convention technology. Therefore, estimates of uncertainty in the copper oxide process are developed first. Then, probabilistic comparisons between the copper oxide process and the wet FGD/SCR systems are made. Both systems are assumed to achieve 90 percent SO_2 and NO_x reduction, and to comply with NSPS for particulate matter emissions. The detailed assumptions regarding performance and cost parameters are described in Chapter 4.

5.1 Running the Models

The models of both the FGD/SCR and copper oxide systems are part of the IECM. The IECM is run interactively on a Macintosh II computer. Running the IECM involves three principal steps. The first is to configure a power plant for analysis. The user specifies the set of pre-combustion, -combustion, and post-combustion technologies of interest, along with the associated waste disposal method. In this study, two configurations are used to represent power plants with the FGD/SCR system and, in separate runs, the fluidized bed copper oxide process. Performance and cost models of each component of the power plant and emission control system are included for each case study.

Next, the user specifies the values of model parameters related to control technology design, power plant characteristics, fuel specifications, and environmental regulatory constraints. Economic and financial parameters also are specified at this stage. For a typical analysis of a specific power plant and emission control system configuration, on the order of 50 parameters must be specified. Default values of all model parameters are included in the model. The user may override these defaults either by editing the computer code, or by interactively making changes while running a simulation. Model parameters may be defined either as single-value deterministic numbers, or as any of several probability distributions available as system functions in the IECM modeling environment.

Once all input parameters are set, the model is then executed interactively. Several standard summary tables may be called by the user, or the user may easily call for any performance or economic parameter of interest. In addition, the user may easily perform sensitivity analysis, or change the uncertainty assumptions of any of the input parameters, by interactively changing the definition of specific variables. Because the modeling environment is interactive, results are not calculated until the user asks for them. Typically, the levelized cost of electricity requires the most time to compute, because it depends on all of the performance and cost parameters and analytical models that influence levelized cost. For a probabilistic simulation involving 100 or 150 samples, the user must wait typically only a few minutes to obtain an answer.

The input assumptions for the case studies here are discussed and summarized in Chapter 4. Assumptions regarding emission constraints, base plant design, and financial parameters summarized in Table 8. Table 9 summarizes the different coals considered, including both unwashed and cleaned (30 percent sulfur reduction on an energy basis) coals. Both medium and high sulfur coals are assumed. Table 10 summarizes key deterministic input values and uncertainties for the conventional wet FGD/SCR emission control system, which is taken as the technological baseline in the comparisons with the copper oxide process. The key deterministic inputs and uncertainties assigned to the copper oxide emission control system are summarized in Table 11.

5.2 Characterizing Uncertainty in Capital Costs

Nearly all deterministic capital cost estimates, whether for a new or existing technology or for a preliminary or detailed cost estimate, include a contingency factor. As discussed in Section 1.1.3, contingency factors are intended to reflect the uncertainties in capital cost estimate. The contingency is often the single largest expense in the cost estimate, and yet it is also the least documented. Contingency factors are typically simple multipliers that are applied to installed equipment costs toward the end of an analysis (e.g., after process area costs have been estimated without regard to their uncertainty). according to a study by the Rand Corporation, contingency factors are often badly under-estimated (Milanese, 1987). This results in misleadingly optimistic cost estimates, particularly for innovative technologies in early stages of development. Such estimates are used for decision making regarding technology selection and research.

A probabilistic modeling approach supplants the traditional contingency factor approach by incorporating expert knowledge about uncertainties explicitly and at a more disaggregated level (e.g., for specific performance and cost parameters). Furthermore, while simple contingency factors provide no explicit insights into the specific performance or cost parameters that contribute most to the process technical and economic risks, a probabilistic approach permits identification and ranking of the uncertain parameters that contribute most to the overall uncertainty, which is discussed in the next section. Because the uncertainties contributing to "contingencies" are considered at a disaggregated level in probabilistic analysis, more realistic estimates of performance and cost will generally result.

The uncertainty in the total capital cost for the fluidized bed copper oxide process, for the case of an unwashed Illinois No. 6 coal and sulfuric acid byproduct recovery, is shown as a cdf in Figure 16. There is a five percent probability that the capital cost would be less than 90 million dollars, and a 95 percent probability that it will be less than 135 million dollars. Thus, the 90 percent probability range for total capital cost encompasses 45 million dollars, or approximately 50 percent of the lower end of the range. However, there is a small probability that costs could go as low as 82 million dollars, or as high as 160 million dollars. The mean value of capital cost obtained from the probabilistic simulation is $111 million dollars. The mean is slightly higher than the median (50th percentile) value, because the uncertainty in capital cost is positively skewed. If the mean value of the probability distribution were used as the budgetary cost estimate for this case, there would be a 45 percent probability that costs could be higher.

By contrast, a deterministic cost estimate, using the "nominal" assumptions regarding performance and cost parameters discussed in Chapter 4, would yield a capital cost of $74 million, excluding contingencies costs. Based on a published design study of the copper oxide process, a project contingency factor of 25 percent and a process contingency factor of 30 percent are assumed (SMC, 1983c). This yields a capital cost estimate of 96 million dollars. Based on the cdf in Figure 16, there is approximately a 90 percent chance that capital cost will be higher than 96 million dollars. Clearly, the deterministic cost estimate using simple contingency factors is not accounting for the interactions among uncertainties that are considered in the probabilistic estimate. A 90 percent chance of cost overrun would be unacceptable to any reasonable decision maker.

A deterministic capital cost estimate can include information developed in a probabilistic estimate through appropriate selection of the contingency factor. Figure 17 illustrates the sensitivity of the deterministic cost estimate to the contingency factor. The contingency factor can be defined as the value that adjusts the deterministic estimate (without contingency) to some specified fractile of the probabilistic estimate. Typically, some "best estimate" value from the probabilistic analysis, such as the mean or the median, would be used. However, if there is significant risk aversion on the part of an investor, who may want to minimize the chance of a cost over-run, then an upper fractile from the probability distribution (e.g., 90th percentile) may be used.

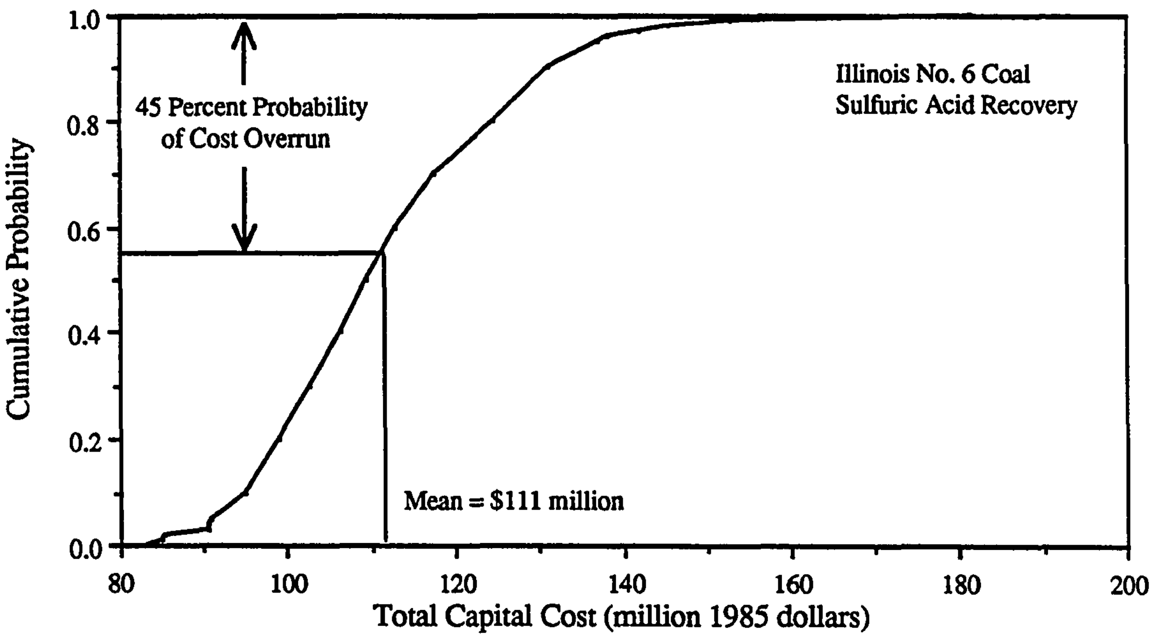

Figure 16. Uncertainty in Total Capital Cost for the Copper Oxide Process

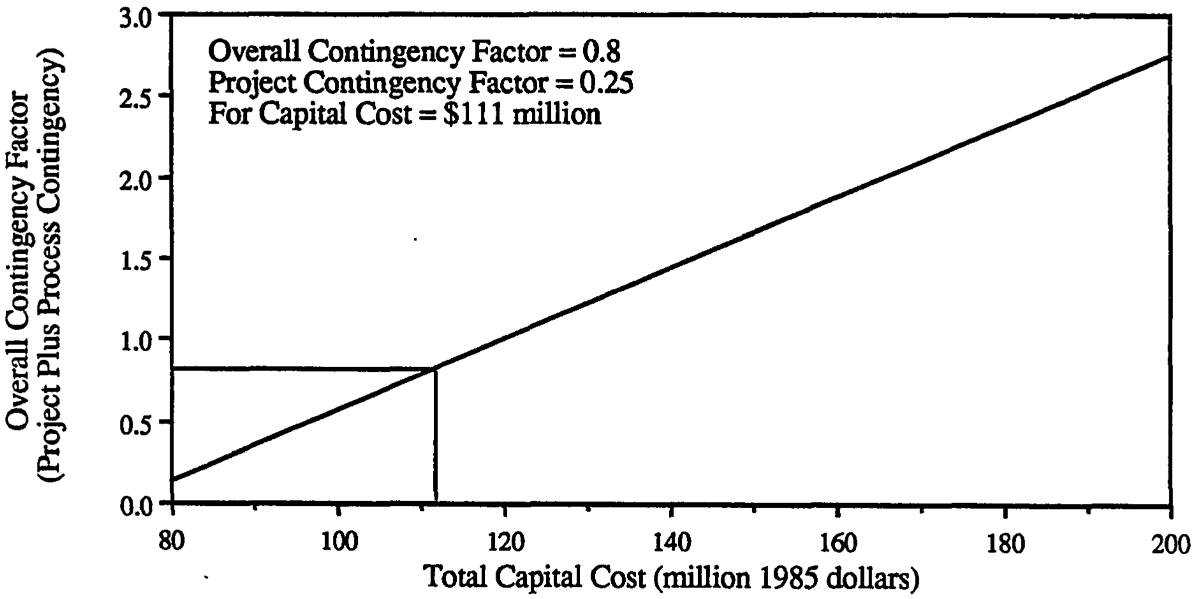

Figure 17. Overall Contingency Factor and Total Capital Cost for the Copper Oxide Process

If the mean value from the probabilistic analysis is selected as the budgetary capital cost estimate, then a overall (project plus process) contingency factor of 80 percent is implied in the deterministic analysis. This overall contingency factor is significantly higher than the 55 percent value (30 percent process contingency, 25 percent project contingency) assumed in previous analyses. Contributing factors to the difference are the uncertainties assigned to the regeneration efficiency and the capital costs for each major process section, which are skewed. The difference is not surprising, since the previous contingency was based on a rule-of-thumb, rather than a detailed probabilistic risk assessment. The fact that the original estimate seems to be low is also supported by the results of the Rand study, which indicates that contingency factors are generally grossly under-estimated, especially early in process development.

5.3 Identifying Key Uncertainties

The primary advantage of probabilistic simulation over traditional sensitivity analysis is the simultaneous incorporation of uncertainties in multiple model inputs. The resulting interactions among uncertain variables results in uncertainties in total costs, which are the basis for comparative analysis. Research can provide additional information about the uncertain input variables, resulting in changes in their uncertainty distributions (such as the mean or standard deviation) and, in turn, in the overall uncertainties of the technology. Therefore, it may be fruitful to reduce the uncertainties of key variables that contribute most to the risk of technology failure.

The key parameter uncertainties have been identified primarily by estimating sample correlations between the primary cost results, such as total levelized revenue requirement, and the copper oxide process input uncertainties included in Table 11. Correlations provide a measure of the linear dependence of one distribution on another; however, there are some non-linear relationships in the model, such as between sorbent flow rate and regeneration efficiency. Scatter plots can be used to visually identify non-linear dependencies that may not be well-characterized by correlation coefficients.

The factors which contributed most to uncertainty in the total levelized process cost were uncertainties in sorbent attrition, regeneration efficiency, and copper-to-sulfur molar ratio, with correlations of 0.55, -0.41, and 0.41, respectively. Uncertainties in sorbent cost and plant capacity factor also were significant. Scatter plots did not reveal any strong non-linear dependencies. These results suggest that further research on the copper oxide process should focus on improving understanding of sorbent attrition, regeneration efficiency, and the variability in the copper-to-sulfur molar ratio required to achieve a given SO_2 removal efficiency.

5.4 Process Design Trade-offs Under Uncertainty

While ultimately we are interested in how the advanced copper oxide process compares with the conventional FGD/SCR technology, it is important first to evaluate the key performance and cost trade-offs which affect the economics of the copper oxide process. Thus, an analysis of performance and cost trade-offs was done to select values of key design parameters such as fluidized bed height, air preheater size, weight percent of copper in the sorbent, and sulfur recovery option. Furthermore, the model was used to identify potential market niches where process costs are likely to be low, such as for certain coal characteristics (including coal cleaning).

The evaluation of design trade-offs must consider performance and cost interactions between the control technology and the balance of the power plant system, in addition to interactions within the technology itself. Thus, comparisons between copper oxide design alternatives were made on the basis of total pollution control system costs, which are exclusive of the base plant and include SO_2, NO_x, and PM removal, solid waste handling, and coal cleaning. Any emission control system-related changes to the base plant are charged to the pollution control system. As a result, interactions between components of the pollution control system and between the pollution control system and the base plant are integrated into the analysis. Furthermore, because design decisions may be affected by process uncertainties, the analysis was based on probabilistic estimates of the costs associated with various design decisions.

5.4.1 Absorber Design

The absorber is the principal vessel for SO_2 and NO_x removal, and the height of the fluidized bed is a key design parameter. "Conventional wisdom" among process developers was that increasing the bed height relative to the nominal design value of 36 inches would increase overall process costs since the higher pressure drop across the bed would increase the flue gas fan energy requirements. However, data indicated that increasing the fluidized bed height also increased the available copper in the bed, reducing the required copper-to-sulfur molar ratio (see Equation 16). This, in turn, reduces the required sorbent flow rate in the system. Because much of the process equipment size is based on the sorbent flow rate, a reduction in sorbent flow yields capital and operating cost savings, offsetting the increased energy costs for the absorber.

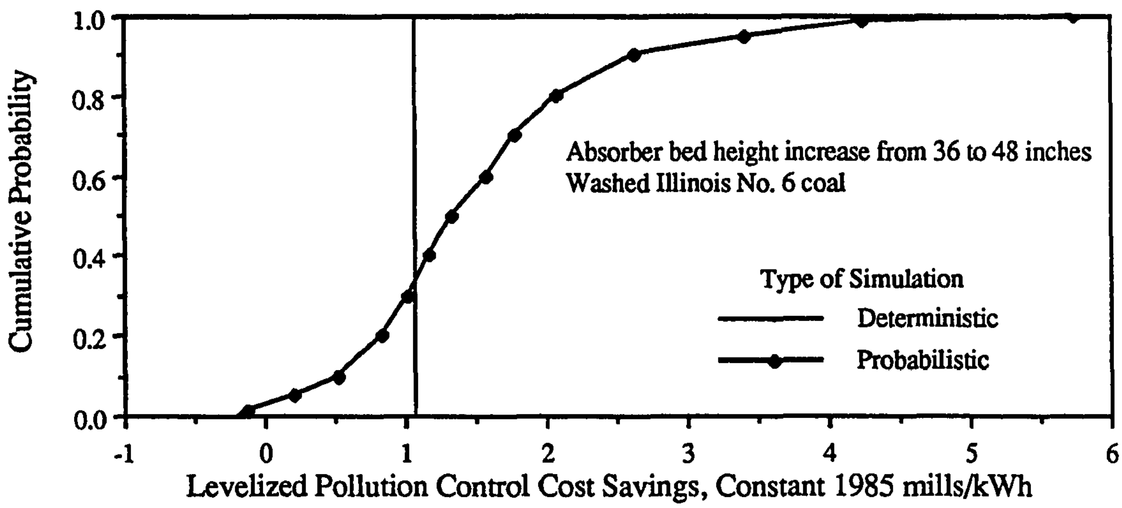

Figure 18. Uncertainty in Levelized Cost Savings Due to Increasing Fluidized Bed Height.

The integrated process model allowed these design tradeoffs to be quantified. Figure 18 shows the results of both a deterministic and probabilistic simulation of the pollution control cost savings from increasing the absorber bed height from 36 to 48 inches using a washed Illinois No. 6 coal. The deterministic case uses the nominal parameter values in Tables 8, 9, and 11, while the probabilistic case uses distributions. The uncertainty in the cost *difference* was calculated using identical paired samples for the input distributions which take into account the underlying correlation between the two cases when only the bed height is changed.

The probabilistic simulation reveals that there is a very small probability (about two percent) that the increase in flue gas fan operating costs could outweigh the cost reductions associated with reduced sorbent circulation rates. On the other hand, there is a 65 percent chance that the cost savings could exceed the deterministic estimate of about 1.1 mill/kWh, with about a 10 percent chance the savings could be over 2.6 mills/kWh. The skewness of the uncertainty distribution in Figure 18 results from assumptions about key uncertain parameters such as the regeneration efficiency. Similar results were obtained for other coals (e.g., run-of-mine Illinois No. 6 and Pittsburgh No. 8) and for other sorbent copper loadings (e.g., 5 and 10 percent).

5.4.2 Sorbent Copper Loading

Increasing the sorbent copper content can reduce the sorbent mass flow. The primary tradeoff is the potential for increased sorbent cost and attrition. An engineering study for DOE, however, reported that copper loadings of from 5 to 11 percent have comparable attrition characteristics and similar manufacturing costs (SMC, 1983). To date,

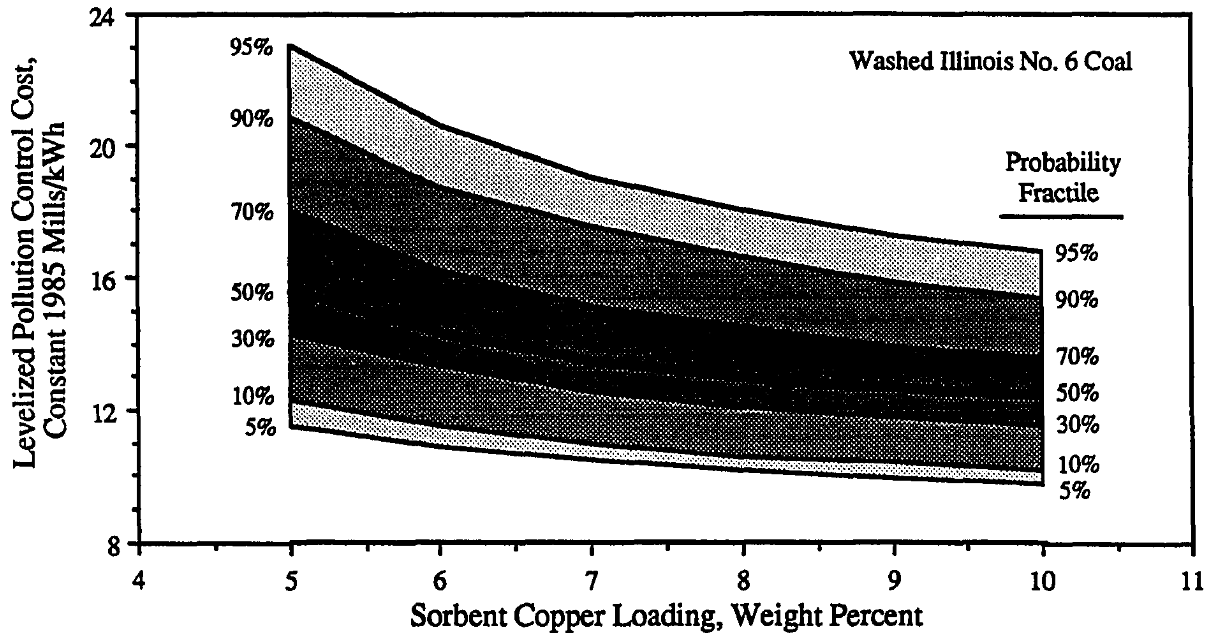

Figure 19. Uncertainty in Levelized Pollution Control Cost vs. Sorbent Copper Loading.

most testing has been with sorbents of 5 to 7 percent copper (PETC, 1984; Plantz et al, 1986; Williamson et al, 1987). Figure 19 shows the effect of increasing the sorbent copper loading from 5 to 10 percent for a washed Illinois No. 6 coal. The figure also shows the associated uncertainty in levelized revenue requirement. As the copper content increases, the median and variance of the revenue requirement decrease. There is a significant cost advantage for the 7 percent sorbent compared to the 5 percent case, with a mean savings of 2.3 mills/kWh. Additional savings may be realized by increasing the copper content to 10 percent. These results indicate that additional research on sorbent attrition at the higher copper loadings is merited.

5.4.3 Energy Recovery System

Another process integration issue is the recovery of energy added to the flue gas by the exothermic reactions in the fluidized bed absorber. For the deterministic parameter assumptions, in Table 11, the incremental capital cost of the enlarged preheater slightly outweighed (by only 0.01 mills/kWh) the cost savings associated with a reduced fabric filter size and an increased energy credit. However, when the same analysis was performed including uncertainties, the likelihood of a small cost advantage from enlarging the air preheater was found (Figure 20). The difference in results was due to the skewness of many of the distributions assigned to key parameters in the probabilistic model.

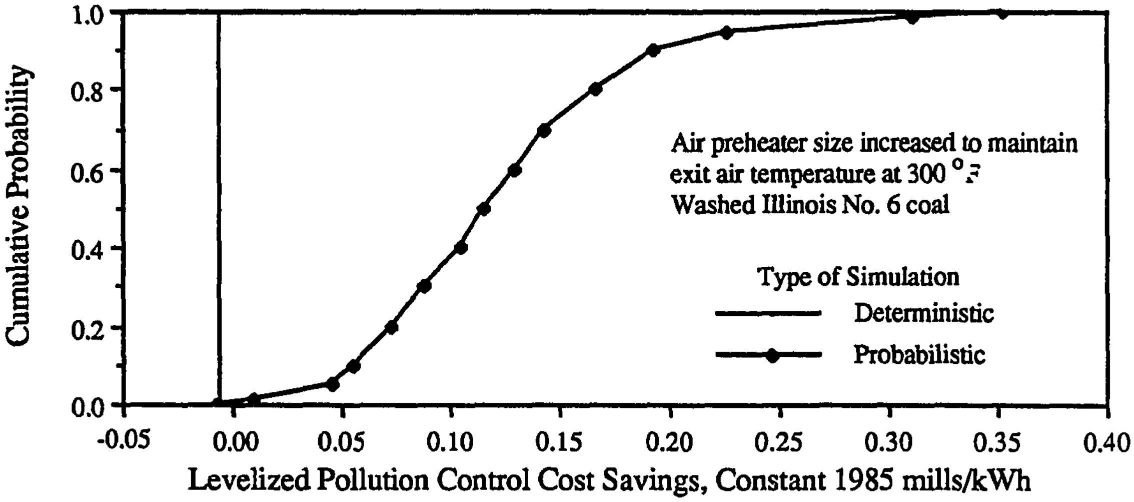

Figure 20. Uncertainty in Levelized Cost Savings Due to Enlarging the Air Preheater.

5.4.4 Integrated Coal Cleaning

Because many of the costs of the copper oxide process are sensitive to the sorbent flow rate, which in turn is proportional to the coal sulfur content, a reduction in sulfur content through coal cleaning can lower the cost of the process. The integrated environmental control model determines how these savings compare to the increased cost of cleaned coal.

As in the air preheater sizing analysis, deterministic and probabilistic simulations yielded qualitatively different results regarding the net cost savings associated with using a washed Illinois No. 6 coal. As shown in Figure 21, the deterministic analysis indicated an overall cost penalty, whereas results of the probabilistic simulation ranged from a net cost penalty to a 55 percent chance of a cost *savings*. The cost penalty outcomes stem from the low copper-to-sulfur molar ratios that are achievable with high regeneration efficiencies. In such instances, the process cost savings from lower sorbent circulation rates are too small to offset the higher coal prices associated with coal preparation. The probabilistic analysis, however, reflects the possibility of lower regeneration efficiencies, requiring higher sorbent circulation rates. In these cases, the benefits of a lower sulfur content outweigh the costs of coal cleaning, yielding a net savings.

The degree of sulfur reduction achieved through coal cleaning also affects the cost results. Figure 22 shows the mean cost of each component of the emissions control system associated with various levels of coal preparation. For the example above, a 30 percent sulfur reduction was assumed. The figure indicates the contribution of each part of the emission control system to the total levelized cost of pollution control. It is clear from the

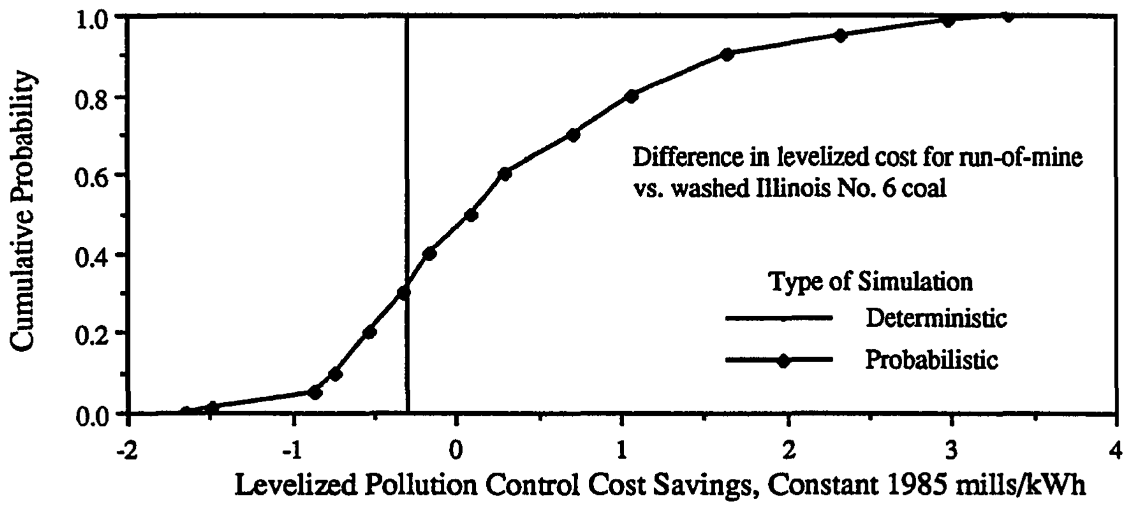

Figure 21. Uncertainty in Levelized Cost Savings Due to Coal Cleaning for the Copper Oxide/Sulfuric Acid Recovery with Illinois No. 6 Coal

figure that the cost of the copper oxide process is sensitive to the coal sulfur content. Figure 23 shows similar results, assuming that sulfur, rather than sulfuric acid, is recovered as a byproduct. In this case, the levelized costs are somewhat higher, but exhibit the same trends as for sulfuric acid recovery. Here, again, 30 percent coal sulfur cleaning is the least cost option, based on probabilistic analysis.

In contrast, the FGD/SCR system is comprised of two separate reactor vessels for SO_2 and NO_x control, both of which are proportional in cost primarily to the flue gas volumetric flow rate, and not significantly influenced by coal cleaning. Thus, the mean pollution control system costs for FGD/SCR with Illinois No. 6 coal increase as the level of coal cleaning is increased. This result is shown in Figure 24. Note that the sum of the FGD and SCR costs decreases only slightly as coal sulfur content is reduced via coal cleaning, in contrast to the more marked cost reduction for the copper oxide systems shown in Figures 22 and 23.

Separate analyses for the Pittsburgh No. 8 coal indicated that the cost of coal cleaning was always larger than the incremental savings for the copper oxide process, for both sulfuric acid and sulfur byproduct recovery. Similarly, results for the conventional FGD/SCR plant indicated that run-of-mine coals always gave the least cost solution. These results are shown graphically in Figures 25, 26, and 27, respectively. In the comparisons that follow, therefore, coal cleaning is assumed only for the copper oxide systems with the Illinois No. 6 coal.

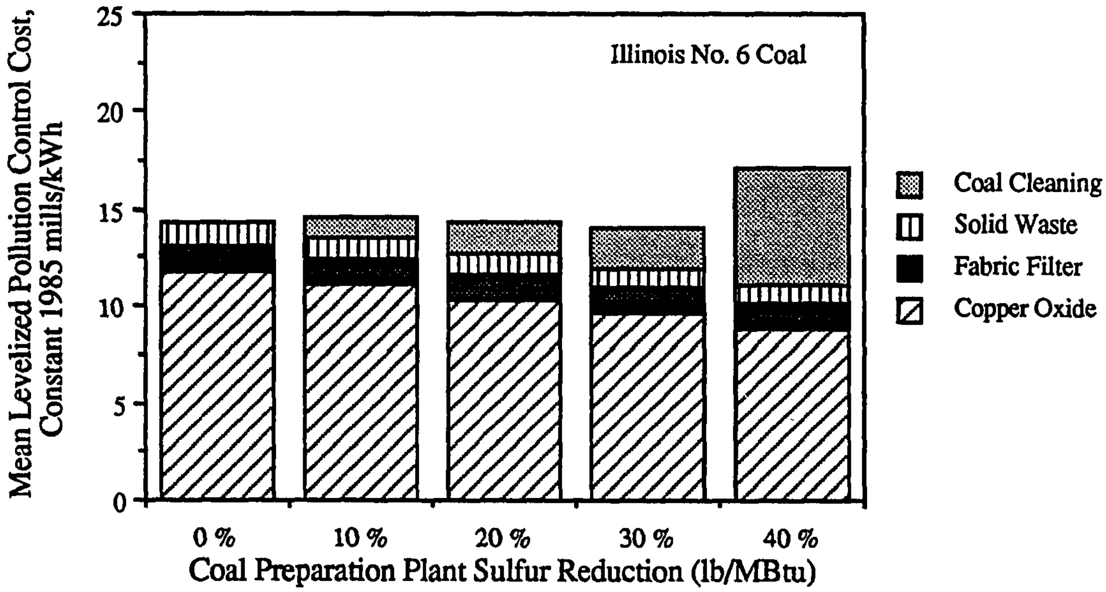

Figure 22. Mean Levelized Pollution Control Cost versus Sulfur Reduction from Coal Cleaning: Copper Oxide/Sulfuric Acid Plant with Illinois No. 6 Coal

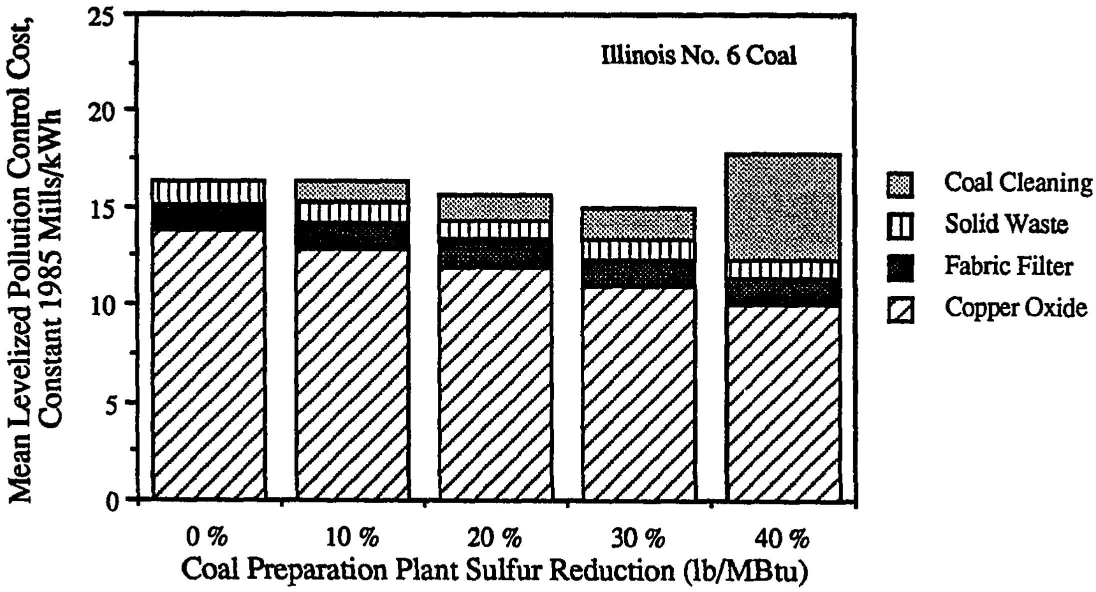

Figure 23. Mean Levelized Pollution Control Cost versus Sulfur Reduction from Coal Cleaning: Copper Oxide/Sulfur Plant with Illinois No. 6 Coal

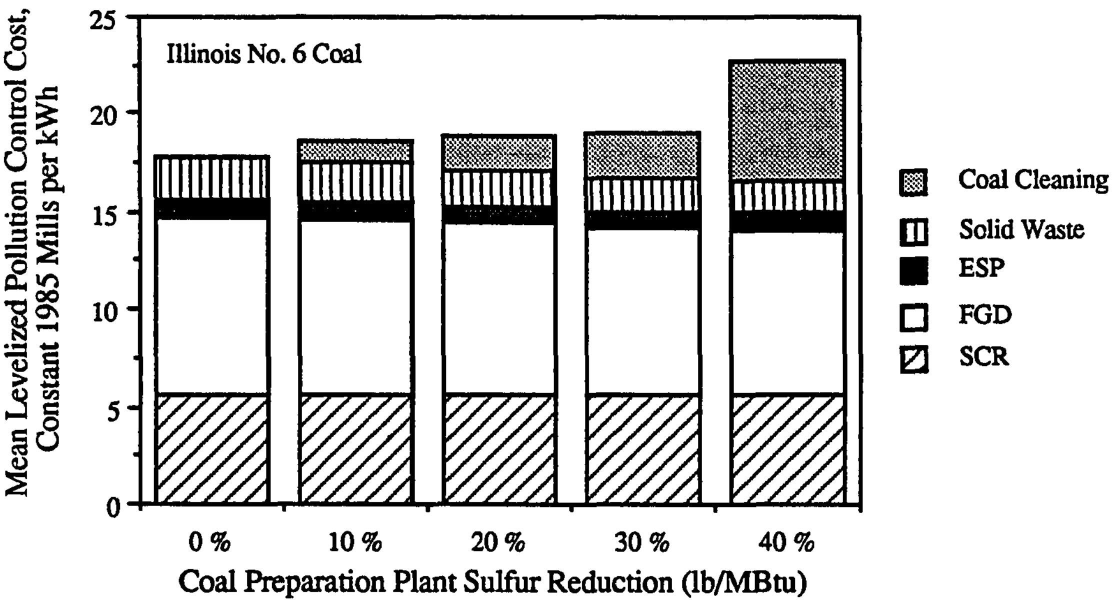

Figure 24. Mean Levelized Pollution Control Cost versus Sulfur Reduction from Coal Cleaning: FGD/SCR with Illinois No. 6 Coal

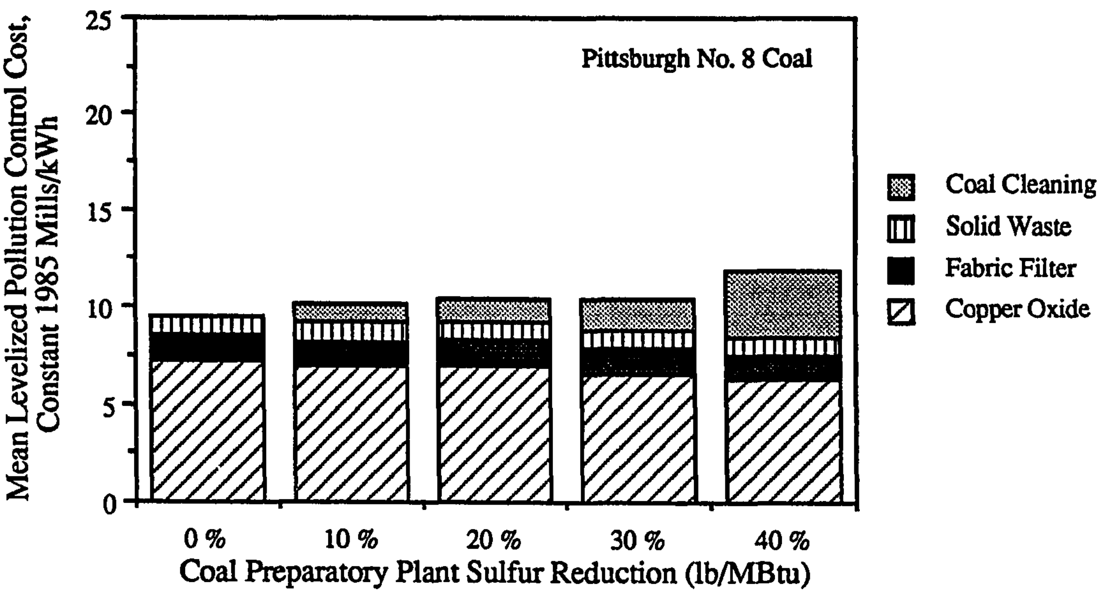

Figure 25. Mean Levelized Pollution Control Cost versus Sulfur Reduction from Coal Cleaning: Copper Oxide/Sulfuric Acid Plant with Pittsburgh No. 8 Coal

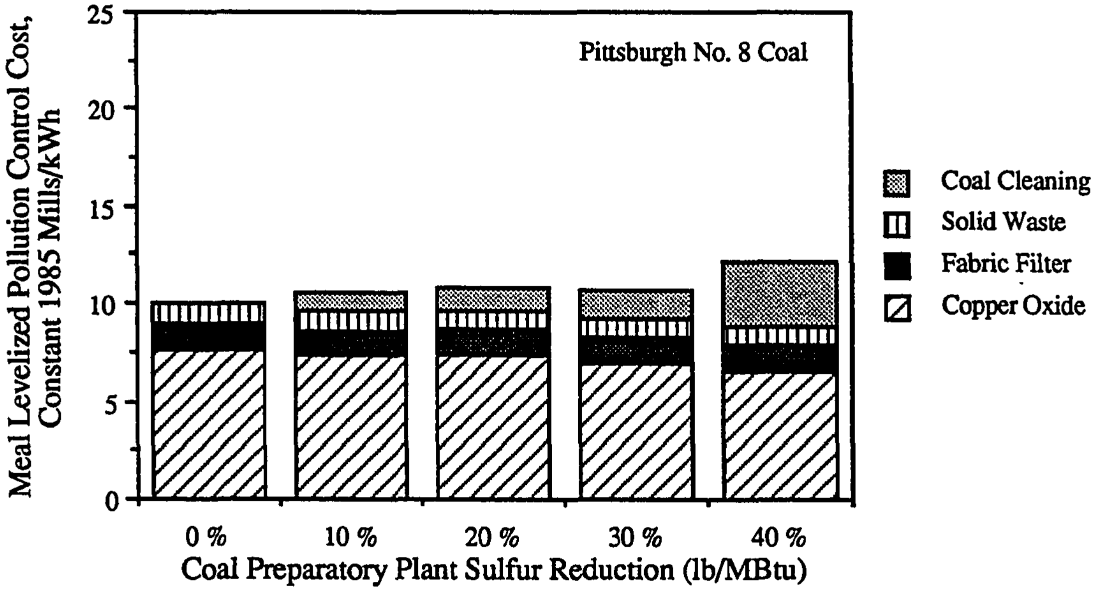

Figure 26. Mean Levelized Pollution Control Cost versus Sulfur Reduction from Coal Cleaning: Copper Oxide/Sulfur Plant with Pittsburgh No. 8 Coal

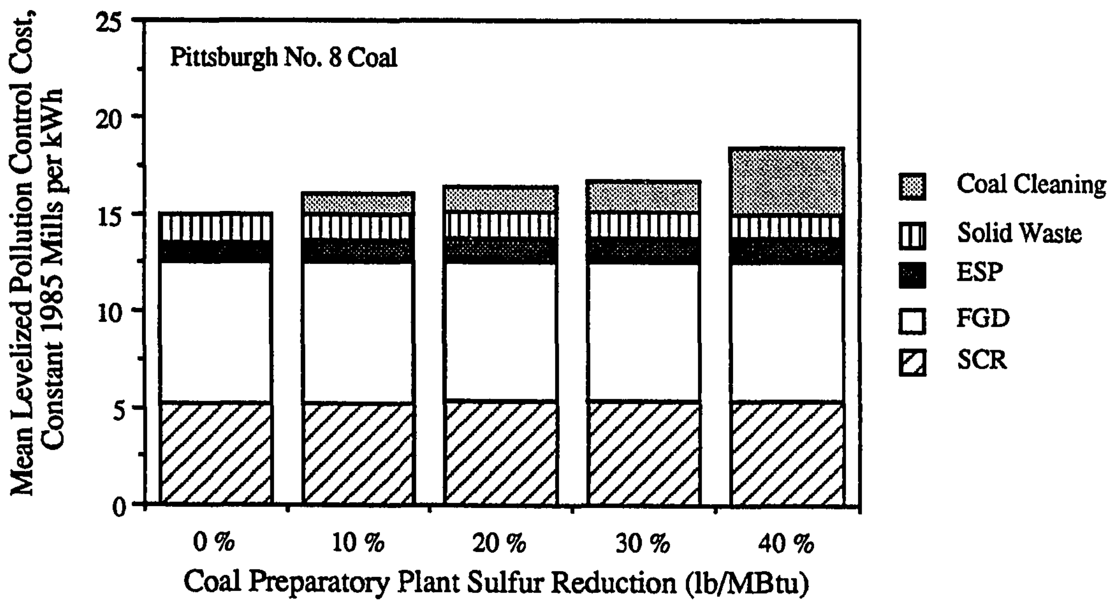

Figure 27. Mean Levelized Pollution Control Cost versus Sulfur Reduction from Coal Cleaning: FGD/SCR with Pittsburgh No. 8 Coal

5.5 Probabilistic Comparisons: Copper Oxide vs. FGD/SCR

A key measure of the viability of advanced environmental control systems is whether they yield a cost savings compared to currently available technology. For comparative analyses, the integrated emission control system including the copper oxide process is compared to the baseline system with FGD and SCR. For this analysis, four separate comparisons are considered. These involve two different coals, which affects both the copper oxide and FGD/SCR systems, and two byproduct recovery options for the copper oxide process.

Because many of the input parameter distributions are common to both systems (e.g., financial parameters, base plant characteristics, solid waste disposal, and ammonia cost), there is, in general, a positive correlation between the cost distributions for the two systems. Therefore, as discussed earlier, probability distributions again have been calculated for the cost *differences* between the copper oxide and FGD/SCR systems using paired samples of input distributions for each system.

Figure 28 shows the uncertainty in the difference in the total capital for pollution control between the copper oxide and FGD/SCR systems. A positive value of cost savings indicates that the copper oxide process is less expensive than the FGD/SCR system. For three of the options considered here, the probabilistic simulation indicates that there is certainty that the copper oxide process would be less expensive than the FGD/SCR system. Furthermore, there is over a 95 percent probability that the cost savings would be $50/kW

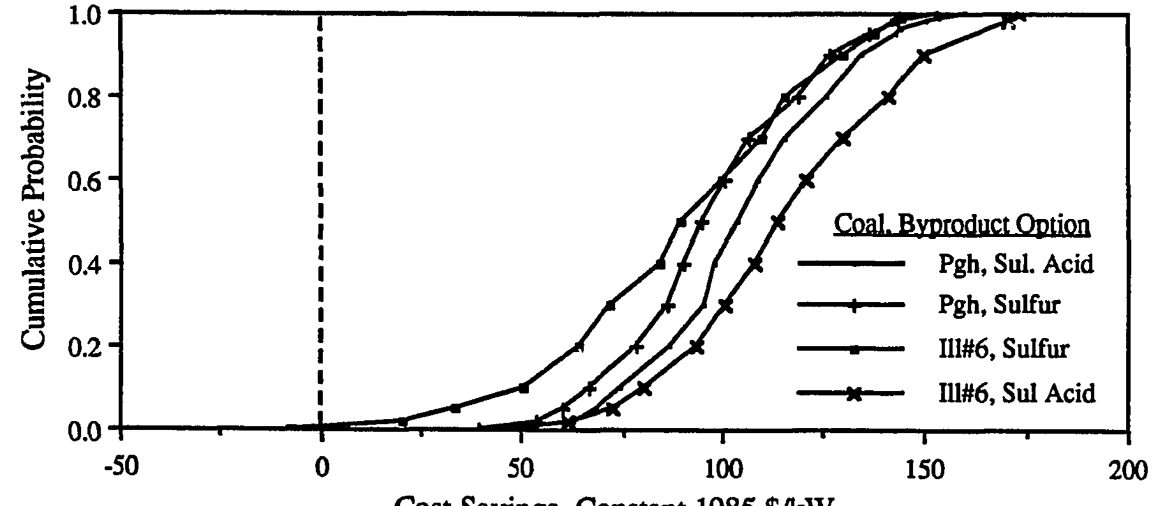

Figure 28. Comparison of Total Capital Cost Savings for Copper Oxide vs. FGD/SCR Systems: Effect of Coal and Byproduct Recovery Options.

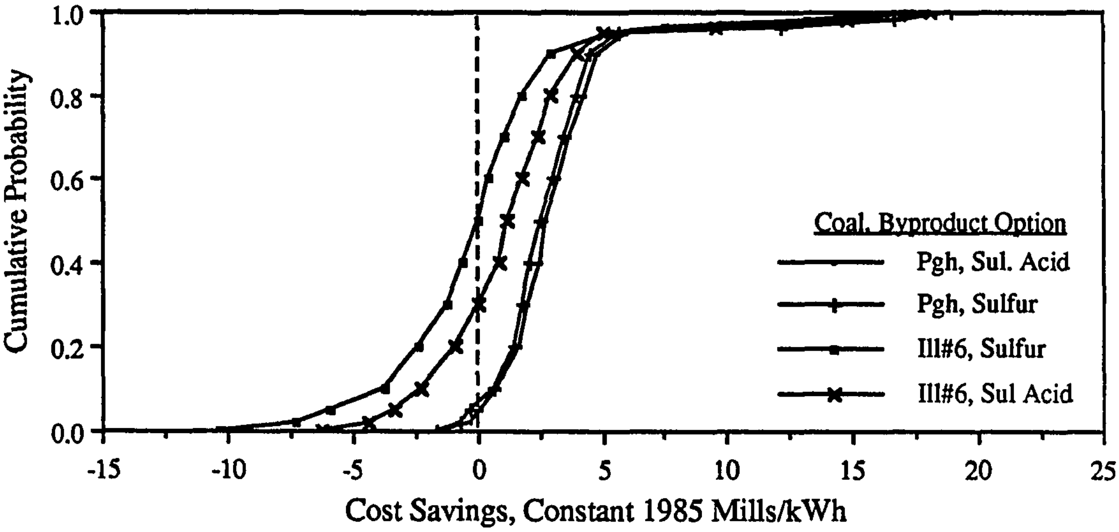

Figure 29. Comparison of Variable Operating Cost Savings for Copper Oxide vs. FGD/SCR Systems: Effect of Coal and Byproduct Recovery Options.

or greater, with median cost savings of around \$100/kW. For the case involving high sulfur Illinois No. 6 coal and sulfur byproduct recovery, the result of the probabilistic simulation indicates about a two percent probability that the copper oxide process would have higher capital cost than the FGD/SCR system. However, there is a 90 percent probability that there would be a cost savings of more than \$50/kW.

Somewhat different results are obtained when comparing the variable operating costs of the advanced and conventional emission control systems, as shown in Figure 29. For the cases involving medium sulfur coal, there is over a 90 percent probability that the copper oxide systems will have lower variable operating costs than the conventional technology. The median cost savings is about 5 mills/kWh for both cases. However, there is about a five percent probability that the variable operating costs will be higher. These outcomes are associated with high sorbent costs, which result from uncertainty in the sorbent circulation rate, attrition rate, and unit cost. The system with sulfur recovery has slightly higher variable costs due to the additional methane that is required to reduce a portion of SO_2 to H_2S prior to the Claus reaction (see Appendix A.1.1.2).

For the two copper oxide systems with high sulfur Illinois No. 6 coal, there is a larger probability that the variable operating costs will be higher than for the conventional system. With sulfuric acid recovery, there is a 30 percent probability of higher variable costs, and with sulfur recovery the probability of higher variable cost is 50 percent. Because the sorbent circulation rate is higher for applications with high sulfur coal, costs associated with sorbent makeup have a more pronounced effect than with the medium

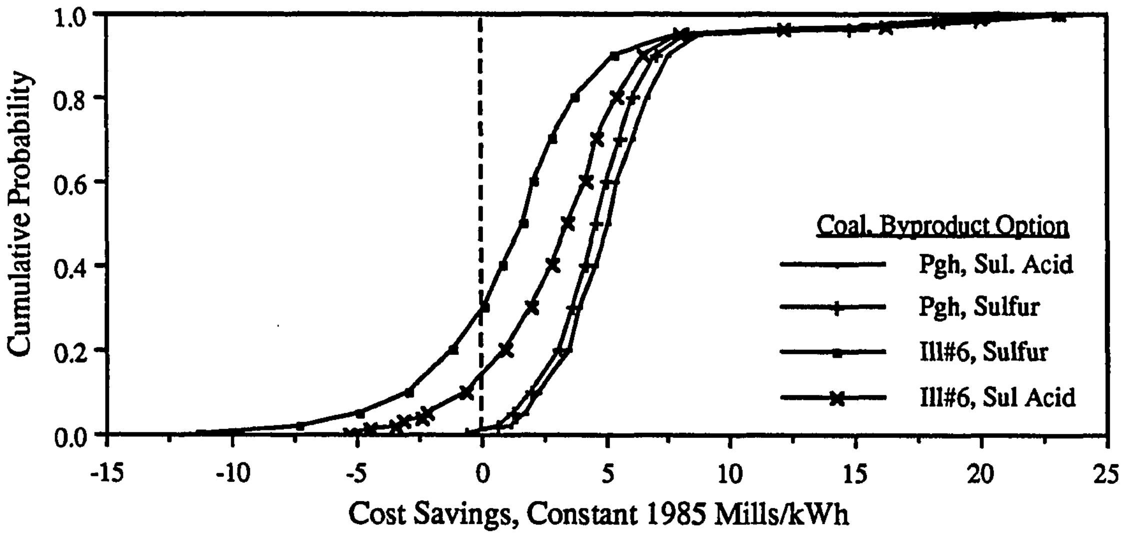

Figure 30. Comparison of Cost of Electricity Savings for Copper Oxide vs. FGD/SCR Systems: Effect of Coal and Byproduct Recovery Options.

sulfur coal. The additional sulfur burden for the case with sulfur recovery also significantly increases the plant methane requirements, which leads to higher variables costs for the copper oxide system.

Figure 30 shows the differences in levelized pollution control costs for the four different copper oxide process cases. When the median results (i.e., 50 percent probability) are considered, the copper oxide process is found to be less expensive than the FGD/SCR system for all cases. As was true for total capital and variable operating cost, the cost savings with copper oxide was higher for the medium sulfur Pittsburgh coal than for the high sulfur Illinois coal.

Figure 30 also shows up to a 30 percent chance that with Illinois No. 6 coal the copper oxide process might be *more* expensive than the FGD/SCR system. This result stems primarily from the possibility of high operating costs (e.g., for sorbent, etc.). In all cases, however, there is considerable uncertainty in the magnitude of potential cost savings using the advanced technology. For the Illinois No. 6 coal with sulfur recovery, the 90 percent probability range (defined by the 5th to 95th percentiles in Figure 9) is -5 to 8 mills/kWh. For the Pittsburgh No. 8 coal the range is 1 to 10 mills/kWh. This indicates that the copper oxide process is likely to be most attractive with medium rather than high sulfur coals. In all cases, there is a small probability that the cost savings could be significantly higher.

5.6 Quantifying Risk of New Technology

The risk that the new technology will be more expensive can be quantified using the partial mean of the cost difference distribution for all negative values. The downward and upward partial means are defined as (Buck and Askin, 1986):

$$\mu_d(x) \equiv \int_{-\infty}^{0} x\, f(x)\, dx \tag{19}$$

$$\mu_u(x) \equiv \int_{0}^{\infty} x\, f(x)\, dx \tag{20}$$

where f(x) is the probability density function for the random variable x. Partitioning of probability distributions is also discussed by Karlsson and Haimes (1988a; 1988b). In the case of the copper oxide system with sulfur recovery and Illinois No 6 coal (the highest risk case), the downward partial mean is -0.8 mills/kWh and the upward partial mean is 2.5 mills/kWh. These sum to the distribution mean of 1.7 mills/kWh. Buck and Askin define the conditional partial mean based on the partial mean and the probability that a loss or gain has occurred. The expected value of a loss, given that a loss has occurred, is:

$$\mu_{d|x<0}(x) \equiv \frac{\mu_d(x)}{P(x<0)} \tag{21}$$

where $P(x<0)$ is the probability that the random variable x has a value less than zero. The expected value of a gain, given that a gain has occurred, is defined similarly. For our example, the expected value of a loss is 2.8 mills/kWh if a loss occurs, and the expected value of a gain is 3.5 mills/kWh if a gain occurs.

The information provided by this analysis can be used to answer questions about the risks and potential pay-offs of the new technology compared to conventional technology. While the copper oxide process is unlikely to be commercialized for another 5 to 15 years, process research will ultimately be used by potential adopters to make a decision about what emission control system to use for a specific application. Therefore, it is reasonable to look at the decision a hypothetical adopter would make with currently available information vis-a-vis information expected to be yielded from research over the next several years.

The opportunity loss from a hypothetical decision to adopt the copper oxide process is given by the downward partial mean (Moore and Chen, 1984). The downward partial mean is the same as the expected value of perfect information (EVPI) for the case where the loss function, L(x), of a potential adopter is represented as linear for all negative outcomes and zero for all positive outcomes, i.e.:

$$E[L(x)] = \int_{-\infty}^{\infty} L(x)\, f_X(x)\, dx \tag{22}$$

where,

$$L(x) = \begin{Bmatrix} x, & x<0 \\ 0, & x>0 \end{Bmatrix} \tag{23}$$

The downward partial mean is the maximum amount that a decision-maker (with the given loss function) would be willing to pay to obtain perfect information that would be used to avoid the downward risk. Although research is unlikely to completely resolve uncertainties, research which leads to a reduction in the probability of a loss through process improvements, or which provides insight into situations in which FGD/SCR systems are less expensive, has value as "information" to a potential process adopter. The value of information is one measure by which to bound the expenditures on research, development, and demonstration.

5.7 Evaluating Additional Research

While additional research may reduce the downward risk of a new technology, it can also lead to incremental improvements in the new technology which would, in turn, increase the expected value of cost savings compared to conventional technology. The value of research may thus be estimated based on the incremental increase in the expected cost savings of the new technology compared to current information, rather than based on the reduction in downside risk.

Several factors must be considered in determining the value of research. First, judgment is required to estimate the likely results from a research effort. The value of research depends also on the circumstances of actual adoption of the new technology, which determines the ultimate cost savings compared to other technology. Judgment is required regarding the likely plant sizes, byproduct markets, coal characteristics, and other influencing factors that will face the new technology. It is unlikely that any single cost estimate can be used for such an analysis; rather, several case studies representative of different applications may be required. A third factor influencing the value of research is the possibility of simultaneous improvement in information about or design of competing

Table 17. Case Studies for Reduction in Copper Oxide Process Uncertainties Due to Research

Model Parameter	Probability Nominal Value	Distribution	Values (or σ as % of mean) Prior to Research	After Research
Regenerator				
Regeneration Efficiency	99.2 %	-1/2 Normal	(20 %)	(5 %)
Regeneration Temp.	900°F	Normal	(2 %)	(1 %)
Regen. Direct Capital Cost	(calc)	Uniform	1.0x - 1.5x	1.1x - 1.4x
Solids Transport				
Sorbent Attrition	0.06 %	Normal	(41 %)	(10 %)
Solids Trans. Dir. Cap. Cost	(calc)	Uniform	1.0x - 2.0x	1.1x - 1.8x
Absorber				
Standard Error, Cu/S Ratio	0	Normal	$\sigma = 0.39$	$\sigma = 0.2$

processes. Therefore, any prior estimate of the value of research is conditioned on the judgments regarding research results, technology diffusion, and improvements in competing processes.

In this analysis, the primary emphasis is on estimating the effect of possible research results on the comparative costs of the copper oxide process versus FGD/SCR systems. It is assumed that research can reduce the uncertainties about several key process variables, and thus provide "imperfect information"[1] about the technology. Because the key uncertainties in process cost have been identified to be related to regeneration, solids transport, and the stoichiometric copper-to-sulfur ratio, it is assumed that new research would be focused on these areas. Table 17 shows illustrative assumptions made about the possible reduction in uncertainties in several variables from new research. More study of the regenerator, solids transport system, and absorber could reasonably be expected to reduce uncertainties regarding regeneration efficiency, regeneration temperature, the equipment costs for the regenerator and solids transport system, sorbent attrition, and the copper-to-sulfur molar ratio.

Four case studies are used to illustrate that the value of research results is conditional on actual applications, although no attempt is made to actually forecast the

[1] As opposed to perfect information, which would remove all uncertainties and would allow a potential process adopter to avoid any loss in selecting between FGD and copper oxide systems.

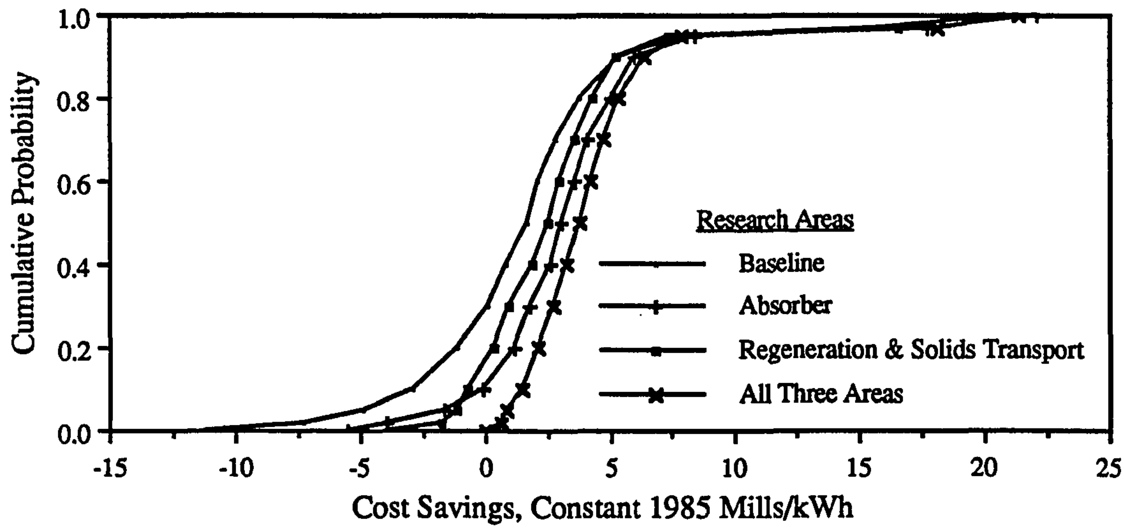

Figure 31. Effect of Illustrative Research Outcomes on Levelized Pollution Control Cost Savings for the Copper Oxide Process with Sulfur Recovery and Illinois No. 6 Coal.

diffusion of the copper oxide process into commercial use. For the sake of simplicity, it is assumed that the FGD/SCR pollution control system is relatively mature, and that, as an approximation, there will be no incremental improvements in FGD/SCR system costs.

Results for the case study involving the Illinois No. 6 coal and elemental sulfur recovery are shown graphically in Figure 31 for levelized total pollution control costs. The figure shows the cost differences for copper oxide versus FGD/SCR systems based on current information, and selected results for the difference based on information from further research. The assumptions about additional research reduced the variance of the cost difference distributions, but also reduced the skewness (due to assumptions about the regeneration efficiency). Thus, the assumed research outcomes have reduced the downside risk of the new technology and increased the expected cost savings.

The results from additional research for all four cases are summarized in Table 18. Note that while the cost difference between the copper oxide process and the FGD/SCR system were obtained as continuous probability distributions, the uncertainties in cost savings were represented in Table 18 as discrete outcomes (i.e. loss or gain) using the statistics discussed previously. These statistics include the probability that the copper oxide process was more expensive than an FGD/SCR system, the downward partial mean, the downward and upward conditional partial means, and the mean for the entire cost difference distribution. The hypothesized research results reduced the

Table 18. Results of Research Information Case Studies: Comparison of Levelized Total Pollution Control Costs for Copper Oxide versus Conventional FGD/SCR

Research Area	Probability of a Loss (%)	Downward Partial Mean (mills/kWh)	Expected Value of a Loss (mills/kWh)	Expected Value of a Gain (mills/kWh)	Mean (mills/kWh)	Reduction in Risk (mills/kWh)	Value of Research (mills/kWh)
Sulfuric Acid Recovery, Washed Illinois No. 6 Coal							
Baseline	15	-0.27	-1.7	4.6	3.6	- -	- -
Solids Transport	10	-0.13	-1.4	4.2	3.6	0.14	0.0
Absorber	9	-0.16	-1.8	4.8	4.2	0.11	0.6
Regeneration	7	-0.09	-1.3	4.8	4.4	0.19	0.8
Regen. and Solids Trans.	2	-0.02	-1.1	4.5	4.4	0.25	0.8
All	0	0	0	5.1	5.1	0.27	1.5
Sulfur Recovery, Washed Illinois No. 6 Coal							
Baseline	29	-0.81	-2.8	3.5	1.7	- -	- -
Solids Transport	17	-0.25	-1.5	3.7	2.8	0.56	1.1
Absorber	11	-0.23	-2.0	4.1	3.4	0.58	1.7
Regeneration	18	-0.28	-1.6	3.8	2.8	0.53	1.1
Regen. and Solids Trans.	17	-0.17	-1.0	3.7	2.9	0.64	1.2
All	0	0	0	4.3	4.3	0.81	2.6
Sulfuric Acid Recovery, Unwashed Pittsburgh Coal							
Baseline	1	> -0.01	-0.3	5.6	5.5	- -	- -
Solids Transport	1	> -0.01	-0.3	5.6	5.5	0	~0.0
Absorber	1	> -0.01	-0.5	5.9	5.9	0	0.4
Regeneration	0	0	0	6.0	6.0	< 0.01	0.5
Regen. and Solids Trans.	0	0	0	6.0	6.0	< 0.01	0.5
All	0	0	0	6.4	6.4	< 0.01	0.9
Sulfur Recovery, Unwashed Pittsburgh Coal							
Baseline	2	> -0.01	-0.4	5.2	5.1	- -	- -
Solids Transport	1	> -0.01	-0.4	5.2	5.1	0	~0.0
Absorber	2	> -0.01	-0.3	5.6	5.5	0	0.4
Regeneration	< 1	> -0.01	> -0.1	5.6	5.6	0	0.5
Regen. and Solids Trans.	0	0	0	5.6	5.6	< 0.01	0.5
All	0	0	0	6.0	6.0	< 0.01	0.9

downward partial mean of the cost differences for all cases, and therefore reduced the risk of an opportunity loss to a potential process adopter.

The mean cost difference with more research for all targeted process areas was higher than for current estimates. It can be seen in Table 18 that the value of research in terms of cost improvements was significantly greater than the reduction in downside risk. Thus, the value of research may be greater than the EVPI discussed previously because of improvement in expected cost savings, as well as reduction in downside risk.

To complete an estimate of the value of research requires some forecasting of technology diffusion. The four case studies indicate the variability of the value of research for different types of applications. Other factors discussed in previous sections, including plant size and capacity factor, will also influence the level of funding that can reasonably be committed to research.

5.8 Selecting Technologies and Research Strategies

The data summarized in Table 18 can be used to answer a number of questions such as:

- Is one technology preferred over another?
- Is additional research merited?
- What should be the research strategy?
- How much is additional research worth?
- Under what conditions does the decision strategy change? (How robust is the decision strategy?)

These questions can be answered using decision analysis as an analytical tool for evaluating alternative technology options and research strategies. The discretization of the continuous probability distributions for the cost savings of the copper oxide process compared to an FGD/SCR system, given in Table 18, facilitates the use of relatively simple decision trees to evaluate research strategies. An example of such a decision tree, based on the case with high sulfur coal and elemental sulfur recovery, is given in Figure 32. In this example, the decision analysis is based on a single attribute of cost savings compared to the conventional FGD/SCR system. Differences in emission rates are not considered here, because both systems are designed to achieve 90 percent SO_2 and NO_x reduction. First, we will consider decisions based on expected cost savings, and then briefly consider a more detailed decision model incorporating the risk attitudes of a decision maker and the time value of research outcomes.

The tree in Figure 32 includes three general decisions. The first is a choice between the copper oxide process based on current knowledge of the process and the FGD/SCR system. In this example, the copper oxide process without additional research is shown to have a positive expected cost savings compared to the conventional FGD/SCR system, based on current information. A second decision is regarding obtaining perfect information that would resolve all downside risks of the new process. The elimination of downside risk increased the expected value by 0.81 mills/kWh, and this is the measure of the EVPI. A third type of decision is that regarding further research and development of the process as discussed previously. As can be seen in Figure 32, and as summarized in Table 18, the expected values of the research options are larger than for the current state of knowledge, indicating that additional research is merited. The most fruitful research strategy in this case appeared to be for all three major process areas considered in the analysis. Such a strategy increased the expected value of the process, compared to current information, by 2.6 mills/kWh; this is the basis for bounding the amount of money that should be spent on further research. These differences are summarized in Table 18 as the "value of research."

The decision model can easily be refined to consider the risk attitude of a particular decision maker using expected utility, rather than expected cost savings, as the basis for decision making. A utility function, such as that discussed in Section 2.5, can be used to represent the personal value a decision maker places on specific outcomes. Furthermore, because the results of research may not be obtained for 5 to 15 years, the time value of the outcomes can be modeled using discounting. One possible utility function for such a decision model was presented in Section 2.5 and is reproduced here:

$$u(x) = \left\{ \frac{x(i,n) - x_l(i,n)}{x_h(i,n) - x_l(i,n)} \right\}^b \tag{24}$$

with variables defined as in Section 2.5. A nominal value of b=0.6 (risk averse) was used in the expected utility analysis.

The effect of discounting the outcomes of research is to reduce the expected utility of these outcomes. For the Illinois No. 6 coal and elemental sulfur recovery case, the research option for all areas is preferred by a risk averse decision maker if the pay-off from research is obtained within 10 years at a discount rate less than about 20 percent. If the pay-off from research is not available for another 20 years, the discount rate would have to

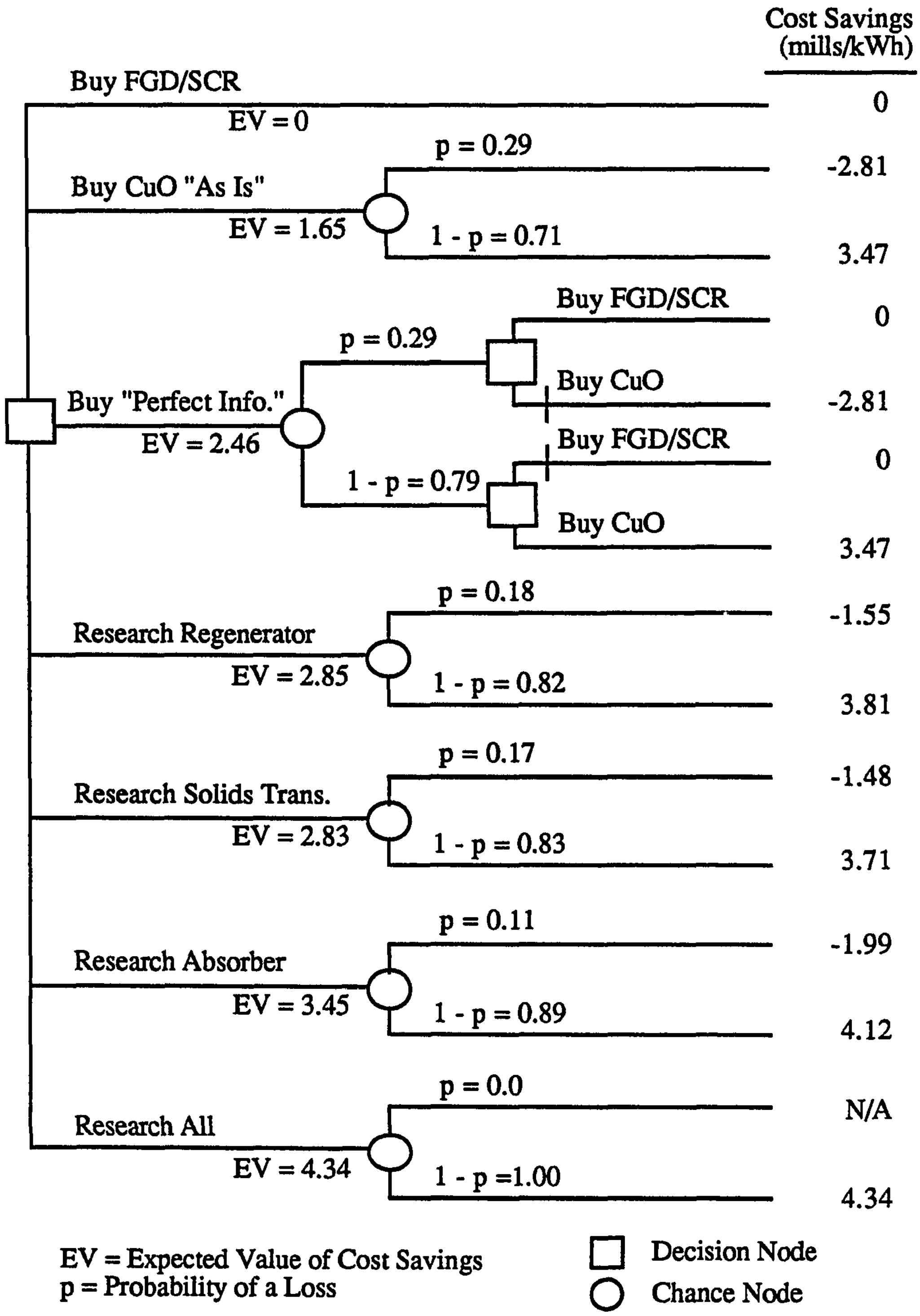

Figure 32. Decision Tree for Copper Oxide Process Research Planning: Example for Illinois No. 6 Coal and Elemental Sulfur Recovery.

be less than 10 percent for the strategy to have the highest expected utility. Research results for the copper oxide process could reasonably be expected in the next 5 to 15 years. Thus, the best research strategy for high sulfur coal applications is to wait for the results of further research. Only an extremely risk seeking decision maker would choose to accept the copper oxide process "as is". For the medium sulfur coal, the robust strategy, considering risk attitude, discount rate, and time until research pay-off, is to accept the copper oxide process "as is".

By delaying construction of a new plant until further research is available, a utility may need to purchase power from neighboring utilities or to take other measures to provide a sufficient power supply. In such site-specific cases, analysis of the benefits of further research must also consider the net costs associated with substitute power supply in the interim. These site-specific considerations are not included in this analysis.

For high sulfur coal/elemental sulfur recovery applications, the decision model can be used to bound research expenditures. Using expected value as the basis for the decision, the expected value of research in all areas is 2.7 mills/kWh higher than the expected value of the process as is. This is equivalent to a savings of about $7 million per year for 500 MW power plant at a capacity factor of 65 percent. For the decision analysis based on expected utility, the equivalent value of research is about $5 million per year. The actual amount to be spent on research depends on how many and what size power plants would be expected to use the copper oxide process with a high sulfur coal and elemental sulfur recovery.

The above example indicates the sensitivity of decisions not only to the outcomes from the engineering process models, but also to the assumptions made in the decision model, which reflect varying preferences of a decision maker.

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6.0 MODELING APPLICATIONS AND RESULTS: INTEGRATED GASIFICATION COMBINED CYCLE SYSTEMS

Three IGCC systems are evaluated using probabilistic engineering models. These models include the air-blown Lurgi, air-blown KRW, and oxygen-blown KRW systems. The most detailed information regarding process uncertainties was obtained for the Lurgi-based system. Therefore, the analysis of this system is presented first. Furthermore, the analysis of the Lurgi-based system is more comprehensive than for the other two systems, due to the availability of alternative expert judgments regarding uncertainties.

The capability and constraints of the modeling environment used for the IGCC case studies influenced the manner in which simulations were developed. These issues are discussed next. Then, the results of probabilistic analyses of the three IGCC systems are presented. Finally, probabilistic comparisons among the three systems are developed. The comparison between the air-blown Lurgi system and the base case oxygen-blown KRW system is explored further from a decision analysis perspective.

6.1 Running the Models

As described in Chapter 3, the IGCC system performance models are implemented in the ASPEN chemical process simulation modeling environment. The IGCC cost models are Fortran subroutines that are called from the performance models. All performance variables are calculated prior to calling the cost model subroutines. Therefore, it is possible to perform cost sensitivity analysis for a given set of performance results. This is done by iteratively calling the cost subroutine for varying assumptions regarding cost model parameters.

The IGCC models were run on a DEC VAXStation 3200 mini-computer using the public U.S. Department of Energy version of ASPEN. Running an ASPEN flowsheet involves several steps. The first is "input translation," in which a performance model, written in ASPEN's keyword-based input language, is read by the ASPEN package and converted to a Fortran program. This step takes approximately 5 to 10 CPU minutes. The ASPEN-generated Fortran program is then compiled and linked, which may take about 5 minutes. After linking, the flowsheet program is executed. The last step in an ASPEN simulation is report generation. The ASPEN simulator writes a report file containing the results of the simulation. Report writing may take several minutes, particularly if the user has requested detailed information regarding the simulation.

For a single run of an IGCC flowsheet, representing either a deterministic analysis or a single repetition during a stochastic analysis, the run time may take approximately 2 to 10 minutes, depending on the flowsheet, initial guesses for key variables, and limits specified in ASPEN design specifications (see MIT (1987) for a description of the structure of ASPEN models). Thus, a deterministic analysis may take approximately 20 to 30 minutes to run, including input translation, compiling, linking, execution, and report generation. In the case of a probabilistic simulation, the flowsheet is executed many times, with a different set of values (samples) assigned to uncertain input parameters each time. Thus, a probabilistic analysis with a sample size of 100 may take 6 to 12 hours to run, depending on the flowsheet. During a probabilistic analysis, the run time for a particular sample varies, depending on the results of the previous simulation (which are used as initial guesses) and the sample values assigned to uncertain variables for a particular sample. It is not unusual for run times to vary from, 2 to 5 or 10 minutes for any given sample in a probabilistic simulation.

6.1.1 Comparing Probabilistic Results with Different Cost Uncertainty Assumptions

There are many cases in which it is instructive to make comparisons between two alternatives when both are uncertain. For example, we may wish to compare the effect that including or excluding a set of input uncertainty assumptions would have on, say, uncertainty in plant efficiency or cost of electricity. Or, we may wish to compare the capital cost of two different technologies. By carefully planning and specifying the input uncertainties in a simulation, it is possible to generate properly paired samples for the uncertainties that are to be directly compared. In particular, when there are input uncertainties that are common to the two alternatives being compared, both alternatives should be analyzed using the same set and ranking of samples for those input uncertainties. Probabilistic analyses that can be made with the ASPEN simulator include comparisons of alternative assumptions regarding cost and/or performance. The first case is discussed here.

A key insight that can be obtained from probabilistic analysis is the effect that interactions among uncertainties in performance and cost parameters can have on total system costs (e.g., total capital cost, levelized cost of electricity). Therefore, it is useful to compare the uncertainty in total cost that is obtained when performance uncertainties only are specified to the case when uncertainties in both performance and cost parameters are considered. To make such a comparison, two results for total capital cost uncertainty are needed. Furthermore, these results should be based on the same set (and ranking) of samples for the uncertain performance variables which are common to both cases. This allows the results for both cases to be directly compared, sample by sample.

Because the cost model subroutines can be called iteratively for a given set of performance variable values, it is possible to perform several probabilistic case studies as part of a single simulation. For example, interactions between performance and cost uncertainties can be examined by generating cost results based on performance uncertainties only, and comparing them to results based on both performance and cost uncertainties. Such a comparisons is done by:

- Calculating flowsheet performance results based on the sample values of the uncertain performance parameters;
- For each repetition, calling the cost model subroutine based on the deterministic "best guess" assumptions for cost model parameters and the performance results obtained based on sampled values of uncertain performance parameters;
- For each repetition, reinitializing the uncertain cost model parameters to their probabilistic sample values, for the same set of performance results, and calling the cost model subroutines a second time.

Thus, several probabilistic "sensitivity" cases involving alternative assumptions for cost-related parameters can be performed as part of a single performance simulation. For example, the uncertainty in the cost of electricity resulting from interactions of uncertainties in performance, capital cost, and operating and maintenance (O&M) cost parameters can be evaluated by calling the cost model subroutine several times, with appropriate assumptions regarding cost model parameters. Furthermore, this approach has the advantage that the same set of samples are used for the performance uncertainty assumptions in all cases. The ability to perform cost-related sensitivity analysis as part of a single performance simulation eliminates the need to run the performance flowsheet several times, at a cost of 6 to 12 hours per run, to obtain the desired case studies.

Alternatively, uncertainties in cost may be evaluated apart from uncertainties in performance. This is done by running the cost model subroutines in a stand-alone mode, with a deterministic set of assumptions for the input performance variables required by the model. The cost model can then be run probabilistically by using the stochastic block in ASPEN to assign sample values to cost model parameters. The ASPEN flowsheet used for such a simulation consists only of the stochastic flowsheet section and a Fortran block to call the cost model. The run time for a cost uncertainty-only simulation is approximately 5 CPU seconds per sample, or about 20 minutes including all steps from input translation to report writing. The cost uncertainty-only simulation can be run using the same set of sample values for cost uncertainties as for the full performance and cost uncertainty simulation. The method for doing this is described in the next section.

6.1.2 Comparing Probabilistic Results from Separate Simulations

Although comparisons of the effect of alternative assumptions regarding input uncertainties in cost can be performed in the context of a single probabilistic simulation of performance, comparison of alternative assumptions regarding performance uncertainties requires separate simulations. Furthermore, as indicated above, analysis of uncertainties in total cost resulting from uncertainties only in cost model parameters requires a separate simulation. However, it is essential that such comparisons take into account the underlying correlation between the cases. For example, if we wish to rerun a performance uncertainty analysis and change the variance in only one input uncertainty, then for all other input uncertainties we should use the same set of sample values and the same ranking of values. In the case of the parameter whose variance is reduced, we should use the same ranking of samples, although the specific values of each sample will differ compared to the base case.

Such comparisons can be made using the newly added probabilistic capability of the ASPEN simulator. As long as the same random seed, the same number of uncertain variables, and the same correlation structure between the uncertain variables are used, then the same ranking of sample values for each uncertain variable will result. This is because probabilistic sampling is based on use of inverse cumulative distribution functions, as described in Section 2.4. Rather than directly sample from a probability distribution, probabilistic simulation techniques generate uniformly distributed random numbers from zero to one for each uncertain variable. These random numbers represent the fractile that is to be sampled. Then, using the inverse cdf, the actual sample value associated with a given fractile is calculated. For a given set of uncertain variables, correlations between variables, and a random seed, the program developed by Iman and Shortencarier (1984), which is the underlying basis of the ASPEN probabilistic modeling capability, will always generate the same set and ranking of fractiles for each uncertain distribution. These fractiles are converted to actual sample values based on the type of probability distribution and the distribution parameters (e.g, variance) specified by the user.

Thus, it is possible to use reproducible samples for uncertain variables when comparing two or more performance uncertainty simulations. When comparing performance uncertainty simulations, the approach is:

- Always specify the same number of random variables to be generated for the simulations to be compared, even if only a subset are actually used in any given simulation. An unused random variable is simply not assigned to any input parameters in the ASPEN IGCC model.

- Always use the same underlying correlation structure. (In this work, comparisons are made assuming independent random variables, although a few sensitivity cases involving correlation structures are considered).
- Always use the same random seed. There is a default random seed which is the same for all cases.

The result of properly specifying uncertainties for comparative analysis is that the samples for any given run can be directly paired between cases. Therefore, it is possible, for example, to estimate the probability distribution for the *difference* in efficiency or cost taking into account uncertainties which are common to both cases. A probability distribution for a difference is estimated by pairing the samples from the two simulations, and then subtracting the paired samples accordingly.

6.2 Probabilistic Analysis of the Air-Blown IGCC System with Hot Gas Cleanup

For the air-blown Lurgi-based IGCC system with hot gas cleanup, judgments were obtained from several experts regarding the performance of both the gasifier and the hot gas cleanup system. The judgments of Expert LG-1 regarding uncertainties in the Lurgi gasifier and of Expert ZF-1 regarding uncertainties in the zinc ferrite desulfurization process are assumed as a base case. The implications of alternative expert judgments are also explored. These analyses developed here include: (1) characterization of uncertainties in key measures of plant performance and cost; (2) identification of the key model uncertainties that are the most important determinants of uncertainty in model outputs; (3) comparison of design trade-offs under uncertainty; (4) evaluation of the reductions in uncertainty that may be obtained from further research; (5) evaluation of the alternative judgments of different experts as they affect model results; and (6) evaluation of the importance of correlation structures on results.

6.2.1 Characterization of Uncertainties in Performance and Cost

The engineering performance and cost models of the Lurgi-based IGCC system were run using the set of judgments regarding uncertainties in process performance and cost shown in Table 16 (see Chapter 4). In addition, a deterministic simulation of the Lurgi-based system was run. The deterministic simulation is based on "best guess" values for the parameters which are treated as uncertain in the probabilistic simulation, as described in Chapter 4. The deterministic simulation is intended to be representative of the estimates for plant performance and cost that would be obtained in lieu of probabilistic

Table 19. Summary of Results from Deterministic and Probabilistic Simulations of a 650 MW Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup: Expert Judgments LG-1 and ZF-1.[a]

Parameter[b]	Units[c]	"Best Guess"[d]	$f_{.50}$	μ	σ	$f_{.05}$ - $f_{.95}$
Plant Performance						
Thermal Efficiency	%, HHV	38.5	37.7	37.5	1.3	35.3 - 39.3
Coal Consumption	lb/kWh	0.789	0.806	0.811	0.029	0.773 - 0.861
Process Water Consump.	lb/kWh	1.604	1.602	1.635	0.261	1.215 - 2.129
ZF Sorbent Charge	10^6 lb	6.54	6.51	7.29	2.95	4.15 - 12.38
Sulfuric Acid Production	lb/kWh	0.085	0.087	0.087	0.003	0.082 - 0.093
Plant Discharges						
SO_2 Emissions	lb/MMBtu	0.042	0.040	0.040	0.001	0.038 - 0.042
NO_x Emissions	lb/MMBtu	2.74	2.19	2.19	0.402	1.53 - 2.84
CO Emissions	lb/kWh	0.003	0.003	0.003	0.003	0.003 - 0.006
CO_2 Emissions	lb/kWh	1.72	1.73	1.73	0.031	1.68 - 1.78
Solid Waste	lb/kWh	0.083	0.096	0.098	0.015	0.079 - 0.125
Plant Costs						
Total Capital Cost	$/kW	1,409	1,463	1,465	127	1,281 - 1,696
Fixed Operating Cost	$/kW-yr	44.8	57.2	59.6	10.6	46.4 - 82.6
Variable Operating	mills/kWh	18.2	19.0	21.9	8.6	16.9 - 36.1
Coal		16.2	16.6	16.7	0.6	15.9 - 17.7
Byproduct		(1.5)	(1.4)	(1.3)	0.5	(0.4) - (2.0)
Other		3.5	3.7	6.5	8.5	2.2 - 18.7
Cost of Electricity	mills/kWh	51.7	56.7	59.0	9.8	49.9 - 73.5

[a] The notation in the table heading is defined as follows: f_n = n^{th} fractile ($f_{.50}$ = median), μ = mean; and σ = standard deviation of the probability distribution. The range enclosed by $f_{.05}$ to $f_{.95}$ is the 90 percent probability range. All costs are January 1989 dollars.

[b] Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, and zinc ferrite steam. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

[c] HHV = higher heating value; MMBtu = million Btu.

[d] Based on a deterministic simulation in which median or modal values of uncertain variables are assumed as "best guess" inputs to the model

analysis. However, a deterministic analysis is not required when doing a probabilistic analysis; it is developed here merely for comparative purposes.

From the probabilistic simulation, frequency distributions for variables calculated in the performance and cost models can be estimated The results of a simulation can be summarized using statistics, such as the mean or standard deviation, or using graphs of the cumulative distribution function (cdf). The results of both a deterministic and probabilistic simulation of a nominal 650 MW Lurgi-based IGCC power plant are summarized in Table 19. The table summarizes selected results for plant performance, environmental discharges, and costs. The "best guess" value is that obtained from deterministic analysis.

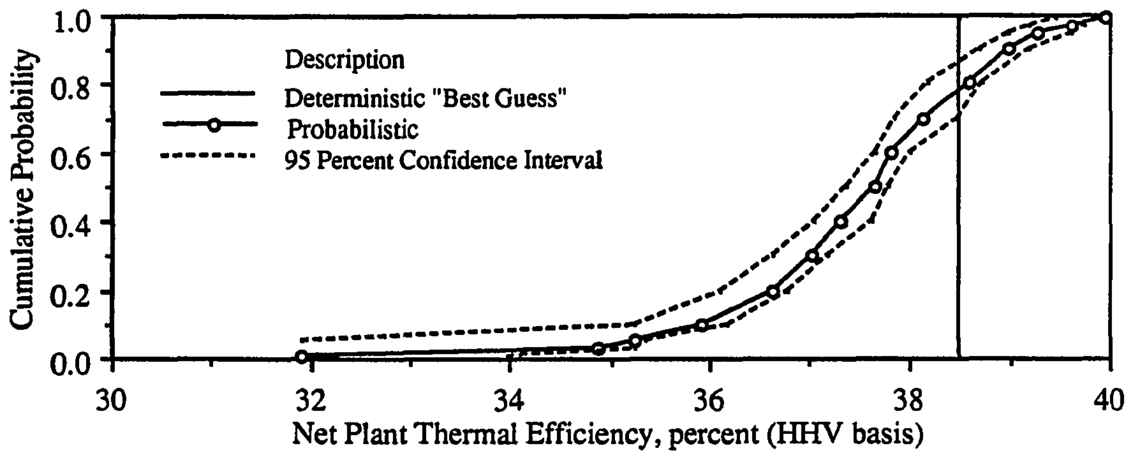

Figure 33. Comparison of Deterministic and Probabilistic Results for the Net Plant Thermal Efficiency of the Lurgi-based System.

The deterministic estimates for capital cost include so-called "process contingency" and "project contingency" factors, based on typical values used in the literature (see Appendices A.6.2 and B.7.2). Several results from Table-19 are graphed as cdfs.

6.2.1.1 Plant Performance

The uncertainty in plant thermal efficiency is shown in Figure 33. The deterministic "best guess" result is shown as a vertical line in the graph. In addition to the cdf obtained from the probabilistic simulation of the ASPEN performance model of the Lurgi-based system, dotted lines are shown to indicate the 95 percent confidence interval for the cdf. The confidence interval is estimated using the technique described in Chapter 2. As the sample size is increased, the confidence interval more tightly approaches the cdf. However, the range of values enclosed by the confidence interval is usually higher at the very low or very high fractiles, particularly in cases where a distribution has a "long" tail. For example, note in Figure 33 that at the 5th percentile, the range of efficiencies enclosed by the 90 percent confidence interval is from 31.9 to 35.3 percent, whereas at the 50th percentile the range is from 37.4 to 37.8 percent. The confidence interval is an indication of the "accuracy" of the probabilistic simulation in estimating the cdf based on limited sample size. It is not, however, a measure of the "accuracy" of the judgments or data analysis that went into developing the model input uncertainties from which the cdf was estimated.

For the analyses of IGCC systems, a sample size of 100 was chosen as a compromise between generating smooth cdfs which could be reasonably reproduced even when using different random seeds, and a need for simulation run times that would permit two or three case studies in a 24 hour period.

The deterministic estimate of plant thermal efficiency for this case is 38.5 percent. However, from the probabilistic simulation, the median (50th percentile) value of efficiency is 37.7 percent, and the mean (average) value is even lower at 37.5 percent (see Table 19). From Figure 33, it is apparent that the probability distribution for efficiency is negatively skewed, with a long tail below the 10th percentile. Thus, there is a 10 percent probability that efficiency could be less than 35.9 percent, and it may go as low as 32 percent. There is only about a 20 percent chance that efficiency would be higher than the deterministic estimate, and it could go as high as 40 percent.

The negative skewness of the uncertainty in plant thermal efficiency results from the assumptions regarding input uncertainties. For example, Expert LG-1, who provided the judgments regarding uncertainties in the gasifier process area used in this example, indicated that the most likely value for coal carbon retention in the bottom ash of the gasifier was 2.5 percent of the carbon in the coal feed. This value was used in the deterministic estimate. However, while the expert indicated that the carbon retention could be as low as 0.75 percent, he also indicated it could be as high as 10 percent. Carbon retained in the bottom ash represents a significant efficiency penalty on the IGCC system, because it is not combusted in the gasifier nor converted to fuel gas. Thus, the positively skewed assumption regarding uncertainty in carbon retention is likely to be a contributing factor to the negatively skewed uncertainty regarding plant thermal efficiency. The identification of key uncertainties in input assumptions is discussed further in the next section.

A brief way to summarize the results of the probabilistic simulation with a few simple numbers would be to use the mean value and the 90 percent probability range to characterize the central tendency and variation in efficiency. For example, from Table 19, the plant efficiency has a mean value of 37.5 percent with a 90 percent probability of being between 35.3 and 39.3 percent. Alternatively, the result might be expressed as an efficiency of 37.5 (+1.8/-2.2) percent . However, such a simple characterization of uncertainty does not provide any indication of the long tail at the lower end of the distribution. Thus, the cdf is the preferred method for communicating results about uncertainty in key variables used in decision making, because it more completely represents the entire range of possible outcomes.

For the Lurgi-based IGCC system with hot gas cleanup, a key performance variable which affects plant costs is the amount of zinc ferrite sorbent that must be charged to all

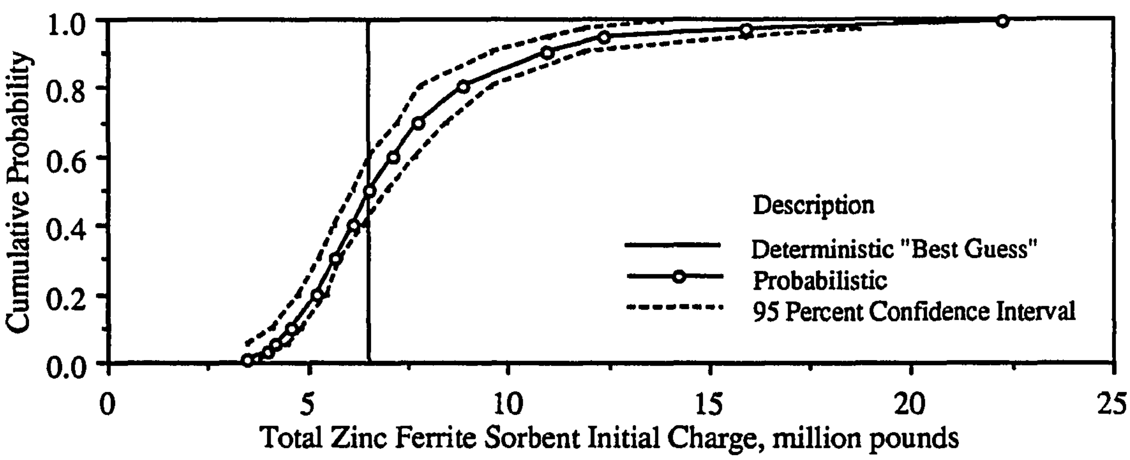

Figure 34. Comparison of Deterministic and Probabilistic Results for the Zinc Ferrite Initial Sorbent Charge of the Lurgi-based System.

fuel gas desulfurization reactor vessels prior to plant startup. The amount of sorbent charge is a key determinant of the number and size of the reactor vessels, and the sorbent can represent a significant portion of the capital cost for initial chemicals and catalysts. The uncertainty in the sorbent charge is shown as a cdf in Figure 34. In this case, the deterministic "best guess" coincides with the median value of the probabilistic result.

The judgment of Expert ZF-1 was that uncertainty in sorbent sulfur loading is normally distributed, with a mean (and median) at 17 weight percent. This value was used in the deterministic analysis. The sorbent charge requirement is a nonlinear function of sorbent sulfur loading (see Appendix A.4.4). Therefore, the resulting uncertainty in sorbent charge is positively skewed. Thus, while the median sorbent charge is almost the same as the deterministic "best guess" value at 6.5 million pounds, the mean value is higher, at 7.3 million pounds. Furthermore, there is a 5 percent chance that the sorbent charge would be more than 12.4 million pounds, and in the worst case the sorbent charge could be over a factor of three greater than the deterministic estimate. Here again, as with the uncertainty in plant efficiency, use of just deterministic or mean values in a performance estimate would mask the risk a process adopter faces that sorbent charge could be substantially higher.

6.2.1.2 Plant Emissions

Uncertainties in plant performance affect plant emissions. Probabilistic results for selected plant environmental discharges are shown as cdfs in Figures 35, 36, and 37 for SO_2, NO_x, and CO_2 emissions, respectively. In the case of SO_2 emissions, the probabilistic simulation indicates that the best guess value obtained from deterministic analysis may in fact be overly pessimistic. There is about an 80 percent probability that

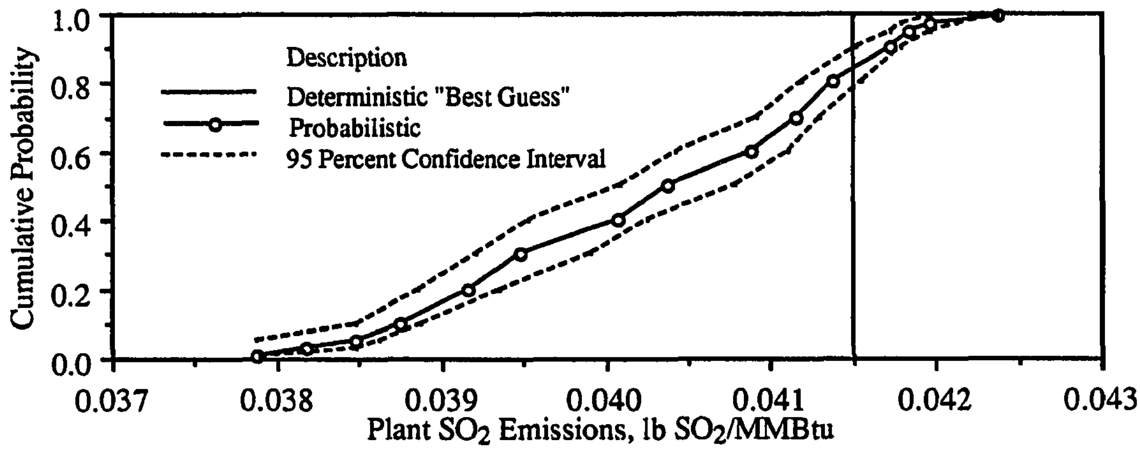

Figure 35. Comparison of Deterministic and Probabilistic Results for the SO_2 Emissions of the Lurgi-based System.

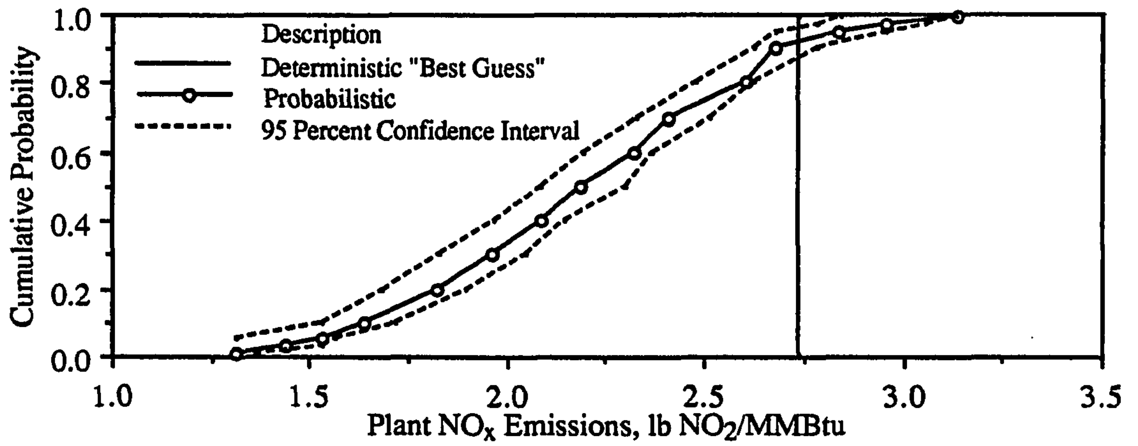

Figure 36. Comparison of Deterministic and Probabilistic Results for the NO_x Emissions of the Lurgi-based System.

SO_2 emissions would be lower than the deterministic estimate. Similarly, there is over a 90 percent probability that NO_x emissions would be less than the deterministic estimate. These results are obtained due to the skewness in several of the model input uncertainties. For example, the NO_x emission rate depends on the ammonia yield in the gasifier and the conversion of ammonia to NO_x in the gas turbine combustor. Expert LG-1's judgment regarding the ammonia yield from the gasifier was negatively skewed. The judgment of the author regarding the fraction of ammonia converted to NO_x in the gas turbine combustor was also negatively skewed. Therefore, the uncertainty in the NO_x emission rate is also negatively skewed.

The CO_2 emission rate is normalized based on plant efficiency, which is penalized for carbon retained in the bottom ash. However, retained carbon retained is not emitted to the atmosphere. Therefore, the CO_2 emission rate does not share the strongly negatively

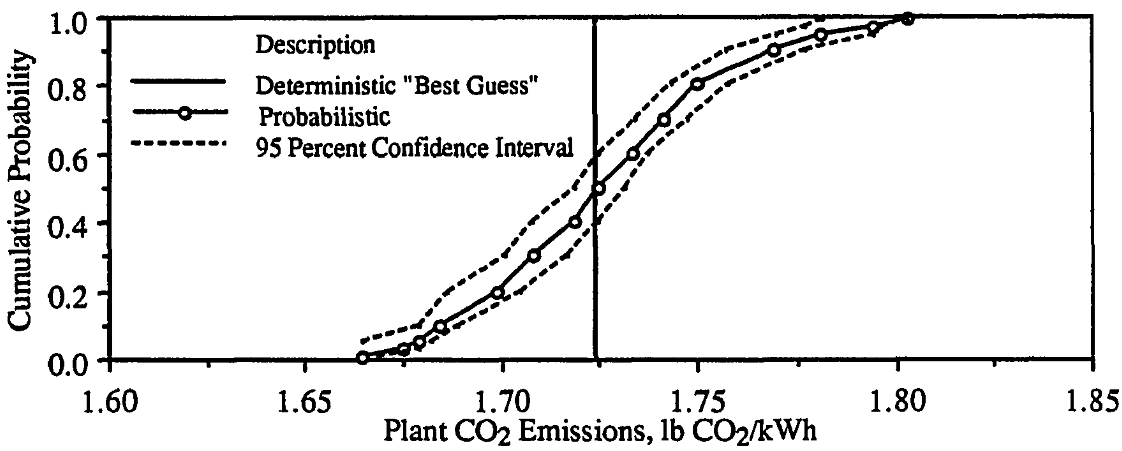

Figure 37. Comparison of Deterministic and Probabilistic Results for the CO_2 Emissions of the Lurgi-based System.

skewed shape of the plant efficiency curve, as would be expected if all carbon were emitted as CO_2.

6.2.1.3 Plant Costs

Uncertainties in plant performance parameters, in interaction with uncertainties in process cost parameters, lead to uncertainties in the key measures of cost often used for process evaluation.

The uncertainty in the plant total capital cost, expressed on a normalized basis of dollars per net kilowatt of plant capacity, is shown in Figure 38. The deterministic estimate of capital cost is also shown. The uncertainty in capital cost covers a wide range, from about $1,200/kW to over $1,800/kW. The mean ($1,465/kW) and median ($1,465/kW) are higher than the deterministic estimate of $1,409/kW. Compared to the deterministic estimate, there is almost a 70 percent probability that the capital cost would be higher. As indicated previously, the deterministic capital cost estimate includes so-called "contingency" allowances, which are intended to account for both performance and project related uncertainties. In this case, the contingency factors appear to be inadequate (too low), and use of the deterministic cost estimate would expose a decision-maker to a substantial chance of cost overrun.

For simulations of limited sample size, it is important to consider whether increasing the sample size could lead to qualitatively different model results. The confidence interval on the cdf can be used to place a confidence interval on the probability of cost overrun. In this case, the 95 percent confidence interval is a 60 to 75 percent probability of cost overrun. Thus, even if the sample size of the probabilistic simulation

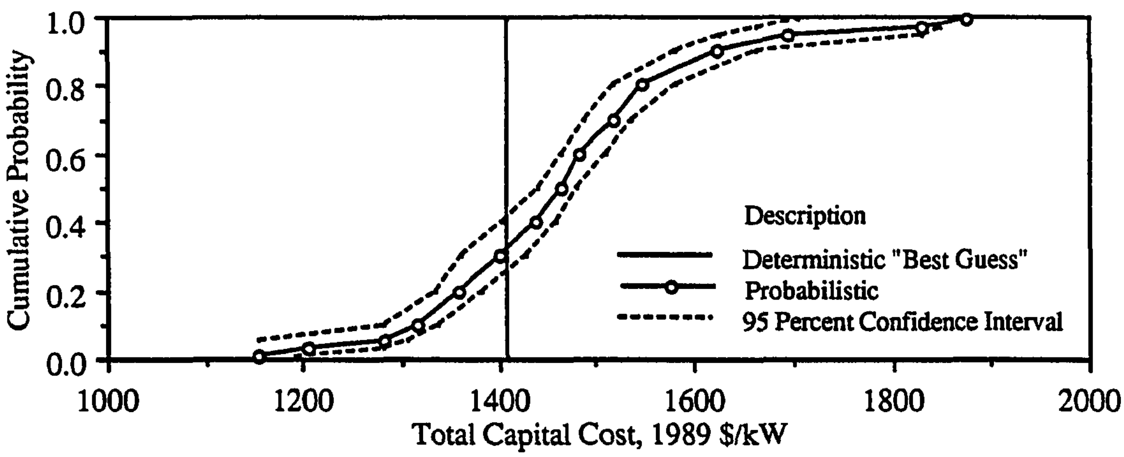

Figure 38. Comparison of Deterministic and Probabilistic Results for the Total Capital Cost of the Lurgi-based System.

were to be increased, we would still expect to find a greater probability of cost overrun than cost overrun, compared to the deterministic estimate.

An alternative representation of uncertainty in capital cost is shown in Figure 39. In this graph, the probability of cost overrun is shown on the x-axis, and the total capital cost is shown on the y-axis. A decision-maker could select an acceptable probability of cost overrun according to his/her own preferences, and then choose a corresponding budgetary value of capital cost. This approach bypasses the need to make a separate deterministic analysis with contingency factors. For example, if a decision maker would accept only a 10 percent probability of cost overrun, the budget estimate should be $1,624/kW (as shown in Figure 39). A capital cost estimate can be uniquely specified, therefore, based on

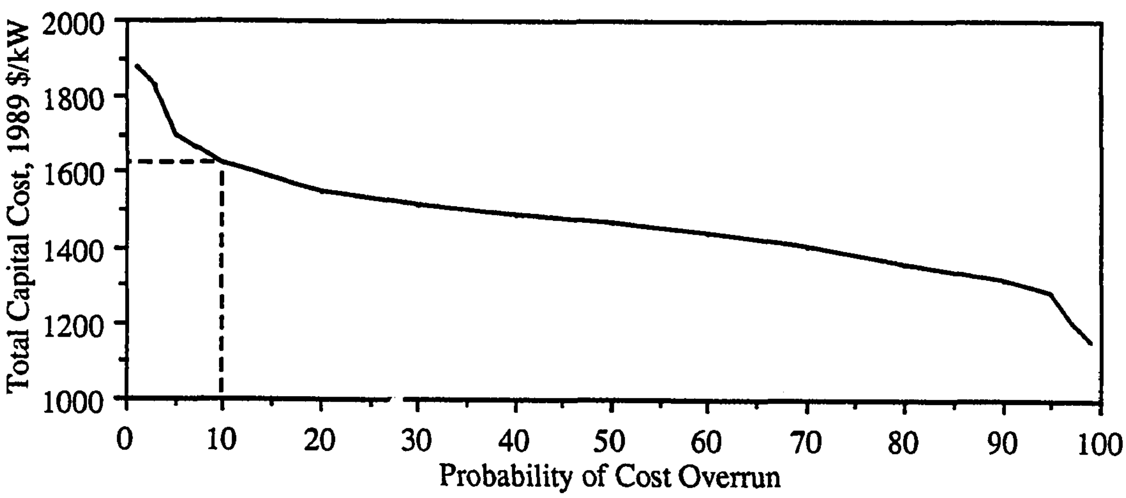

Figure 39. Comparing Total Capital Cost and the Probability of Cost-Overrun for the Lurgi-based System.

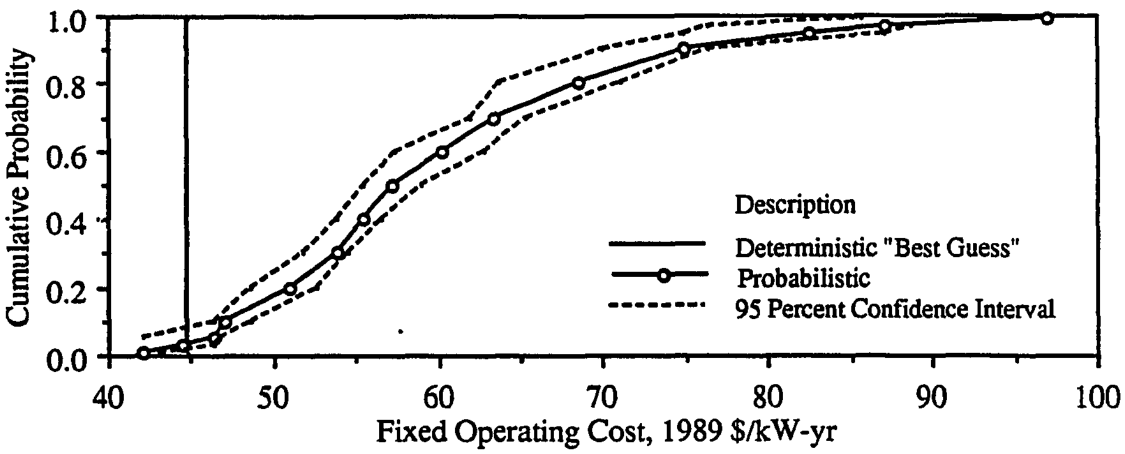

Figure 40. Comparison of Deterministic and Probabilistic Results for the Fixed Operating Cost of the Lurgi-based System.

the probability of cost overrun associated with it. However, with a scant few exceptions, conceptual cost estimates provide no indication of the risk of cost overrun associated with the assumed contingency factors.

The notion of uncertainty is extended here to operating and maintenance (O&M) cost estimates. Early in the development of a process technology, maintenance costs may be poorly anticipated. The deterministic and uncertainty estimates of fixed operating costs are shown in Figure 40. Fixed operating costs include maintenance materials, maintenance labor, operating labor, and administrative and supervisory labor. These costs are incurred regardless of the operating schedule of the power plant, because maintenance and operating staff are required to be on-site on a regular schedule. For this reason, the cost is reported on a normalized basis of annual dollars per kilowatt of plant capacity. Of the cost-related variables, the fixed operating cost manifests the largest discrepancy between the deterministic and probabilistic cost estimates. There is about a 95 percent probability that the fixed operating cost would be higher than the "best guess" estimate.

Uncertainty in fixed operating costs stems from uncertainties in both the capital cost of each process area and the annual maintenance costs expressed as a percentage of process area capital cost. While the process area capital cost uncertainties are symmetrically distributed around the deterministic contingency factor values (see Appendix B.7.2), several of the maintenance cost factors are positively skewed. For example, the maintenance cost factor for the Lurgi gasification process area is based on an elicited expert judgment. The most likely annual maintenance cost is 3 percent of the process area capital cost. The lowest possible maintenance cost is estimated to be 2 percent, while the highest is estimated to be 12 percent. Similarly, the best guess maintenance cost factor for the zinc

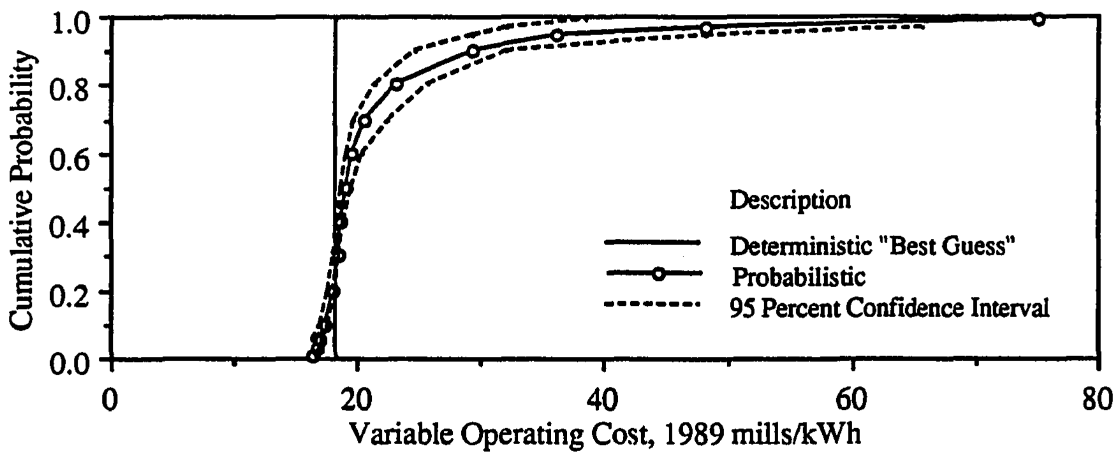

Figure 41. Comparison of Deterministic and Probabilistic Results for the Variable Operating Cost of the Lurgi-based System.

ferrite process area is taken to be a lower bound, with a chance that the cost could be higher. See Appendix B for more detail regarding judgments about uncertainty in maintenance cost factors.

The best guess and uncertainty estimates for variable operating cost are shown in Figure 41. A striking feature of this graph is the extremely long tail of the probability distribution. The variable operating cost includes the cost of consumable materials (e.g., chemicals, catalyst, coal), disposal costs for ash, and a byproduct credit for the sale of sulfuric acid. The cost is expressed on the basis of kilowatt-hours of plant output, because these costs are incurred only if the plant is operating. The deterministic best guess estimate from Table 19 is 18.2 mills/kWh (a mill is one-thousandth of a dollar). There is about a 25 percent probability that the variable operating cost could be lower than this estimate. However, there is a 10 percent probability that the cost could be greater than 30 mills/kWh, and it could go as high as 75 mills/kWh.

The extreme skewness of this distribution is the result of interactions among uncertainties in performance and cost parameters. One example is the uncertainty in makeup zinc ferrite sorbent cost. The annual requirement for makeup sorbent depends on both the size of the sorbent charge and the percentage of sorbent that is lost due to attrition per absorption and regeneration cycle. As already discussed, the uncertainty in sorbent charge shown in Figure 34 is the result of uncertainty in the zinc ferrite sorbent sulfur loading capacity, and it is positively skewed. The judgment regarding uncertainty in sorbent attrition obtained from Expert ZF-1 (see Appendix B.5.3.1) is also positively skewed. The interaction of these two uncertain variables contributes to the characterization of uncertainty in variable operating cost. The extremely high values for variable operating

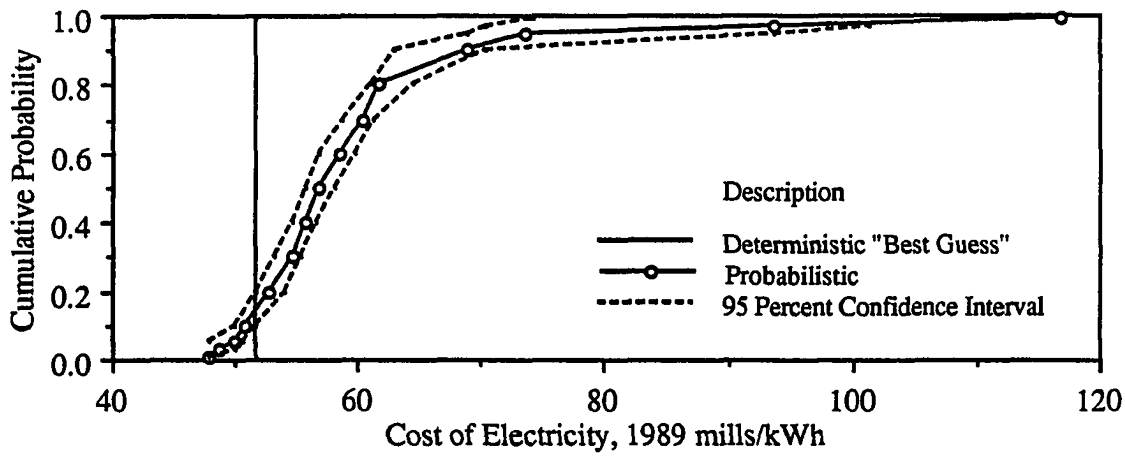

Figure 42. Comparison of Deterministic and Probabilistic Results for the Cost of Electricity of the Lurgi-based System.

cost are associated with the possibility of low sorbent sulfur loading capacity and high sorbent attrition rates.

The levelized cost of electricity is the single most comprehensive measure of plant cost, because it is based on (and sensitive to) all of the factors which affect capital, fixed operating, and variable operating costs. Because it is expressed on a net electricity production basis, it is also sensitive to the plant thermal efficiency. For the deterministic cost estimate, the contribution to the cost of electricity from levelized capital cost is 50 percent, from fixed operating cost is 15 percent, and from variable operating cost is 35 percent. For the probabilistic cost estimate, the relative contribution of the three varies. For example, for the sample where variable operating cost was 75 mills/kWh, the contribution of variable operating cost to the cost of electricity was 64 percent.

Like the uncertainties in fixed and variable operating cost, the uncertainty in the cost of electricity is positively skewed, as shown in Figure 42. In addition, the central values of the probability distribution are higher than the "best guess" estimate. The median value is 5 mills/kWh higher than the deterministic estimate. There is only a 15 percent probability that the cost of electricity could be less than the deterministic estimate. There is a 20 percent probability that the cost could be higher than 62 mills/kWh, and it could go over 100 mills/kWh. The range of uncertainty in the cost of electricity varies by a factor of 2.5 from the lowest to the highest values.

This section has focused on characterization of uncertainties in key measures of plant performance, emissions, and cost. In some cases, the uncertainty in the Lurgi-based system is shown to be quite large, particularly for the variable operating cost and the cost of electricity. Because of interactions among the input uncertainty assumptions, many of which are positively skewed, the central values of the probabilistic results, such as the median and the mean, tend to be higher than the deterministic estimate. A research planner is interested in knowing what factors contribute most to the uncertainties described here. Thus, identifying and prioritizing key input uncertainties, in order to better understand the model output uncertainties, is the next step.

6.2.2 Identifying Key Uncertainties

Several approaches to identifying key uncertainties are possible in probabilistic analysis. One approach involves statistical analysis using regression techniques. Regression techniques can be used to help identify input variables which are most highly correlated with output variables. Another approach is probabilistic sensitivity analysis, in which alternative assumptions about uncertainties are compared. From this approach, it is often possible to gain insights into the interaction between different subsets of uncertain input variables as they affect uncertainty in an output variable. A third approach, which is similar to the second, is to confirm the results of a regression or sensitivity analysis by deleting uncertainties from the model which are not believed to be important. This is an uncertainty screening study. The results of a screening study can be compared to the results obtained from the original probabilistic analysis. If the results are similar, then the deleted uncertainties need not be considered probabilistically in further studies. The development of improved judgments regarding uncertainties can then focus on the key uncertainties remaining after the screening study.

6.2.2.1 Regression Analysis

As part of the probabilistic modeling capability in the ASPEN simulator, four alternative approaches to regression analysis are available for analyzing model results. These were discussed earlier in Section 2.4. The output analysis capability utilizes a program developed by Iman et al (1985). For the Lurgi system, all four techniques will be compared. These are: (1) partial correlation coefficients (PCC); (2) standardized regression coefficients; (3) partial rank correlation coefficients (PRCC); and (4) standardized rank regression coefficients (SRCC).

When running a stochastic simulation in ASPEN, the user may specify which type of output analysis is desired. The results are reported in the form of a table. For each output, a series of coefficients is reported representing either the partial correlation or standardized regression coefficient between the output variable and each of the uncertain input variables. In addition, the ranks of the magnitudes of the coefficients are also given. Thus, the user can use the output as one basis for ranking the relative importance of input

Table 20. Comparison of Rankings of Uncertain Parameters Affecting Plant Efficiency for the Lurgi-based System.

	Type of Output Analysis[a,b]			
Rank[c]	PCC	SRC	PRCC	SRRC
1	Carbon to ash	Carbon to ash	Carbon to ash	Carbon to ash
2	Air/Coal ratio	Air/Coal ratio	Air/Coal ratio	Air/Coal ratio
3	Fines capture	Fines capture	Fines capture	Fines capture
4	Fines carryover	Fines carryover	Fines carryover	Fines carryover
5	ZF resid. sulfate	ZF resid. sulfate	(see note)[d]	(see note)[d]
6	NH_3 yield	NH_3 yield		

[a]Abbreviations for type of analysis: PCC = partial correlation coefficients; SRC = standardized regression coefficients; PRCC = partial rank correlation coefficients; SRRC = standardized rank regression coefficients.
[b]Abbreviations for uncertain parameters: CH = coal handling; DC = direct cost; HRSG = heat recovery steam generator; ICC = indirect construction cost; SE = standard error; STG = steam turbine-generator; Unc. = uncertainty; ZF = zinc ferrite.
[c]At about the seventh most important parameter, the correlation or regression coefficients become sufficiently small to no longer be statistically significant. Therefore, rankings below six are not shown.
[d]The partial rank correlation coefficients become sufficiently small to no longer be statistically significant at the fifth-ranked parameter. Therefore, rankings below four are not shown. The same result is assumed applicable for the standardized rank regression coefficients.

uncertainties. However, such results must be interpreted with care. The regression analysis is based on a linear model. There may be significant non-linearities that might bias the results from a linear regression analysis. Thus, regression analysis on the sample ranks, rather than the sample values, of the input and output variables may be preferred.

A comparison of the rankings obtained from the four approaches of key uncertainties affecting plant thermal efficiency is shown in Table 20. Only those uncertainties for which the coefficients were found to be statistically significant (see Section 2.4) are included in the list. Insignificant correlations are often easily recognized. For example, the seventh ranked input uncertainty according to PCC analysis, which was statistically insignificant and is not shown in the table, was zinc ferrite unit cost. Clearly, the uncertainty in the unit cost of zinc ferrite has no relationship to uncertainty in plant efficiency.

According to Table 20, all four regression techniques yield the same ranking for the top four uncertainties affecting plant efficiency. Thus, regardless of whether partial correlation or standardized rank regression coefficients are used, and regardless of whether sample values or sample ranks are used, the same relative result is obtained. However, the

Table 21. Comparison of Rankings of Uncertain Parameters Affecting Capital Cost for Lurgi-based System.

	Type of Output Analysis[a,b]			
Rank[c]	PCC	SRC	PRCC	SRRC
1	Coal throughput	Coal throughput	Coal throughput	Coal throughput
2	Project Unc.	Project Unc.	Project Unc.	Project Unc.
3	Gas Turbine DC	Gas Turbine DC	Gas Turbine DC	Gas Turbine DC
4	ZF Loading	ZF Loading	ZF Loading	ZF Loading
5	Gasifier DC	Gasifier DC	ICC	ICC
6	ICC	ICC	SE HRSG	SE HRSG
7	Carbon to Ash	Carbon to Ash	Gasifier DC	Gasifier DC
8	SE STG	SE STG	Carbon to Ash	Carbon to Ash
9	SE HRSG	ZF Attrition	ZF DC	ZF DC
10	SE CH	SE HRSG	SE CH	SE CH

[a]Abbreviations for type of analysis: PCC = partial correlation coefficients; SRC = standardized regression coefficients; PRCC = partial rank correlation coefficients; SRRC = standardized rank regression coefficients.
[b]Abbreviations for uncertain parameters: CH = coal handling; DC = direct cost; HRSG = heat recovery steam generator; ICC = indirect construction cost; SE = standard error; STG = steam turbine-generator; Unc. = uncertainty; ZF = zinc ferrite.
[c]At about the tenth most important parameter, the correlation or regression coefficients become sufficiently small to no longer be statistically significant. Therefore, rankings below 10 are not shown.

fifth and sixth ranked uncertainties obtained from sample regression are not significant in the rank regressions. There are two contributing reasons for this. One is that the sample correlations are relatively weak for these two input uncertainties (-0.45 and -0.33, respectively). The other is that these uncertainties are skewed. Thus, in the sample regression, there a few extreme values near the tails that become "compressed" in the rank regression and, therefore, are less influential in the regression model.

The key uncertainties affecting uncertainty in plant thermal efficiency are shown to be associated with the gasification process area. There is a weak relationship between uncertainty in the zinc ferrite sorbent residual sulfate content after oxidative regeneration and plant thermal efficiency. Thus, the gasification process is the primary source of uncertainty in efficiency, for the assumptions used in this case study.

The results of the alternative regression analyses for plant total capital cost are shown in Table 21. As in the case of plant efficiency, the four approaches agree with respect to the ranking of the first four input uncertainties. Furthermore, the two analyses based on rank regression agree on the relative ordering of all ten input uncertainties shown in the table. The two analyses based on sample regression agree on the relative ordering of the first eight input uncertainties shown. However, they disagree on the remaining ones, with the SRC analysis indicating a stronger influence from zinc ferrite sorbent attrition than PCC analysis. In the capital cost model, sorbent attrition affects costs because it influences the quantity of zinc ferrite sorbent that must be stored in reserve and how much must be used for plant startup. The difference in result indicates that the partial derivative of capital cost with respect to sorbent attrition is higher than with respect to the standard error of the HRSG direct capital cost model. However, the PCC analysis indicates that removing the HRSG standard error from a linear regression model would have a larger effect on the coefficient of determination than removing the sorbent attrition rate.

The sample and rank regression analyses disagree after the fourth ranked uncertainty. For example, the gasifier direct cost uncertainty drops from fifth to seventh in the rankings, while the indirect construction cost uncertainty jumps from sixth to fifth. The gasifier direct cost is uniformly distributed, whereas the indirect capital cost (ICC) factor is triangularly distributed. However, the ranks for any input variable are uniformly distributed. Thus, the underlying basis for the probability distributions used in rank regression may differ significantly from that in sample regression. Variables which are relatively "peaky" (having values concentrated near the mode) will tend to become more influential in rank regression, where they are treated as uniformly distributed. The effect here for the ICC is to more heavily weigh the high and low outcomes in rank regression than is the case for sample regression. Similarly, the standard error for the HRSG is normally distributed. Rank regression treats this variable as uniformly distributed, leading to more influence in the rank regression than in the sample regression.

The results in Table 21 illustrate the importance of considering uncertainties in both performance and cost parameters when estimating uncertainty in capital cost. Uncertainties in performance parameters of both the gasification and zinc ferrite process area are shown to be important determinants of uncertainty in capital cost. These performance parameters, such as gasifier coal throughput and zinc ferrite sorbent sulfur loading, affect the sizing and number of vessels for the respective process areas. Uncertainties in cost parameters interact with the uncertainties in performance. The sources of cost-related uncertainties include direct process area costs, project-related capital costs, indirect construction costs, and the standard error of several of the process area direct cost regression models. The latter indicates that uncertainty in the cost estimate could be reduced, to some degree, by developing better process area cost models. The explicit characterization of standard errors

Table 22. Comparison of Rankings of Uncertain Parameters Affecting Cost of Electricity for the Lurgi-based System.

Rank[c]	Type of Output Analysis[a,b] PCC	SRC	PRCC	SRRC
1	ZF Attrition	ZF Attrition	ZF Attrition	ZF Attrition
2	ZF Loading	ZF Loading	Coal throughput	Coal throughput
3	Coal throughput	Coal throughput	Project Unc.	Project Unc.
4	Gas Turbine DC	Gas Turbine DC	Carbon to Ash	Carbon to Ash
5	Gasifier MC	Gasifier MC	Gasifier MC	Gasifier MC
6	Project Unc.	Project Unc.	ZF Loading	ZF Loading
7	ZF Unit Cost	ZF Unit Cost	ZF Unit Cost	ZF Unit Cost
8	Gasifier DC	Gasifier DC	Gas Turbine DC	Gas Turbine DC

[a]Abbreviations for type of analysis: PCC = partial correlation coefficients; SRC = standardized regression coefficients; PRCC = partial rank correlation coefficients; SRRC = standardized rank regression coefficients.
[b]Abbreviations for uncertain parameters: CH = coal handling; DC = direct cost; HRSG = heat recovery steam generator; ICC = indirect construction cost; SE = standard error; STG = steam turbine-generator; Unc. = uncertainty; ZF = zinc ferrite.
[c]At about the ninth most important parameter, the correlation or regression coefficients become sufficiently small to no longer be statistically significant. Therefore, rankings below eight are not shown.

thus allows the analyst to quantitatively identify specific models for which further development effort is warranted.

A third example of regression analysis is shown in Table 22 for the influence of input uncertainties on uncertainty in the levelized cost of electricity. In this example, both sample regression approaches produce the same rankings. Also, both rank regression approaches produce the same rankings. However, the rankings obtained from sample and rank regression do not agree, and the differences are quite strong. The four approaches agree only with respect to the top-ranked uncertainty. However, the second most important uncertainty according to sample regression, zinc ferrite sorbent sulfur loading, is only the sixth most important uncertainty obtained from rank regression.

The difference between the two results is attributable to the nonlinearity of the zinc ferrite performance model. As shown in Figure 34, the uncertainty in the zinc ferrite sorbent charge is positively skewed, with the extremely high values almost a factor of three greater than the median. The uncertainty in sorbent charge is driven primarily by uncertainty in sorbent sulfur loading, and it affects both capital and operating costs through the initial and annual makeup sorbent requirements. However, in rank regression, the use of rank values eliminates the influence that the extreme sample values, associated with low sorbent loading, have in the linear regression model. Thus, the relative importance of sorbent loading is much less according to rank than sample regression.

For other switches in the ordering of uncertainties, the effects are similar to those observed for total capital cost. "Peaky" distributions, such as coal throughput and project uncertainty, tend to rise through the rankings when comparing rank to sample regression results. Uniform distributions, such as gas turbine direct cost, tend to fall through the rankings, because they are represented as uniform distributions in both approaches, while peaky distributions in sample regression are represented as uniform in rank regression. Thus, their relative influence in the linear regression model tends to decrease.

The uncertainty in levelized cost is shown to be influenced by both performance and cost uncertainties, and by uncertainties in all three major process areas. The key performance-related uncertainties include zinc ferrite attrition rate and sorbent sulfur loading, that result in uncertainty in the requirements for chemicals and other consumables. The key cost-related uncertain parameters include direct capital costs (e.g., gas turbine direct cost), maintenance costs (e.g, gasifier maintenance cost), and variable operating costs (e.g., sorbent unit cost).

In spite of the differences in rankings that can be obtained using different approaches, it is often possible nonetheless to obtain a robust list of key uncertainties. For example, in the cases where all four analysis approaches agree on rankings, the results can be considered to be robust. Groups of variables may be identified as important in common between the techniques, but their rankings may differ slightly from one to the other. In these cases, the entire grouping of variables may be assumed to important, even if it is not possible to find agreement among the approaches regarding the ordering within the group.

6.2.2.2 Probabilistic Sensitivity Analysis

Another approach to identifying key uncertainties is probabilistic sensitivity analysis. Insight into the sensitivity of output variable uncertainties to the assumptions regarding uncertainties in input variables can be obtained by comparing the effect that different assumptions have on the result. One type of useful insight is the relative importance of uncertainties in performance parameters versus cost parameters. Another is the relative contribution to uncertainty from different process areas. Through probabilistic sensitivity analysis, it is possible to characterize the effect that specific uncertainties or groups of uncertainties have on specific output variables. In cases were uncertainties are

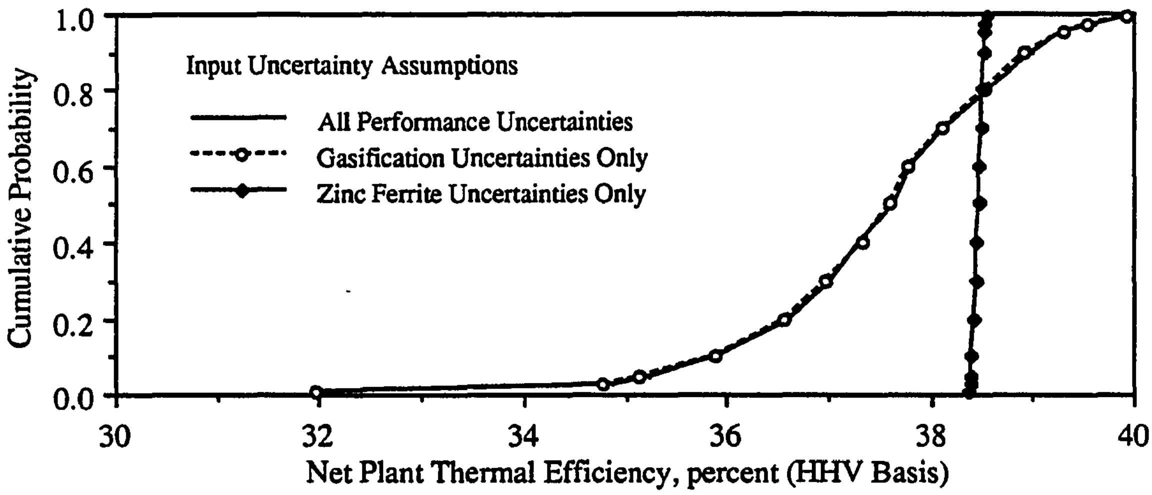

Figure 43. Sources of Uncertainty for Plant Thermal Efficiency.

excluded from a case study, the probability distribution for the parameter is replaced by its deterministic value.

An example of probabilistic sensitivity analysis is shown in Figure 43. The uncertainty in plant thermal efficiency resulting from all of the performance uncertainties from Table 16 is compared to that resulting from uncertainties in the gasification and zinc ferrite process areas, considered separately. The figure clearly indicates that uncertainty in the gasification process area is almost completely responsible for the uncertainty in plant thermal efficiency. The range of uncertainty in efficiency resulting from the zinc ferrite process area is very small.

Figure 44 compares the uncertainty in total capital cost attributable to either performance or cost parameter uncertainties alone with the result from including all uncertain parameters in the simulation. Uncertainty in cost alone results in a 90 percent probability range of approximately a $250/kW, while uncertainty in performance alone results in a range of approximately $325/kW. However, in addition to the effect on variance, the assumptions regarding performance uncertainties also shift the central value of the distribution upward by about $100/kW, compared to the results from cost uncertainties alone. The reasons for the shift can be seen in Figures 33 and 34. The performance-related uncertainties tend to be skewed, resulting in the negatively skewed distribution for plant efficiency and the positively skewed distribution for zinc ferrite sorbent charge. Both of these results tend to increase capital costs, compared to the deterministic values and compared to the case with cost-related uncertainties only. In contrast, the cost-related uncertainties are more symmetric. Thus, the uncertainty due to cost parameters leads to an

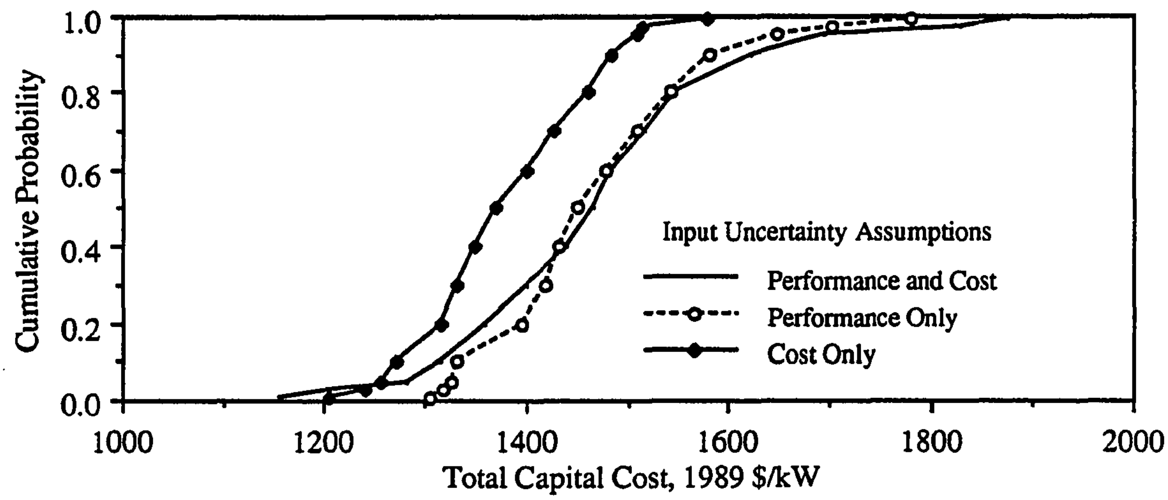

Figure 44. Sources of Uncertainty for Total Capital Cost.

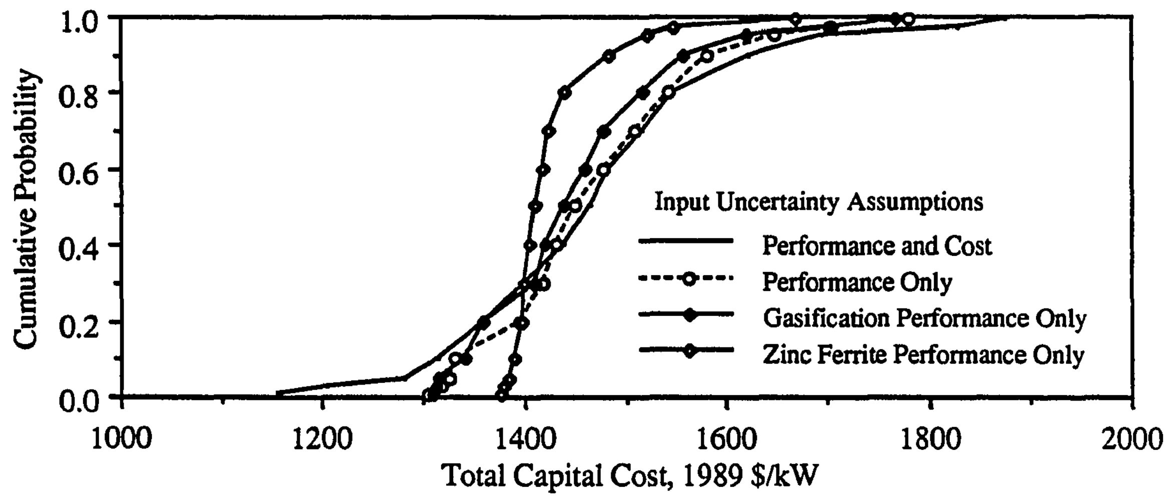

Figure 45. Performance-Related Sources of Uncertainty for Total Capital Cost.

increase in the variance of total capital cost when considered simultaneously with uncertainty in performance.

The effect of assumptions regarding performance uncertainties are examined in more detail in Figure 45. In the case where performance uncertainty in the gasification process area only is considered, the resulting uncertainty in capital cost is similar to that obtained when all performance-related uncertainties are considered. Performance uncertainties in the zinc ferrite process alone result in a positively skewed uncertainty in capital cost. The interactions between uncertainties in the zinc ferrite and gasification process areas result in the positively skewed distribution for capital cost resulting from all performance related uncertainties.

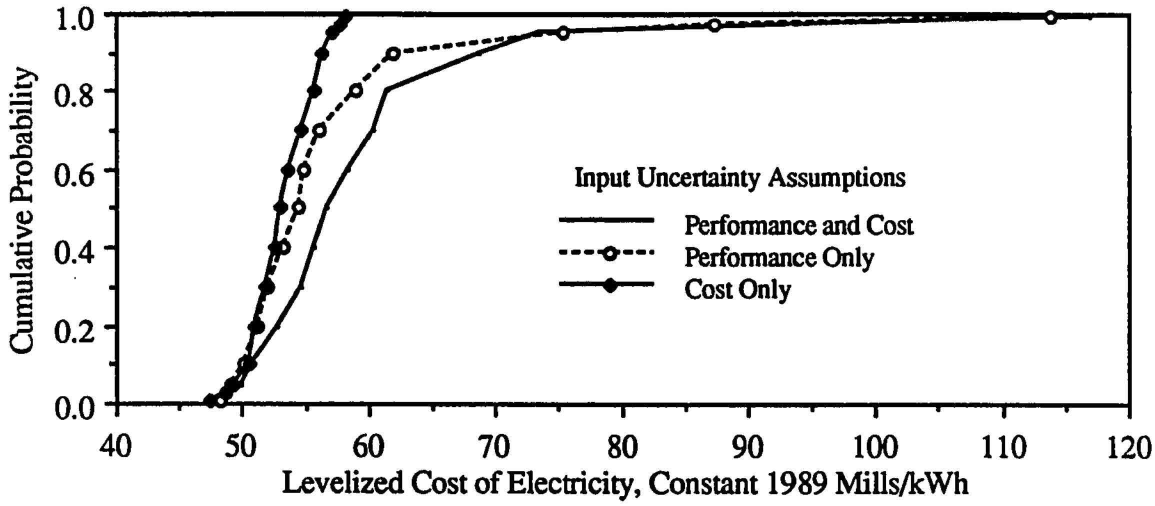

Figure 46. Sources of Uncertainty for Cost of Electricity.

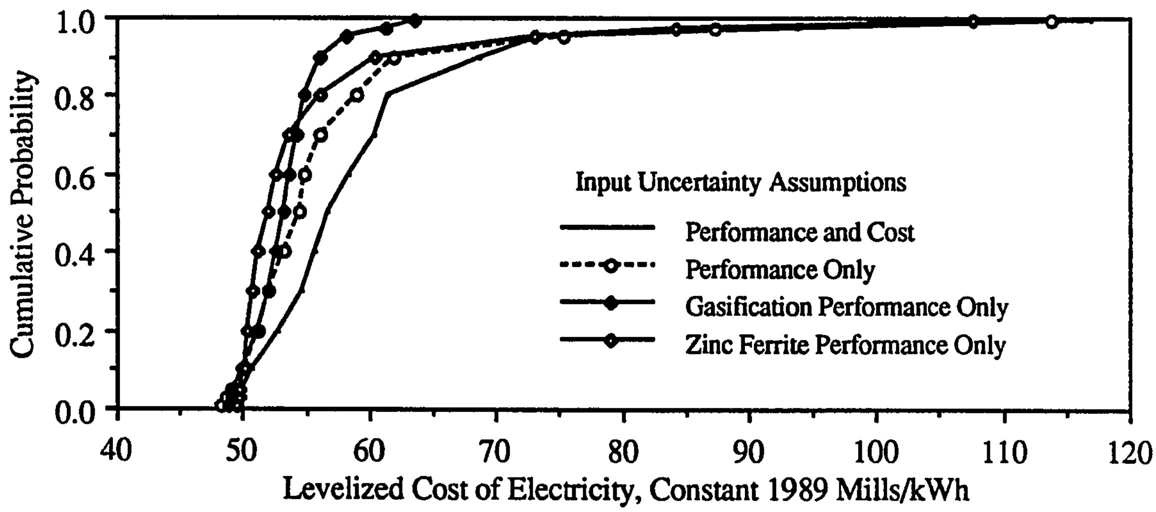

Figure 47. Performance-Related Sources of Uncertainty for Cost of Electricity.

A comparison of uncertainty estimates for the cost of electricity is shown in Figure 46. The graph shows the effect of specifying uncertainties in performance and cost either separately or combined. Because the cost of electricity is sensitive to most of the 47 uncertainties assumed in Table 16, the interactions among uncertainties are complex. In this case, neither performance nor cost uncertainties taken alone are shown to adequately describe the overall uncertainty in the cost of electricity. However, performance-related uncertainties are responsible for the positive skewness of the distribution.

The interactions of uncertainties in performance parameters are explored further in Figure 47. The graph clearly illustrates that uncertainties in the zinc ferrite process area are primarily responsible for the positive skewness of the uncertainty in the cost of electricity. The uncertainty resulting from gasification performance parameters is also positively

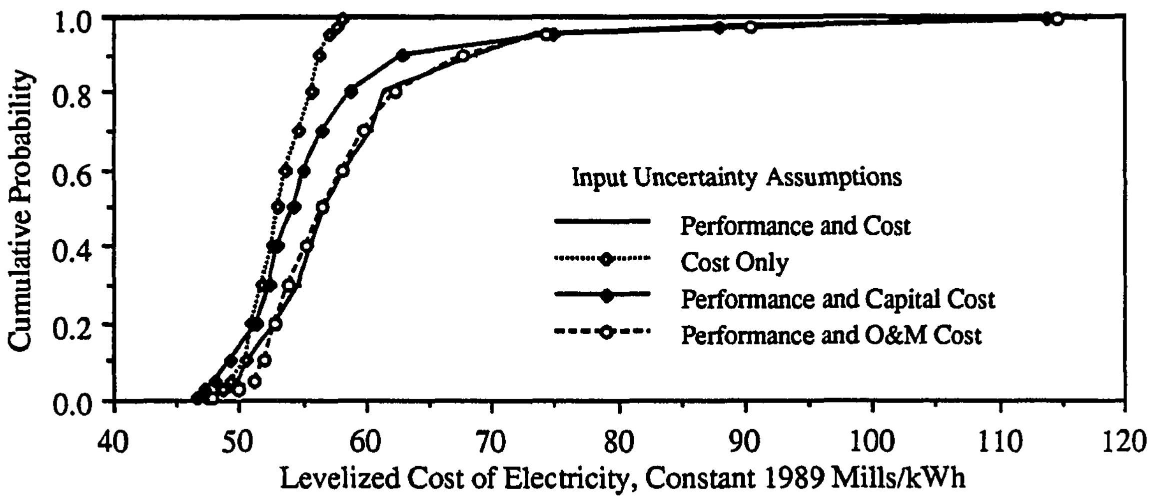

Figure 48. Cost-Related Sources of Uncertainty for Cost of Electricity.

skewed. However, the 90 percent probability range resulting from gasification uncertainties alone is only approximately 8 mills/kWh, whereas the corresponding range resulting from zinc ferrite uncertainties alone is approximately 23 mills/kWh. The extremely high cost outcomes are driven by uncertainty in the performance of the zinc ferrite sorbent.

The interactions between uncertainties in performance and cost are detailed in Figure 48. Uncertainties in cost parameters alone results in a modest range of values for the cost of electricity. If performance and capital cost uncertainties are considered simultaneously, the probability distribution begins to shift toward higher costs, in addition to becoming positively skewed. However, it is clear that uncertainties in operating and maintenance costs, simultaneous with uncertainties in performance, have a more pronounced effect than uncertainties in capital cost on the cost of electricity. This indicates that uncertainties in operating costs are key determinants of uncertainty in levelized annual costs.

6.2.2.3 Screening Analysis

A final approach considered here for identifying key uncertainties is a screening analysis. In this approach, uncertainties which were found to be insignificant in the regression analyses are removed from the model. The model is then run again, and the results compared to the simulation in which all uncertainties were included. The uncertainties which were found to be statistically insignificant, or which were found to have very weak relationships with uncertainties in key output variables, are listed in Table 23. The key output variables considered were efficiency, total capital cost, cost of electricity, and SO_2, NO_x, and CO_2 emissions. A total of 19 uncertainties were screened

Table 23. Uncertainties Screened Out of Case Studies for Air-Blown Lurgi-based IGCC System with Expert Judgments LG-1 and ZF-1.

Fines Carbon Content
Residual Sulfide in Zinc Ferrite After Reductive Regeneration
Zinc Ferrite Absorber Pressure Drop
Thermal NO_x
Coal Handling Direct Capital Cost
Cyclone Direct Capital Cost
Oxidant Feed Direct Capital Cost
HRSG Direct Capital Cost
Steam Turbine Direct Capital Cost
General Facilities Direct Capital Cost
Standard Error of Oxidant Feed Direct Capital Cost Model
Standard Error of Sulfuric Acid Direct Capital Cost Model
Standard Error of Boiler Feedwater Direct Capital Cost Model
Standard Error of Auxiliary Power Model for Coal Handling
Maintenance Cost for Oxidant Feed
Maintenance Cost for Cyclones
Maintenance Cost for Zinc Ferrite
Byproduct Marketing Cost
Operating Labor Rate

from further case studies, leaving 28 of the original 47 uncertain parameters. The uncertain parameters removed from the screening case study were assigned their respective deterministic "best guess" values.

The results of the screening analysis are shown graphically for the plant efficiency, capital cost, and cost of electricity in Figures 49, 50, and 51, respectively. For all three output variables, the results obtained from the screened uncertainties are virtually indistinguishable from the results obtained when all 47 uncertainties are included in the simulation. This verifies that the 19 uncertainties listed in Table 23 are unimportant. These uncertainties can be eliminated from further data collection efforts.

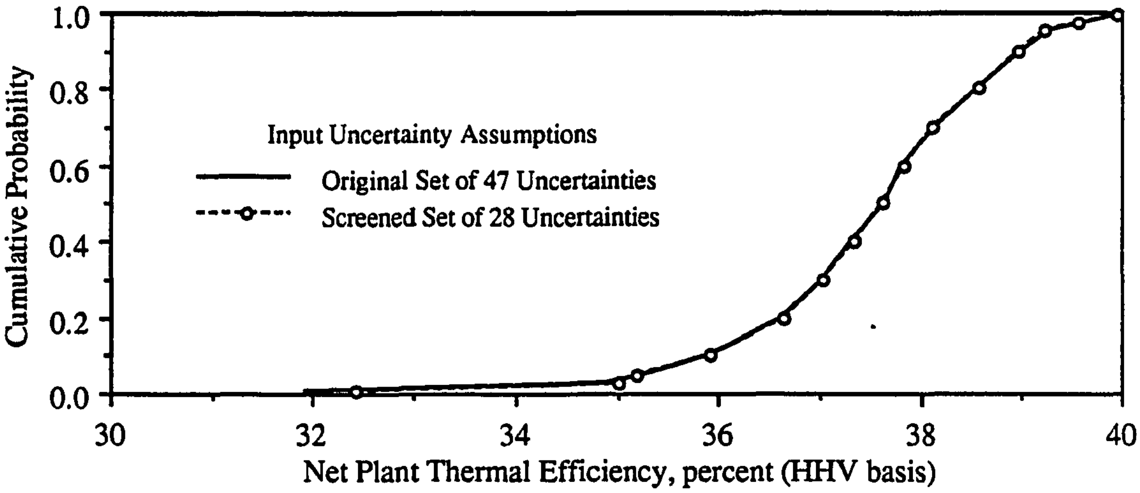

Figure 49. Comparison of Uncertainty in Plant Efficiency for Original and Screened Sets of Uncertainties.

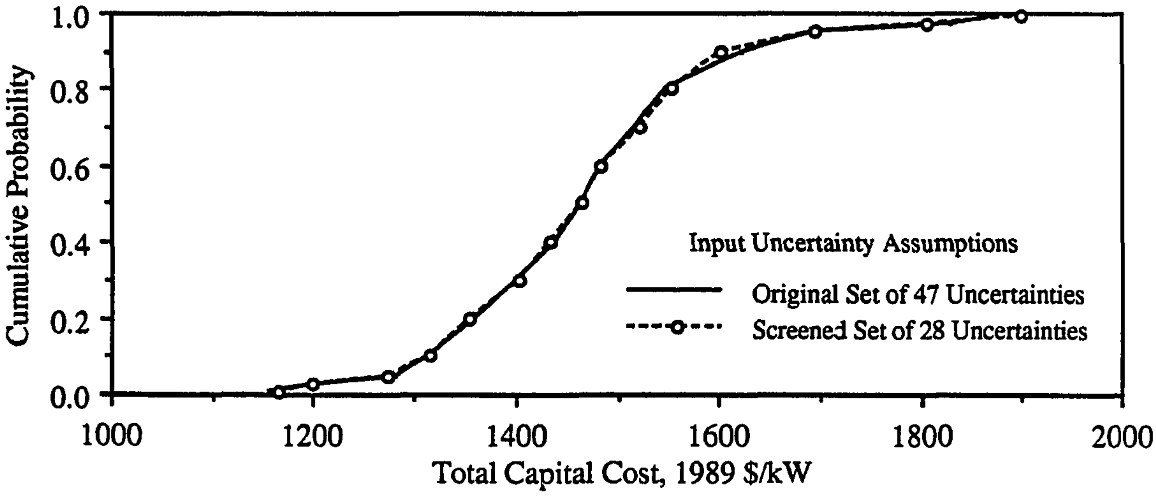

Figure 50. Comparison of Uncertainty in Total Capital Cost for Original and Screened Sets of Uncertainties.

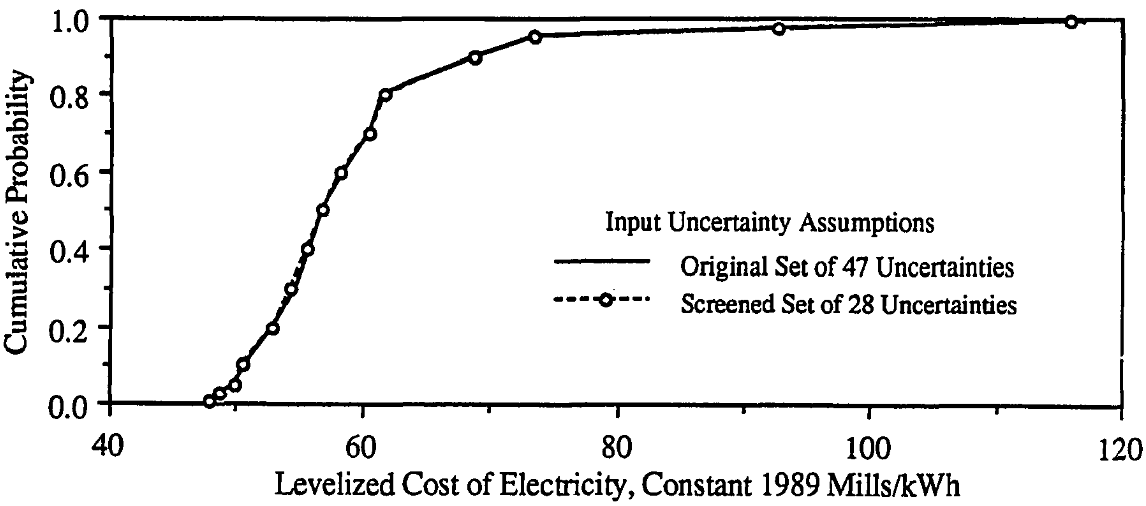

Figure 51. Comparison of Uncertainty in the Cost of Electricity for Original and Screened Sets of Uncertainties.

6.2.3 Probabilistic Design Analysis

In this section, several examples of probabilistic design analysis are considered. In each case, two alternative design assumptions are compared probabilistically. The design issues include the gas turbine fuel valve pressure drop, rich/lean gas turbine combustion for NO_x control, an incremental improvement in gas turbine design, and zinc ferrite sorbent regeneration temperature. The effect of these design modifications on plant performance and cost are evaluated.

6.2.3.1 Gas Turbine Fuel Valve

One process development that is being sought by DOE and others is the development of a low pressure drop fuel valve for gas turbine applications. As discussed in Appendix B.6, current gas turbine models commonly assumed for IGCC systems have fuel gas pressure drops of about 70 psi. This pressure drop represents an energy penalty to the IGCC system, because gasifier blast air must be pressurized to overcome the pressure drop between the gasifier and the gas turbine combustor. Process engineers expect future fuel gas valves to have pressure drops of 20 psi or less. Therefore, an IGCC system with a low pressure drop (20 psi) fuel valve was compared to on with a conventional fuel valve (70 psi pressure drop). Differences in gas turbine cost, if any, associated with the advanced fuel gas valve are not considered.

The 50 psi reduction in pressure drop between the gasifier exit and the gas turbine combustor results in a mean efficiency savings of 0.8 percentage points, or a 2.2 percent improvement in plant efficiency compared to the base case design. There is very little uncertainty regarding the efficiency savings; it varies from a low of about 0.7 to a high of 0.9 percentage points.

However, the effect of the reduced pressure drop on process costs is quite different. The gasifier coal throughput is a function of gasifier pressure. As pressure is reduced, the gasifier coal throughput is reduced. Therefore, more gasifier vessels, which are of a standard size, must be utilized to accommodate the total coal flow. Furthermore, as system pressure is reduced, the fuel gas volumetric flow rate increases. This results in increased vessel size requirements for the cyclone and zinc ferrite desulfurization process areas. These interactions are considered in the performance and cost model.

The combined effect of these trade-offs is shown in Figure 52, which shows a probability distribution for the difference in cost between the systems with the advanced and conventional fuel valves. A positive number indicates that the advanced fuel valve reduces levelized cost, neglecting any incremental costs associated with the new fuel valve

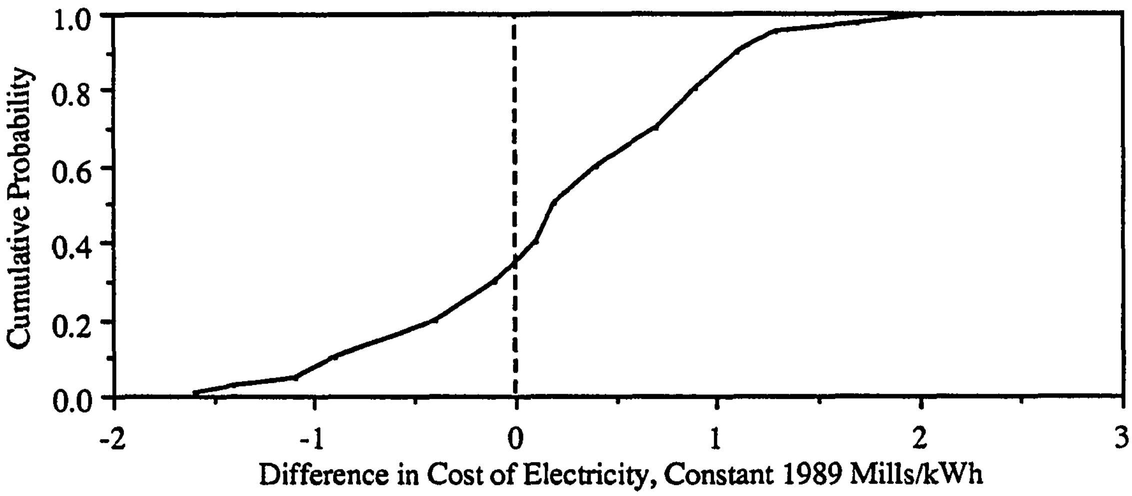

Figure 52. Uncertainty in the Difference in Cost of Electricity Between Base Case and Advanced Gas Turbine Fuel Valves.

itself. The mean cost savings is 0.2 mills/kWh. However, there is approximately a 35 percent probability that levelized costs will be higher for the advanced option, even in spite of the efficiency savings. For many of the outcomes in the probabilistic analysis, the increased costs associated with larger or more numerous process vessels offset the cost savings associated with higher plant efficiency.

6.2.3.2 Gas Turbine Rich/Lean Combustion

NO_x emissions from air-blown IGCC systems are a concern, because hot gas cleanup systems typically do not remove fuel bound-nitrogen species from the fuel gas prior to combustion in the gas turbine. The primary fuel-bound nitrogen specie in the fuel gas is ammonia, which may react almost completely to form NO_x during combustion. In contrast, cold gas cleanup systems remove essentially all of the ammonia from the fuel gas prior to combustion. Uncontrolled fuel-NO_x emissions from air-blown systems may be unacceptable, requiring the use of post-combustion emission control using SCR or the development of advanced combustors that minimize fuel NO_x formation.

As discussed in Appendix B.6, there has been research on "rich/lean" combustors. These combustors burn the fuel gas in two stages. In the first stage, fuel is combusted in an oxygen-deficient atmosphere. Most of the ammonia in the fuel gas is converted to diatomic nitrogen during rich combustion. Then the products from the rich combustion stage, along with additional fuel gas, enter a lean combustion stage. Here, conditions are such that only a portion of the diatomic nitrogen is converted to "thermal NO_x." Overall, the ammonia conversion rate to NO_x in commercial rich/lean combustors, which have yet to be developed, is predicted to be as high as 20 percent to as low as less than one percent. In

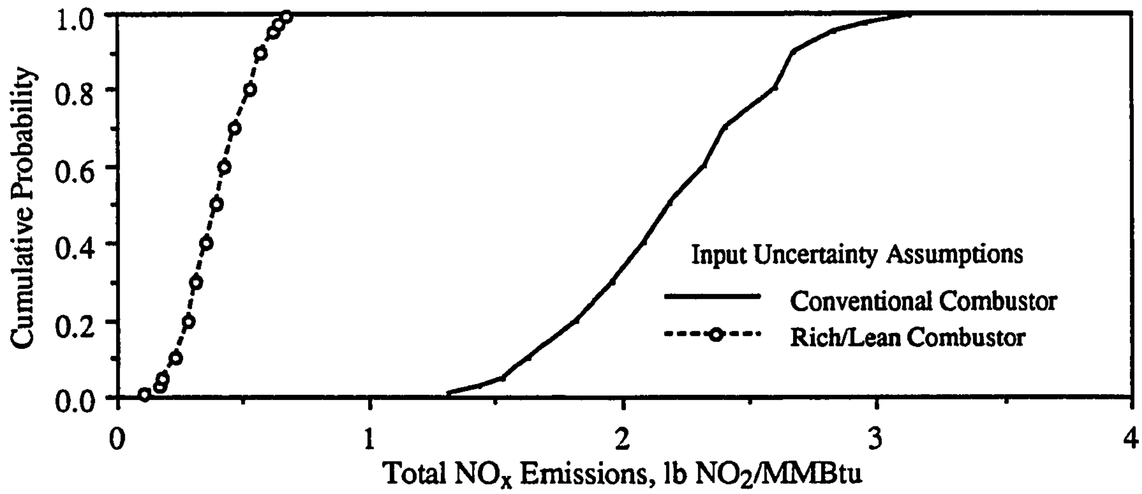

Figure 53. Uncertainty in NO_x Emissions for Conventional and Rich/Lean Gas Turbine Combustors.

contrast, conventional gas turbine combustors may convert 50 to 100 percent of ammonia to NO_x. See Appendix B.6.7.4 for more discussion of uncertainty for the rich/lean combustor.

The NO_x emission rates for conventional and rich/lean combustors are compared in Figure 53. The uncertainty in emission rates for both cases is attributable to uncertainty in the performance of the combustor as well as uncertainty in the ammonia yield from the gasifier. For the conventional combustor, even the most optimistic outcome is approximately 1.3 lb NO_2/MMBtu, which significantly exceeds typical allowable emission rates for natural gas and distillate oil-fired gas turbines. The results for the rich/lean combustor indicate a 90 percent probability range from 0.2 to 0.6 lb NO_2/MMBtu. Because of increasingly stringent environmental permitting practices, it is not clear if emission rates of this magnitude will be acceptable for an IGCC system.

6.2.3.3 Advanced Performance Gas Turbine

Gas turbine manufacturers are continually improving their products, and developing increasingly efficient machines. One design sensitivity study of interest, therefore, is the effect of an incremental improvement in gas turbine performance on the performance and cost of an IGCC system. Here, as an illustrative case study, it is assumed that an incremental improvement to the assumed gas turbine would feature an increase in pressure ratio, from 13.5 to 14.4, and turbine inlet temperature, from 2,300 °F to 2,350 °F. The incremental costs of such a machine, compared to the base case, are not considered.

The advanced gas turbine design would yield improvements in both plant efficiency and cost, without any risk that the advanced system would have lower performance or higher cost. However, there is uncertainty in the actual improvement that would be obtained. The mean efficiency improvement would be 0.45 percentage points, with a very tight uncertainty range; efficiency would not improve more than 0.5 or less than 0.4 percentage points. There is more uncertainty in the cost differences between the base case and advanced gas turbine assumptions. The mean capital cost improvement would be $70/kW, with a 90 percent probability range from $21/kW to $98/kW. The mean levelized cost savings would be 2.1 mills/kWh, with a 90 percent range from 0.7 to 3.0 mills/kWh. In making estimates of the differences in performance and cost, the correlation between the two cases is considered by properly paired the samples, as discussed in Section 6.1.

6.2.3.4 Zinc Ferrite Regeneration Temperature

The zinc ferrite regeneration reactions are highly exothermic. To prevent sintering of the sorbent, steam is proposed as a thermal diluent to keep temperatures below 1,450 °F (Kasper, 1988). The steam required for regeneration is taken from the plant steam cycle. The regeneration steam is thus diverted from use in energy conversion, and it is an efficiency penalty on the system. If a lower regeneration temperature is found to be needed to promote longer sorbent life and prevent sorbent life, what would be the effect on system performance? Here, we consider a maximum regeneration temperature of 1,400 °F.

Decreasing the regeneration temperature increases the requirement for dilution steam, resulting in a mean efficiency decrease of 0.4 points. Neglecting the effects of the lower regeneration temperature on sorbent durability, capital cost would increase, within a 90 percent probability range, from $12/kW to $17/kW. The cost of electricity would increase from 0.4 to 0.7 mills/kWh. Thus, in all cases, a decrease in regeneration temperature would lead to modest penalties on efficiency and cost. However, further study is needed to evaluated the effect of lower regeneration temperature on sorbent performance. Because sorbent performance has been shown to play an influential in both capital and operating cost, improved sorbent performance may offset the cost increase associated with reduced thermal efficiency.

6.2.4 Additional Research

Additional research is likely to reduce the uncertainties in specific process performance and cost parameters. Therefore, it is instructive to analyze the effect on total performance or cost from reductions in the uncertainty in key parameters that could result from further research.

Table 24. Illustrative Assumptions Regarding Reduction in Uncertainty in Key Process Areas

Description	Units	Distribution	Base Case Uncertainty Ranges	Reduced Case Uncertainty Ranges
GASIFIER PROCESS AREA				
Fines Carryover from Gasifier	wt-% of Coal Feed	Fractile	5%: 0 to 1	0.25 to 3
			20%: 1 to 3.5	3 to 4.25
			25%: 3.5 to 5	4.25 to 5
			25%: 5 to 8	5 to 6.5
			15%: 8 to 15	6.5 to 10
			5%: 15 to 20	10 to 12.5
			5%: 20 to 30	12.5 to 17.5
Fines Capture in Recycle Cyclone	% of Carryover	Fractile	25%: 50 to 90	72.5 to 92.5
			25%: 90 to 95	92.5 to 95
			25%: 95 to 97	95 to 96
			25%: 97 to 98	96 to 96.5
Carbon Retention in Bottom Ash	wt-% of coal feed carbon	Triangular	0.75 to 10 (2.5)	1.625 to 6.25 (2.5)
Gasifier Coal Throughput				
250 psia	lb DAF/(hr-ft^2)	Triangular	133 to 333 (266)	199.5 to 299.5 (266)
300 psia	lb DAF/(hr-ft^2)	Triangular	152 to 381 (305)	228.5 to 343 (305)
350 psia	lb DAF/(hr-ft^2)	Triangular	170 to 426 (341)	255.5 to 383.5 (341)
Gasifier Air/Coal Ratio	lb air/lb DAF	Triangular	2.7 to 3.4 (3.1)	2.9 to 3.25 (3.1)
Gasifier Steam Requirement				
Air/coal = 2.7	lb H_2O/lb DAF	Uniform	0.54 to 1.08	0.675 to 0.945
Air/coal = 3.1	lb H_2O/lb DAF	Uniform	1.24 to 1.86	1.395 to 1.705
Air/coal = 3.4	lb H_2O/lb DAF	Uniform	2.04 to 2.72	2.21 to 2.55
Direct Cost	% of DC	Uniform	10 to 30	15 to 25
Maintenance Cost	% of TC	Triangular	2 to 12 (3)	2.5 to 7.5
ZINC FERRITE DESULFURIZATION PROCESS AREA				
Sorbent Sulfur Loading	wt-% S in sorbent	Normal	2.16 to 31.84 (17)	9.58 to 24.42 (17)
Sorbent Attrition Rate	wt-% sorbent loss/cycle	Fractile	5%: 0.17 to 0.34	0.585 to 0.67
			20%: 0.34 to 0.5	0.67 to 0.75
			25%: 0.5 to 1	0.75 to 1
			25%: 1 to 1.5	1 to 1.25
			20%: 1.5 to 5	1.25 to 3
			5%: 5 to 25	3 to 13
Direct Cost	% of DC	Uniform	0 to 80	20 to 60
Maintenance Cost	% of TC	Triangular	3 to 6 (3)	3 to 4.5 (3)
GAS TURBINE PROCESS AREA				
Direct Cost	% of DC	Uniform	0 to 50	12.5 to 37.5
Maintenance Cost	% of TC	Triangular	1.5 to 6 (2)	1.75 to 4 (2)

As an illustrative case study, hypothetical reductions in uncertainty in performance and cost parameters for three major process areas are considered. These assumptions are shown in Table 24. For each uncertain parameter shown in the table, the "reduced"

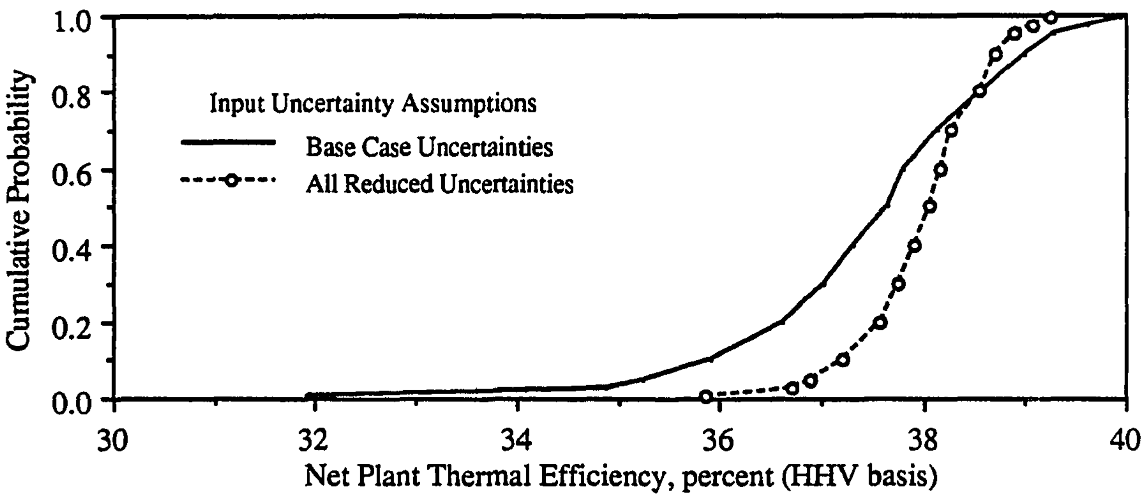

Figure 54. Effect of Reductions in Uncertainty on Uncertainty in the Plant Thermal Efficiency.

uncertainty assumes no change in the central value of the probability distribution, as represented by the median for fractile and uniform distributions, or by the mode for normal and triangular distributions. However, the ranges between the central value and the upper and lower limits are reduced by 50 percent. For fractile distributions, corresponding adjustments were made for fractiles between the median and the upper and lower limits. Thus, in all cases, the reduced uncertainties retain the qualitative features of the original assumptions; e.g., if the original distribution was positively skewed, the reduced uncertainty distribution is also positively skewed.

The effect of reducing uncertainties in performance parameters on plant efficiency is shown in Figure 54. The uncertainty in efficiency based on the reduced input parameter uncertainties has a smaller variance than for the base case. In addition, the mean efficiency increases from 37.5 in the base case to 38.0 in the reduced uncertainty case, due to the skewness of uncertainty assumptions in both the base and reduced uncertainty cases. Thus, based on the assumptions regarding results obtainable from further research, both a reduction in the risk of poor performance and an increase in the expected value of performance are obtained. The reduced uncertainties in the gasification process area play a predominate role, with uncertainties in the zinc ferrite process area having a negligible effect. This result is consistent with that obtained in the identification of key uncertainties.

The interaction among alternative groupings of reduced uncertainties as they affect capital cost is shown in Figure 55. Here, the base case uncertainties are compared to cases where reduced uncertainties in only the gasification or zinc ferrite process areas are considered, and to the case when all of the reduced uncertainties, including those of the gas

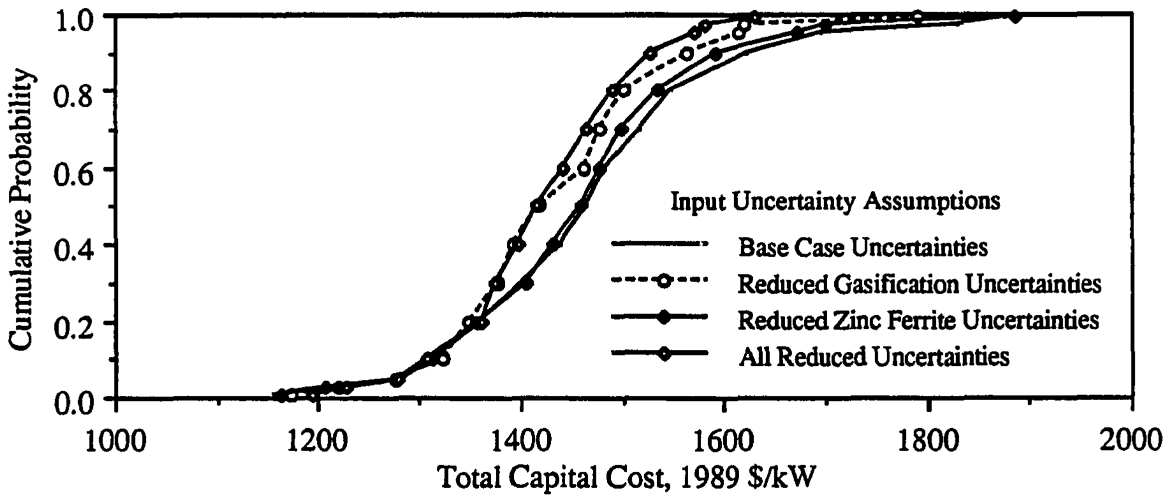

Figure 55. Effect of Reductions in Uncertainty in Specific Process Areas on Uncertainty in the Total Capital Cost.

turbine process area, are included. Reductions in uncertainty in the gasification process area alone yield a greater effect than reductions in uncertainty in the zinc ferrite process area. The mean total capital cost is reduced from $1,465/kW to $1,431/kW due to assumptions in the gasification process area. When all reduced uncertainties are considered, the mean is decreased further to $1,420/kW. Thus, a mean cost savings of $45/kW would result if reduced uncertainties could be obtained from further research.

Furthermore, the risk of high capital cost is also reduced. For example, in the base case there is a 10 percent probability that cost could exceed $1,625/kW, whereas in the case where all reduced uncertainties are assumed, the corresponding value is $1,525/kW. Further research can be expected to reduce both the expected costs and the risks of high costs. This result will be explored later when the IGCC systems are compared.

A similar result is obtained for the cost of electricity, as shown in Figure 56. However, the effect of reducing uncertainties in the zinc ferrite process area on the extreme high values obtained in the base case is more pronounced. While reducing uncertainty in the zinc ferrite process has only a modest effect on the median (a reduction less than 1 mill/kWh), it has a more pronounced effect on the mean and particularly on the upper fractiles of the distribution. The mean is reduced by 1.8 mills/kWh. At the 95th percentile, the cost is reduced by 7.3 mills/kWh, and at the 97th fractile the reduction is 21 mills/kWh. These reductions are associated with the differences in assumptions regarding sorbent sulfur loading and attrition rate.

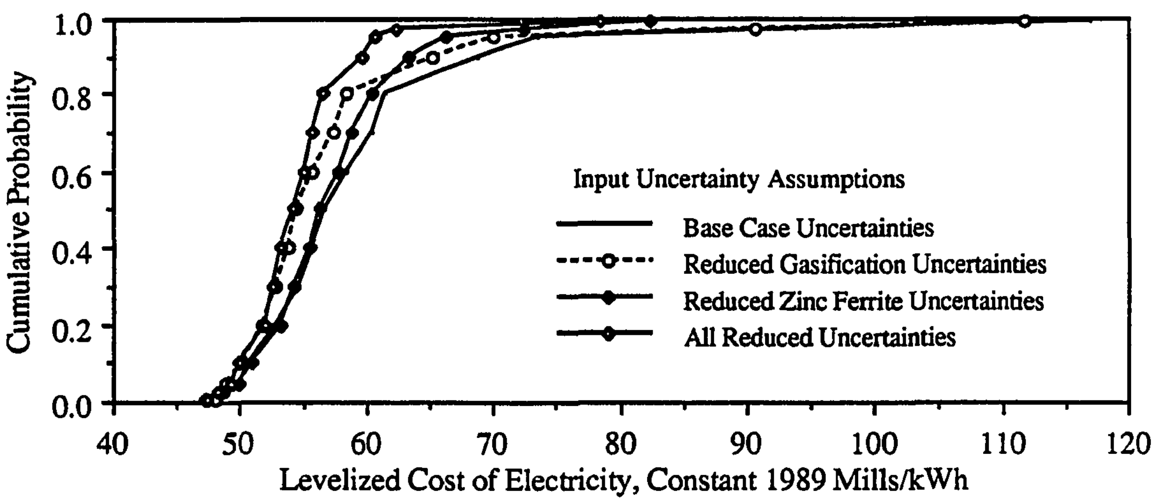

Figure 56. Effect of Reductions in Uncertainty in Specific Process Areas on Uncertainty in the Cost of Electricity.

When all reduced uncertainties are considered, the mean cost of electricity is reduced by over 4 mills/kWh, and there is only a five percent probability that cost would exceed 60.6 mills/kWh. In the base case, the probability that cost would exceed 60.6 mills/kWh is over 20 percent. Thus, as with capital cost, reduced process performance and cost uncertainties are shown to reduce the mean cost and the risk associated with high cost outcomes.

The preceding comparisons of base case and reduced uncertainties can be viewed as a form of probabilistic sensitivity analysis. The technique used here to characterize the possible effects of further research illustrates the interactions among uncertainties in several process areas. These types of insights are not obtained from traditional deterministic sensitivity analysis.

6.2.5 Multiple Experts

In many cases, there may be more than one expert from whom judgments could be elicited regarding uncertainties in a process technology. Experts may have differing beliefs regarding the range and probability of possible outcomes for specific parameters. They may have different opinions as to the set of parameters that should be treated probabilistically.

For the Lurgi-based IGCC systems, judgments from more than one expert were obtained for the zinc ferrite and gasification process areas. The implications of these alternative judgments are investigated in the following two sections.

The approach taken here is to prepare separate case studies of process performance and cost based on alternative sets of judgments from different experts. Then, the results of the case studies are compared to obtain insights into whether the experts "agree" and to identify the uncertainties that result in significant differences. This approach is preferred to combining the judgments of several experts for each input parameter and then running a single case study. In the latter case, little insight would be obtained regarding whether the differences in expert judgments would lead to different results if considered separately. Furthermore, combining judgments is problematic, because it requires a comparative weighting of the judgments. This introduces an additional element of subjectivity into the analysis.

6.2.5.1 Zinc Ferrite Process Area

Judgments from three experts were obtained regarding uncertainties in the zinc ferrite process area. In this section, these judgments are compared, with all other uncertainties held at their base case values as given in Table 16. The base case assumptions include the judgment of Expert ZF-1 regarding the zinc ferrite process areas.

Expert ZF-2 provided two sets of judgments, based on different design assumptions. These are discussed in detail in Appendix B.5.3.2. The two sets of judgments are labeled ZF-2P and ZF-2R. For ZF-2P, the expert accepted the assumptions given in the elicitation briefing package. These included the use of high efficiency cyclones for upstream particulate removal, a maximum zinc ferrite process vessel length-to-diameter ratio of 4 and a maximum diameter of 12.5 feet. However, Expert ZF-2 recommended an alternative set of assumptions. These include the use of barrier filters, rather than cyclones, for particulate control, a maximum vessel diameter of 14 feet, a maximum length-to-diameter ratio of 2.5, and the use of a chloride guard upstream of the zinc ferrite sorbent to remove hydrochloric acid from the fuel gas.

Barrier filters are believed to result in higher particulate collection efficiencies, particularly for smaller size particles. The higher particulate removal efficiency is believed to result in lower pressure drops in the zinc ferrite absorber associated with the deposition of particles in the bed. Chlorine "attacks" the sorbent, reducing its sorbent absorption capacity and leading to higher sorbent makeup requirements. Because cost data regarding barrier filters and the chloride guard were not available, the case study for judgment ZF-2R does not include any incremental costs associated with these options. However, it is not expected that they would add significantly to capital or operating costs.

The judgment of Expert ZF-3 is described in Appendix B.5.3.3. While Experts ZF-1 and ZF-2 indicated that the central value of the sorbent sulfur loading capacity would be between 15 and 18 weight percent, Expert ZF-3's judgment was that the median sorbent sulfur loading capacity would be 10 percent for a commercial scale system. However, Expert ZF-3 expects a lower sorbent replacement rate than Experts ZF-1 and ZF-2.

Four probabilistic case studies were made in which all uncertainty assumptions were the same except for the differences between the zinc ferrite experts. The results of the base case study with the judgments of Expert ZF-1 were given previously in Table 19. The results of the case studies with the alternative judgments regarding zinc ferrite process area uncertainties are given in Tables 25, 26, and 27 for judgment sets ZF-2P, ZF-2R, and ZF-3, respectively. The results for three key measures of plant performance and cost are also compared graphically. These are zinc ferrite sorbent requirement, total capital cost, and cost of electricity, as shown in Figures 57, 58, and 59, respectively.

Of the four sets of results shown in Figure 57, three are relatively close together, while the fourth indicates substantially higher sorbent requirements than for the other three. Based on the judgments of Experts ZF-1, ZF-2P, and ZF-2R, the sorbent charge is less than 20 million pounds, and the mean values are between 7 and 10 million pounds. Comparing cases ZF-2P and ZF-2R, the effect of including barrier filtration and a chloride guard is a reduction in mean sorbent charge of 1.7 million pounds. However, even in case ZF-2R, the sorbent requirement is higher than for case ZF-1, reflecting differences in the assumptions regarding zinc ferrite sorbent sulfur capacity. The sorbent requirement based on Expert ZF-2R tends to be about one to two million pounds greater than that estimated based on Expert ZF-1.

In contrast, the results based on Expert ZF-3 are a mean value of 32.5 million pounds, which is 22 to 25 million pounds greater than for the other three cases. In addition, there is a five percent probability that the requirement could be over 55 million pounds. As indicated previously, Expert ZF-3 assumed lower values for sorbent sulfur loading than the other experts. In addition, the expert also assumed a negatively skewed distribution, with a chance of obtaining very low loadings that the other experts did not consider possible. Because of the nonlinearity of the relationship between sorbent sulfur loading and the sorbent requirement, the low loading outcomes result in large sorbent requirements. These outcomes are revealed as a long tail in Figure 57.

Table 25. Summary of Results from Deterministic and Probabilistic Simulations of a 650 MW Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup: Expert Judgments LG-1 and ZF-2P.[a]

Parameter[b]	Units[c]	"Best Guess"[d]	$f_{.50}$	μ	σ	$f_{.05}$ - $f_{.95}$
Plant Performance						
Thermal Efficiency	%, HHV	38.6	37.7	37.5	1.3	35.1 - 39.4
Coal Consumption	lb/kWh	0.786	0.805	0.811	0.029	0.770 - 0.865
Process Water Consump.	lb/kWh	1.60	1.600	1.634	0.261	1.209 - 2.119
ZF Sorbent Charge	10^6 lb	9.31	9.15	10.25	3.00	6.82 - 16.43
Sulfuric Acid Production	lb/kWh	0.085	0.087	0.087	0.003	0.083 - 0.093
Plant Discharges						
SO_2 Emissions	lb/MMBtu	0.042	0.040	0.040	0.001	0.038 - 0.042
NO_x Emissions	lb/MMBtu	2.74	2.19	2.19	0.402	1.53 - 2.84
CO Emissions	lb/kWh	0.003	0.003	0.003	0.003	0.003 - 0.006
CO_2 Emissions	lb/kWh	1.72	1.72	1.72	0.032	1.67 - 1.78
Solid Waste	lb/kWh	0.083	0.096	0.098	0.015	0.080 - 0.125
Plant Costs						
Total Capital Cost	$/kW	1,442	1,511	1,522	131	1,306 - 1,739
Fixed Operating Cost	$/kW-yr	45.1	57.2	59.6	10.5	46.9 - 81.7
Variable Operating	mills/kWh	20.88	30.1	33.2	11.5	20.7 - 59.5
Coal		16.2	16.6	16.7	0.6	15.9 - 17.7
Byproduct		(1.5)	(1.4)	(1.3)	0.5	(0.4) - (2.0)
Other		6.3	14.8	17.9	11.4	5.4 - 43.1
Cost of Electricity	mills/kWh	55.0	67.6	71.3	12.6	55.5 - 100.5

[a] The notation in the table heading is defined as follows: f_n = n^{th} fractile ($f_{.50}$ = median), μ = mean; and σ = standard deviation of the probability distribution. The range enclosed by $f_{.05}$ to $f_{.95}$ is the 90 percent probability range. All costs are January 1989 dollars.

[b] Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, and zinc ferrite steam. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

[c] HHV = higher heating value; MMBtu = million Btu.

[d] Based on a deterministic simulation in which median or modal values of uncertain variables are assumed as "best guess" inputs to the model

Table 26. Summary of Results from Deterministic and Probabilistic Simulations of a 650 MW Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup: Expert Judgments LG-1 and ZF-2R.[a]

Parameter[b]	Units[c]	"Best Guess"[d]	$f_{.50}$	μ	σ	$f_{.05}$ - $f_{.95}$
Plant Performance						
Thermal Efficiency	%, HHV	38.8	37.9	37.7	1.3	35.5 - 39.6
Coal Consumption	lb/kWh	0.783	0.801	0.806	0.029	0.767 - 0.856
Process Water Consump.	lb/kWh	1.59	1.59	1.62	0.258	1.206 - 2.114
ZF Sorbent Charge	10^6 lb	7.66	7.69	8.54	2.16	6.15 - 13.06
Sulfuric Acid Production	lb/kWh	0.085	0.086	0.087	0.003	0.082 - 0.093
Plant Discharges						
SO_2 Emissions	lb/MMBtu	0.042	0.040	0.040	0.001	0.038 - 0.042
NO_x Emissions	lb/MMBtu	2.74	2.19	2.19	0.402	1.53 - 2.84
CO Emissions	lb/kWh	0.003	0.003	0.003	0.002	0.003 - 0.006
CO_2 Emissions	lb/kWh	1.71	1.71	1.71	0.030	1.67 - 1.77
Solid Waste	lb/kWh	0.082	0.096	0.098	0.015	0.079 - 0.124
Plant Costs						
Total Capital Cost	$/kW	1,409	1,475	1,477	123	1,297 - 1,696
Fixed Operating Cost	$/kW-yr	44.6	57.7	60.0	10.6	46.4 - 84.9
Variable Operating	mills/kWh	17.9	19.4	19.9	1.9	17.4 - 23.9
Coal		16.1	16.5	16.6	0.6	15.8 - 17.6
Byproduct		(1.5)	(1.4)	(1.3)	0.5	(0.4) - (2.0)
Other		3.4	4.2	4.7	1.7	2.8 - 8.2
Cost of Electricity	mills/kWh	51.4	56.9	57.3	4.3	50.7 - 65.4

[a] The notation in the table heading is defined as follows: f_n = n^{th} fractile ($f_{.50}$ = median), μ = mean; and σ = standard deviation of the probability distribution. The range enclosed by $f_{.05}$ to $f_{.95}$ is the 90 percent probability range. All costs are January 1989 dollars.

[b] Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, and zinc ferrite steam. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

[c] HHV = higher heating value; MMBtu = million Btu.

[d] Based on a deterministic simulation in which median or modal values of uncertain variables are assumed as "best guess" inputs to the model

Table 27. Summary of Results from Deterministic and Probabilistic Simulations of a 650 MW Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup: Expert Judgments LG-1 and ZF-3.[a]

Parameter[b]	Units[c]	"Best Guess"[d]	$f_{.50}$	μ	σ	$f_{.05}$ - $f_{.95}$
Plant Performance						
Thermal Efficiency	%, HHV	38.6	37.6	37.5	1.3	35.2 - 39.3
Coal Consumption	lb/kWh	0.786	0.807	0.811	0.029	0.773 - 0.864
Process Water Consump.	lb/kWh	1.599	1.601	1.636	0.261	1.212 - 2.124
ZF Sorbent Charge	10^6 lb	25.1	26.2	32.5	21.1	16.4 - 54.7
Sulfuric Acid Production	lb/kWh	0.085	0.087	0.087	0.003	0.083 - 0.094
Plant Discharges						
SO_2 Emissions	lb/MMBtu	0.042	0.040	0.040	0.001	0.038 - 0.042
NO_x Emissions	lb/MMBtu	2.74	2.19	2.19	0.402	1.53 - 2.84
CO Emissions	lb/kWh	0.003	0.003	0.003	0.002	0.003 - 0.006
CO_2 Emissions	lb/kWh	1.71	1.73	1.73	0.031	1.68 - 1.78
Solid Waste	lb/kWh	0.083	0.096	0.098	0.015	0.079 - 0.125
Plant Costs						
Total Capital Cost	$/kW	1,568	1,685	1,735	273	1,392 - 2,161
Fixed Operating Cost	$/kW-yr	46.0	63.0	64.7	11.7	49.0 - 89.0
Variable Operating	mills/kWh	16.7	18.0	18.6	2.2	16.3 - 23.5
Coal		16.2	16.6	16.7	0.6	15.9 - 17.8
Byproduct		(1.5)	(1.4)	(1.3)	0.5	(0.4) - (2.0)
Other		2.1	2.6	3.2	2.0	1.8 - 6.1
Cost of Electricity	mills/kWh	53.3	60.3	61.5	7.6	51.7 - 74.7

[a] The notation in the table heading is defined as follows: f_n = n^{th} fractile ($f_{.50}$ = median), μ = mean; and σ = standard deviation of the probability distribution. The range enclosed by $f_{.05}$ to $f_{.95}$ is the 90 percent probability range. All costs are January 1989 dollars.

[b] Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, and zinc ferrite steam. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

[c] HHV = higher heating value; MMBtu = million Btu.

[d] Based on a deterministic simulation in which median or modal values of uncertain variables are assumed as "best guess" inputs to the model

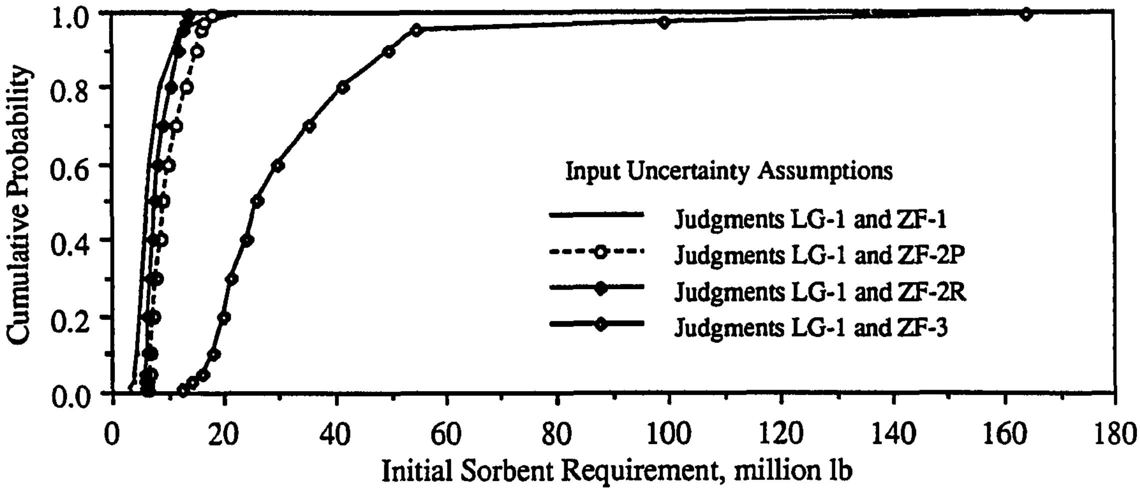

Figure 57. Comparison of the Uncertainty in Initial Zinc Ferrite Sorbent Requirement based on the Judgments of Different Experts Regarding the Zinc Ferrite Process Area.

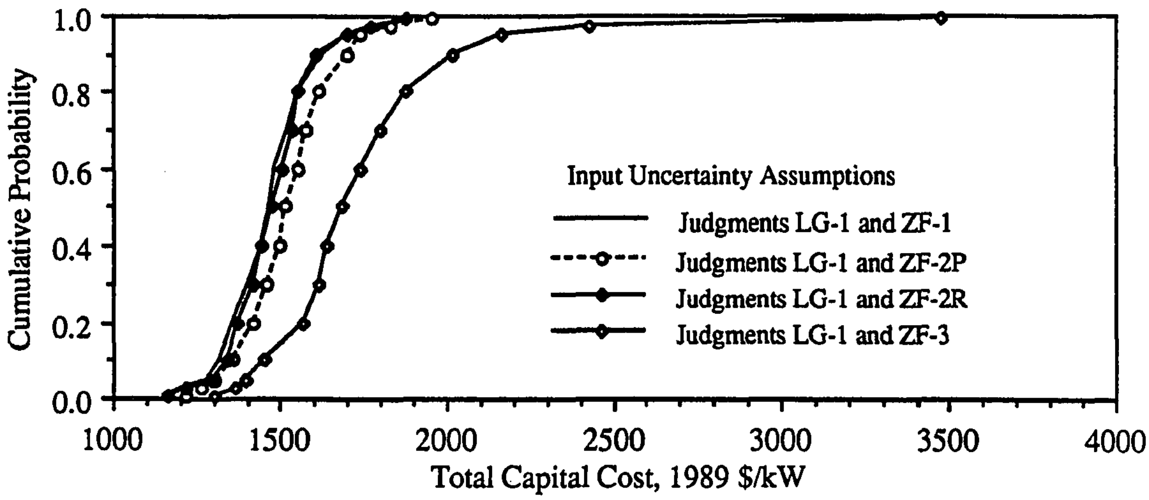

Figure 58. Comparison of the Uncertainty in Total Capital Cost based on the Judgments of Different Experts Regarding the Zinc Ferrite Process Area.

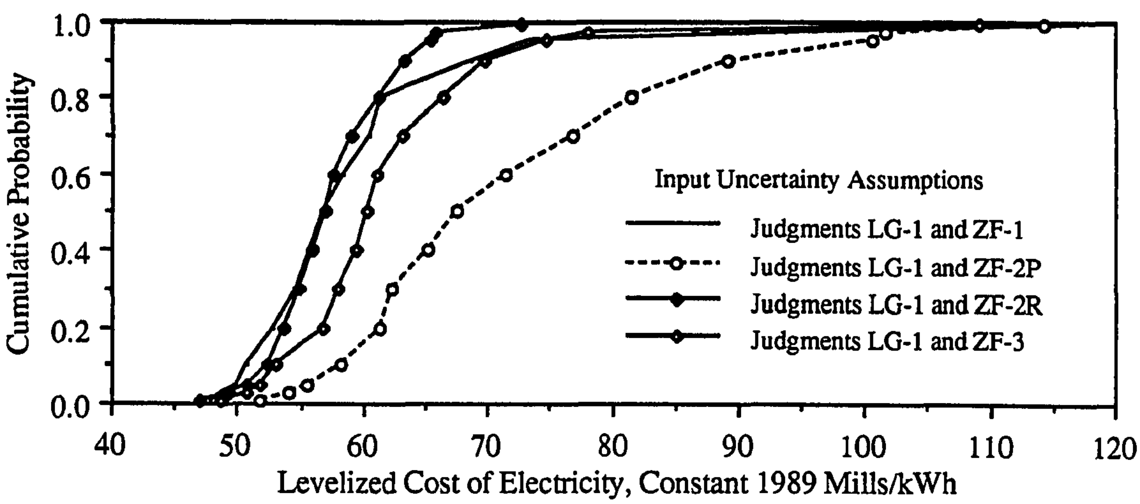

Figure 59. Comparison of the Uncertainty in Cost of Electricity based on the Judgments of Different Experts Regarding the Zinc Ferrite Process Area.

The results obtained for zinc ferrite sorbent sulfur loading directly affect capital cost, as shown in Figure 58. Qualitatively, the same results are obtained: the results based on the judgments of Experts ZF-1 and ZF-2 are clustered together, while those based on Expert ZF-3 lead to higher capital costs than for the other three cases.

The differences between the four sets of judgments become more evident when comparing results for the cost of electricity, as shown in Figure 59. Unlike the results for capital cost, the judgment of Expert ZF-3 does not result in the highest cost of electricity. Instead, the highest cost result is obtained based on the judgment set ZF-2P.

The results for the cost of electricity are influenced both by the sorbent sulfur loading requirement and the annual makeup sorbent requirement due to sorbent attrition and deactivation. As shown graphically in Chapter 4, the judgment of Expert ZF-3 regarding the sorbent replacement rate is lower than for the other experts. Although Expert ZF-3 generally predicted lower sorbent sulfur loadings for a commercial scale zinc ferrite process, the expert also predicted lower sorbent attrition rates than assumed in the other three cases. Therefore, in spite of the higher initial sorbent requirement obtained using the judgments of Expert ZF-3, the net effect of the interactions of uncertainties in sorbent loading and sorbent replacement leads to a lower cost of electricity uncertainty estimate than obtained from Expert ZF-2P.

The judgments ZF-1 and ZF-2R result in similar central tendencies. However, there is less of a risk of extremely high costs based on ZF-2R than for the other three cases. This is because Expert ZF-2 indicated that sorbent performance would be improved with the use of a chloride guard and barrier filtration, allowing for high sorbent sulfur loadings and increased sorbent life. Without a chloride guard or barrier filtration, the judgments of Expert ZF-2 lead to an upward shift in the central values of the distribution for cost of electricity, and a long tail, due to the possibility of low sorbent capacity, loss of sorbent due to chloride attack, and a short life cycle.

The mean and median values for the cost of electricity for cases ZF-1, ZF-2P, and ZF-3 fall between 55 and 60 mills/kWh. For case ZF-2P, the mean value is 71.3 mills/kWh. Within a tolerance of about 5 mills/kWh, the first three cases are in approximate agreement, while case ZF-2P represents a substantial departure from the other three. With respect to extreme values, the case based on ZF-2R is a departure from the other three. The other three cases indicate that outcomes of well over 70 mills/kWh are possible, while such outcomes are not indicated based on ZF-2R. The implication here is that the use of barrier filtration and a chloride guard could substantially reduce the risk of poor sorbent performance.

Do the experts agree? There appears to be reasonable agreement among three of the cases with respect to zinc ferrite sorbent charge and total capital cost. However, all four cases lead to differences in either the central tendency or high fractiles of the cost of electricity, representing the more complex interactions among assumptions that affect sorbent charge and makeup sorbent. Furthermore, although the results from the judgment ZF-2R agrees with the central tendency of the result based on Expert ZF-1, the design assumptions differ substantially. The results suggest that the experts do not agree with respect to the factors that influence the sorbent replacement rate, although most of the judgments result in similar results for the initial sorbent charge and capital cost.

6.2.5.2 Gasification Process Area

Two sets of judgments were obtained regarding uncertainties in the Lurgi gasification process area. The judgments of Expert LG-2 are discussed in detail in Appendix B.3.4.2. Here, the results obtained based on the judgments of Experts LG-1 and LG-2 are compared. The judgments of Expert ZF-2P are used for the both cases in the comparison. The uncertainties in all other process variables are the same as given in Table 16.

The experts differ primarily in the assessment of uncertainty in the air-to-coal ratio of the gasifier. Expert LG-1 estimated that uncertainty in the air/coal ratio is triangularly distributed, with minimum and maximum values of 2.7 and 3.4 lb air/lb dry ash-free (DAF) coal and a mode at 2.9. In contrast, Expert LG-2 estimated that the mode of the uncertainty in the air/coal ratio required to achieve a "good fuel gas" is 2.41 lb air/lb coal, and that the range of uncertainty is from 0.4 to 2.9. Thus, the modal value assumed by Expert LG-2 is substantially lower than that assumed by Expert LG-1. Furthermore, the judgment of Expert LG-2 is negatively skewed toward low values, while that of Expert LG-1 is symmetric about the mode. Because the air flow rate to the gasifier directly affects the plant auxiliary power requirement for boost air compression, high air/coal ratios result in a significant reduction in net plant efficiency.

The results of the probabilistic case study based on LG-2 are summarized in Table 28. In addition, the results for plant efficiency, total capital cost, and the cost of electricity are shown in Figures 60, 61, and 62, respectively. With respect to plant efficiency, the results based on Expert LG-2 indicate both a higher expected value (mean) and higher

Table 28. Summary of Results from Deterministic and Probabilistic Simulations of a 650 MW Air-Blown Lurgi-based IGCC System with Hot Gas Cleanup: Expert Judgments LG-2 and ZF-2P.[a]

Parameter[b]	Units[c]	"Best Guess"[d]	$f_{.50}$	μ	σ	$f_{.05}$ - $f_{.95}$
<u>Plant Performance</u>						
Thermal Efficiency	%, HHV	41.9	38.9	39.0	3.3	32.7 - 44.7
Coal Consumption	lb/kWh	0.725	0.781	0.785	0.068	0.680 - 0.929
Process Water Consump.	lb/kWh	1.170	1.222	1.320	0.304	0.992 - 1.928
ZF Sorbent Charge	10^6 lb	7.72	8.01	8.84	2.8	5.8 - 14.5
Sulfuric Acid Production	lb/kWh	0.073	0.073	0.075	0.008	0.062 - 0.090
<u>Plant Discharges</u>						
SO_2 Emissions	lb/MMBtu	0.036	0.035	0.035	0.002	0.031 - 0.039
NO_x Emissions	lb/MMBtu	1.60	1.43	1.43	0.325	0.901 - 1.98
CO Emissions	lb/kWh	0.003	0.003	0.003	0.002	0.003 - 0.006
CO_2 Emissions	lb/kWh	1.59	1.61	1.61	0.11	1.41 - 1.79
Solid Waste	lb/kWh	0.077	0.112	0.115	0.029	0.079 - 0.184
<u>Plant Costs</u>						
Total Capital Cost	$/kW	1,250	1,317	1,313	105	1,152 - 1,516
Fixed Operating Cost	$/kW-yr	37.6	46.4	47.6	6.5	39.0 - 59.4
Variable Operating	mills/kWh	19.09	28.5	31.1	10.2	19.0 - 52.2
Coal		14.9	16.1	16.1	1.4	14.0 - 19.1
Byproduct		(1.3)	(1.1)	(1.1)	0.4	(0.3) - (1.8)
Other		5.5	13.8	16.1	9.9	5.1 - 36.4
Cost of Electricity	mills/kWh	48.4	61.7	63.3	11.4	49.4 - 87.0

[a] The notation in the table heading is defined as follows: f_n = n^{th} fractile ($f_{.50}$ = median), μ = mean; and σ = standard deviation of the probability distribution. The range enclosed by $f_{.05}$ to $f_{.95}$ is the 90 percent probability range. All costs are January 1989 dollars.

[b] Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, and zinc ferrite steam. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

[c] HHV = higher heating value; MMBtu = million Btu.

[d] Based on a deterministic simulation in which median or modal values of uncertain variables are assumed as "best guess" inputs to the model

variance than for Expert LG-1. The mean efficiency obtained is 1.5 percentage points higher. There is more probability of obtaining an efficiency below 35 percent based on Expert LG-2, but also about a 40 percent probability of obtaining an efficiency greater than 40 percent. In contrast, based on Expert LG-1, there is no chance of an efficiency greater than 40 percent. Efficiencies as high as 45 percent may be attainable, based on the judgment of Expert LG-2. These differences are strongly dependent on the the differences in assumptions regarding the air/coal ratio.

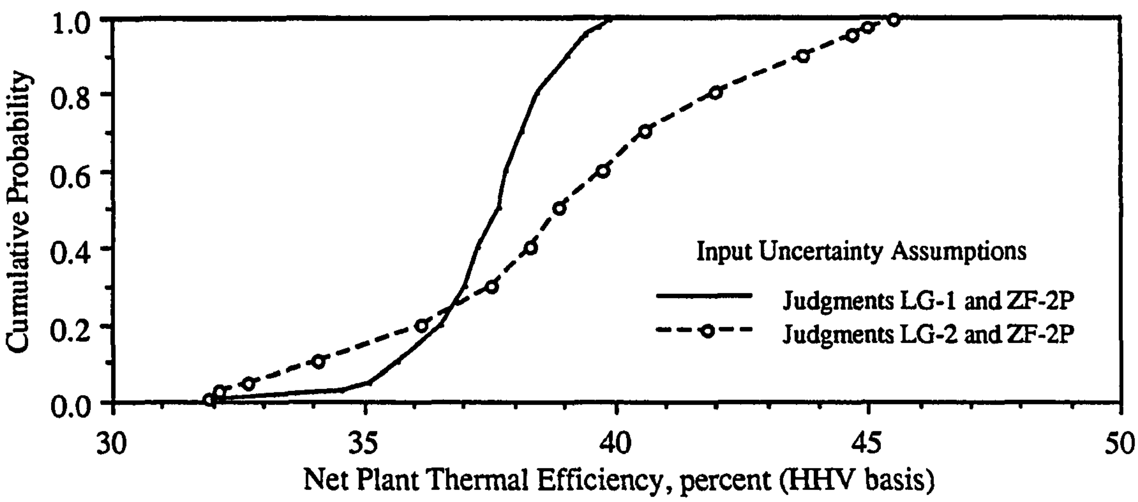

Figure 60. Comparison of the Uncertainty in Plant Efficiency based on the Judgments of Different Experts Regarding the Gasification Process Area.

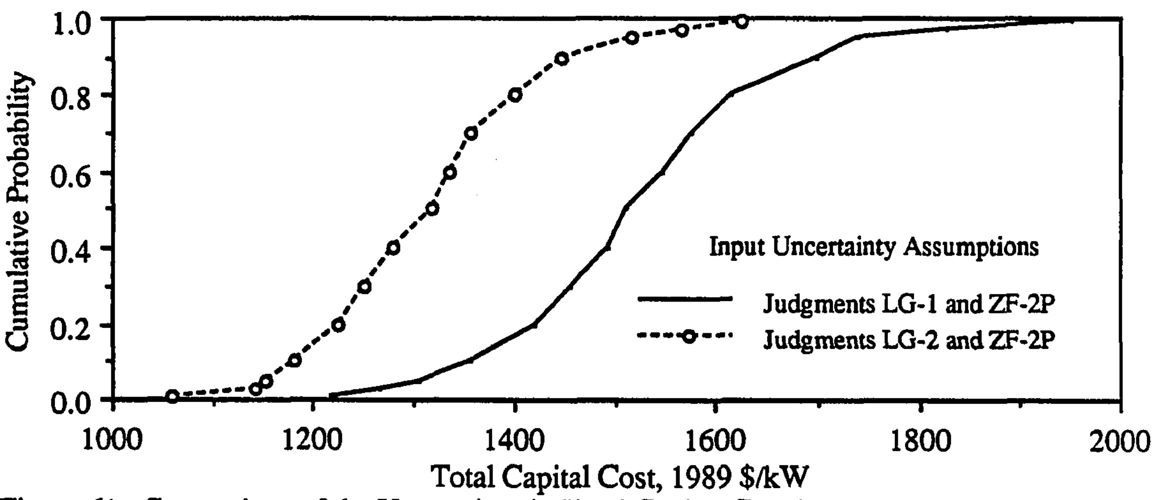

Figure 61. Comparison of the Uncertainty in Total Capital Cost based on the Judgments of Different Experts Regarding the Gasification Process Area.

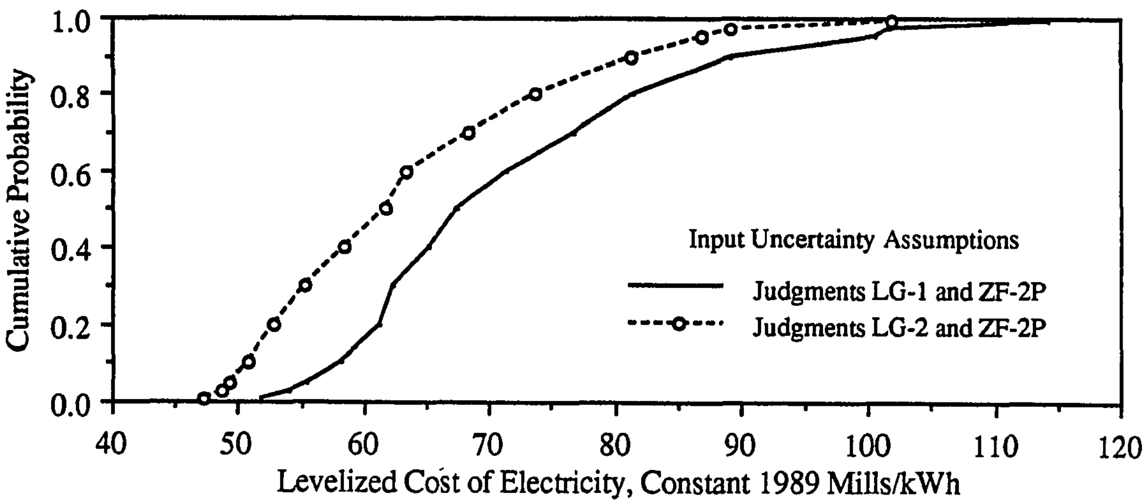

Figure 62. Comparison of the Uncertainty in Cost of Electricity based on the Judgments of Different Experts Regarding the Gasification Process Area.

Using the judgments of Expert LG-2, substantially lower capital and levelized costs are obtained. The mean capital cost is about \$200/kW lower, while the mean cost of electricity is about 8 mills/kWh lower. These results are strongly affected by the air/coal ratio assumption.

Thus, the judgments of Experts LG-1 and LG-2 have significantly different implications for process performance and cost. With respect to performance, the judgment of Expert LG-2 leads to a larger magnitude of uncertainty, but also a greater probability of favorable outcomes. With respect to cost, the variance of the capital and levelized costs are slightly reduced; however, the outcomes are generally substantially lower than those based on Expert LG-1.

6.2.6 Correlation Structures

The preceding analyses, except in the cases of probabilistic comparisons of two alternative systems, have assumed that the model input probability distributions are uncorrelated. While this assumption is often reasonable, there may be cases when correlations are known or believed to exist between input uncertainties. In this section, the effect of possible correlations in input uncertainties is considered.

A number of possible correlation structures are discussed in Appendix B. The identification of these correlations was based on the explanations given by the experts of the reasons for obtaining various outcomes for performance parameters. In many cases, the same mechanism may be responsible for high or low outcomes in two or more variables. For example, uncertainty carbon and sulfur retention in the gasifier bottom ash may vary together. Correlations in the gasification process area are discussed further in Appendix B.3.4, and for the zinc ferrite process area in B.5.3.

The correlations assumed for the gasification process area tend to mitigate both high and low outcomes, but only slightly. For example, the carbon retention in the bottom ash is assumed to be negatively correlated with the air/coal ratio. As the air/coal ratio is increased, the carbon retention would be expected to tend to decrease, as more carbon would tend to be either combusted or gasified. Thus, there is less chance of obtaining high carbon retentions and high air/coal ratios simultaneously when correlations are imposed. The effect of correlation on the pairing of samples for these two parameters is revealed in a comparison of Figures 63 and 64. In the first figure, the pairing of the 100 samples of both the air/coal ratio and the carbon retention in the bottom ash is shown when no correlation is assumed. Note that both uncertain variables are assumed to have triangular distributions. Thus, we expect a concentration of samples near the intersection of the model values, which are an air/coal ratio of 3.1 and a carbon retention of 2.5 percent. In the second figure, a correlation of -50 percent is imposed. The pairing is thus more ordered than in the previous case.

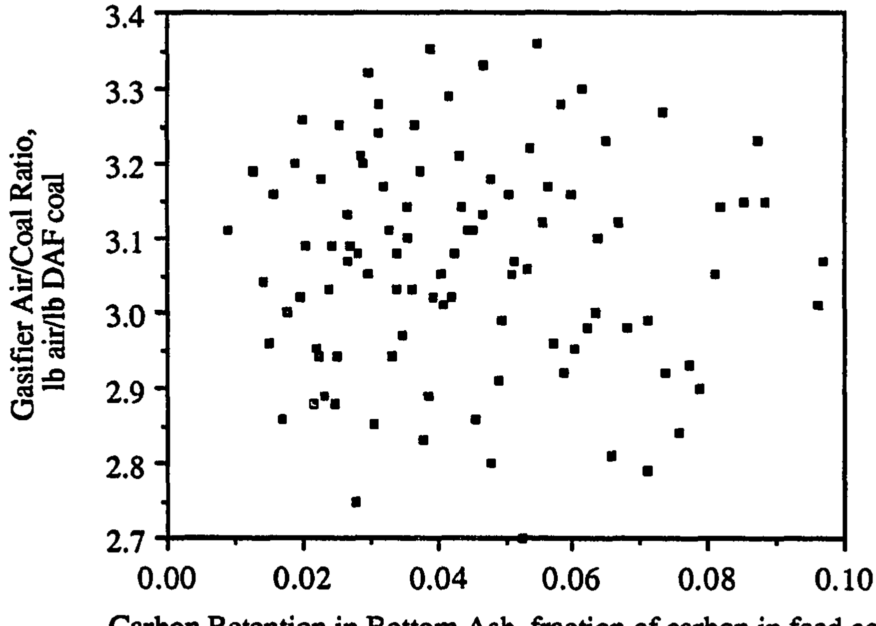

Figure 63. Pairing of Samples for Air/Coal Ratio and Carbon Retention in Bottom Ash with Rank Correlation = 0

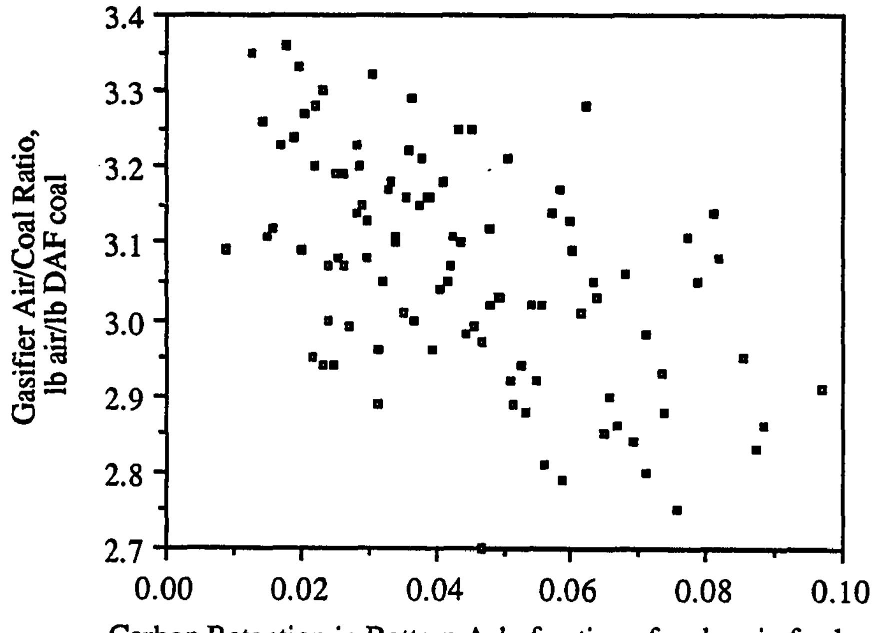

Figure 64. Pairing of Samples for Air/Coal Ratio and Carbon Retention in Bottom Ash with Rank Correlation = -0.5

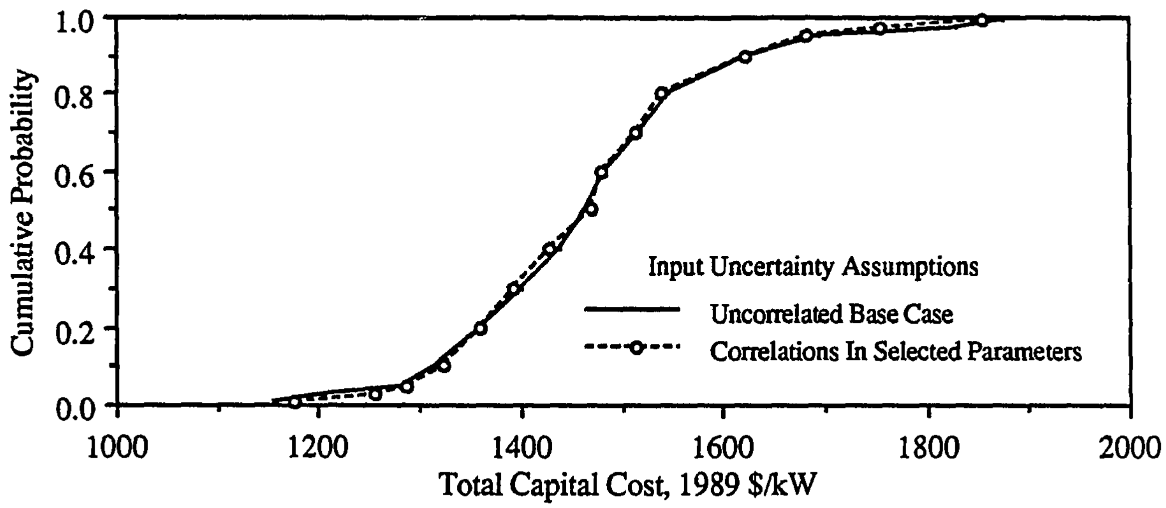

Figure 65. Effect of Assumed Correlations on Uncertainty in Capital Cost.

The effect of the assumed correlation between the air/coal ratio and carbon retention, and the other correlations assumed also, is shown for the capital cost in Figure 65. The results from the correlated case are approximately the same as from the base case. Although many of the performance parameters for which correlations were assumed are key uncertainties in the model, the correlations apparently were not of sufficient magnitude to result in a significant difference in results.

Other correlation cases were evaluated, including for case studies with the judgments of Expert LG-2. Only minor effects on the results were obtained with respect to performance. In addition, correlations among the direct and maintenance cost uncertainties in the gasification, zinc ferrite, and gas turbine process areas were also considered. These, too, produced only minor effects on the results. Therefore, for convenience, in later comparative case studies, the results based on uncorrelated sampling are used. However, in most cases where correlations were assumed, only modest correlation coefficients were used, such as of magnitude 0.5 or less. In cases where an analyst believes that strong correlations exist, their effect on the result may be stronger and should to be evaluated.

6.3 Probabilistic Analysis of the Oxygen-Blown KRW-Based IGCC System with Cold Gas Cleanup

The probabilistic case studies of the oxygen-blown KRW-based IGCC system with cold gas cleanup are based upon the assumptions regarding values and uncertainties in key model parameters given in Table 14. The analysis of this system included: (1) characterization of uncertainties in model output variables; (2) identification of key uncertainties; (3) evaluation of the effect that additional research might have on output

Table 29. Summary of Results from Deterministic and Probabilistic Simulations of a 650 MW Oxygen-Blown KRW-based IGCC System with Cold Gas Cleanup.[a]

Parameter[b]	Units[c]	"Best Guess"[d]	$f_{.50}$	μ	σ	$f_{.05}$ - $f_{.95}$
Plant Performance						
Thermal Efficiency	%, HHV	38.5	36.2	36.0	1.6	33.2 - 38.2
Coal Consumption	lb/kWh	0.788	0.839	0.845	0.038	0.794 - 0.914
Process Water Consump.	lb/kWh	0.779	0.803	0.812	0.051	0.733 - 0.896
Sulfur Production	lb/kWh	0.021	0.023	0.023	0.001	0.021 - 0.025
Plant Discharges						
SO_2 Emissions	lb/MMBtu	0.342	0.335	0.334	0.012	0.311 - 0.352
NO_x Emissions	lb/MMBtu	0.142	0.132	0.131	0.034	0.077 - 0.186
CO Emissions	lb/kWh	0.0001	0.0001	0.0001	2.2×10^{-5}	0.0001 -0.0001
CO_2 Emissions	lb/kWh	1.68	1.68	1.67	0.021	1.63 - 1.70
Solid Waste	lb/kWh	0.079	0.084	0.084	0.004	0.079 - 0.091
Plant Costs						
Total Capital Cost	$/kW	1,738	1,796	1,806	99	1,645 - 1,985
Fixed Operating Cost	$/kW-yr	54.4	57.1	56.8	4.4	49.4 - 64.2
Variable Operating	mills/kWh	16.2	17.6	17.8	08	16.6 - 19.4
Coal		16.2	17.3	17.4	0.8	16.3 - 18.8
Byproduct		(1.2)	(1.1)	(1.1)	0.2	(0.8) - (1.3)
Other		1.2	1.4	1.4	0.2	1.2 - 1.7
Cost of Electricity	mills/kWh	57.3	60.5	60.5	2.6	56.4 - 65.6

[a] The notation in the table heading is defined as follows: f_n = n^{th} fractile ($f_{.50}$ = median), μ = mean; and σ = standard deviation of the probability distribution. The range enclosed by $f_{.05}$ to $f_{.95}$ is the 90 percent probability range. All costs are January 1989 dollars.

[b] Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, and zinc ferrite steam. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

[c] HHV = higher heating value; MMBtu = million Btu.

[d] Based on a deterministic simulation in which median or modal values of uncertain variables are assumed as "best guess" inputs to the model

uncertainties; and (4) evaluation of the effect that a correlation structure might have on the results.

6.3.1 Characterization of Uncertainties in Performance and Cost

Results of the simulation with the base case uncertainty assumptions are given in Table 29. The results for plant thermal efficiency, SO_2 emissions, total capital cost, and the cost of electricity are shown as cdfs in Figures 66, 67, 68, and 69, respectively.

Because of the negative skewness of the assumption regarding uncertainty in the gasifier carbon conversion efficiency, the plant thermal efficiency is also negatively skewed. The mode of the uncertainty in carbon conversion was taken to be at the upper

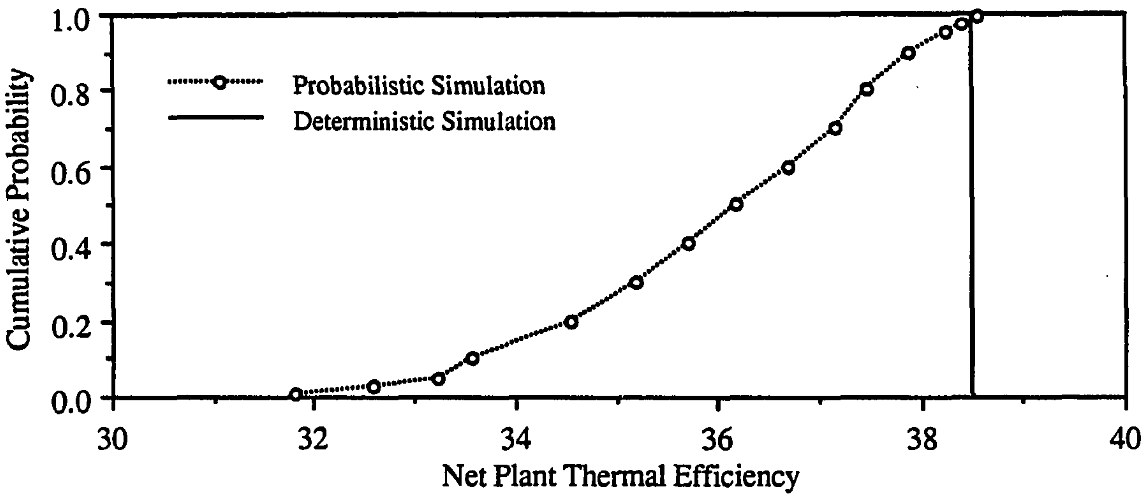

Figure 66. Comparison of Deterministic and Probabilistic Results for the Net Plant Thermal Efficiency of the Oxygen-blown KRW-based System.

bound of the distribution, and the modal value was used as the "best guess" in the deterministic analysis. The modal value of 95 percent carbon conversion is also widely assumed in conceptual design studies (e.g., Dawkins et al, 1985; Gallaspy et al, 1990). However, scale-up risks and inherent design limitations for the KRW gasifier may lead to lower carbon conversions and, hence, lower plant efficiencies, than commonly assumed (Shinnar et al, 1987). See Appendix B.4 for a more detailed discussion of uncertainties in the KRW gasifier.

The KRW gasifier may retain a significant portion of coal sulfur in the bottom ash, thereby reducing the sulfur burden to the cold gas cleanup system and reducing SO_2 emissions. The uncertainty in SO_2 emissions is illustrated in Figure 67, indicating that emissions may be lower than the "best guess" assumption. The SO_2 emissions are normalized to the inlet heating value flow associated with the coal feed. The higher probability of obtaining values lower than the "best guess" is associated with the interactions between the symmetric distribution assumed for sulfur retention and the negatively skewed distribution assumed for carbon conversion. The latter results in a positively skewed distribution for coal feed rate and, hence, total heating value entering the IGCC plant. Higher heating value flows are associated with low carbon conversions and low plant efficiencies. As heating value increases, the normalized emission rate tends to decrease.

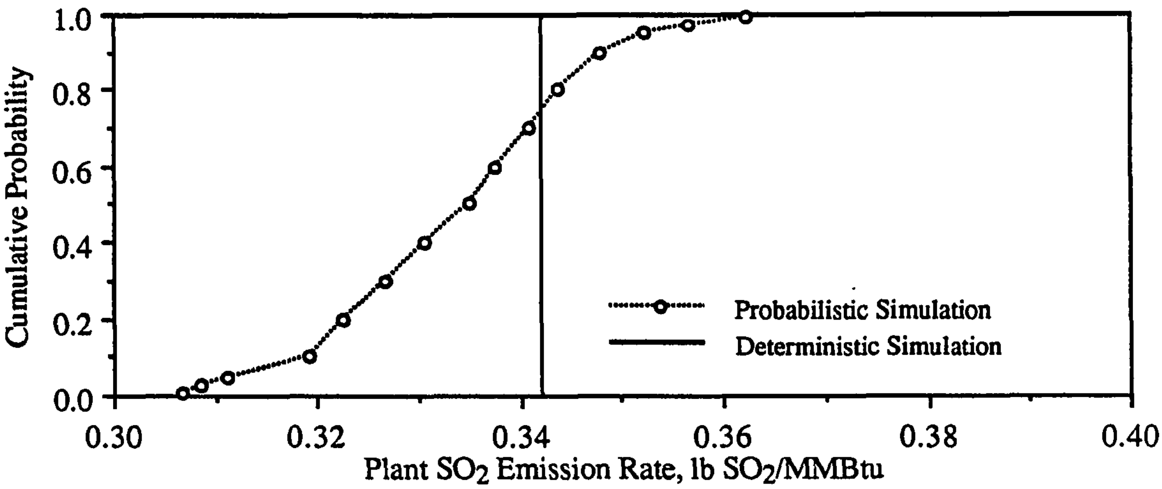

Figure 67. Comparison of Deterministic and Probabilistic Results for the Sulfur Dioxide Emissions of the Oxygen-blown KRW-based System.

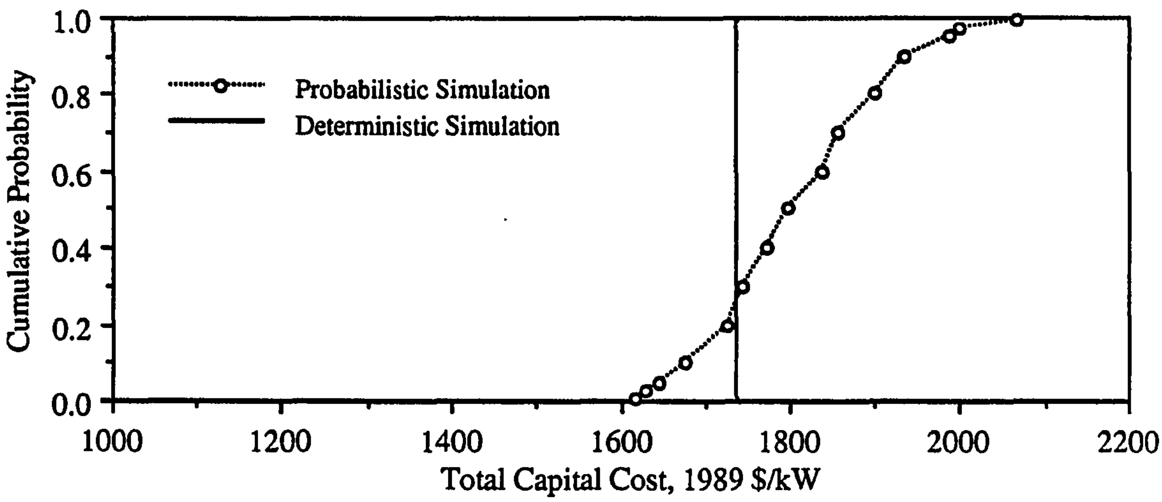

Figure 68. Comparison of Deterministic and Probabilistic Results for the Total Capital Cost of the Oxygen-blown KRW-based System.

Although the oxygen-blown KRW system considered here represents elements of "conventional" IGCC technology, particularly the cold gas cleanup system, there is still considerable performance and cost risk associated with the gasification process area. Uncertainty in plant performance and capital cost-related parameters result in the uncertainty in total capital cost shown in Figure 68. Compared to the deterministic "best guess" estimate, which includes values of both process and project contingency factors typically assumed in the literature, there is a 70 percent probability of cost overrun. While estimates of uncertainties in capital cost parameters, including process area direct costs, were based on symmetric probability distributions, the underlying negative skewness of the major measures of plant performance, such as efficiency and coal consumption, shift the resulting

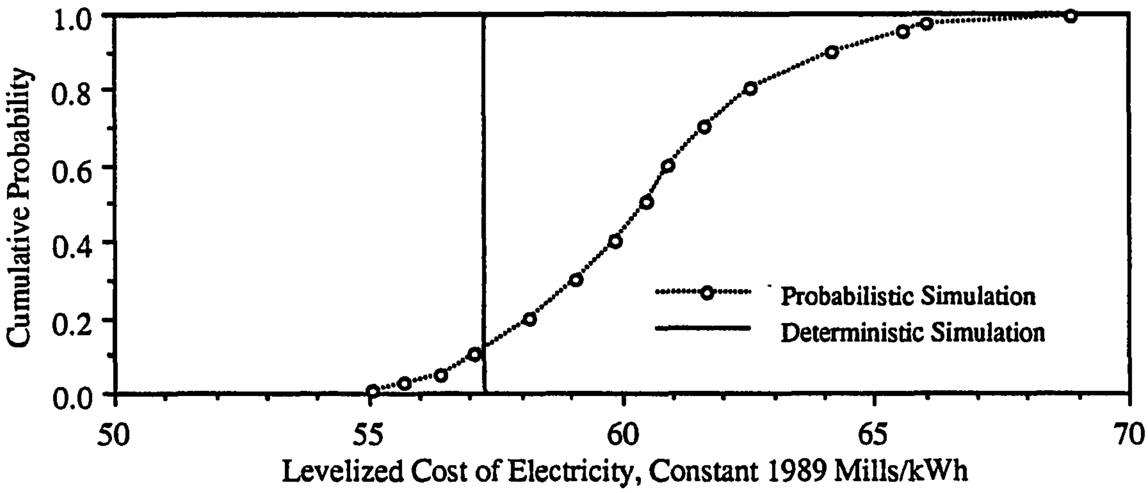

Figure 69. Comparison of Deterministic and Probabilistic Results for the Cost of Electricity of the Oxygen-blown KRW-based System.

capital cost uncertainty toward higher values than the "best guess." Thus, the interactions among performance and cost uncertainties, considered here, are shown to have important implications for capital cost.

The difference between the deterministic and probabilistic estimates of cost are more pronounced for the levelized cost of electricity, as shown in Figure 69. Recall that while typical cost estimating practices include capital cost contingency factors, there is no accepted systematic notion of contingencies with respect to fixed and variable operating costs. Here, the possibility of additional maintenance cost requirements for the gas turbine process area is represented by a positively skewed distribution for maintenance cost. Performance uncertainties play a key role also. The negative skewness of the carbon conversion rate leads to positive skewness in consumable requirements such as fuel (coal) and process water, and in the ash disposal rate. Furthermore, the unit costs associated with both ash disposal and byproduct recovery were assumed to have skewed distributions. In the case of ash disposal, it was assumed that costs could go up, but not down, due to increasingly stringent landfill requirements and associated difficulty in siting and complying with regulations. In the case of byproduct sale price, a negative skewness was assumed, representing the likelihood that market conditions at any given location in the U.S. may not be favorable to obtaining the maximum world market price.

The interactions among uncertainties in performance, capital cost, maintenance cost, and unit cost uncertainties result in the difference between the deterministic and probabilistic estimates for cost of electricity. Here, the deterministic estimate has an associated 90 percent probability of cost overrun. Furthermore, while the cost of electricity could be perhaps 2 mills/kWh less than the "best guess," it could be over 10 mills/kWh higher.

6.3.2 Identifying Key Uncertainties

As for the previous case study of the Lurgi-based IGCC system, the key uncertainties in the oxygen-blown KRW system were identified using regression analysis and probabilistic sensitivity analysis. These techniques were used to identify uncertainties which were unimportant in influencing uncertainties in key output variables. A screening analysis was then done to confirm that the removal of the unimportant uncertainties would not significantly change the results.

6.3.2.1 Regression Analysis

The results of a regression analysis using partial correlation coefficients are shown in Table 30. Three key measures of plant performance and cost were selected for detailed analysis. These are plant efficiency, total capital cost, and the cost of electricity. Other output variables that were considered, but not shown in the table, include SO_2, NO_x, and CO emissions.

The output analysis indicated that only one performance parameter was significantly correlated with uncertainty in plant efficiency. Uncertainty in carbon conversion dominated the other uncertain parameters which were expected to affect plant efficiency. Because the variance of the carbon conversion uncertainty was large relative to the uncertainties in gasifier reagent feed ratios (oxygen/carbon, steam/oxygen), the ratios had little effect on the uncertainty in plant efficiency.

For both the total capital cost and cost of electricity, uncertainty in several of the cost model parameters (e.g., project cost uncertainty, indirect construction cost, engineering and home office fees) were found to be influential. Total capital cost uncertainty was found to have significant linear correlation to direct cost uncertainties in several process areas and also to performance uncertainties that affect vessel sizing. An example of the latter is the oxygen/carbon ratio, which affects the capacity requirement for the air separation plant. Several regression model standard errors were found to be significant, indicating that addition regression model development should be focused on the air separation plant, HRSG, steam turbine, and coal handling direct cost models. The standard error of the air separation auxiliary power model was also found to be significantly correlated with total capital cost.

The uncertainty in the cost of electricity is significantly correlated with uncertainties in performance, capital cost, maintenance cost, and unit costs, as well as with several

Table 30. Key Uncertainties for Oxygen-Blown KRW-based IGCC Based on Partial Correlation Coefficients.

Partial Correlation Coefficients		
Efficiency	TCC	COE
Carbon Conv. (.99)	Project Unc. (.99)	Project Unc. (.99)
--	Gasifier DC (.95)	Carbon Conv. (-.98)
--	SE Gasification (.93)	Gasifier DC (.96)
--	Indirect Const. (.92)	Gasifier Maint (.94)
--	SE Air Separation (.89)	SE Gasification (.94)
--	SE HRSG (.89)	Indirect Const. (.92)
--	Carbon Conv. (-.89)	Engr & Home (.89)
--	Gas Turbine DC (.89)	Gas Turbine DC (.88)
--	Engr & Home Fees (.89)	SE Air Sep. (.88)
--	SE Steam Turbine (.85)	SE HRSG (.87)
--	SE Coal Handling (.82)	Steam/Oxygen (.86)
--	Steam/Oxygen Ratio (.76)	SE Steam Turb. (.83)
--	General Facilities DC (.66)	SE Coal Hd DC (.82)
--	Oxygen/Carbon (.61)	SE Air Sep. Pwr (.71)
--	SE Air Sep. Power (.61)	Oxygen/Carbon (.68)
--	SE Selexol DC (.57)	General Fac. DC (.63)
--	Air Separation DC (.53)	SE Selexol (.56)
--	--	Gas Turb. Maint (.52)
--	--	Sulfur Price (-.52)
--	--	Air Sep. DC (.50)
--	--	Ash Disp. Cost (.40)
--	--	SE Coal Hd Pwr (.33)

regression model standard error terms. Thus, interactions among among many uncertainties involving multiple aspects of performance and cost are shown to have an important influence on uncertainty in the cost of electricity.

6.3.2.2 Probabilistic Sensitivity Analysis

An example of probabilistic sensitivity analysis is shown in Figure 70 to illustrate the interaction among uncertainties in different groups of parameters in the performance and cost model. Uncertainty in performance only leads to a 90 percent probability range in the

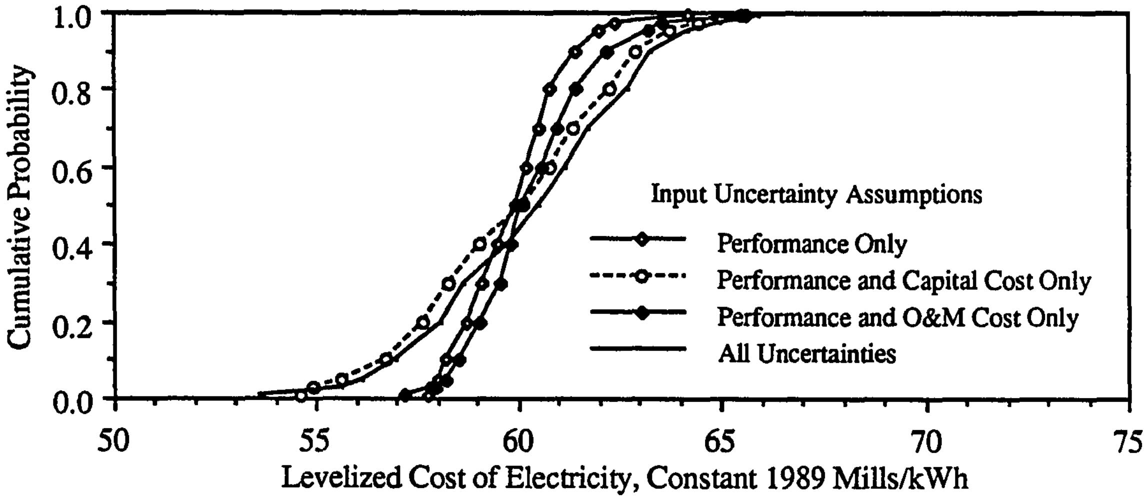

Figure 70. Sources of Uncertainty for Cost of Electricity.

cost of electricity from 58 to 62 mills/kWh, a range of 4 mills/kWh. When uncertainties in capital cost parameters, including process area direct costs and indirect capital costs, are considered in addition, the 90 percent range encompasses the values between 55.6 and 63.7 mills/kWh, a range of 8.1 mills/kWh. Because the uncertainties in capital cost are symmetric, the mean of the uncertainty for the cost of electricity remains near 59.9 mills/kWh for both cases.

However, the skewness of some of the assumptions regarding the unit costs of consumables, and regarding several maintenance cost factors, results in a shift in the central value of the uncertainty in the cost of electricity when uncertainties in performance and operating cost are considered. The mean increases by 0.4 mills/kWh. The simultaneous interactions among all uncertainties in performance and cost result in a 90 percent probability range for the cost of electricity extending from 56.2 to 64.2 mills/kWh. This range has the same magnitude as the uncertainty resulting from interaction between performance and capital cost; however, it is shifted upward in value due to the uncertainties in operating costs.

6.3.2.3 Screening Analysis

Based on the regression analysis and sensitivity analysis, a set of 16 uncertainties were preliminarily identified as being unimportant to the results of the probabilistic simulation. These uncertainties are listed in Table 31. A sensitivity case study was then run to compare the results obtained with a screened set of uncertainties to those from the original set of uncertainties. In the screening study, the uncertainties shown in Table 31 were assigned their deterministic "best guess" values.

Table 31. Uncertainties Screened Out of Case Studies for Oxygen-Blown KRW-based IGCC System.

Gas Turbine CO Conversion
Coal Handling Direct Capital Cost
Beavon-Stretford Direct Capital Cost
Process Condensate Direct Capital Cost
Standard Error KRW Gasification Power
Standard Error Selexol Power
Low Temperature Gas Cooling Direct Capital Cost
Selexol Direct Capital Cost
Claus Plant Direct Capital Cost
HRSG Direct Capital Cost
Steam Turbine Direct Capital Cost
Process Condensate Maintenance Cost
Selexol Maintenance Cost
Low Temperature Gas Cooling Maintenance Cost
Claus Plant Maintenance Cost
Operator Labor Rate

Several output variables were analyzed to compare the results from the original and screened uncertainty cases. The comparison for the cost of electricity is shown in Figure 71. There is little difference between the two cases. Thus, the uncertainties screened out of the case studies need not be the subject of any further study.

6.3.3 Additional Research

As discussed for previous case studies, additional research would be likely to result in a reduction in the variance in the uncertainties in specific model parameters. Therefore, an evaluation of the results of additional research may be based on probabilistic sensitivity analysis, in which the range of uncertainty in selected parameters is reduced. An illustrative set of assumptions for reductions in uncertainty for the oxygen-blown KRW system are shown in Table 32. As for the case study of the Lurgi-based system, a nominal 50 percent reduction in the range above and below the central value of each distribution is assumed. In the case of carbon conversion, the reduction in uncertainty is one-sided, because the mode of the distribution is at the upper end-point.

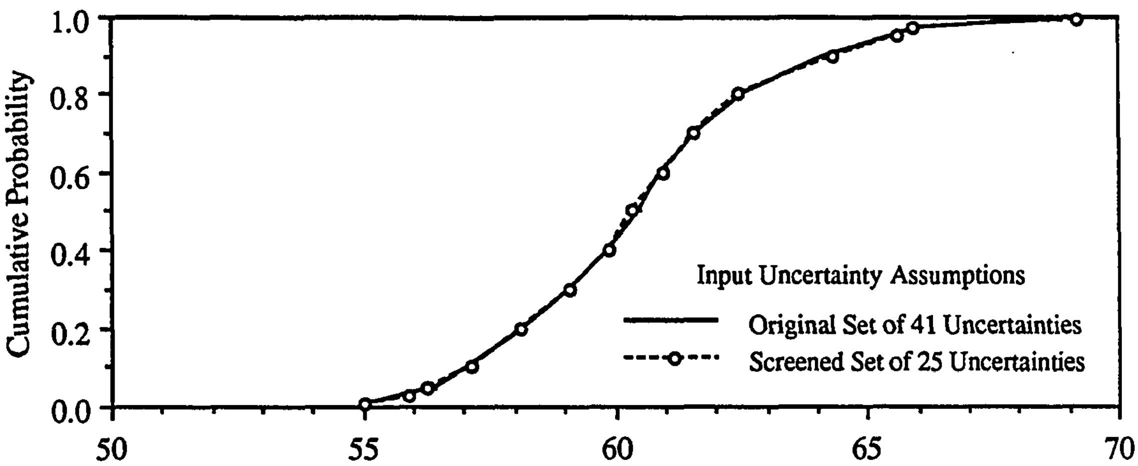

Figure 71. Comparison of Uncertainty in the Cost of Electricity for Original and Screened Sets of Uncertainties.

Table 32. Illustrative Assumptions Regarding Reduction in Uncertainty in Key Process Areas for the Oxygen-blown KRW-based System

Description	Units	Distribution	Base Case Uncertainty Ranges	Reduced Case Uncertainty Ranges
GASIFIER PROCESS AREA				
Overall Carbon Conversion	wt-% of feed coal carbon	Triangular	75 to 95 (95)	85 to 95 (95)
Oxygen/Carbon Ratio	lbmole O_2/C	Uniform	0.33 to 0.35	0.335 to 0.345
Steam/Oxygen Ratio	lbmole H_2O/O_2	Uniform	1.1 to 1.6	1.225 to 1.475
Sulfur Retention in Bottom Ash	mol-% of inlet sulfur	Triangular	10 to 20 (15)	12.5 to 17.5 (15)
Direct Cost	% of DC	Triangular	0 to 40 (20)	10 to 30 (20)
Maintenance Cost	% of TC	Triangular	3 to 6 (4.5)	3.75 to 5.25 (4.5)
GAS TURBINE PROCESS AREA				
Direct Cost	% of DC	Uniform	0 to 25	6.25 to 18.75
Maintenance Cost	% of TC	Triangular	1.5 to 2.5 (2)	1.75 to 2.25 (2)

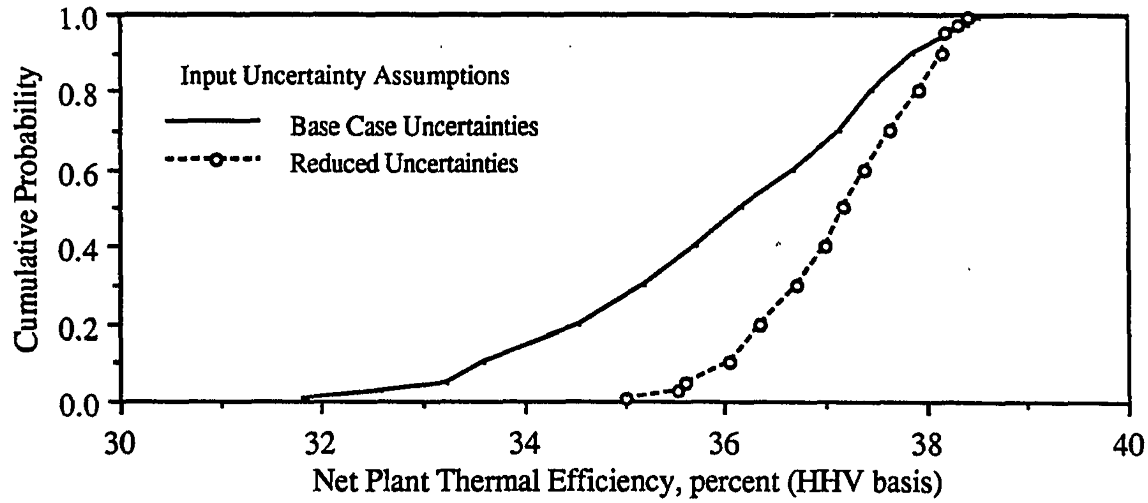

Figure 72. Effect of Reductions in Selected Input Uncertainties on Uncertainty in the Net Plant Thermal Efficiency.

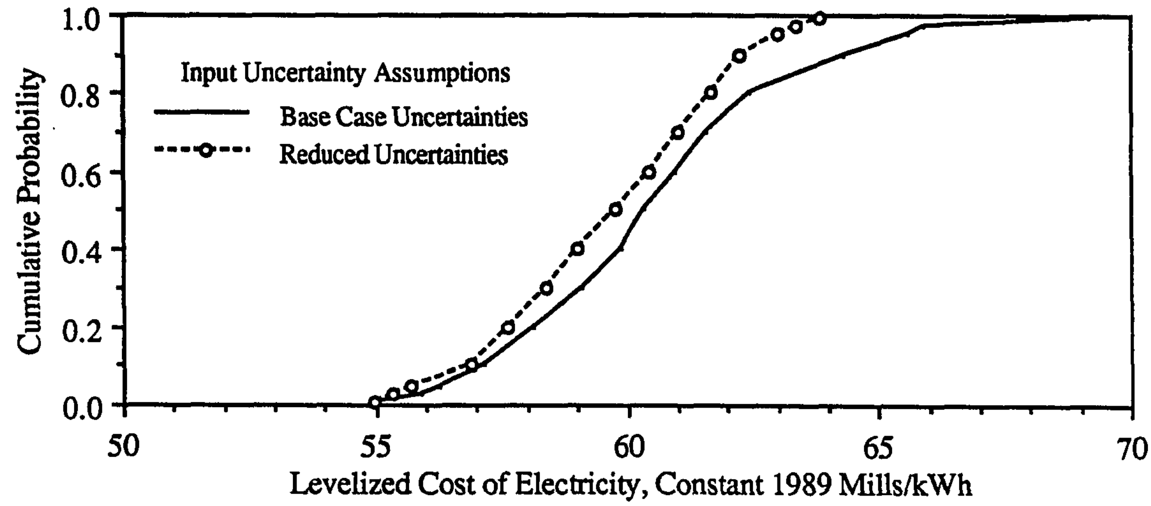

Figure 73. Effect of Reductions in Selected Input Uncertainties on Uncertainty in the Cost of Electricity.

The effects of these assumptions on the uncertainty in plant thermal efficiency and the cost of electricity are shown in Figures 72 and 73, respectively. As expected, the uncertainty in the plant thermal efficiency responded directly to the change in the uncertainty in carbon conversion. The range of uncertainty, and particularly the risk of high costs, is reduced for the cost of electricity.

6.3.4 Correlation Structures

A probabilistic sensitivity analysis was made to identify the effect that a possible correlation between the carbon conversion rate and steam/oxygen ratio would have on modeling results. A correlation coefficient of 0.75 was assumed. The comparison of the correlated and uncorrelated cases for the cost of electricity is given in Figure 74. The two results are statistically indistinguishable.

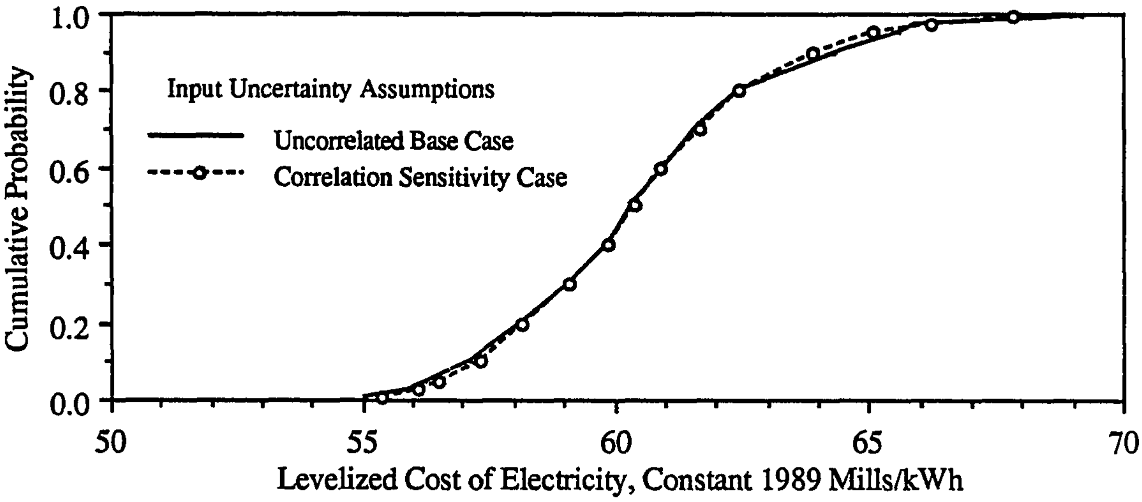

Figure 74. Effect of Correlation in a Selected Input Uncertainty on Uncertainty in the Cost of Electricity.

6.4 Probabilistic Analysis of the Air-Blown KRW-Based IGCC System with Hot Gas Cleanup

For the probabilistic case study of the air-blown KRW-based IGCC system, uncertainties in selected output variables were characterized, the key model input uncertainties were identified, the potential pay-offs from further process research were considered, and a possible correlation structure among input uncertainties was evaluated.

6.4.1 Characterization of Uncertainties in Performance and Cost

A summary of the deterministic and probabilistic results for key performance, emissions, and cost variables is given in Table 33. The analyses are based on the parameter values given in Table 15 in Chapter 4. For many of the results, the deterministic, median, and mean values are similar, indicating that uncertainties in this process are not strongly skewed, as for some of the previous case studies. A few of the results are discussed here briefly.

The uncertainty in the plant thermal efficiency is shown in Figure 75, and it is compared to the deterministic estimate. The uncertainty in efficiency covers a 90 percent probability range of less than 2 percentage points, and the mean, median, and deterministic values approximately coincide. The distribution is slightly skewed toward lower values. This result is expected due to the negative skewness of the uncertainty in carbon conversion.

Table 33. Summary of Results from Deterministic and Probabilistic Simulations of a 730 MW Air-Blown KRW-based IGCC System with Hot Gas Cleanup.[a]

Parameter[b]	Units[c]	"Best Guess"[d]	$f_{.50}$	μ	σ	$f_{.05}$ - $f_{.95}$
Plant Performance						
Thermal Efficiency	%, HHV	41.1	41.1	41.0	0.5	40.1 - 41.8
Coal Consumption	lb/kWh	0.739	0.739	0.741	0.009	0.727 - 0.758
Process Water Consump.	lb/kWh	0.727	0.738	0.740	0.016	0.716 - 0.771
ZF Sorbent Charge	10^6 lb	4.57	4.63	4.63	0.104	4.47 - 4.82
Byproduct	N/A					
Plant Discharges						
SO_2 Emissions	lb/MMBtu	0.013	0.014	0.014	0.001	0.013 - 0.016
NO_x Emissions	lb/MMBtu	0.714	0.507	0.487	0.137	0.269 - 0.714
CO Emissions	lb/kWh	0.005	0.005	0.005	0.003	0.004 - 0.009
CO_2 Emissions	lb/kWh	1.71	1.71	1.71	0.024	1.68 - 1.75
Solid Waste	lb/kWh	0.228	0.228	0.227	0.012	0.205 - 0.247
Plant Costs						
Total Capital Cost	$/kW	1,381	1375	1,376	75	1,251 - 1,516
Fixed Operating Cost	$/kW-yr	45.1	48.2	48.9	4.2	42.6 - 56.6
Variable Operating	mills/kWh	19.4	20.2	20.2	0.6	19.3 - 21.3
Coal		15.2	15.2	15.2	0.6	15.0 - 15.6
Byproduct	N/A					
Other		4.3	4.9	4.9	0.5	4.2 - 5.9
Cost of Electricity	mills/kWh	52.5	53.8	53.8	2.0	50.3 - 57.1

[a] The notation in the table heading is defined as follows: f_n = n^{th} fractile ($f_{.50}$ = median), μ = mean; and σ = standard deviation of the probability distribution. The range enclosed by $f_{.05}$ to $f_{.95}$ is the 90 percent probability range. All costs are January 1989 dollars.

[b] Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, and zinc ferrite steam. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

[c] HHV = higher heating value; MMBtu = million Btu.

[d] Based on a deterministic simulation in which median or modal values of uncertain variables are assumed as "best guess" inputs to the model

The median NO_x emission rate is lower than the deterministic estimate (see Figure 76). As with the Lurgi-based system, this result is obtained because of the negative skewness of the uncertainty in the conversion rate of fuel-bound nitrogen (ammonia) to NO_x in the gas turbine combustor. Although the median is shifted downward, the uncertainty in NO_x emissions is only slightlv negatively skewed, because the uncertainty in the ammonia yield from the gasifier is approximately symmetric.

In Figure 77, the uncertainty in total capital cost is compared to the deterministic estimate. Unlike previous case studies, the deterministic estimate, which includes process and project contingency factors, coincides with the median value of the probabilistic

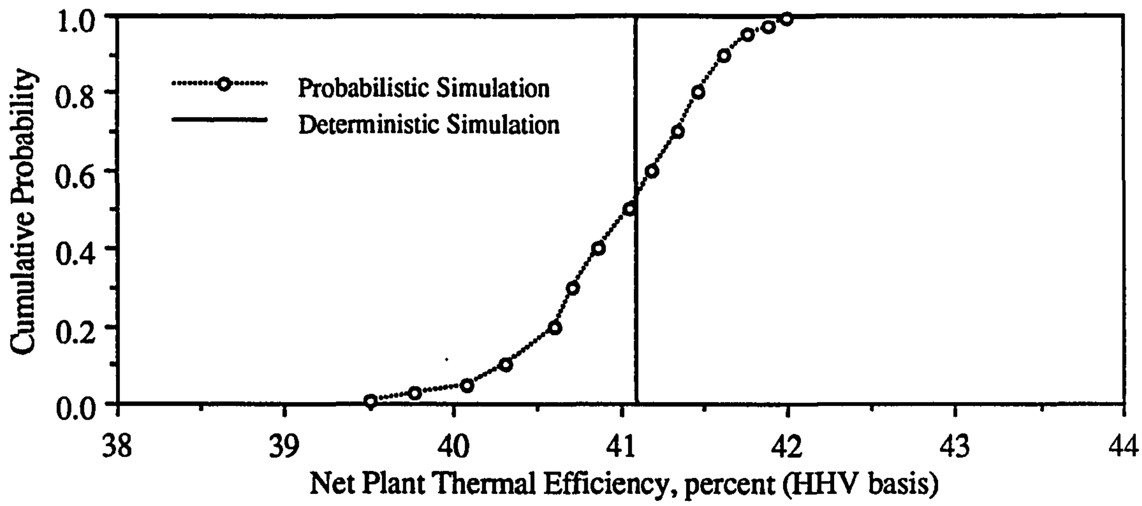

Figure 75. Comparison of Deterministic and Probabilistic Results for the Net Plant Thermal Efficiency of the Air-blown KRW-based System.

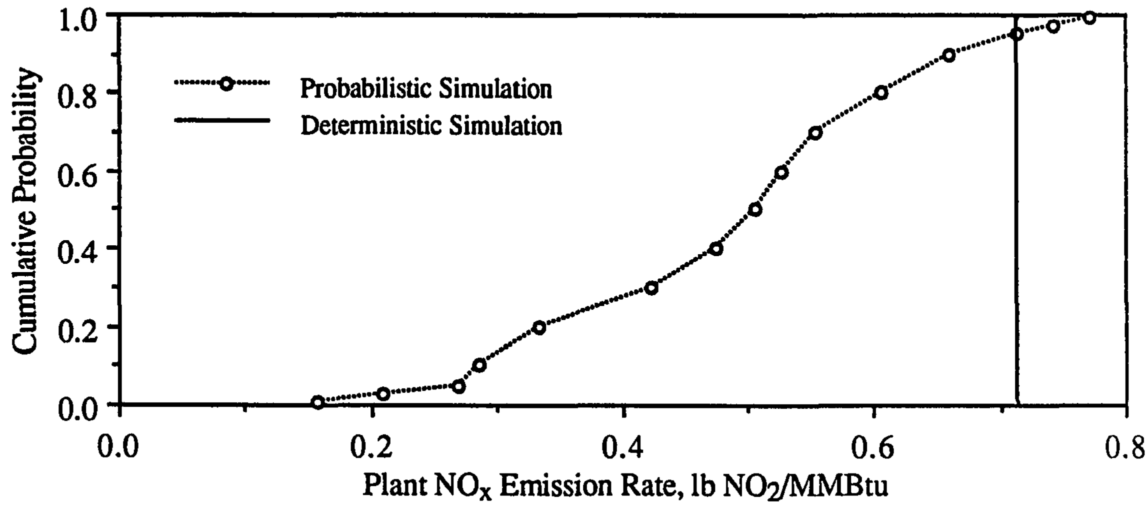

Figure 76. Comparison of Deterministic and Probabilistic Results for the NO_x Emission Rate of the Air-blown KRW-based System.

simulation. Thus, there is a 50 percent chance of cost overrun associated with the deterministic estimate of $1,380/kW. Because the performance parameter uncertainties were symmetric or only moderately skewed, and because all of the cost related uncertain parameters affecting capital cost were assumed to be symmetrically distributed, the uncertainty in capital cost is approximately symmetric. The 90 percent probability range for capital cost is $255/kW.

In spite of the agreement between the deterministic and probabilistic results for capital cost, the two analyses do not agree with respect to the cost of electricity, as seen in Figure 78. There is a 75 percent probability that the cost will be higher than the deterministic estimate. In the probabilistic analysis, the uncertainties in the maintenance

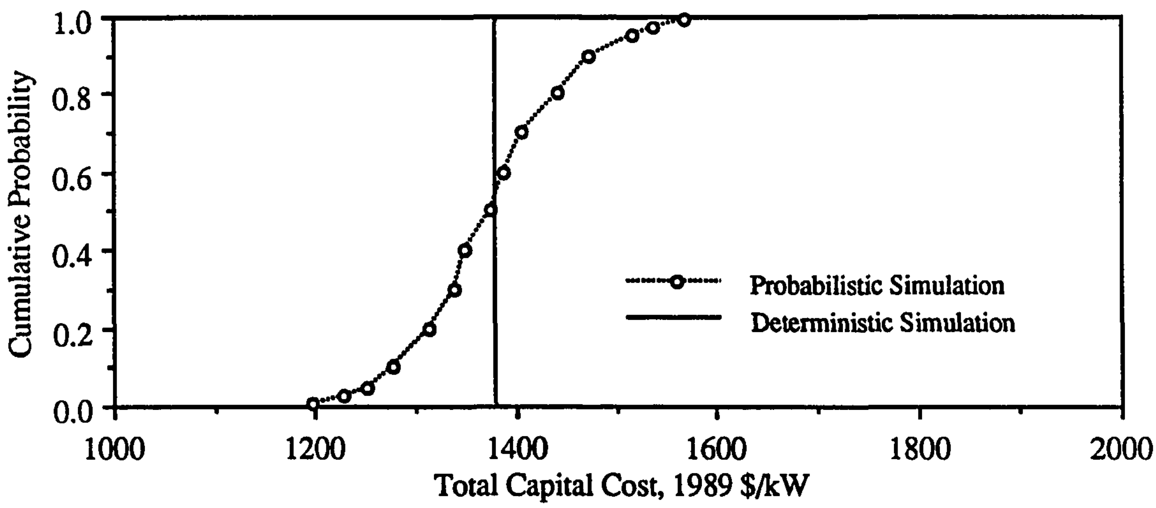

Figure 77. Comparison of Deterministic and Probabilistic Results for the Total Capital Cost of the Air-blown KRW-based System.

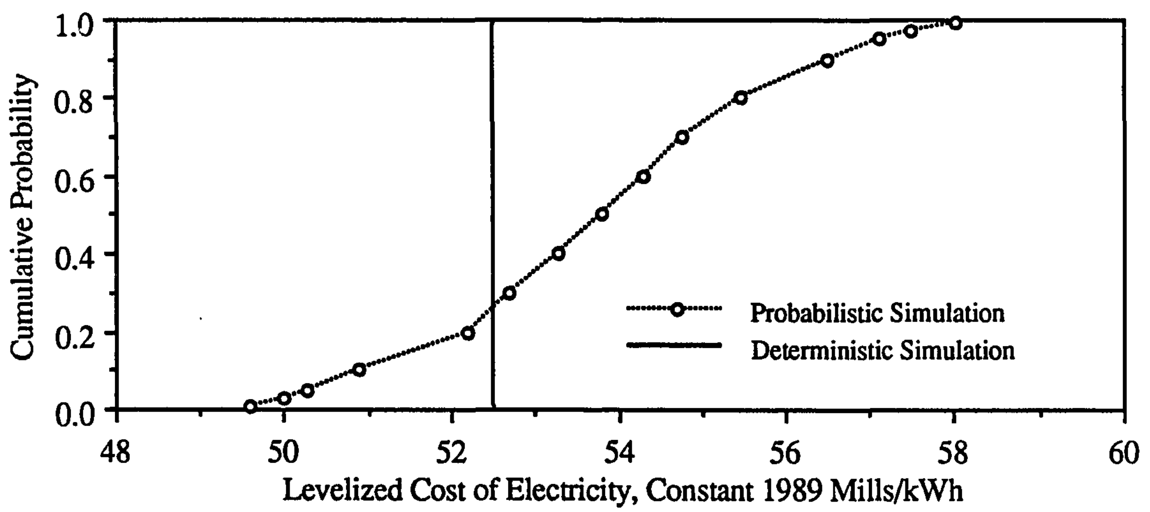

Figure 78. Comparison of Deterministic and Probabilistic Results for the Cost of Electricity of the Air-blown KRW-based System.

cost of the gas turbine, zinc ferrite, and sulfation process areas were assumed to be positively skewed. Also, the unit costs of limestone and ash disposal were assumed to be positively skewed. These assumptions affect fixed and variable operating cost and, in turn, the cost of electricity.

6.4.2 Identifying Key Uncertainties

The key uncertainties in the air-blown KRW-based case study are identified using regression analysis and probabilistic sensitivity analysis. The results are confirmed using a probabilistic screening analysis.

Table 34. Key Uncertainties for Air-Blown KRW-based IGCC Based on Partial Correlation Coefficients.

Partial Correlation Coefficients		
Efficiency	TCC	COE
Gasifier Carbon Conv. (.99)	Project Unc. (.99)	Project Unc. (.99)
CaS Sulfation Rate (.99)	Gas Turbine DC (.99)	Gas Turbine DC (-.98)
Oxygen/Carbon Ratio (-.99)	Indirect Const. (.98)	Gasification DC (.98)
Gasifier Temperature (.97)	Gasification DC (.98)	Gas Turbine Maint. (.98)
Gas Turbine CO Conv (.92)	SE HRSG (.98)	Indirect Const. (.98)
Ca/S Ratio (.64)	SE Steam Turbine (.97)	SE HRSG (.97)
Sulfator Carbon Conv. (.55)	Engr & Home Fees (.96)	Ash Disposal Cost (.97)
Gasifier NH_3 Yield (-.39)	Zinc Ferrite DC (.86)	SE Steam Turbine (.96)
--	General Facilities DC (.86)	Gasifier Maintenance (.96)
--	Zinc Fer. Unit Cost (.83)	Engr & Home Office (.96)
--	Sulfation DC (.82)	Gasif. Carbon Conv. (-.96)
--	CaS Sulfation Rate (-.77)	Ca/S Ratio (.95)
--	Gasifier Carbon Conv. (-.75)	CaS Sulf. Rate (-.93)
--	Ca/S Ratio (.73)	Zinc Ferrite Unit Cost (.89)
--	Oxygen/Carbon Ratio (.70)	Oxygen/Carbon Ratio (.88)
--	Coal Handling DC (.47)	Zinc Ferrite DC (.86)
--	Steam Turbine DC (.37)	Limestone Unit Cost (.85)
--	--	Sulfation DC (.82)
--	--	General Facilities DC (.80)
--	--	Sulfation Maintenance (.78)
--	--	Gasifier Sulfur Cap. (-.75)
--	--	Gasifier Temperature (-.62)

6.4.2.1 Regression Analysis

The key input uncertainties resulting in uncertainty in plant efficiency, total capital cost, and the cost of electricity are shown in Table 34. These results are based on sample PCC analysis.

The plant efficiency is most strongly influenced by uncertainty in the gasifier carbon conversion efficiency. However, uncertainties related to the sulfation unit are also significantly correlated with efficiency. In the air-blown KRW system, limestone is used as a sorbent to remove sulfur during gasification. Because of the reducing atmosphere in

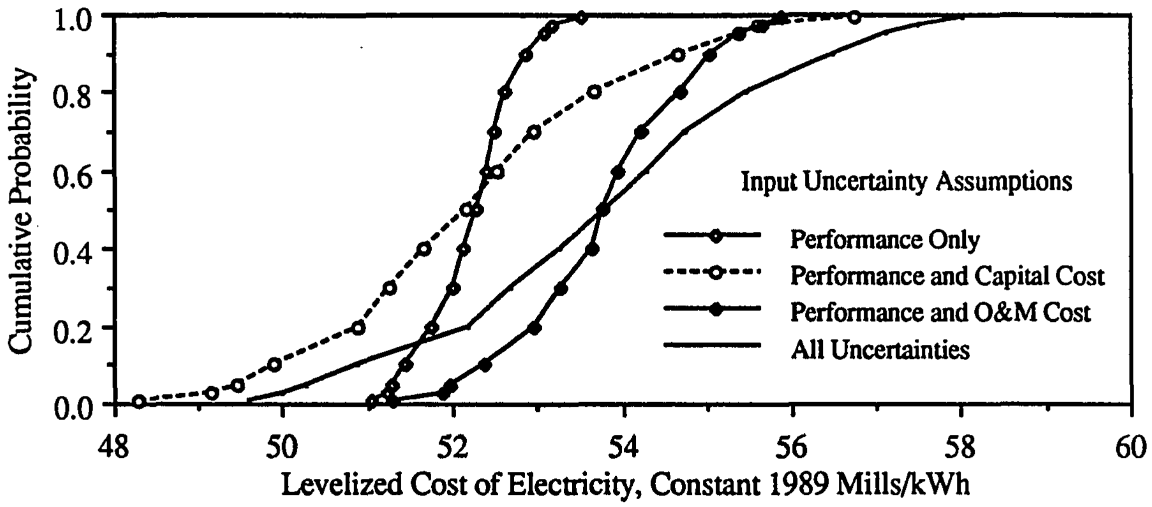

Figure 79. Sources of Uncertainty for Cost of Electricity.

the gasifier, the spent sorbent contains sulfur in the form of calcium sulfide. The calcium sulfide must oxidized to calcium sulfate in a fluidized bed boiler prior to landfilling the spent sorbent. A high sulfide concentration would result in a solid waste which would be classified as hazardous under RCRA. However, the conversion rate of calcium sulfide to calcium sulfate in the boiler is uncertain. The energy released from this exothermic reaction is used to generate steam for the plant steam cycle, and is thus recovered as an energy credit.

The total capital cost is strongly influenced by uncertainties in indirect capital costs, process area direct costs, and the error terms of two direct cost regression models. For this technology, uncertainty in capital cost is not strongly influenced by performance uncertainties. Similarly, uncertainty in the cost of electricity is influenced primarily by capital, fixed operating, and variable operating cost uncertainties, with weaker influences from performance-related uncertainties. Of the skewed uncertainties affecting the cost of electricity, the gas turbine maintenance and ash disposal costs are the most highly correlated. The gasifier maintenance and limestone costs are also significantly correlated.

6.4.2.2 Probabilistic Sensitivity Analysis

An example of a probabilistic sensitivity analysis is shown in Figure 79 for the cost of electricity. As shown in the figure, the range of uncertainty in the cost of electricity solely attributable to uncertain performance-related parameters is narrow. The 90 percent probability range attributable to performance only is less than 2 mills/kWh, from 51.2 to 53 mills/kWh. Uncertainties in capital cost are shown to substantially increase the variance in the cost of electricity without shifting the central value of the distribution significantly. The

Table 35. Uncertainties Screened Out of Case Studies for Air-Blown KRW-based IGCC System.

Gasifier Steam/Oxygen Ratio
Residual Sulfide in Zinc Ferrite After Reductive Regeneration
Zinc Ferrite Absorber Pressure Drop
Limestone Handling Direct Capital Cost
Oxidant Feed Direct Capital Cost
HRSG Direct Capital Cost
Steam Turbine Direct Capital Cost
Oxidant Feed Maintenance Cost
Zinc Ferrite Maintenance Cost
Standard Error of Oxidant Feed Direct Capital Cost Model
Standard Error of Boiler Feedwater Direct Capital Cost Model
Operator Labor Rate

90 percent probability range in this case covers 5.8 mills/kWh from 49.5 to 55.3 mills/kWh.

The interaction between uncertainty in O&M costs with performance uncertainties results in an upward shift in the central value of almost 2 mills/kWh and an increase in the variance. As seen in Figure 79, the difference between the case with uncertainties in performance and O&M costs and the "all uncertainties" case is in the variance, not the median. The difference between these two is the inclusion of the symmetric capital cost uncertainties in the "all uncertainties" case.

From the probabilistic sensitivity analysis, it is clear that performance-related uncertainties are a relatively minor component of overall uncertainty in cost for this technology. Furthermore, while the variance in the result is strongly influenced by uncertainties in capital cost, it is the uncertainties in O&M costs that are responsible for the shift in the central tendency of the distribution for the cost of electricity.

6.4.2.3 Screening Analysis

Based on the results of the regression and sensitivity analyses, a set of model input uncertainties were identified as not significantly contributing to uncertainty in key model outputs. The key output variables considered in the screening study were the same as for the previous IGCC technology case studies. A total of 12 uncertainties were screened

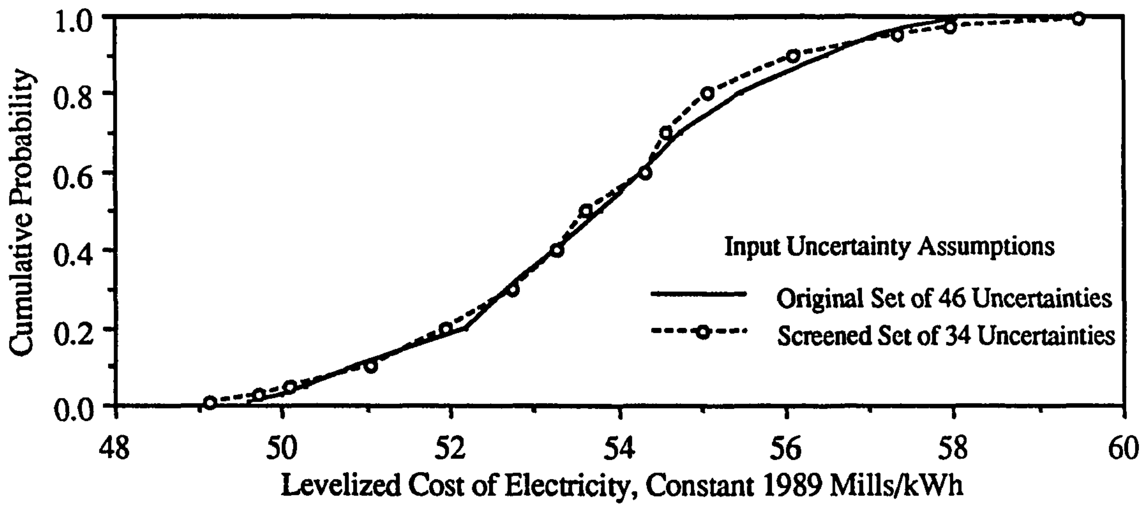

Figure 80. Comparison of Uncertainty in the Cost of Electricity for Original and Screened Sets of Uncertainties.

from further case studies, leaving 34 of the original 46 uncertain parameters. The uncertainties identified as unimportant are listed in Table 35.

The results of the screening analysis are shown graphically in Figure 80 for the cost of electricity. The results of the two cases differ slightly, but within the 95 percent confidence interval for the cdfs. Therefore, the results are statistically indistinguishable.

6.4.3 Additional Research

The effect of hypothetical reductions in the uncertainties that might be obtained from focused research in specific model parameters is the basis here for identifying the pay-off from additional research. As an illustrative case study, it is assumed that the range of uncertainties in selected performance and cost parameters for four major process areas would be reduced by 50 percent above and below their central values. The process areas included are gasification, sulfation, zinc ferrite desulfurization, and gas turbine. The base case uncertainty assumptions, from Table 15, and the reduced uncertainties are compared in Table 36.

The effect of the hypothetical research outcomes on the plant thermal efficiency is shown in Figure 81. From the regression analysis previously described, the uncertainty in the carbon conversion rate was found to be the input uncertainty most strongly correlated with plant efficiency. This uncertainty is negatively skewed. Therefore, the reduction in the range of uncertainty above and below its mode results in the asymmetric effect on uncertainty in efficiency in Figure 81.

Table 36. Illustrative Assumptions Regarding Reduction in Uncertainty in Key Process Areas for the Air-blown KRW-based System

Description	Units	Distribution	Base Case Uncertainty Ranges	Reduced Case Uncertainty Ranges
GASIFIER PROCESS AREA				
Overall Carbon Conversion	wt-% of feed coal carbon	Triangular	90 to 97 (95)	92.5 to 96 (95)
Oxygen/Carbon Ratio	lbmole O_2/C	Triangular	0.45 to 0.47 (46)	0.455 to 0.465 (46)
Sulfur Retention in Bottom Ash	mol-% of inlet sulfur	Triangular	85 to 95 (90)	87.5 to 92.5 (90)
Limestone Calcium-to-Sulfur Ratio	lbmole Ca/S	Triangular	2 to 2.8 (2.6)	2.3 to 2.7 (2.6)
Direct Cost	% of DC	Triangular	0 to 40 (20)	10 to 30 (20)
Maintenance Cost	% of TC	Triangular	3 to 6 (4.5)	3.75 to 5.25 (4.5)
SULFATION PROCESS AREA				
Conversion of CaS to $CaSO_4$	%	Uniform	30 to 60	45 to 75
Carbon Conversion	%	Triangular	90 to 98 (95)	92.5 to 96.5 (95)
Direct Cost	% of DC	Triangular	20 to 60 (40)	30 to 50 (40)
Maintenance Cost	% of TC	Triangular	3 to 6 (4)	3.5 to 5 (4)
ZINC FERRITE DESULFURIZATION PROCESS AREA				
Sorbent Sulfur Loading	wt-% S in sorbent	Normal	2.16 to 31.84 (17)	9.58 to 24.42 (17)
Sorbent Attrition Rate	wt-% sorbent loss/cycle	Fractile	5%: 0.17 to 0.34	0.585 to 0.67
			20%: 0.34 to 0.5	0.67 to 0.75
			25%: 0.5 to 1	0.75 to 1
			25%: 1 to 1.5	1 to 1.25
			20%: 1.5 to 5	1.25 to 3
			5%: 5 to 25	3 to 13
Direct Cost	% of DC	Uniform	0 to 80	20 to 60
GAS TURBINE PROCESS AREA				
Direct Cost	% of DC	Uniform	0 to 25	6.25 to 18.75
Maintenance Cost	% of TC	Triangular	1.5 to 6 (2)	1.75 to 4 (2)

The assumed reductions in uncertainty from process research have only a modest effect on the uncertainty in the cost of electricity, as shown in Figure 82. This is because many of the uncertainties that strongly affect the cost of electricity, such as project capital cost uncertainty, indirect construction costs, and the standard error of the HRSG direct cost model, would not be affected by further process research. Instead, these sources of uncertainty would be reduced by the development of detailed finalized cost estimates for a particular site-specific construction project. Other key uncertainties, such as the ash

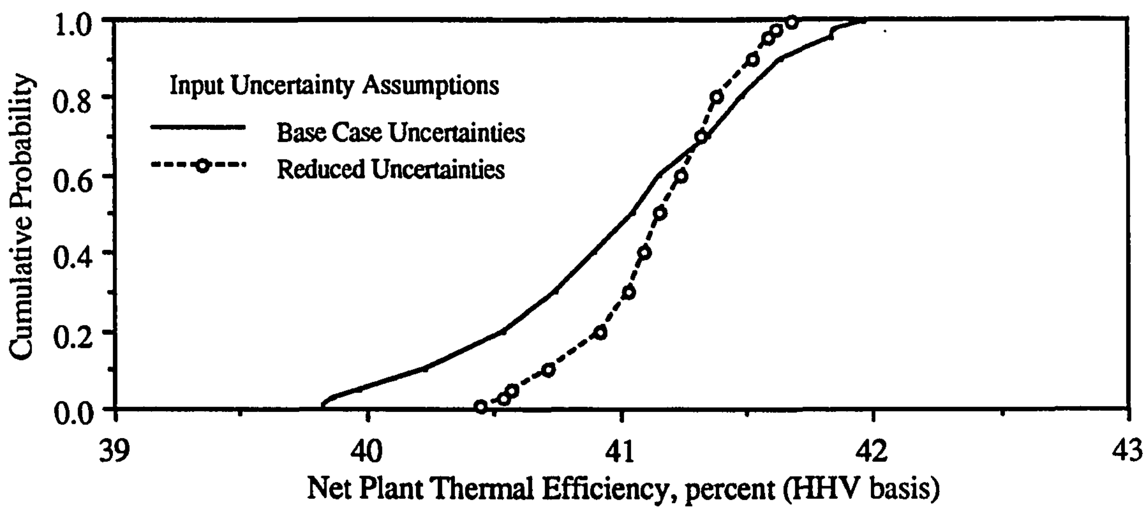

Figure 81. Effect of Reductions in Selected Input Uncertainties on Uncertainty in the Net Plant Thermal Efficiency.

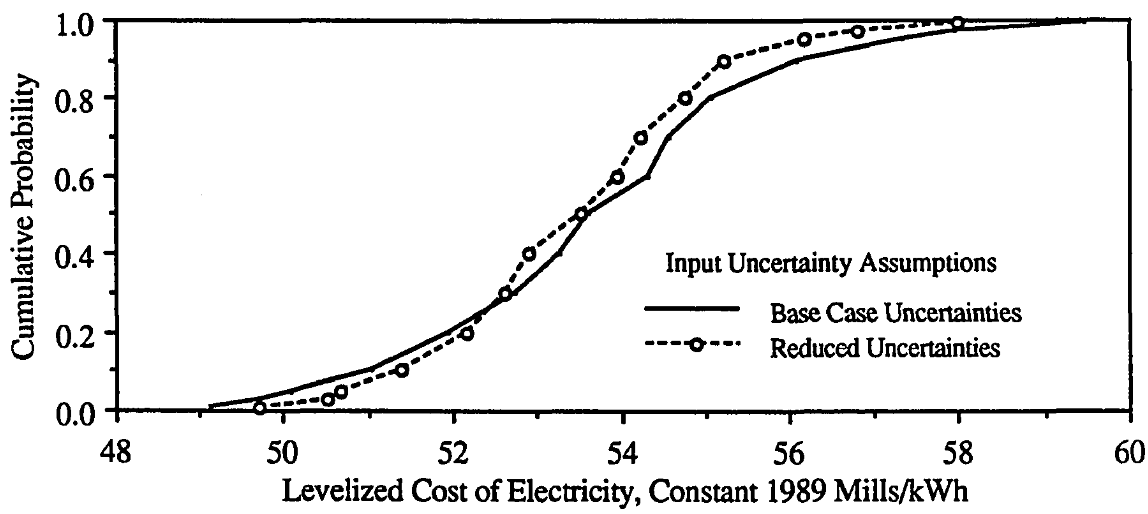

Figure 82. Effect of Reductions in Selected Input Uncertainties on Uncertainty in the Cost of Electricity.

disposal cost, are also intended to represent both uncertainty and variance from one site to another.

6.4.4 Correlation Structures

A probabilistic sensitivity case including correlations among key performance parameters in the gasification and zinc ferrite process areas was developed. The correlations for the gasification process area are discussed in Appendix B.4.5, and those for the zinc ferrite process area are discussed in Appendix B.5.3.1.

The correlation analysis is compared to the base case analysis for the cost of electricity in Figure 83. Although correlations among six performance parameters were

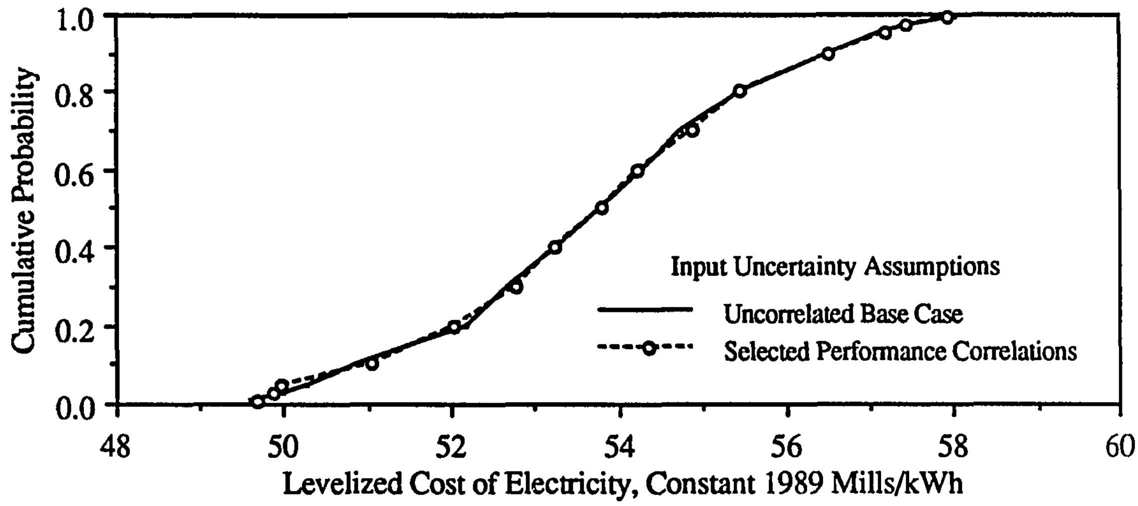

Figure 83. Effect of Correlation in Selected Input Uncertainties on Uncertainty in the Cost of Electricity.

considered, there was negligible effect on the probabilistic results. Thus, the assumed correlation structure had no significant effect on model results.

6.5 Comparative Analysis of the IGCC Systems

The preceding sections have focused on case studies of individual IGCC technologies. In this section, the three systems will be compared in the face of uncertainty. A total of three comparisons are made. These include: (1) the air-blown Lurgi versus oxygen-blown KRW systems; (2) the air-blown versus oxygen-blown KRW systems; and (3) the air-blown Lurgi versus air-blown KRW systems. These comparisons are based on key measures of plant performance, emissions, and cost.

Because the most detailed information regarding uncertainties was obtained for the Lurgi-based system, and because the oxygen-blown KRW system is most representative of conventional technology employing cold gas cleanup, the most detailed comparative case study is that of the Lurgi versus oxygen-blown KRW system.

For all three technologies, additional research is likely to reduce uncertainties in both performance and cost. Therefore, sensitivity cases based on alternative assumptions regarding process uncertainties are considered.

6.5.1 Lurgi Vs. Oxygen-Blown KRW System

For comparative analysis of the Lurgi and oxygen-blown KRW systems, the key uncertainties identified in the screening studies for the respective technologies were used. Furthermore, in cases where uncertainties are assumed to be common to both systems, the same set of sample values and ranking of values were used in the probabilistic simulation of both technologies. Comparisons between the two technologies are based on probability distributions for the *differences* in performance, emissions, and cost. Because parameters common to both systems are given the same sample values in corresponding repetitions of the probabilistic simulation, the probability distributions of the differences account for any underlying correlations between the two systems.

The pairing of uncertain parameters between the two technologies for the probabilistic comparisons is shown in Table 37. Parameters which are similar or the same between the two technologies are shown on the same line. Parameters which are assumed to have the same set and ranking of sample values between the two simulations are indicated. These parameters include the gas turbine direct cost, standard errors of regression models common to both systems, ash disposal cost, and indirect capital cost uncertainties. The gas turbine direct cost is partly a function of modifications required for application to the IGCC process environment and of market conditions. The modifications and market conditions faced by the two cases are assumed to be sufficiently similar to treat the uncertainty in capital cost as 100 percent correlated.

The regression model error terms are generic to the HRSG and steam turbine direct cost models. If a high result is obtained for one system, it would be expected for the other as well. Therefore, the error terms are assumed to be 100 percent correlated between the two cases.

Ash disposal cost depends on the plant siting. The assumption here is that either technology could be selected for the same plant site. Therefore, uncertainty in ash cost between the two systems is taken to be the same.

The uncertainty in indirect capital costs depend in part on the particular architect/engineer firm and construction team involved in building an actual plant. These uncertainties may be partly resolved by developing a more finalized, site-specific estimate of the cost of constructing a particular IGCC system at a particular site. However, a substantial portion of the uncertainty may remain unresolved until "all the bills are in." It is assumed here that, regardless of which type of IGCC system is constructed, the same uncertainties regarding indirect costs are faced.

Table 37. Pairing of Uncertain Parameters for Comparative Study of Oxygen-Blown KRW- and Air-Blown Lurgi-based IGCC Systems.

Description of Uncertain Parameter[a]		
Oxygen-blown KRW (Case OKC)	Air-blown Lurgi (Case ALH)	Correlated[b]
	Fines Carryover	
	Fines Capture	
	Carbon Retention in Ash	
	Gasifier Coal Throughput	
	Gasifier Ammonia Yield	
Carbon Conversion		
Oxygen/Carbon Ratio	Air/Coal Ratio	No
Steam/Oxygen Ratio	Steam/Coal Ratio	No
Sulfur Retention in Ash		
	Zinc Ferrite Residual Sulfate	
	Zinc Ferrite Sorbent Loading	
	Zinc Ferrite Sorbent Attrition	
	Fuel NO_x Conversion	
	Gas Turbine CO Conversion	
Thermal NO_x		
SE Air Separation Aux. Power		
SE Coal Hndg Aux. Power		
Air Separation DC		
Gasification DC	Gasification DC	No
	Zinc Ferrite DC	
	Sulfuric Acid Plant DC	
Gas Turbine DC	Gas Turbine DC	Yes
General Facilities DC		
SE Coal Hndg DC	SE Coal Hndg DC	No
SE Air Separation DC		
SE Gasification DC		
SE Selexol DC		
SE HRSG DC	SE HRSG DC	Yes
SE Steam Turbine DC	SE Steam Turbine DC	Yes
Gasification Maintenance	Gasification Maintenance	No
Gas Turbine Maintenance	Gas Turbine Maintenance	No
	Unit Cost of Zinc Ferrite Sorb.	
Ash Disposal Cost	Ash Disposal Cost	Yes
Sulfur Byproduct		
	Sulfuric Acid Byproduct	
Byproduct Marketing		
Engr & Home Office Fees	Engr & Home Office Fees	Yes
Indirect Construction Cost	Indirect Construction Cost	Yes
Project Cost Uncertainty	Project Cost Uncertainty	Yes

[a] Uncertain parameters which are analogous or the same between the two technologies are listed on the same line.

[b] The term "correlation" is used here to indicate parameters for which the same vector and ranking of samples was used in uncertainty analysis of both technologies. Parameters which are unique to one technology are not correlated with any other uncertain parameters. Parameters which are similar across technologies may be uncorrelated, as indicated, because of differences in the underlying technologies. See text for explanation of basis for correlations.

Other uncertain parameters that are similar between the systems are assumed to be uncorrelated. For example, although the performance of the two gasifiers can be characterized using similar parameters, the systems are sufficiently different that no correlations are assumed to exist among them. The direct cost of the coal handling systems are calculated using different regression models. Therefore, the standard errors for these two models are assumed to be uncorrelated. As a third example, even though gas turbine direct costs are assumed to be correlated, the maintenance costs are not. This is because the factors that would lead to high maintenance cost (e.g., alkali deposition) would not be expected to occur simultaneously for the two systems.

Based on the pairing of input uncertainties in Table 37, properly paired probabilistic simulations of both technologies were run using the ASPEN simulator. The results of the simulations for several key measures of plant performance, emissions, and cost, were then paired, sample by sample. Each pair of samples was subtracted, and the resulting set of sample differences were used to construct cdfs for the performance, emissions, or cost savings of the advanced technology compared to the conventional technology.

As described in previous sections, additional research on both the Lurgi and oxygen-blown KRW systems can be expected to reduce the uncertainties in these technologies. Reduction in the uncertainty in one or both technologies affects the probability distribution for the differences between the two. Therefore, several comparisons are made for each key variable, based on alternative combinations of base case and reduced uncertainties for the two technologies. The multiple set of comparisons provides insight into whether the advantage seen for one technology is robust when the underlying assumptions change.

For the base case uncertainties of both systems, the Lurgi-based system is likely to be superior to the oxygen-blown KRW-based system with respect to capital cost, levelized cost of electricity, net plant thermal efficiency, and coal consumption. The Lurgi system is certain to have lower SO_2 emissions. However, the KRW system is likely to have lower fixed and variable operating costs, and lower CO_2 emissions. In addition, the KRW system is certain to have lower NO_x emissions and lower water consumption. The results of comparisons for plant efficiency, capital cost, fixed operating cost, variable operating cost, and the cost of electricity are given in Table 38. The table includes the same summary statistics used for the comparison of the copper oxide and FGD/SCR systems in the previous chapter. See Section 5.6 for a discussion of the statistics used here. Several of the results are discussed here and shown graphically.

Table 38. Results of Research Information Case Studies: Comparison of Air-blown Lurgi and Oxygen-blown KRW Systems

Research Area[a]	Probability of a Loss (%)	Downward Partial Mean	Expected Value of a Loss	Expected Value of a Gain	Mean	Reduction in Risk	Value of Research
Plant Efficiency, percent							
Baseline	20	0.25	1.2	2.2	1.5	- -	- -
Gasification	9	0.06	0.6	2.3	2.0	0.19	0.5
Zinc Ferrite Desulfurization	20	0.25	1.2	2.2	1.5	0	0
All	9	0.06	0.6	2.3	2.0	0.19	0.5
Base Case ALH vs. All OKC	35	0.40	1.1	1.2	0.4	- -	- -
All ALH vs. All OKC	15	0.09	0.6	1.2	0.9	0.31	0.5
Total Capital Cost, 1989 $/kW							
Baseline	1	1.4	144	342	337	- -	- -
Gasification	0	0	0	371	371	1.4	34
Zinc Ferrite Desulfurization	1	1.3	132	353	348	0.1	11
All	0	0	0	383	383	1.4	46
Base Case ALH vs. All OKC	1	1.7	172	331	331	- -	- -
All ALH vs. All OKC	0	0	0	372	372	1.7	41
Fixed Operating Cost, 1989 $/kW-yr							
Baseline	55	5.8	10.5	6.7	-2.7	- -	- -
Gasification	32	1.4	4.3	6.9	3.3	4.4	6.0
Zinc Ferrite Desulfurization	55	5.7	10.5	6.8	-2.6	0.1	0.1
All	16	0.5	3.7	7.2	5.4	5.3	8.1
Base Case ALH vs. All OKC	51	5.3	10.4	5.7	-2.5	- -	- -
All ALH vs. All OKC	15	0.4	2.4	7.1	5.7	4.9	8.2
Variable Operating Cost, 1989 mills/kWh							
Baseline	82	4.3	5.3	0.8	-4.2	- -	- -
Gasification	77	4.0	5.2	0.8	-3.8	0.3	0.4
Zinc Ferrite Desulfurization	87	2.8	3.2	0.8	-2.7	1.5	1.5
All	83	2.5	3.0	0.8	-2.3	1.8	1.9
Base Case ALH vs. All OKC	92	4.7	5.1	0.5	-4.7	- -	- -
All ALH vs. All OKC	94	2.9	3.1	0.4	-2.9	1.8	1.8
Levelized Cost of Electricity, Constant 1989 mills/kWh							
Baseline	27	2.6	9.5	5.5	1.5	- -	- -
Gasification	16	1.9	11.7	6.4	3.5	0.7	2.0
Zinc Ferrite Desulfurization	21	1.0	5.0	5.4	3.2	1.6	1.7
All	6	0.4	6.8	6.4	5.6	2.2	4.1
Base Case ALH vs. All OKC	32	2.7	8.4	5.1	0.8	- -	- -
All ALH vs. All OKC	9	0.4	4.9	5.9	4.9	2.3	4.1

[a] For each parameter used for comparison, the results are grouped separately for comparisons of the air-blown Lurgi system with hot gas cleanup (ALH) to the base case and "All" reduced uncertainties case for the oxygen-blown KRW-based system with cold gas cleanup (OKC).

The higher efficiency and lower capital cost of the Lurgi system is expected, because it is a "simplified" IGCC system featuring hot gas cleanup. Therefore, it does not have the expensive fuel gas cooling and cleanup equipment associated with cold gas cleanup in the KRW system, leading to substantial capital cost savings. The mean capital

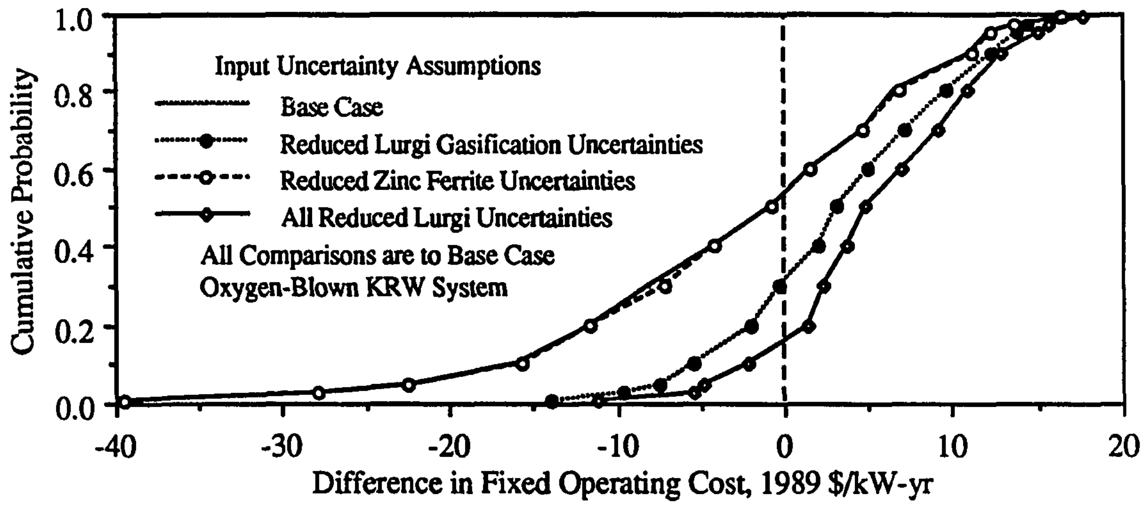

Figure 84. Effect of Illustrative Research Outcomes on Fixed Operating Cost Savings for Air-blown Lurgi System Compared to Oxygen-blown KRW System

cost savings is $337/kW, for the base case uncertainties. For this case, there is a 46 percent correlation between the capital costs of the two systems, due to the assumed pairing of input uncertainties between them. Additional research on the Lurgi system is likely to improve the cost savings. For the assumptions used in this study, the capital cost difference increases to $383/kW. However, research may simultaneously reduce uncertainties in the KRW system. The cost savings for this scenario is relatively unchanged, at $372/kW. Thus, regardless of whether research is conducted on the KRW system, research on the Lurgi system is expected to yield both a reduction in the downside risk that capital cost could be more expensive, and an increase in the mean cost savings.

However, the Lurgi system is more risky from a maintenance standpoint. Cold gas cleanup will remove many of the trace contaminants that result in uncertainties for the hot gas cleanup system. For example, alkali and particulate matter are removed to a high degree in cold gas cleanup. However, the performance of high efficiency cyclones may not be as good for alkali and particulate control. Therefore, downstream equipment, such as the zinc ferrite process and gas turbine, may be faced with high maintenance costs.

The probability differences for the fixed operating cost for several alternative assumptions regarding Lurgi system uncertainties is shown in Figure 84. All of 'he comparisons in the figure are based on the base case uncertainty assumptions for the KRW system. Based on current understanding of uncertainty for the Lurgi system, there is a 55 percent probability that the Lurgi system will have higher fixed operating costs than the KRW system. Furthermore, the conditional expected value of higher fixed operating costs

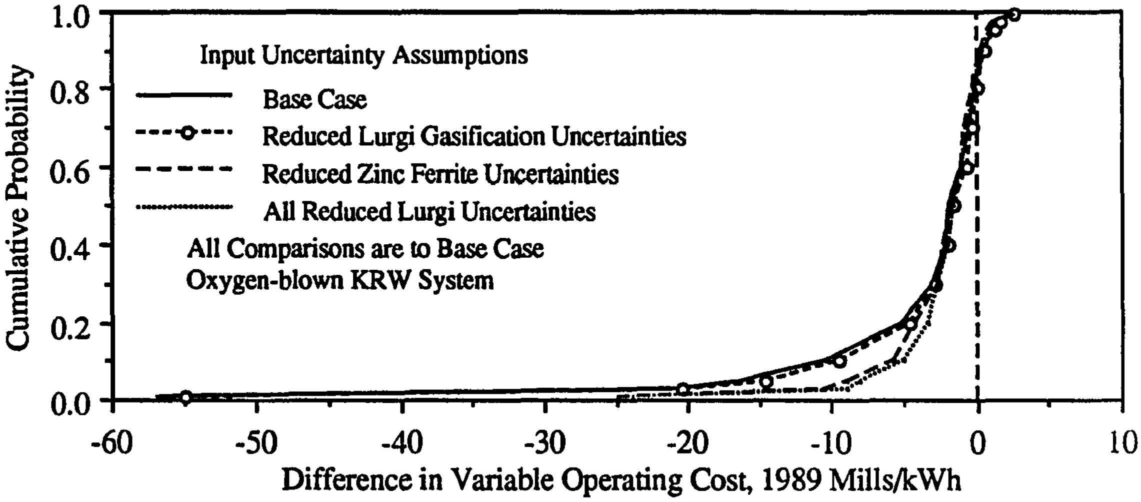

Figure 85. Effect of Illustrative Research Outcomes on Variable Operating Cost Savings for Air-blown Lurgi System Compared to Oxygen-blown KRW System

(see Table 38) is higher than the conditional expected value of lower costs, and the mean of the distribution is a net loss of \$3/kW-yr for the Lurgi system.

The outcomes of further research in the Lurgi system would reduce the probability of higher fixed operating cost, and would decrease the risk associated with higher costs. In the "all reduced uncertainties" case, there is only a 16 percent chance of higher fixed operating costs, and a mean cost *savings* of \$5/kW-yr. Thus, the effect of additional research in this case is to yield an advantage for the Lurgi system. As indicated in Table 38, a similar result is obtained even if further research reduces uncertainties in the KRW system.

The Lurgi system suffers from high variable operating costs associated with the zinc ferrite sorbent. Thus, regardless of the assumptions regarding research outcomes, there is a high probability, around 80 percent, that the Lurgi system will have higher variable operating costs than the KRW system, as indicated in Figure 85. Furthermore, there is a risk that the cost could be over 10 mills/kWh higher, even with a reduction in uncertainties from further process research. While further research will not eliminate the probability of more expensive costs, it will reduce the downward partial mean of higher costs and increase the mean of the distribution, thereby reducing the expected value of the higher costs, as indicated in Table 38. Thus, there is value obtained from research with respect to reducing the difference in variable operating costs between the two systems.

Perhaps the single most important variable for comparative analysis is the levelized cost of electricity. The probability distributions for the difference in cost between the Lurgi and KRW systems are shown in Figure 86, assuming the base case uncertainty values for

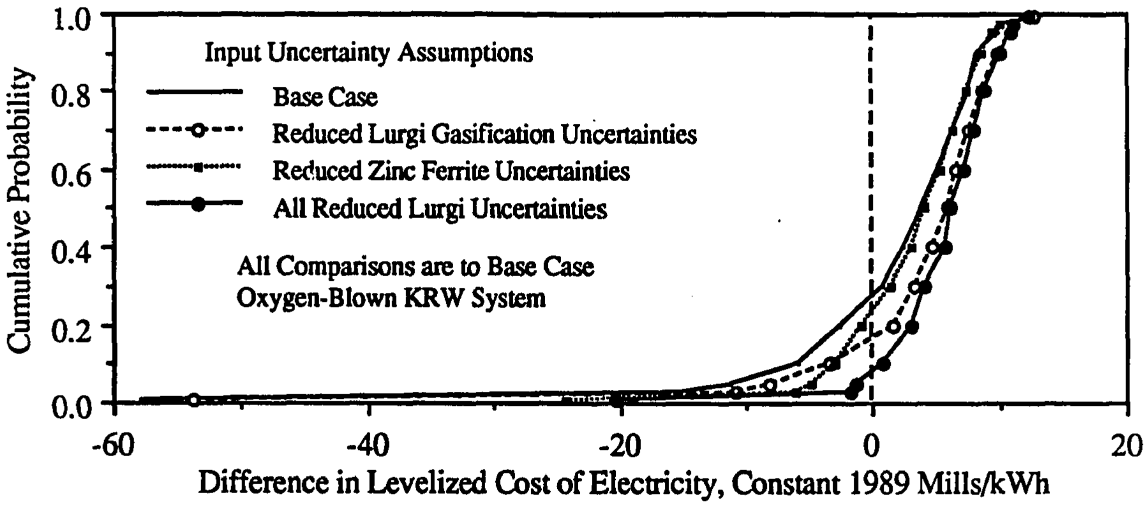

Figure 86. Effect of Illustrative Research Outcomes on Cost of Electricity Savings for Air-blown Lurgi System Compared to Oxygen-blown KRW System

the KRW system. For all cases, the Lurgi system is likely to be less expensive than the KRW system. However, the risk of higher costs for the base case is 27 percent, and the expected value of a loss, given that a loss has occurred, is 9.5 mills/kWh. Research in the gasification process area reduces the probability of a loss to 16 percent, and it reduces the downward partial mean by 0.7 mills/kWh. However, it does not eliminate the long tail associated with the loss. Therefore, the conditional expected value of a loss actually goes up. Reduced gasification uncertainties result in an upward shift in the central values of the distribution, due to the skewness of uncertainties in the model. Therefore, the expected net cost savings increases by 2 mills/kWh.

Research on the zinc ferrite process area is expected to reduce the risks associated with poor sorbent performance. Therefore, the long tail of the distribution is substantially reduced. Although zinc ferrite research only reduces the probability of a loss to 21 percent, it has a more substantial effect on the other measures of downside risk, compared to gasification research outcomes. The downward partial mean is reduced by 1.6 mills/kWh. Reduction in zinc ferrite uncertainty does not lead to a major change in the central values of the cost difference. For this reason, the expected net cost savings is less than for gasification research, in spite of the stronger effect on downside risk.

When all research areas are combined, the probability of higher cost is reduced to six percent, and the expected net cost savings increases by 4.1 mills/kWh. Thus, additional research on the Lurgi system may yield substantial pay-offs, both in terms of risk reduction and net cost savings. Similar results are obtained even if uncertainties in the

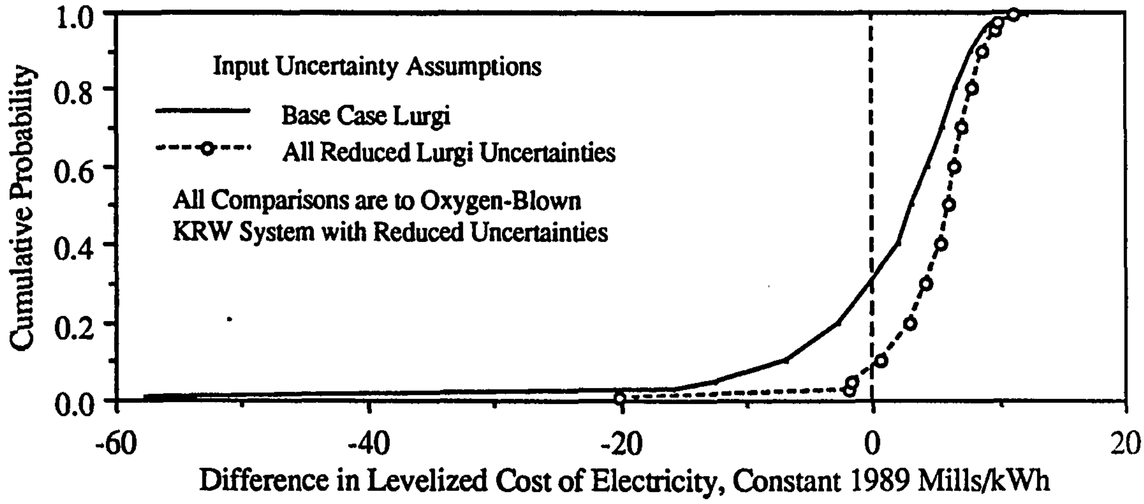

Figure 87. Effect of Illustrative Research Outcomes on Cost of Electricity Savings for Air-blown Lurgi System Compared to Oxygen-blown KRW System

KRW system are reduced, as shown in Figure 87 and given in Table 38, indicating that the results for the Lurgi system are robust.

The cost of electricity depends on the plant capacity factor. As already discussed, although the Lurgi holds substantial capital cost advantages, it is likely to have higher operating costs. Therefore, for high capacity factors, which lead to a relatively higher weighting of operating costs in the cost of electricity, the advantage of the Lurgi system is reduced. For a capacity factor of 90 percent, the Lurgi system enjoys a 65 percent probability of cost savings. However, because of the long tail associated with poor zinc ferrite sorbent performance, the mean value of the distribution for the difference in cost is negative, indicating that the Lurgi system is expected to be 0.08 mills/kWh more *expensive* than the KRW system. For capacity factors slightly below 90 percent, the Lurgi system has an expected cost savings. Capacity factors of 90 percent are not often achieved in the electric utility industry. Therefore, this represents perhaps an overly stringent comparison point for the Lurgi system.

The Lurgi system will have lower SO_2 but higher NO_x emissions than the KRW system. The lower SO2 emissions are due to the very high efficiency (e.g., 99.8 percent) of the hot gas cleanup system. The higher NO_x emissions are due to conversion of fuel bound nitrogen (ammonia) in the conventional gas turbine combustor. In spite of the higher efficiency of the Lurgi system, the KRW system will tend to have lower CO_2 emissions. This is because the KRW system tends to have higher carbon retention in the bottom ash than the Lurgi system, leading to sequestering of carbon in the solid waste stream that otherwise would be emitted as CO_2. The KRW system also has a substantially lower process water requirement, because the higher temperature KRW gasifier does not require as much steam for the purpose of temperature control, as compared to the Lurgi gasifier.

6.5.2 Comparing the Air- and Oxygen-Blown KRW Systems

The pairing of model input variables between the air-blown KRW system with hot gas cleanup and the oxygen-blown KRW system with cold gas cleanup is shown in Table 39. The pairing of uncertainties between the two systems was based on the screened uncertainties. Several of the variables common to, or similar between, both flowsheets are assumed to be completely correlated. For example, cost-related parameters such as engineering and home office fees are taken to be the same for both cases. The basis of the assumed correlations is the same as described in the previous section.

The results of the paired simulations of the two flowsheets were obtained and analyzed, accounting for the underlying correlation between the two cases. For base case uncertainty assumptions, the correlation between the total capital costs uncertainties of the two systems was 0.75, and the correlation between levelized costs was 0.54. These correlations significantly affect the comparative results.

The air-blown KRW-based system compares favorably with the oxygen-blown KRW-based system on several attributes. Here, the oxygen-blown system is assumed to represent "conventional" IGCC technology embodying cold gas cleanup. The air-blown system featuring hot gas cleanup system holds a clear advantage with respect to plant efficiency, SO_2 emissions, total capital cost, and cost of electricity. It is likely to have lower water consumption and improved fixed operating cost. However, it is certain to have higher variable operating cost and higher NO_x emissions. It is also likely to have higher CO_2 emissions in spite of its higher efficiency.

The comparisons for several of these attributes are discussed in more detail. The statistics associated with the probability distributions of the differences between the two technologies are summarized in Table 40 for plant efficiency, CO_2 emissions, ash disposal, total capital cost, fixed operating cost, variable operating cost, and levelized cost of electricity. From the table, it is clear that the air-blown system is either better or worse than the conventional system for a given attribute; there is little question regarding the probability of a loss.

Table 39. Pairing of Uncertain Parameters for Comparative Study of Oxygen-Blown KRW- and Air-Blown KRW-based IGCC Systems.

Description of Uncertain Parameter[a]		
Oxygen-blown KRW (Case OKC)	Air-blown KRW (Case AKH)	Correlated[b]
Carbon Conversion	Carbon Conversion	No
Oxygen/Carbon Ratio	Oxygen/Carbon Ratio	No
Steam/Oxygen Ratio		
Sulfur Retention in Ash	Sulfur Retention in LASH	No
	Gasifier Ca/S Ratio	
	Gasifier Temperature	
	Gasifier Ammonia Yield	
	Sulfation SO_2 Emission Rate	
	Sulfation NO_x Emission Rate	
	Sulfation Carbon Conversion	
	Sulfation Sulfide Conversion	
	Zinc Ferrite Residual Sulfate	
	Zinc Ferrite Sorbent Loading	
	Zinc Ferrite Sorbent Attrition	
Thermal NO_x	Thermal NO_x	Yes
	Fuel NO_x Conversion	
Gas Turbine CO Emissions	Gas Turbine CO Emissions	No
SE Air Separation Aux. Power		
SE Coal Hndg Aux. Power		
Air Separation DC		
	Coal Handling DC	
Gasification DC	Gasification DC	No
	Sulfation DC	
	Zinc Ferrite DC	
Gas Turbine DC	Gas Turbine DC	Yes
General Facilities DC	General Facilities	Yes
SE Coal Hndg DC	SE Coal Hndg DC	No
SE Air Separation DC		
SE Gasification DC		
SE Selexol DC		
SE HRSG DC	SE HRSG DC	Yes
SE Steam Turbine DC	SE Steam Turbine DC	Yes
Gasification Maintenance	Gasification Maintenance	No
	Sulfation Maintenance	
Gas Turbine Maintenance	Gas Turbine Maintenance	No
	Unit Cost of Zinc Ferrite Sorb.	
	Unit Cost of Limestone Sorbent	
Ash Disposal Cost	Ash Disposal Cost	Yes
Sulfur Byproduct		
Byproduct Marketing		
Engr & Home Office Fees	Engr & Home Office Fees	Yes
Indirect Construction Cost	Indirect Construction Cost	Yes
Project Cost Uncertainty	Project Cost Uncertainty	Yes

[a] Uncertain parameters which are analogous or the same are listed on the same line.
[b] The term "correlation" is used here to indicate parameters for which the same vector and ranking of samples was used in uncertainty analysis of both technologies.

The air-blown system with hot gas cleanup holds a substantial efficiency advantage over the conventional system. This advantage is attributable to the reduction in fuel gas cooling, a lower auxiliary power requirement for oxidant feed, and combustion of unconverted carbon in the gasifier ash to generate steam in the sulfation unit.

However, in spite of substantially higher efficiency, the air-blown system has over an 80 percent probability of higher CO_2 emissions. This result is obtained because of the use of a limestone sorbent for desulfurization in the gasifier. The calcium carbonate in the limestone is calcined in the gasifier, releasing CO_2. The carbon retained in the bottom ash that is combusted in the sulfation unit is an additional source of CO_2 emissions. In the conventional system, unconverted carbon is sequestered in the bottom ash.

Because of the additional burden of spent limestone sorbent in the air-blown system, the ash disposal requirement will be higher than for the conventional system. The oxygen-blown KRW system using cold gas cleanup converts sulfur in the fuel gas to a elemental sulfur, thereby reducing the solid waste burden and generating a byproduct revenue stream.

The air-blown system will have lower capital cost, due to the reduction in equipment cost associated with fuel gas cooling and cleanup. The expected cost savings is over $400/kW. The 90 percent probability range for the cost savings, assuming base case uncertainties for both systems, is from $330/kW to $538/kW. If reduced uncertainties are assumed for both cases, the 90 percent range becomes $339/kW to $495/kW.

There is a chance that the fixed operating cost of the air-blown system could be higher than for the conventional system, due to the risks of contaminant related problems in the hot gas cleanup system. However, the downside risk of more expensive fixed operating cost is small compared to the mean of the cost difference, as seen in Table 40.

Regardless of assumptions regarding uncertainties, the air-blown system will have higher operating costs than the conventional system. These higher costs are associated with gasifier and external desulfurization in the advanced system. The advanced system has higher costs associated with limestone sorbent, ash disposal, and zinc ferrite sorbent.

Table 40. Results of Research Information Case Studies: Comparison of Air-blown KRW and Oxygen-blown KRW Systems[a]

Research Area[b]	Probability of a Loss (%)	Downward Partial Mean	Expected Value of a Loss	Expected Value of a Gain	Mean	Reduction in Risk	Value of Research
Plant Efficiency, percent							
AKH vs. OKC Base Cases	0	0	0	5.0	5.0	--	--
AKH Reduced vs. OKC Base	0	0	0	5.1	5.1	0	0.1
AKH Base vs. OKC Reduced	0	0	0	3.9	3.9	--	--
AKH vs. OKC Reduced Cases	0	0	0	4.0	4.0	0	0.1
CO2 Emissions, lb/kWh							
AKH vs. OKC Base Cases	89	0.042	0.048	0.009	-0.041	--	--
AKH Reduced vs. OKC Base	95	0.039	0.041	0.006	-0.039	0.003	0.002
AKH Base vs. OKC Reduced	83	0.031	0.037	0.009	-0.029	--	--
AKH vs. OKC Reduced Cases	89	0.027	0.030	0.004	-0.027	0.004	0.002
Ash Disposal Rate, lb/kWh							
AKH vs. OKC Base Cases	100	0.153	0.153	0	-0.153	--	--
AKH Reduced vs. OKC Base	100	0.153	0.153	0	-0.153	0	0
AKH Base vs. OKC Reduced	100	0.155	0.155	0	-0.155	--	--
AKH vs. OKC Reduced Cases	100	0.155	0.155	0	-0.155	0	0
Total Capital Cost, 1989 $/kW							
AKH vs. OKC Base Cases	0	0	0	428	428	--	--
AKH Reduced vs. OKC Base	0	0	0	425	425	0	-3
AKH Base vs. OKC Reduced	0	0	0	417	417	--	--
AKH vs. OKC Reduced Cases	0	0	0	414	414	0	-3
Fixed Operating Cost, 1989 $/kW-yr							
AKH vs. OKC Base Cases	9	0.16	1.77	8.84	7.88	--	--
AKH Reduced vs. OKC Base	1	0.009	0.9	9.73	9.63	0.007	1.75
AKH Base vs. OKC Reduced	3	0.03	1.1	8.39	8.10	--	--
AKH vs. OKC Reduced Cases	0	0	0	9.85	9.85	0.03	1.75
Variable Operating Cost, 1989 mills/kWh							
AKH vs. OKC Base Cases	100	2.5	2.5	0	-2.5	--	--
AKH Reduced vs. OKC Base	100	2.5	2.5	0	-2.5	0	0
AKH Base vs. OKC Reduced	100	3.0	3.0	0	-3.0	--	--
AKH vs. OKC Reduced Cases	100	3.0	3.0	0	-3.0	0	0
Levelized Cost of Electricity, Constant 1989 mills/kWh							
AKH vs. OKC Base Cases	0	0	0	6.6	6.6	--	--
AKH Reduced vs. OKC Base	0	0	0	6.9	6.9	0	0.3
AKH Base vs. OKC Reduced	0	0	0	5.9	5.9	--	--
AKH vs. OKC Reduced Cases	0	0	0	6.2	6.2	0	0.3

[a] The comparison is from the perspective of the air-blown KRW-based system (AKH). Thus, "loss" in this case is a probability that system AKH will be worse (lower efficiency, higher emissions, higher cost) in a given attribute than the oxygen-blown KRW-based system (OKC).

[b] For each parameter used for comparison, the results are grouped separately for comparisons of the air-blown KRW system with hot gas cleanup (AKH) to the base case and "All" reduced uncertainties case for the oxygen-blown KRW-based system with cold gas cleanup (OKC).

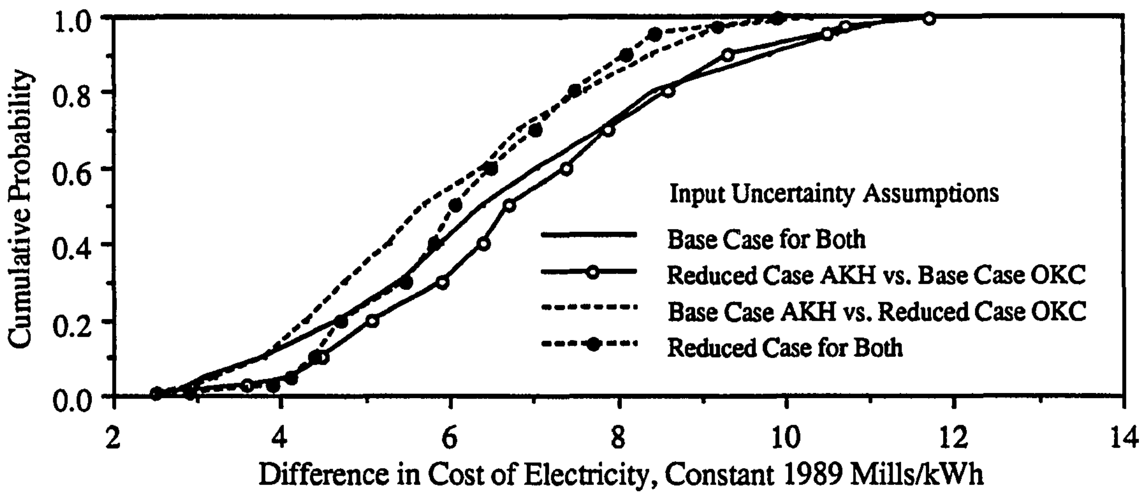

Figure 88. Effect of Illustrative Research Outcomes on Cost of Electricity Savings for Air-blown KRW System Compared to Oxygen-blown KRW System

Overall, however, the advanced system will enjoy levelized cost savings over the conventional system, for all four case studies shown in Table 40. The levelized cost savings is shown graphically in Figure 88. The solid lines represent comparisons to the base case uncertainties for the conventional system, while the dashed lines represent comparisons to the reduced uncertainties case for the conventional system. While research on the conventional system will tend to reduce the expected cost savings of the advanced systems, the cost savings remain substantial nonetheless. Savings of over 2 mills/kWh are obtained from the analysis for all cases. The mean cost savings are typically 6 mills/kWh, with a chance that cost savings could be 10 mills/kWh or higher.

Similar to the previous comparative case study, the advantage of the air-blown system with hot gas cleanup is diminished as the plant capacity factor increases, because of its higher variable operating costs. However, in the case of the air-blown KRW system, the levelized cost savings is sufficiently large that it continues to have a 100 percent probability of cost savings even at a 90 percent capacity factor.

6.5.3 Comparing the Air-Blown Lurgi and KRW Systems

In the previous comparisons to the oxygen-blown KRW system, both air-blown systems with hot gas cleanup were shown to compare favorable. How do the air-blown systems compare to each other?

The pairing of uncertain parameters for the comparative case studies of the two systems is shown in Table 41. The rationale for correlations between uncertain variables is the same as discussed in Section 6.5.1.

The results of the comparative case study of the air-blown KRW and Lurgi systems are summarized in Table 42. The summary includes several key performance, emissions, and cost variables. Not shown in the table are SO_2 and NO_x emissions, and water consumption. The KRW system is certain to have lower SO_2 and NO_x emissions and lower water consumption.

The difference in SO_2 emissions is attributable to the byproduct sulfuric acid plant in the Lurgi system. The KRW system uses a combination of gasifier in-bed desulfurization with limestone and external zinc ferrite desulfurization. The regeneration off-gas from the zinc ferrite unit is recycled to the gasifier for capture of the evolved SO_2. Approximately 90 percent of incoming sulfur is captured in the gasifier, which substantially reduces the sulfur loading to the zinc ferrite system. In contrast, all of the sulfur released in the Lurgi gasifier enters the zinc ferrite unit in "bulk" desulfurization mode. However, both the KRW and Lurgi systems result in a fuel gas containing 10 ppmv of sulfur (primarily as H_2S) Thus, the overall sulfur capture efficiency is approximately the same for both systems. However, a portion of the sulfur captured in the Lurgi system is emitted in the tail gas of the sulfuric acid plant. In the KRW system, all captured sulfur is contained in the limestone sorbent. A small fraction of sulfur is emitted in the sulfation unit. The SO_2 emissions from the sulfuric acid plant in the Lurgi system outweigh those from the sulfation unit in the KRW system. These interaction are considered in the performance models.

The lower NO_x emissions of the KRW system are attributable to the lower ammonia yield from the KRW gasifier compared to the Lurgi gasifier. The lower water consumption of the KRW system is an advantage of the higher temperature gasifier. The The dry-ash Lurgi gasifier requires steam to prevent the temperature in the combustion zone from exceeding the ash fusion temperature.

The KRW system with hot gas cleanup has an advantage over the Lurgi system with respect to plant efficiency. Based on probabilistic simulations with four combinations of uncertainties in the two technologies, no outcomes were obtained in which the Lurgi system had higher efficiency. The KRW system is expected to have an efficiency 3 to 3.7 percentage points higher than the Lurgi system, as shown in Table 42.

Table 41. Pairing of Uncertain Parameters for Comparative Study of Air-Blown KRW- and Air-Blown Lurgi-based IGCC Systems.

Description of Uncertain Parameter[a]		
Air-blown KRW (Case AKH)	Air-blown Lurgi (Case ALH)	Correlated[b]
	Fines Carryover	
	Fines Capture	
	Carbon Retention in Ash	
	Gasifier Coal Throughput	
Gasifier Ammonia Yield	Gasifier Ammonia Yield	No
Carbon Conversion		
Oxygen/Carbon Ratio	Air/Coal Ratio	No
	Steam/Coal Ratio	
CaS Sulfation Rate		
Gasifier Temperature		
Limestone Ca/S Ratio		
Gasifier Sulfur Capture		
Sulfator Carbon Conversion		
Zinc Ferrite Residual Sulfate	Zinc Ferrite Residual Sulfate	Yes
Zinc Ferrite Sorbent Loading	Zinc Ferrite Sorbent Loading	Yes
Zinc Ferrite Sorbent Attrition	Zinc Ferrite Sorbent Attrition	Yes
Fuel NO_x Conversion	Fuel NO_x Conversion	Yes
Gas Turbine CO Conversion	Gas Turbine CO Conversion	Yes
Thermal NO_x		
Gasification DC	Gasification DC	No
Zinc Ferrite DC	Zinc Ferrite DC	Yes
	Sulfuric Acid Plant DC	
Sulfation DC		
Gas Turbine DC	Gas Turbine DC	Yes
General Facilities DC		
	SE Coal Hndg DC	No
SE HRSG DC	SE HRSG DC	Yes
SE Steam Turbine DC	SE Steam Turbine DC	Yes
Gasification Maintenance	Gasification Maintenance	No
Gas Turbine Maintenance	Gas Turbine Maintenance	No
Unit Cost of Zinc Ferrite Sorb.	Unit Cost of Zinc Ferrite Sorb.	Yes
Unit Cost of Limestone		
Ash Disposal Cost	Ash Disposal Cost	Yes
	Sulfuric Acid Byproduct	
Engr & Home Office Fees	Engr & Home Office Fees	Yes
Indirect Construction Cost	Indirect Construction Cost	Yes
Project Cost Uncertainty	Project Cost Uncertainty	Yes

[a] Uncertain parameters which are analogous or the same between the two technologies are listed on the same line.

[b] The term "correlation" is used here to indicate parameters for which the same vector and ranking of samples was used in uncertainty analysis of both technologies. Parameters which are unique to one technology are not correlated with any other uncertain parameters. Parameters which are similar across technologies may be uncorrelated, as indicated, because of differences in the underlying technologies. See text for explanation of basis for correlations.

Table 42. Results of Research Information Case Studies: Comparison of Air-blown KRW and Air-blown Lurgi Systems[a]

Research Area[b]	Probability of a Loss (%)	Downward Partial Mean	Expected Value of a Loss	Expected Value of a Gain	Mean	Reduction in Risk	Value of Research
Plant Efficiency, percent							
AKH vs. ALH Base Cases	0	0	0	3.5	3.5	- -	- -
AKH Reduced vs. ALH Base	0	0	0	3.7	3.7	0	0.2
AKH Base vs. ALH Reduced	0	0	0	3.0	3.0	- -	- -
AKH vs. ALH Reduced Cases	0	0	0	3.1	3.1	0	0.1
CO2 Emissions, lb/kWh							
AKH vs. ALH Base Cases	29	0.009	0.030	0.031	0.014	- -	- -
AKH Reduced vs. ALH Base	23	0.006	0.027	0.030	0.017	0.003	0.003
AKH Base vs. ALH Reduced	28	0.006	0.022	0.024	0.011	- -	- -
AKH vs. ALH Reduced Cases	16	0.003	0.017	0.021	0.015	0.003	-0.004
Ash Disposal Rate, lb/kWh							
AKH vs. ALH Base Cases	100	0.129	0.129	0	-0.129	- -	- -
AKH Reduced vs. ALH Base	100	0.129	0.129	0	-0.129	0	0
AKH Base vs. ALH Reduced	100	0.137	0.137	0	-0.137	- -	- -
AKH vs. ALH Reduced Cases	100	0.137	0.137	0	-0.137	0	0
Total Capital Cost, 1989 $/kW							
AKH vs. ALH Base Cases	21	10	49	127	90	- -	- -
AKH Reduced vs. ALH Base	15	10	65	115	88	0	-2
AKH Base vs. ALH Reduced	23	10	42	70	44	- -	- -
AKH vs. ALH Reduced Cases	22	8	36	64	42	2	-2
Fixed Operating Cost, 1989 $/kW-yr							
AKH vs. ALH Base Cases	14	0.6	4.2	13.0	10.6	- -	- -
AKH Reduced vs. ALH Base	7	0.2	3.1	13.5	12.3	0.4	1.7
AKH Base vs. ALH Reduced	34	1.3	3.9	5.7	2.4	- -	- -
AKH vs. ALH Reduced Cases	20	0.4	2.0	5.7	4.2	0.9	1.8
Variable Operating Cost, 1989 mills/kWh							
AKH vs. ALH Base Cases	66	1.2	1.8	8.4	1.7	- -	- -
AKH Reduced vs. ALH Base	68	1.2	1.7	8.9	1.7	0	0
AKH Base vs. ALH Reduced	70	1.2	1.7	3.4	-0.2	- -	- -
AKH vs. ALH Reduced Cases	72	1.2	1.7	3.7	-0.2	0	0
Levelized Cost of Electricity, Constant 1989 mills/kWh							
AKH vs. ALH Base Cases	22	0.5	2.1	7.2	5.2	- -	- -
AKH Reduced vs. ALH Base	23	0.4	1.5	7.5	5.4	0.1	0.2
AKH Base vs. ALH Reduced	45	0.8	1.8	3.4	1.0	- -	- -
AKH vs. ALH Reduced Cases	41	0.6	1.5	3.3	1.3	0.2	0.3

[a] The comparison is from the perspective of the air-blown KRW-based system (AKH). Thus, "loss" in this case is a probability that system AKH will be worse (lower efficiency, higher emissions, higher cost) in a given attribute than the air-blown Lurgi-based system (ALH).

[b] For each parameter used for comparison, the results are grouped separately for comparisons of the air-blown KRW system with hot gas cleanup (AKH) to the base case and "All" reduced uncertainties case for the air-blown Lurgi-based system with hot gas cleanup (ALH).

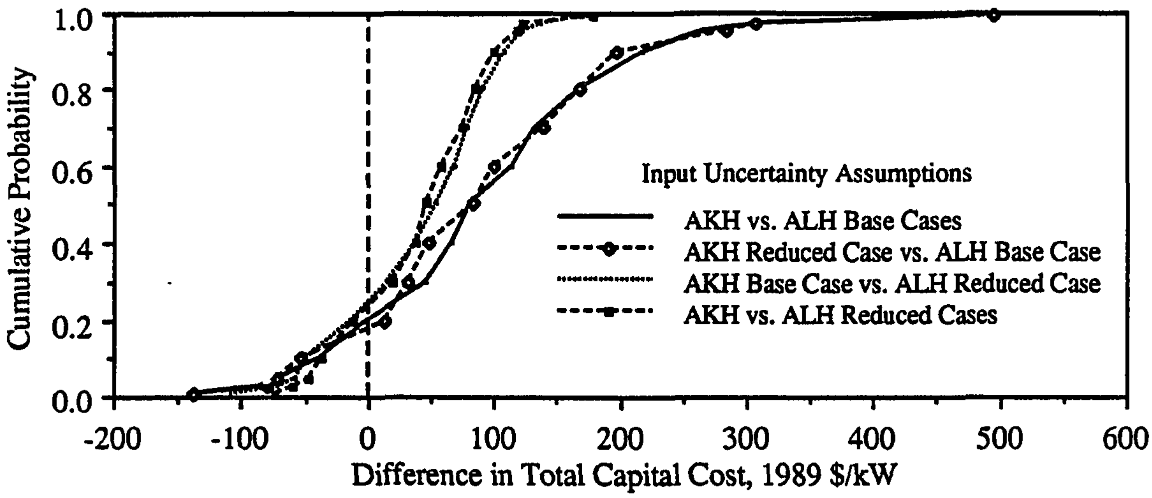

Figure 89. Effect of Illustrative Research Outcomes on Total Capital Cost Savings for Air-blown KRW System Compared to Air-blown Lurgi System

In spite of its higher efficiency, the KRW system has a substantial probability of higher CO_2 emissions. This results is associated with the calcination of limestone in the gasifier, as discussed previously. The KRW system also has a higher ash disposal burden associated with the throw-away limestone sorbent.

The air-blown KRW system is likely to have lower capital cost than the Lurgi-based system. As indicated in Table 42 and shown in Figure 89, there is a 15 to 23 percent probability that the capital cost of the KRW system would be higher, depending on the uncertainty sets assumed for both technologies. Because performance related uncertainties were shown to have a major influence on uncertainty in Lurgi capital cost, reductions in performance-related uncertainties of the Lurgi system substantially reduce the mean cost savings of the KRW system. However, the KRW system enjoys a high probability of cost savings for all cases, and a mean cost savings of at least $40/kW. The correlation between the capital cost for the two cases is 0.54.

One of the most interesting results obtained is for the comparison of variable operating cost between the KRW and Lurgi systems with hot gas cleanup, shown in Figure 90. As indicated in Table 42, the KRW has approximately a 70 percent probability of higher variable costs than the Lurgi system. However, for the base case uncertainties in the Lurgi system, the KRW system has a mean cost *savings* of 1.7 mills/kWh. Although the probability of cost savings is about 30 percent, the probability distribution is sufficiently skewed that that the mean lies above zero. When reduced uncertainties in the Lurgi system are considered, the tail of the distribution is shortened. In these cases, the mean of the

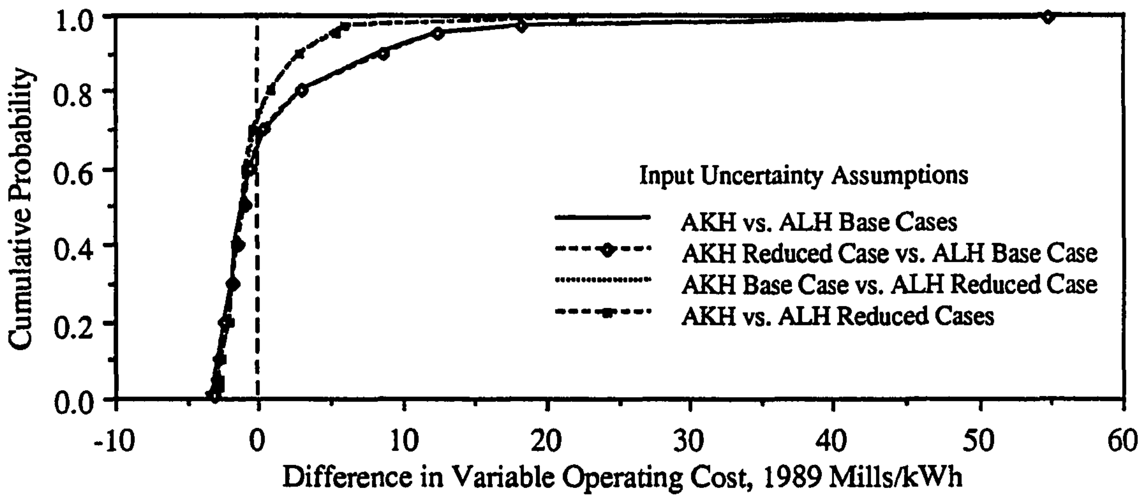

Figure 90. Effect of Illustrative Research Outcomes on Variable Operating Cost Savings for Air-blown KRW System Compared to Air-blown Lurgi System

distribution is negative, indicated that the KRW system is expected to be more expensive than the Lurgi system.

While both systems feature zinc ferrite desulfurization, the KRW system has a much lower sulfur loading to the zinc ferrite process area. Therefore, the sorbent costs for the KRW system are substantially less. However, the KRW has added costs for limestone sorbent and for the higher ash disposal burden associated with in situ gasification desulfurization.

The results for the difference in the cost of electricity for the two systems are shown in Figure 91. In all four cases, representing combinations of base case and reduced uncertainties for both technologies, the KRW system has greater than 50 percent probability of cost savings. The mean cost savings for all four cases is positive, as indicated in Table 42. However, for the cases where uncertainties in the Lurgi system are reduced, there is over a 40 percent probability that the Lurgi system would be less expensive, although there is an expected cost savings of about 1 mill/kWh for the KRW system. The skewness of the distributions for the cost differences are attributable to the uncertainties in the zinc ferrite process area.

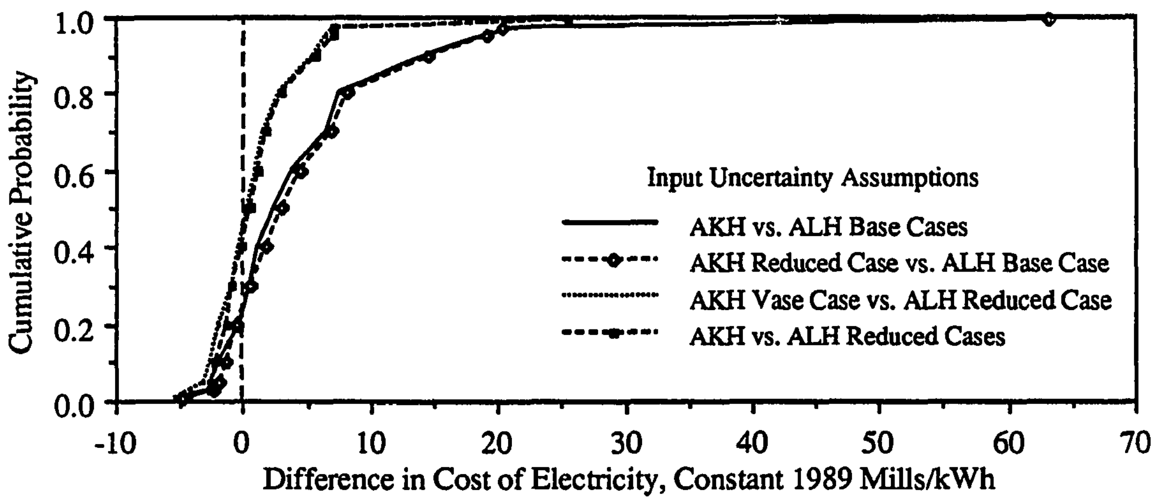

Figure 91. Effect of Illustrative Research Outcomes on Cost of Electricity Savings for Air-blown KRW System Compared to Air-blown Lurgi System

The probabilistic comparative case studies have indicated that both air-blown systems with hot gas cleanup hold significant performance and cost advantages over the oxygen-blown KRW system, although they have higher NO_x emissions. The air-blown KRW system appears to be preferred over the Lurgi system; however, there is a substantial probability that the Lurgi system would yield cost savings.

6.6 An Illustrative Decision Model

In this section, the results of the comparative case study of the air-blown Lurgi and oxygen-blown KRW systems are used as inputs to a decision model. The purpose of the decision model is to represent the preferences of a decision maker, such as a research planner or process adopter, who is faced with choices between competing innovative technologies under uncertainty. The decision model accounts for the decision maker's attitude toward risk. In addition, the model used here is sensitive to the timing of research outcomes.

A simple conceptual diagram of a decision tree representing the alternatives of selecting a conventional technology or one of several research strategies for an advanced technology is shown in Figure 92. The conventional technology represents the status quo, and has an associated cost savings of zero. If the advanced technology is selected based on current knowledge, it is expected to have a cost savings of 1.47 mills/kWh. However, there is a risk that the Lurgi system would be more expensive than the conventional technology, as discussed in Section 6.5.1. Instead of choosing the technology "as is," a

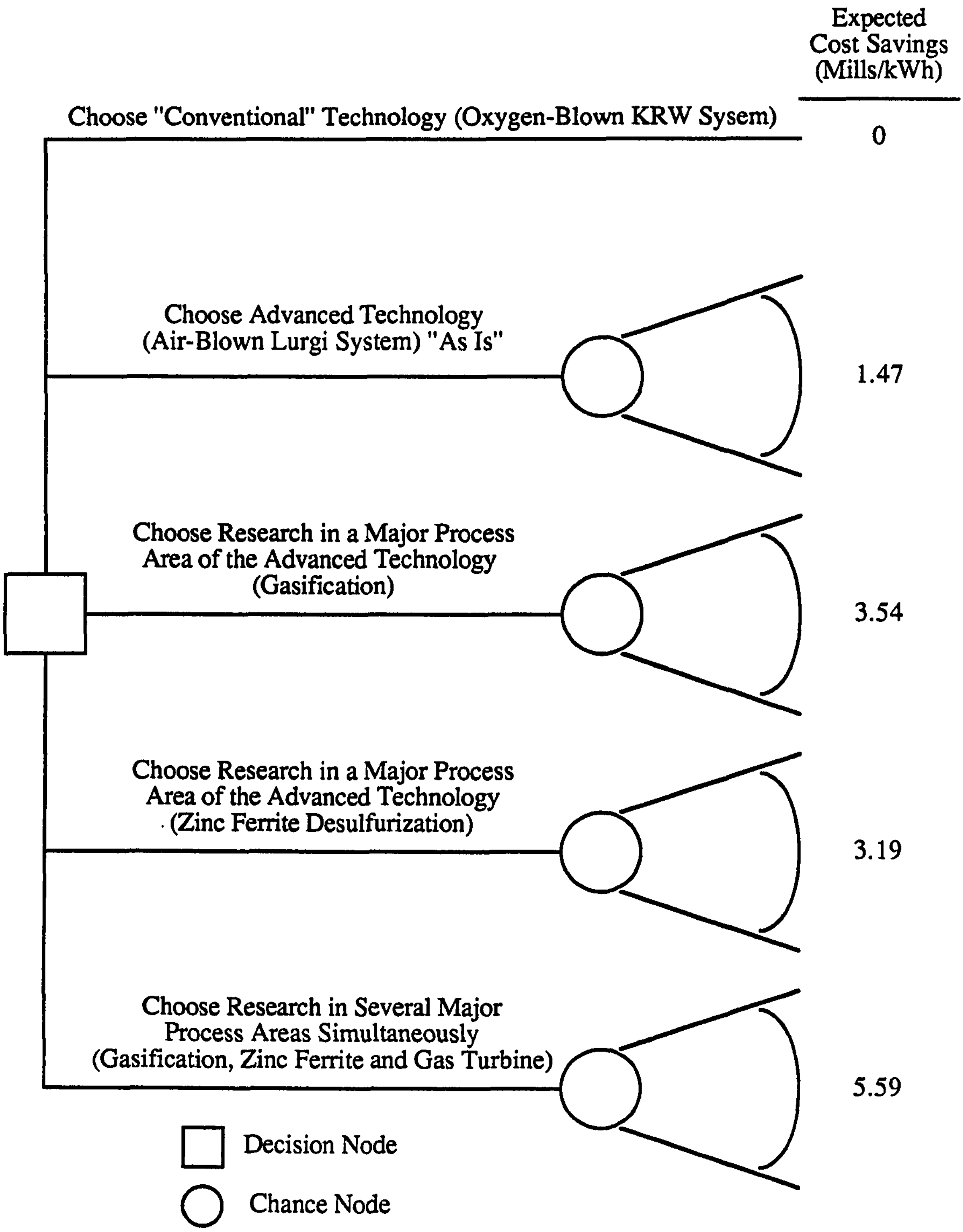

Figure 92. Conceptual Decision Tree for Decisions Regarding Advanced Technology Selection and Research Planning

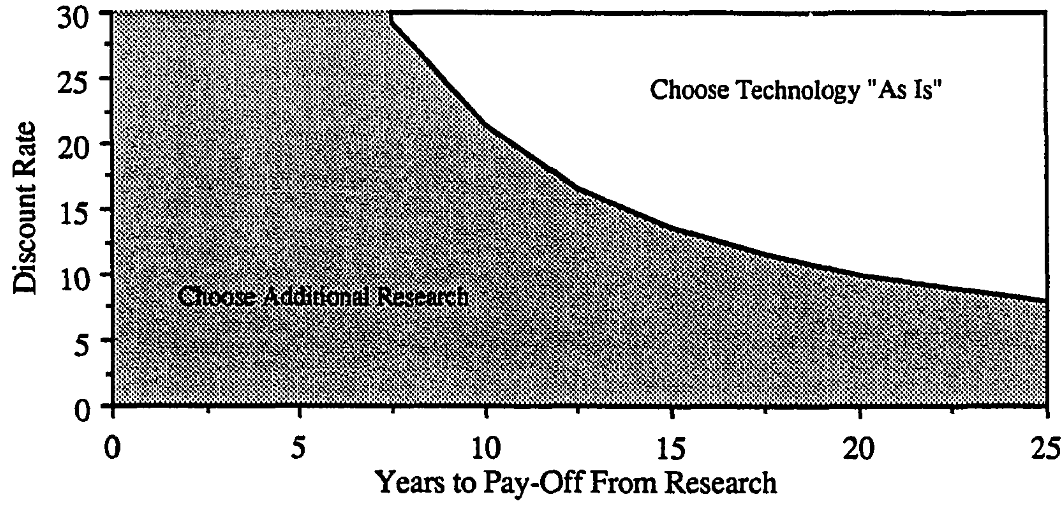

Figure 93. Sensitivity of Decision to Pursue Further Research to Discount Rate and Time for a Risk-Averse Decision-Maker

process adopter could opt to wait for the results of further research. Depending on the specific process areas which are targeted for research, the expected cost savings would increase to as high as 5.6 mills/kWh. However, these outcomes may not be available for another 5 to 20 years. A decision maker may prefer an outcome this year to the same outcome obtained next year. Therefore, the time value of the outcomes are modeled here using discounting.

A decision model based on the utility function given in Section 2.5 was used to model the preferences of a decision maker. Assuming that the results of further research could be obtained instantaneously, only an extremely risk averse decision maker would choose the conventional technology over the advanced technology. In all other cases, the decision maker would prefer the advanced technology "as is" to the conventional technology. Furthermore, the decision maker would obtain the highest expected utility from pursuing research in several major process areas simultaneously.

The timing of research outcomes affects the expected utility of the three research options shown in Figure 92. For a moderately risk averse decision maker, the expected utility of the research option for all process areas remains higher than all other alternative as long as the research outcomes are obtained within 15 years at a discount rate lower than 14 percent. The sensitivity of the decision to pursue further research to the time to pay-off and the discount rate is given in Figure 93. In all cases, the expected utility of research into all process areas is higher than for research in a single process area. As noted in Chapter 5, decision analysis of research outcomes for a particular utility may need to consider interim power supply measures in cases where the outcomes are not immediately available. Such measures may alter the net benefit obtained from additional research. Moreover, the costs of research also would need to be considered.

In addition to a decision analysis based on the single-attribute of levelized cost, a multi-attribute analysis was considered that included differences in SO_2, NO_x, and CO_2 emissions between the two technologies. Both the SO_2 and NO_x emission rates were converted to a plant energy output basis for the comparison. Thus, the environmental impact and cost associated with 1 kWh of plant output was considered for both technologies. The emission rates were converted to a mills/kWh basis by assuming a unit cost per amount of emitted pollutant. As a nominal assumption, the external costs of SO_2, NO_x, and CO_2 were assumed to be \$2,000/ton, \$1,000/ton, and \$25/ton, respectively. However, including these pollutants in the decision analysis with these assumed external prices did not significantly change the model results.

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7.0 DISCUSSION

The generic features of the probabilistic modeling approach are discussed here in comparison to traditional approaches to technology evaluation and to the work of the Rand Corporation. Rand has conducted a number of quantitative studies regarding the development of performance and cost estimates of innovative process technologies.

7.1 Information Requirements

Compared to deterministic analysis, the probabilistic modeling approach requires that more detailed judgments be made regarding the values assigned to performance and cost parameters in an engineering model. Thus, the time required to develop estimates of uncertainty is usually higher than the time that would be required to make a "best guess" estimate. However, by systematically thinking about uncertainties in specific parameters, an analyst is more likely to uncover potential sources of cost growth or performance shortfalls that would otherwise be overlooked.

A tendency that has been observed in this research and by others is for best guess estimates to be more similar to the median or mode than to the mean of a distribution. For symmetric distributions, there is no difference. However, for skewed distributions, the mean may differ substantially from the median and the mode. People tend to think more in terms of the single "most likely" outcome (mode) or the outcome for which there is a 50-50 chance of higher or lower outcomes (median). It is much more difficult to evaluate the average value of a distribution (mean), particularly if, in making a best guess judgment, the analyst does not consider both the range and probability of outcomes for an uncertain parameter. As shown in many of the case studies, the influence of skewed distributions on model results can be important. They tend to shift the central tendency of resulting uncertainties in performance and cost, and can lead to long tails representing unfavorable outcomes. These types of interactions cannot be evaluated systematically in deterministic analysis.

Thus, while the information requirements may be more demanding for probabilistic analysis, there is a benefit obtained in return. The benefit is more realistic estimates of performance and cost. Also, thinking about uncertainties is an important way to gain understanding into the key factors that drive the risk of failure for an innovative process technology.

7.2 Computational Requirements

The computational requirements for probabilistic modeling depend on the complexity of the engineering model and the probabilistic modeling environment. For the pulverized coal-fired power plants evaluated here, the models were run on a desktop computer in a manner of minutes for sample sizes of 150. For the IGCC systems, the case studies took 6 to 12 hours for a sample size of 100, and required a powerful mini-computer. However, because probabilistic case studies provide a systematic means to capture the effects of simultaneous variations in many input parameters, they eliminate the need for a combinatorial explosion of "sensitivity" case studies that would be typical of deterministic analysis.

7.3 Cost Estimating and Risk Assessment

As shown in the case studies of the copper oxide process and IGCC systems, probabilistic analysis provides explicit insights into the range and likelihood of outcomes for key measures of plant performance and cost. In many cases, there is a probability of obtaining extreme outcomes, such as low performance or high cost, that would result in technology failure. The characterization of uncertainties in performance and cost results from the simultaneous interaction of uncertainties in many input parameters. These types of insights cannot be obtained from deterministic analysis.

Furthermore, although the notion of uncertainty is claimed to be imbedded in so-called "contingency factors," in most cases contingency factors are inadequate for providing reliable cost estimates for innovative process technologies. This is due, in part, to the tendency for judgments about deterministic point values to correspond to the median or modal values that would be revealed by an uncertainty analysis. In the case of skewed distributions, or where there are nonlinearities in an engineering model, the mean outcomes may often tend to be "worse" than the best guess outcomes. Thus, there is often a high probability of cost overrun compared to the deterministic analysis. Uncertainties in key measures of plant performance and cost result from often complex interactions among uncertain input parameters in the model. These types of interactions simply cannot be captured using a multiplier applied to a cost estimate after it has been developed without regard to uncertainty.

Contingency factors provide no insight into the variance of the capital cost estimate. They also do not allow identification of the specific performance or cost parameters which are the source of potential cost growth and variance in the cost estimate. By suppressing information about uncertainties, deterministic estimates may give a false sense of confidence regarding the certainty of the estimate.

Uncertainties in performance, capital cost, and O&M costs have been shown to contribute significantly to uncertainty in the levelized cost of electricity. However, in most cost estimates, the notion of contingency is not extended to O&M costs, nor to the levelized cost of electricity.

Through regression and probabilistic sensitivity analysis, it is possible to isolate the key input uncertainties which drive the variance in key measures of plant performance and cost. Therefore, it is possible to identify the specific areas for further process research that would lead to significant reductions in the risk for the innovative process.

The results of a probabilistic engineering analysis can be used as inputs to a decision model that represents the preferences of a decision maker. The effect that a risk of low performance or high cost, compared to conventional technology, has on decisions regarding technology adoption and research planning can be evaluated quantitatively.

7.4 Comparison of Probabilistic Approach to RAND Cost Growth Model

The Rand Corporation has conducted a number of studies for DOE regarding the difference between cost estimates and the actual costs incurred for a technology project. Results of these studies, based on a database of cost estimates and actual costs for chemical process plants, were first reported in 1981 (Merrow, Phillips, and Myers). A more recent report applies the Rand cost growth estimation method to a first-of-a-kind magneto-hydrodynamic power station (Hess, Merrow, and Pei, 1987). The method is based on a regression analysis of a database prepared by Rand that contains 106 cost estimates for 44 pioneer chemical process plants. The cost estimates, of which there may be several prepared for one plant during various stages of development, range in scope and detail.

In this section, the results obtained from probabilistic analyses of the copper oxide process and the IGCC system are compared to the results from the Rand cost growth model. First, the Rand model is briefly summarized.

7.4.1 Rand First-of-a-Kind Plant Cost Growth Model

The cost growth model is summarized as follows:

$$\begin{aligned}\text{Cost Growth Factor} = 1.12 &- 0.00297 \times \text{Percent New}\\ &- 0.0213 \times \text{Impurities}\\ &- 0.0114 \times \text{Complexity}\end{aligned}$$

+ 0.00111	x	Inclusiveness
- 0.0401	x	Project Definition
- 0.0235	x	(Project Definition x Process Development)

The coefficient of determination (R^2) of the equation is 0.83.

The terms in the equations are:

Cost Growth Factor:	ratio of actual to estimated capital cost.
Percent New:	Percent of capital cost associated with new
Impurities:	Scale of 0 to 5 assessment of level of difficulty encountered with process stream impurities.
Complexity:	Number of linked process blocks.
Inclusiveness:	Completeness of cost estimate
Project Definition:	Scale of 2 (well-defined) to 8 (poorly defined).
Process Development:	0 well understood, 1 significant R&D issues.

7.4.2 A Case Study For the Copper Oxide Process

An assessment of the cost growth for a first-of-a-kind copper oxide process was made on the basis of regression model given above. Approximately 70 percent of the direct capital cost for the copper oxide process was identified as associated with new, commercially undemonstrated equipment, including the fluidized bed absorber, solids heater, regenerator, and solids transport system. The level of difficulty assessed for impurities was taken to be 2, based on the possibility of some slag build-up in the absorber or dust collection in the regenerator. The number of continuously linked steps was taken to be 10, including parts of the sulfur recovery plant. The cost estimate includes land costs, initial plant inventories, and pre-operating personnel costs, and so the inclusiveness is 100 percent, based on the definition of the variable. The project definition is generally poor, and a value of 7 for this parameter is assumed, based mainly on the fact that the estimate is not site-specific and that there are only limited data from which to draw conclusions about a commercial scale design. The process development variable is given a value of 1.

The resulting cost growth factor is 2.3. This compares reasonably with an average cost growth factor of 2.6 for capital cost estimates made during the R&D phase for the plants in the Rand database.

The value^ assigned to each parameter are within the range of values contained in the pioneer plant database, and so it is not unreasonable to use these models. The regression models do not imply causality; however, they are an statistically-derived representation of the data base regarding chemical process plants. As such, the models can be expected to reasonably represent, at least in a qualitative sense, a new chemical process such as the copper oxide process.

7.4.2.1 Technology Demonstration Costs

The Rand model has implications for the cost of a demonstration plant. Two cost estimates were made to illustrate the effect of different applications on first-of-a-kind plant demonstration costs. One possible demonstration is assumed to be a 125 MW module associated with a coal-fired power plant burning an unwashed Illinois No. 6 coal (4.36 percent sulfur) and sulfuric acid recovery. The other case is also a 125 MW module, but with an unwashed Pittsburgh coal (2.15 percent sulfur) and sulfur recovery. The 125 MW size is selected because it is the likely size of a copper oxide train; for a 500 MW plant, the nominal assumption is that four 125 MW trains would be used, with a fifth serving as a spare. In the case of the demonstration plant, it is assumed that there is no spare as a cost saving measure (if the demonstration is done at an existing site, there may be an alternative existing FGD system).

Not surprisingly, the medium sulfur coal application results in lower costs than the high sulfur coal application. The base total capital costs (without contingency) for the systems are $24 and $20 million, for the Illinois and Pittsburgh coals, respectively. With a factor 2.3 cost growth, these estimates are increased to $56 and $47 million, respectively. The difference in application is estimated to result in a $9 million capital cost difference for a first-of-a-kind demonstration plant.

7.4.2.2 Comparison of First-of-a-Kind and Mature Plant Costs

The construction and operation of a pioneer first-of-a-kind plant can lead to significant cost reductions in subsequent plants, assuming that an effort is made to incorporate information about the pioneer plant into the design and operating philosophy of the subsequent plants. Improvements in capital costs can be substantial for subsequent plants with similar site conditions. For example, the demonstration of a relatively standard technology can result in 10-15 percent cost reduction in later plants, while demonstration of a highly innovative process can result in perhaps a 30 percent subsequent cost reduction (Hess, 1985).

An interesting exercise is to compare the results of the cost growth method for the copper oxide process with the capital costs estimated for a mature (say, "fifth-of-a-kind") plant. The probabilistic analysis of the copper oxide process in Chapter 5 was intended to represent uncertainty in a fifth-of-a-kind plant. In the case study used for that analysis, the mean capital cost of the copper oxide process was found to be $111 million (representing an 80 percent overall contingency factor). By comparison, the cost of a first-of-a-kind plant cost, based on a 2.3 cost growth factor, is about $170 million. A 30 percent reduction from this amount, which could be expected to result if an effort is made to apply

Table 43. Analysis of Cost Growth and Cost Improvement for the IGCC Systems

Description	Oxygen-blown KRW with Cold Gas Cleanup	Air-blown KRW with Hot Gas Cleanup	Air-blown Lurgi with Hot Gas Cleanup
Deterministic Est. (10% contingency)	1,490	1,130	1,170
Percent New	23.0	34.5	32.5
Impurities	2	3	3
Complexity	12	10	11
Inclusiveness	100	100	100
Process Development	1	1	1
Cost Growth Factor			
Proj Def = 7	1.86	1.98	2.00
Proj Def = 6	1.66	1.76	1.77
1st Plant Cost			
Proj Def = 7	2,770	2,230	2,330
Proj Def = 6	2,475	1,980	2,065
5th Plant Cost			
Proj Def = 7	1,940	1,560	1,632
Proj Def = 6	1,730	1,390	1,445
"Best Guess"	1,730	1,380	1,410
Mean	1,806	1,376	1,465

experience gained in the pioneer application to later plants, results in a cost of $120 million, which is reasonably close to the estimated mature plant cost of $111 million from the probabilistic analysis. This result indicates that the probabilistic risk assessment is consistent with a deterministic assessment based on pioneer plant cost growth and subsequent cost improvement for later plants.

7.4.3 Case Studies for the IGCC Systems

Analyses of cost growth for a first-of-a-kind plant and cost improvement to a fifth-of-a-kind plant were performed for the three IGCC systems using the same approach as for the study of the copper oxide process. The results are shown in Table 43. A 10 percent contingency factor is used for the base capital cost estimate, because this is the average contingency Rand reports for their cost database (Hess et al, 1987). As for the copper oxide case study, a 30 percent cost improvement is assumed from the first- to the fifth-of-a-kind plant.

The results obtained from the Rand model are highly sensitive to the assumption regarding "project definition." Hess and Myers (1989) point this out in a study prepared by Rand for GRI regarding cost growth in coal-to-SNG plants. The project definition may only take discrete integer values, according to the Rand studies. As indicated in Table 43, the difference in results between a project definition of 6 and 7 is on the order of several hundred dollars per kilowatt for the cost of a first-of-a-kind plant. When a project definition of 7 is used, the Rand cost growth model, combined with an assumption of 30 percent cost improvement between the first and fifth plants, yields a cost estimate $130/kW to $190/kW higher than the mean value from the probabilistic simulation. However, when a project definition of 6 is assumed, the results agree with $20/kW for the two air-blown systems, and with $80/kW for the oxygen-blown KRW-based system.

The probabilistic modeling approach and the Rand cost growth model are shown to yield qualitatively similar results. However, because the probabilistic approach is based on a disaggregated analysis of specific risk factors for innovative technologies, it provides a more detailed means to identify the potential sources of cost growth. For example, although the Rand model and assumptions regarding cost improvement may provide similar estimates of the central tendency of capital cost as the probabilistic approach, it does not provide the insight into the risk of extremely high costs for the Lurgi-based system associated with zinc ferrite sorbent performance. Furthermore, the Rand approach does not provide a technology-specific indication of the range of uncertainty associated with the cost estimate.

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8.0 CONCLUSIONS

Conclusions specific to the case studies of the innovative clean coal technologies are discussed briefly, followed by some closing remarks with respect to the methodological approach employed in this research.

8.1 Technology-Specific Conclusions

An integrated model of the copper oxide process has permitted the evaluation of interactions involving components of the copper oxide process, the pollution control system, and the power plant. These interactions, which can be overlooked if not included in a systematic modeling framework, significantly influence process costs. Identification of important interactions provides the basis for determining research priorities, such as evaluating the effects of increased bed height on sorbent circulation rate. The explicit characterization of uncertainty in the model provides additional insights that may be overlooked in deterministic analysis, as demonstrated with the air preheater sizing analysis. Integration of pre- and post-combustion pollution control measures can lead to significant cost savings with high sulfur coals, although results indicate that the copper oxide process has an increased comparative advantage over FGD/SCR systems on medium sulfur coals.

While the magnitude of cost savings may be greatest on medium sulfur coals, the copper oxide system appears to dominate FGD/SCR systems for all cases considered. Cost savings appear to be larger for sulfuric acid byproduct recovery systems, although the available markets may be more limited than for elemental sulfur. The availability of byproduct markets will be significant in determining the extent of process application and the pay-off from R&D. Probabilistic comparisons of innovative and conventional technologies provides quantitative information about the risk that a new technology may be more expensive, and the potential pay-off of process research. By explicitly considering uncertainties and key process interactions, the probabilistic engineering models can be used to improve research planning and ultimately to assist potential process adopters in decision making.

For the air-blown Lurgi-gasifier based IGCC system with hot gas cleanup, uncertainties in gasifier performance are the most significant factor affecting plant efficiency. Uncertainty regarding the performance of the zinc ferrite sorbent strongly affects variable operating costs. These results are based on characterizations of uncertainty in model parameters obtained from several technical experts. Nonlinearities in the engineering model, as well as the skewness of the uncertainties elicited from the technical experts, lead to skewed uncertainties in key measures of performance and cost. These factors also lead to higher cost outcomes than obtained from deterministic analysis. Furthermore, uncertainties in performance, capital cost, and operating costs all contributed significantly to uncertainty in the levelized cost of electricity.

With sufficiently detailed and integrated performance and cost models, probabilistic analysis of design trade-offs can reveal insights not easily obtained from point-estimate techniques. For example, while decreasing the pressure drop across the gas turbine fuel valve is certain to improve plant efficiency, there is a significant probability that the cost of electricity will also be higher, due to uncertainties regarding the capacity of the gasifier as a function of pressure.

Further research may reduce uncertainties in specific model parameters. For cases in which skewed input distributions substantially affect model results, reductions in the uncertainty in specific model parameters lead to reductions in the downside risk of the technology and an increase in the expected value for key measures of performance and cost. These outcomes can be used to bound research expenditures, and to identify specific process areas for which research would yield pay-offs. For example, reductions in uncertainties in performance and cost parameters of three process areas would reduce the mean cost of electricity and decrease the probability of extremely high cost outcomes for the Lurgi system.

The implications of the judgments of several experts were evaluated. Three experts provided judgments regarding uncertainty in the zinc ferrite process area, and two regarding the Lurgi gasifier process area. The implications of these alternative judgments were used to identify areas where robust conclusions regarding performance and cost could be made, and, conversely, areas where disagreement among the experts exists. The sources of disagreement in the zinc ferrite process area include sorbent sulfur loading capacity and sorbent replacement requirements. Substantial disagreement was found between the two Lurgi gasifier experts, based primarily on differing judgments regarding the gasifier air/coal ratio. These specific parameters should be the focus of further elicitations.

Correlation structures among input uncertainties for individual process technologies were not found to be important in the case studies evaluated here. However, when comparing two technologies or alternatives for a given technology, correlations between them due to uncertainties which are common to both may be important. Thus, probabilistic estimates of differences between alternatives must be based on proper pairing of samples from the probabilistic simulations.

For the oxygen-blown KRW-based system with cold gas cleanup, the major source of performance-related uncertainty is carbon conversion in the gasifier. While hot gas cleanup systems are often viewed as a major source of performance and cost risk, uncertainties in the gasification process area for both the oxygen-blown KRW and air-blown Lurgi systems were important contributors to uncertainty in overall process performance and cost. However, because the Lurgi system relies on the zinc ferrite desulfurization process for "bulk" desulfurization, it is more exposed to the risks associated with poor sorbent performance than the KRW system with hot gas cleanup. The air-blown KRW system utilizes the zinc ferrite process for "polishing" desulfurization, and thus imposes a lower sulfur removal burden on this process area. These risks were characterized quantitatively in a series of case studies. The air-blown KRW system exhibited typically less variance in performance and cost uncertainties than the other two systems. The air-blown KRW system is expected to have better gasification performance than the oxygen-blown KRW system. This is due to the expectation that the limestone sorbent used for in-bed desulfurization in the gasifier also acts as a catalyst, improving gasifier performance.

Both air-blown systems are likely to have higher variable operating costs associated with the hot gas cleanup systems. For the KRW system with hot gas cleanup, higher costs are due to the cost of limestone sorbent, the higher ash disposal burden, and the lack of a saleable byproduct. The Lurgi system may incur high zinc ferrite sorbent replacement costs associated with potentially poor sorbent performance, in interaction with other process uncertainties.

Both of the IGCC systems with hot gas cleanup offer significantly lower SO_2 emissions than the system with cold gas cleanup. The SO_2 emissions of all three systems are below current NSPS for coal-fired power plants. However, the two air-blown systems, and the Lurgi-based system in particular, suffer from unacceptably high NO_x emissions. The presence of ammonia in the fuel gas for both systems with hot gas cleanup leads to high fuel NO_x emissions. Emission control strategies to reduce the NO_x emissions from these systems were not evaluated quantitatively here. Two possible approaches include the used of post-combustion SCR or the development of staged rich/lean combustors to minimize NOx formation during combustion. SCR would increase plant costs and lead to a slight reduction in plant efficiency associated with the incremental increase in gas turbine backpressure and additional auxiliary power requirements. Published cost and performance information for the rich/lean combustor is scarce or nonexistent because this technology is in a very early stage of development. The NO_x emissions from IGCC systems featuring hot gas cleanup must be reduced prior to commercialization. It appears likely that SCR will be required for such systems, at least in the near term.

The Lurgi-based system is likely to have higher efficiency and lower capital and levelized costs than the oxygen-blown KRW system, but it will tend to have higher variable operating costs attributable to the zinc ferrite process area. However, further process research can improve the cost savings for the Lurgi system in all areas, even if further research also reduces uncertainties in the oxygen-blown KRW system.

The air-blown KRW system has higher efficiency, lower capital, fixed operating, and levelized costs, and higher variable operating costs than the oxygen-blown system with cold gas cleanup. The air-blown KRW system compares similarly to the Lurgi-based system, although the probabilities of cost savings are reduced than when compared to the oxygen-blown KRW system. Furthermore, if uncertainties in the Lurgi-based system are reduced from further research, the cost advantage of the KRW system is decreased substantially, although it still would be likely to have lower levelized costs.

8.2 Methodological Conclusions

Significant uncertainties inevitably surround advanced environmental control technologies in the early stages of development. Thus, engineering performance and cost models developed to evaluate process viability must be capable of adequately analyzing and displaying the consequences of these uncertainties. Toward this end, the probabilistic modeling capability described here allows the effect of uncertainties in multiple performance and cost parameters to be evaluated explicitly and systematically. The results give a measure of overall uncertainty in key model outputs, such as cost, and serve to identify the key process variables that contribute most to overall uncertainty.

A detailed approach to the elicitation of expert judgments was employed for the IGCC system case studies. The results of the elicitations indicate that process engineers are able to make detailed judgments regarding both the range and likelihood of outcomes for specific parameters. They also were able to provide detailed explanations for the basis of the judgments. In many cases, the judgments were skewed, representing the risks often associated with innovative process technologies.

Using probabilistic modeling techniques, explicit and quantitative characterizations of uncertainty in key measures of plant performance, emissions, and cost may be obtained, based on the judgments regarding model input uncertainties. The probabilistic results indicate the range of possible outcomes, the likelihood of obtaining particular outcomes, the risk of unfavorable outcomes, and the probability of pay-offs associated with favorable outcomes.

The probabilistic approach is shown to be superior to deterministic estimating approaches using contingency factors for estimating process performance and cost. The probabilistic approach is sensitive to the skewness of uncertainties that may exist in key performance and cost parameters. The implications of skewed distributions are not easily captured in deterministic estimates, due to the observed tendency to use most likely or median values, instead of mean values, as the "best guess" inputs to a point-estimate.

Probabilistic analysis is recommended as a replacement to deterministic cost estimating approaches. When explicitly evaluating uncertainties, contingency factors are not needed. Instead, budgetary cost estimates for capital and O&M costs can be selected based on the acceptable probability of cost overrun.

Using probabilistic modeling techniques, the key uncertainties that drive uncertainty in model output variables can be identified using regression techniques. Also, key groups of uncertain parameters, such as for performance uncertainties in individual process areas, can be identified using probabilistic sensitivity analysis. Conversely, unimportant uncertainties can be eliminated from further consideration using probabilistic screening analysis. This allows further research or estimating work to be prioritized where reductions in uncertainty would most significantly lead to reduced uncertainty in key measures of performance and cost.

Probabilistic comparisons between conventional and advanced technology can be used to estimate the likely cost savings and the risks of a new technology. Judgments about the outcome of further process research can be combined with probabilistic modeling to estimate the value of research information to a potential process adopter, and to estimate the cost-savings from process improvement. The value of the research, coupled with judgment about the extent and nature of technology diffusion, can be used to bound research expenditures. Whether research is feasible depends also on the costs of the first commercial scale demonstration plant. Because these costs are potentially large, care must be exercised in the selection of an appropriate first application.

Decisions regarding research strategies when faced with uncertainties may be quantitatively evaluated using decision models. Such models are used to capture the preferences of a decision maker with respect to risk and the timing of outcomes. Thus, robust research strategies can be identified based on the results of both the engineering and decision models.

When there are multiple experts, the implications of their judgments regarding uncertainties should be evaluated separately. Then, the results may be compared to determine if there are robust conclusions resulting from agreement, or if there are key areas of disagreement that warrant further study.

The results of probabilistic analysis are shown to be qualitatively similar to those obtained from the use of Rand's cost growth models. However, because probabilistic analysis is based on a disaggregated consideration of technology-specific uncertainties, it may yield different results on an absolute basis than the Rand approach. Furthermore, it allows identification of specific sources of uncertainty. Such insight is difficult to obtain with the Rand model.

The probabilistic modeling approach developed here has been applied to detailed case studies of selected IGCC systems. However, the approach is applicable to the evaluation of any process technology for which characterizations of uncertainty can be obtained.

Probabilistic modeling is shown here to be a versatile tool for technology evaluation, cost estimating, process design, risk assessment, research planning, and technology selection. Of course, as with any other modeling approach, probabilistic methods rely on data and judgments that must be provided by the user. To be sure, different judgments or assumptions can alter the results. But forcing process developers and evaluators to consider uncertainties explicitly (rather than ignore them) in probabilistic engineering models can help improve research planning and management by allowing the implications of alternative judgments to be tested. Indeed, the process of thinking about key parameter uncertainties, as inputs to a model, often is the most valuable component of this approach that fosters improved understanding of the systems being modeled.

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PROBABILISTIC MODELING OF INNOVATIVE CLEAN COAL TECHNOLOGIES: Implications for Technology Evaluation and Research Planning

VOLUME 2: Appendices

A DISSERTATION

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A.0 MODEL DOCUMENTATION

Appendix A contains details regarding the performance and cost models of the clean technologies discussed in the main body of the dissertation. A total of five such technologies are analyzed, and they include:

1) Pulverized coal (PC) power plant with flue gas desulfurization (FGD) for SO_2 control and selective catalytic reduction (SCR) for NO_x control;
2) PC power plant with the fluidized bed copper oxide process for simultaneous SO_2/NO_x control;
3) Oxygen-blown fluidized bed gasifier-based integrated gasification combined cycle (IGCC) system with cold gas cleanup;
4) Air-blown fluidized bed gasifier-based IGCC system with hot gas cleanup; and
5) Air-blown fixed bed gasifier-based IGCC system with hot gas cleanup.

The performance and cost models of the PC plant with FGD/SCR used in this study were previously developed by Rubin et al (1986) and are not discussed in this Appendix. The performance and cost models for the fluidized bed copper oxide process were originally developed by Frey (1987). These models are described in Chapter 3. Two major modifications to the model were made for this research. These include: (1) the development of a more detailed representation of the reaction by which the copper oxide process removes SO_2 from the flue gas of a coal-fired power plant; and (2) the development of a performance and cost model of a Claus sulfur recovery plant as an alternative to sulfuric acid recovery assumed in previous studies.

The major focus of model development in this research has been the adaptation of a set of three previously existing IGCC performance models and the development of new cost models for each system. As described in Chapter 3, the U.S. Department of Energy (DOE) had previously developed performance models for the three IGCC systems selected for evaluation in this research. The performance models were developed in the ASPEN chemical process simulation environment. These models were obtained and adapted for use here. A number of modifications were made to the ASPEN performance models to include aspects of process performance not previously modeled, to improve some of the existing process area models, and to cleanup and organize the models.

Cost models for each of the three IGCC systems were developed based on a detailed review of approximately 30 published performance and cost studies of IGCC and coal-to-substitute natural gas systems, as well as review of other related design studies. The models characterize capital, operating and maintenance (O&M), and levelized costs.

Section A.1 provides information on the modifications made to the performance models of the copper oxide process and IGCC systems. The remaining sections provide detailed documentation of the newly developed IGCC cost models.

A.1 Performance Models

This section describes newly developed features added to the performance models of the innovative clean coal technologies evaluated in this study. In particular, two new features were added to the performance model of the fluidized bed copper oxide process, and a number of new features were added to the models of the three IGCC systems.

A.1.1 Fluidized Bed Copper Oxide Process

The two major modifications to the fluidized bed copper oxide process, compared to the model reported by Frey (1987), include the development of a more detailed representation of the SO_2 absorption reaction by which flue gas is desulfurized, and the addition of a Claus plant model to investigate the costs of byproduct sulfur recovery.

A.1.1.1 Sulfation Model

Yeh and Drummond (1986) developed a model of the sulfation reaction in the fluidized bed copper oxide process which may be written as:

$$p = \frac{\left(1 - \frac{p_o V_o M}{G\ F C_o}\right) p_o}{\exp\left[\frac{k D A G Z C_o}{M V_o}\left(1 - \frac{p_o V_o M}{G\ F C_o}\right)\right] - \frac{p_o V_o M}{G\ F C_o}} \tag{A-1}$$

where,

p = fractional partial pressure of SO_2 at absorber outlet
p_o = fractional partial pressure of SO_2 at absorber inlet
V_o = inlet flue gas volumetric flow rate, m^3/hr
M = molecular weight of CuO, kg/kgmole
G = molar volume of gas at reaction temperature, m^3/kgmole
F = sorbent feed rate, kg Al_2O_3/hr
C_o = initial CuO content, kg CuO/kg Al_2O_3
k = reaction rate constant, 1/hr
D = fluidized bed (expanded) density, kg Al_2O_3/m^3
A = reactor cross-sectional area, m^2
Z = expanded bed depth, m.

The terms on the right-hand side of the model may be redefined as follows:

$$r = \frac{p_o V_o M}{G\ F C_o} = \frac{\text{Inlet } SO_2\text{, kgmole/hr}}{\text{Inlet CuO, kgmole/hr}} \tag{A-2}$$

$$B = k\left(\frac{k\,D\,A\,G\,Z\,C_o}{M\,V_o}\right) = k\left(\frac{\text{kgmole CuO in fluidized bed}}{\text{kgmole/hr flue gas flow rate}}\right) \quad \text{(A-3)}$$

Equation (A-2) is the inverse of the available copper-to-sulfur molar ratio. Equation (A-3) is the ratio of the copper oxide resident in the bed to the incoming flue gas molar flowrate, multiplied by the reaction rate constant. Equation (A-1) may be rewritten in terms of Equations (A-2) and (A-3) and the SO_2 removal efficiency as follows:

$$\eta_s = \frac{p_o - p}{p_o} = \frac{\exp[B(1-r)] - 1}{\exp[B(1-r)] - r} \quad \text{(A-4)}$$

This is a convenient formulation in the case where we wish to calculate the removal efficiency in the case when regeneration efficiency is 100 percent and B and r are known. However, more typically, we desire to calculate the value of r required to meet a specified removal efficiency, for arbitrary regeneration efficiencies.

The term C_o in Equation (A-1), which appears in both r and B, is interpreted to be the available copper content (as copper oxide) of the sorbent. In the case where regeneration of the sorbent is complete, the available copper content will be the same as the sorbent copper loading. However, in the more likely case where regeneration is incomplete, the available copper will be less than the sorbent copper loading. Thus, an expression for C_o was developed which includes regeneration efficiency as a parameter. The derivation of this expression is based on equations for the alumina oxide mass flow rate in the sorbent and the available copper flow rate in the sorbent. Recall from Frey (1987) that the sorbent flow rate entering the absorber is given by:

$$m_s = \left\{\frac{MW_{Cu}}{x_{Cu}}\left[\frac{1}{r} + \left(\frac{\eta_s}{\eta_r}\right)(1 - \eta_r)(1 + 1.260\,x_{Cu})\right]\right\} M_{SO_2} \quad \text{(A-5)}$$

where,

m_s = mass flow rate of sorbent entering the absorber, lb/hr
MW_{Cu} = molecular weight of copper, lb/lbmole
x_{Cu} = weight fraction of available copper in sorbent (kg Cu/kg fresh sorbent)
η_r = regeneration efficiency, fraction
M_{SO_2} = Sulfur dioxide flow rate entering the absorber, lbmole/hr

This equation represents the total mass flow rate of three species in the sorbent: copper oxide, unregenerated copper sulfate, and the alumina oxide substrate. The mass flow rate of copper oxide is given by:

$$m_{CuO} = \frac{MW_{CuO}}{r} M_{SO_2} \quad \text{(A-6)}$$

and the mass flow of copper sulfate, which results from incomplete regeneration, is given by:

$$m_{CuSO_4} = \left[MW_{CuSO_4}\left(\frac{\eta_s}{\eta_r}\right)(1 - \eta_r)\right] M_{SO_2} \tag{A-7}$$

The mass flow of the alumina substrate is given by the difference between the total sorbent mass flow rate and the mass flow rates of copper oxide and copper sulfate in the sorbent:

$$m_{Al_2O_3} = \left\{\frac{MW_{Cu}}{x_{Cu}}\left[\frac{1}{r} + \left(\frac{\eta_s}{\eta_r}\right)(1 - \eta_r)(1 + 1.260\, x_{Cu})\right] - \frac{MW_{CuO}}{r} - M_{CuSO_4}\left(\frac{\eta_s}{\eta_r}\right)(1 - \eta_r)\right\} M_{SO_2} \tag{A-8}$$

The mass ratio of the available copper (copper oxide) to the alumina oxide substrate, which is defined as C_O, is then given by the ratio of Equations (A-7) and (A-9). After expanding, and then collecting terms, and approximating the ratio of the molecular weights of copper sulfate to copper oxide to be 2, rather than 2.006, the following expression results:

$$C_o = \frac{x_{Cu}}{(R_c - x_{Cu})\left[1 + r\left(\frac{\eta_s}{\eta_r}\right)(1 - \eta_r)\right]} \tag{A-9}$$

where,

R_C = ratio of the molecular weights of Cu and CuO

In the limit where the regeneration efficiency is 100 percent, Equation (A-9) reduces to:

$$C_o = \frac{x_{Cu}}{(R_c - x_{Cu})} \tag{A-10}$$

However, when regeneration is less than complete, the weight ratio of actual copper oxide to alumina substrate is shown to depend on the sulfur-to-available copper molar ratio, r, and the sulfur dioxide removal efficiency, as well as the weight percent copper in fresh sorbent.

Substituting Equation (A-9) into the value of C_O in B, defining r as the molar ratio of sulfur flow rate in the flue gas to *available* copper (copper as copper oxide) flow rate entering the absorber, and rewriting, we obtain the following expression:

$$\eta_s = \frac{\exp\left\{\alpha\left[\frac{1-r}{1+r\left(\frac{\eta_s}{\eta_r}\right)(1-\eta_r)}\right]\right\} - 1}{\exp\left\{\alpha\left[\frac{1-r}{1+r\left(\frac{\eta_s}{\eta_r}\right)(1-\eta_r)}\right]\right\} - r} \tag{A-11}$$

where,

$$\alpha = \frac{k\,D\,A\,G\,Z}{M\,V_o}\left(\frac{x_{Cu}}{R_c - x_{Cu}}\right) \tag{A-12}$$

In most applications, we wish to solve Equation (A-11) for r as a function of η_s. However, an interative numerical technique is required to obtain the solution. A simple technique is Newton's method.

The rate constant used in the calculations is estimated as follows:

$$k = A\exp\left(-\frac{E}{R\,T}\right) \tag{A-13}$$

where,

A	=	frequence factor, 1/hr
E	=	activation energy, KJ/gmole
R	=	universal gas constant, 0.008314 KJ/(gmole-K)
T	=	absolute reaction temperature, K

Equation (A-13) is the reaction rate constant as a function of frequency factor, activation energy, and temperature. From a separate paper (Yeh, Strakey, and Joubert), the activation energy of a UOP copper oxide sorbent was reported to be 20.1 KJ/gmole.

Yeh and Drummond (1986) report three values of the frequency factor as a function of the sorbent copper loading in terms of the percent copper in fresh sorbent. An equation for the frequency factor was developed using regression analysis. The three data points indicate a non-linear relationship between frequency factor and sorbent copper loading. The following equation was found to provide good agreement with the data:

$$A = 94{,}400\ 10^{-6.18x_{Cu}} \qquad R^2 = 0.998 \tag{A-14}$$

In Equation (A-14), the sorbent copper loading is the weight fraction of copper as copper oxide in the sorbent, to maintain consistency with the model.

A.1.1.2 Sulfur Byproduct Recovery Plant

A performance and cost model of a Claus plant was developed to estimate sulfur byproduct recovery costs and process performance for the PETC copper oxide SO_2/NO_x control system. The performance and cost model of the copper oxide process developed by Frey (1987) originally included a sulfuric acid plant for byproduct recovery. The Claus plant is used to convert sulfur dioxide in the copper oxide process offgas to elemental sulfur, instead of sulfuric acid.

The Claus plant model is based on an Allied Chemical SO_2 Reduction process design documented by Ratafia-Brown (1983). The performance and cost information for this design was submitted by Allied Chemical Corporation in response to a specified copper oxide process offgas composition, temperature, pressure, and flow rate. A standard Claus plant usually processes a gas stream containing hydrogen sulfide, a portion of which is combusted to form sulfur dioxide. The hydrogen sulfide and sulfur dioxide are converted to elemental sulfur via the Claus reaction:

$$2\ H_2S + SO_2 \rightarrow 3\ S + 2\ H_2O \qquad \text{(A-15)}$$

However, the copper oxide process offgas does not contain hydrogen sulfide. Therefore, a portion of the SO_2 in the offgas must be reduced with natural gas to produce the required quantity of hydrogen sulfide. This is accomplished via the following reaction:

$$2\ CH_4 + 3\ SO_2 \rightarrow S + 2\ H_2S + 2\ CO_2 + 2\ H_2O \qquad \text{(A-16)}$$

Thus, some of the elemental sulfur is obtained via the reducing reaction, while the remainder is obtained via the Claus reaction. The overall reaction is:

$$CH_4 + 2\ SO_2 \rightarrow 2\ S + CO_2 + 2\ H_2O \qquad \text{(A-17)}$$

Thus, the required molar flow rate of methane is one-half the molar flow rate of sulfur dioxide in the offgas.

The Allied Chemical sulfur recovery plant design includes a reduction stage using two packed-bed, cyclic heat exchangers and a catalyst packed reactor. The gas stream then flows through a two-stage Claus plant, where sulfur is recovered for byproduct sale. Allied recommended that the water content of the copper oxide offgas must be reduced prior to treatment in the sulfur recovery plant. Therefore, gas cooling and water removal prior to the sulfur recovery plant is assumed in the design reported by Ratafia-Brown.

The performance model estimates the molar flow rate of the offgas entering the sulfur recovery plant, based on an assumed reduction in the water vapor content of the gas to about 6 percent of the total gas volume flow rate. For the specific case reported by Ratafia-Brown, this implies about 92 percent removal of water vapor from the offgas. The amount of methane required for the reducing reaction is one-half of the inlet molar sulfur dioxide flow rate, as previously discussed. Because the copper oxide sorbent is regenerated using methane, any residual methane in the offgas is deducted from the total methane requirement to determine the flow rate of methane into the reducing stage. The electric power consumption of the Claus plant is scaled based on the copper oxide process offgas flow rate. Power is consumed in an offgas compressor and in air blowers. The amount of sulfur recovered by the process is estimated based on the sulfur entering in the copper oxide offgas and the sulfur recovery efficiency of the byproduct plant. This efficiency is approximately 95 percent. This contrasts with the efficiency of over 99 percent achievable for the sulfur acid plant byproduct recovery system. The lower efficiency for the elemental sulfur recovery system requires that the copper oxide process must be operated at a higher flue gas SO_2 removal efficiency to maintain overall sulfur emissions at a given level. Alternatively, Claus plant tail gas treating can be used to reduce sulfur emissions from the Claus plant and increase the sulfur recovery efficiency. This latter option was acknowldegded but not considered in the study by Ratafia-Brown.

The cost of the sulfur recovery system is estimated based on the direct cost of the sulfur plant gas pretreatment equipment and the Allied Chemical SO_2 Reduction and Claus plant system. The cost of the pretreatment section is based on an "exponential scaling rule" with the offgas flow rate as the predictive parameter. The cost of the Allied Chemical SO_2 Reduction and Claus plant system is scaled to the inlet flow of gas to the reduction unit of the plant after gas treating. The cost of Claus plants has been shown to scale with mass flow rates using an exponent of approximately 0.7 in other studies (e.g., EPA, 1983).

The total capital cost is estimated based on the direct costs discussed above and indirect costs. The indirect costs include general facilities, engineering and home office fees, and project and process contingency.

The annual costs of the sulfur recovery plant include methane and power consumption and sulfur byproduct credit. The model currently does not include costs for catalyst replacement, although the initial catalyst charge is included in the capital costs.

A.1.2 Integrated Gasification Combined Cycle Systems

The newly added features for the ASPEN IGCC simulation models are summarized here. Some of the features are general to all three IGCC systems, while a few are specific to just one or two flowsheets. These cases are noted. The changes are discussed by process area. The affected process areas include: gasification, external desulfurization, and gas turbine. In addition, a few modifications were made with respect to the way design parameters are initialized and how performance results are summarized.

A.1.2.1 Gasification Process Area

The modifications to the gasification process include specification of carbon and sulfur conversion, and specification of ammonia yield.

Oxygen-Blown KRW-based System

The design pressure of the gasification process area, which affects a number of unit operation blocks in the ASPEN input file for this system, are initialized in one Fortran block. Thus, changes in the assumptions regarding system pressures can easily be made. A new design specification is used to adjust the amount of carbon in the inlet coal that is allocated to fines and bottom ash. The amount of carbon that is not retained in fines or bottom ash is fully consumed in the gasifier for fuel gas production. Therefore, carbon conversion is specified by appropriately selecting the fraction of carbon in the coal feed that is retained in fines and ash. Similarly, a portion of the sulfur in the coal is not gasified. A design specification is used to specify the fraction of sulfur in the coal feed that is retained in the bottom ash and fines.

Air-Blown KRW-based System

As with the oxygen-blown KRW system, gasification system pressures for a number of unit operation blocks are centrally specified using a Fortran block. In addition, gasification temperature is also initialized in the Fortran block, as are parameters regarding the molar Ca/S ratio for limestone sorbent used for in-bed desulfurization, the fraction of sulfur captured during in-bed desulfurization, overall carbon conversion in the gasifier, and the gasifier oxidant and steam ratios. A new Fortran block is used to set the carbon conversion in the gasifier, by adjusting a mass flow split fraction in a unit operation block.

A new Fortran block was added to specify the amount of ammonia that is produced during gasification. The Fortran block sets the value of a parameter used in the gasifier unit operation reaction model to achieve an ammonia yield according the the equivalent fraction of coal-bound nitrogen that would be converted, as specified by the user.

A new set of unit operation blocks was added to represent the sulfation unit, which is a process area required for proper treatment of spent limestone sorbent prior to disposal. The DOE flowsheet provided no characterization of this system, and improperly assumed that calcium in the limestone which reacts with sulfur in the coal and recycle gas from the zinc ferrite desulfurization unit would be converted to calcium sulfate. However, in the reducing atmosphere of the gasifier, calcium would react with sulfur to form calcium sulfide. Therefore, the in-bed desulfurization reaction model was adjusted accordingly, and calcium sulfide (CaS), which was not previously included as a chemical species in the model, had to be added to the chemical species component list of the input file.

Spent sorbent containing calcium sulfide, and gasifier bottom ash containing unconverted carbon, would be sent to a circulating fluidized bed combustor to convert at least a portion of the calcium sulfide to calcium sulfate and to burn unconverted carbon. The energy released in these reactions can be used to generate steam for the plant steam cycle. A portion of the energy released would not be available for steam generation, because of losses within the boiler and because of sensible heat losses in the hot flue gas leaving the boiler.

Therefore, to characterize the sulfation unit, new unit operation blocks were added to represent the reaction of calcium sulfide to calcium sulfate and conversion of carbon in the bottom ash to CO_2. The conversion rates of these reactions are specified in a Fortran block. A fraction of the heat released in the sulfation unit, representing the boiler efficiency, is then sent, via a heat stream, to the plant steam cycle to generate high pressure steam.

Air-Blown Lurgi Gasification Process Area

The gasification process area pressures are initialized in a central Fortran block, to facilitate sensitivity analysis. The gasifier pressure for the Lurgi-based system is estimated based on the gas turbine combustor pressure and the pressure losses between the gasifier and the gas turbine. As with the air-blown KRW-based system, a new Fortran block was added to specify the ammonia yield from the gasifier.

A.1.2.2 Zinc Ferrite Desulfurization Process Area

For both of the air-blown IGCC systems, fixed bed zinc ferrite desulfurization is used to remove most of the hydrogen sulfide from the fuel gas. However, the original DOE model for the Lurgi-based system included a model of the moving bed zinc ferrite process. Cost data could not be obtained for the moving bed process, and therefore it was decided to replace the moving bed model with the fixed bed model. Therefore, both of the air-blown IGCC systems use the same model for zinc ferrite desulfurization.

In addition, a modificaiion to the model was made to more completely represent the regeneration process. The original DOE model was based on a single step oxidative regeneration in which oxygen in an air stream reacts with the sorbent to generate an off-gas containing SO_2. Steam is also added as a thermal diluent to prevent the reaction temperature from becoming too high and sintering the sorbent. However, oxidative regeneration is expected to leave some sulfur still retained in the sorbent as sulfates. These sulfates must be reduced prior to the next absorption cycle. Thus, a second reductive regeneration step is required. The reactants for this step may be either hydrogen or carbon monoxide, which are both available in the coal-derived fuel gas. Therefore, fuel gas has been proposed as the reactant for the reductive regeneration step.

The fuel gas requirement for reductive regeneration is estimated based on the amount of residual sulfate that would remain in the sorbent after oxidative regeneration, which is a new parameter added to the model. For each mole of sulfate in the sorbent, four moles of either hydrogen or carbon monixide are required as a reactant in reductive regeneration (Kasper, 1988). Therefore, the fuel gas requirement is estimated based on the fraction of absorbed sulfur that remains as sulfate, and the molar requirement for hydrogen and carbon monoxide. New unit operation blocks were added to represent reductive regeneration by converted the required amounts of hydrogen and carbon monoxide to water vapor and carbon dioxide, respectively, as would occur during reductive regeneration. Thus, the efficiency of the IGCC plant is sensitive to the energy penalty associated with reductive regeneration, which reduces the heating value of the fuel gas entering the gas turbine combustor. The effect, however, is slight in most cases.

A.1.2.3 Gas Turbine Process Area

The gas turbine process area models for all three IGCC systems were substantially modified. All key design and performance parameters affecting the gas turbine process area are initialized in a single Fortran block, rather than in individual unit operation blocks. This greatly simplifies the specification of sensitivity and probabilistic case studies. These parameters include inlet turbine ("firing") temperature, pressure ratio (which affects about 10 unit operation blocks), NO_x emission parameters, CO emissions, pressure drops in inlet and exit ducting, pressure drop in the gas turbine combustor, and the assumed isentropic efficiencies for the gas turbine model compressor and turbine stages.

Performance model changes include specification of choked conditions at the turbine inlet nozzle, which affects the gas turbine compressor inlet air requirement, characterization of NO_x and CO emissions, and the addition of more detail regarding cooling air circuity.

Estimating Compressor and Turbine Mass Flows

In Appendix B.6.3, the assumption regarding choked conditions at the turbine inlet nozzle is discussed. The DOE models previously assumed that the gas turbine compressor inlet air was fixed. Therefore, changes in fuel heating value, or in the amount of compressor air extracted for gasification blast air, did not affect the compressor air demand. However, they substantially affected the mass flow of gases through the turbine. Because of the low heating value of coal-derived fuel gases, the mass flow of gases through the turbine will usually be greater than that through the compressor, even with gasifier blast air extraction. Therefore, specifying the compressor air flow as fixed will usually lead to an unrealistically large flow rate of gases in the turbine. Furthermore, the gas turbine cost model is based on a particular type of gas turbine. Therefore, for the performance and cost models to be properly matched, it is important to obtain proper sizing of the flows in the gas turbine.

As discussed in Appendix B.6.3, the expected operating practice for gas turbines in IGCC service is to adjust the air flow through the gas turbine compressor such that the flow at the turbine inlet nozzle is (approximately) choked. This usually involves the use of compressor inlet guide vanes to adjust the compressor air flow based on fuel flow and compressor air extraction (if any) to obtain design flow in the turbine.

Therefore, the approach taken here is to add a new design specification to the ASPEN IGCC performance models which adjusts the compressor inlet air flow rate to obtain choked air flow at the turbine inlet. The turbine inlet nozzle air flow is referenced to that of available data for the General Electric MS7001F operating on natural gas, based on published values of gas turbine inlet air and exhaust gas flow rates, gas turbine pressure ratio, and turbine inlet temperature (Allen, 1990; Brandt, 1989). However, the flow rate at the turbine inlet nozzle is less than the exhaust flow rate, because a portion of the compressor air is introduced in later turbine stages as part of the cooling air circuits. Therefore, the reference turbine inlet nozzle mass flow was estimated assuming that 12 percent of the compressor air was diverted for downstream turbine blade and vane cooling. The compressor air diverted for cooling is based on an estimate provided by a DOE engineer (Geiling, 1991).

Because a reference mass flow is assumed, the critical area of the nozzle does not need to be explicitly estimated. Instead, the choked mass flow rate for IGCC applications is estimated based on the reference mass flow adjusted for differences in gas pressure, temperature, and molecular weight. The adjustment is made according to Equation (B-1) presented in B.6.3, which is reproduced here for convenience:

$$m_{max} = P\,A^{*}\sqrt{\frac{MW}{T}}\sqrt{\frac{\gamma}{R}\left(\frac{2}{\gamma+1}\right)^{\frac{\gamma+1}{\gamma-1}}} \qquad \text{(B-1)}$$

where,

m_{max} = maximum mass flow rate
P = total pressure
A^* = critical area where flow is choked
MW = molecular weight of gas
T = total temperature
R = universal gas constant
γ = ratio of specific heats for the gas

Assuming that the term under the radical is approximately constant for both the reference case with natural gas firing and the IGCC cases, the mass flow of gas entering the turbine nozzle can be estimated for different pressure, temperature, and gas molecular weight:

$$m_{act} = m_{ref}\left(\frac{P_{act}}{P_{ref}}\right)\sqrt{\left(\frac{MW_{act}}{MW_{ref}}\right)\left(\frac{T_{ref}}{T_{act}}\right)} \quad \text{(A-18)}$$

The new design specification adjusts the compressor air flow so that the ratio of the actual turbine inlet gas flow to the reference value, adjusted for temperature, pressure, and gas molecular weight, approaches unity to within a specified tolerance.

The effect of this new design specification is that the turbine inlet nozzle mass flow rate remains relatively constant even for varying values of fuel gas heating value and compressor air extraction. Thus, the gas turbine is more properly sized compared to the cost model.

Gas Turbine Emissions

The original DOE flowsheets did not characterize the emissions from the gas turbine process area. In particular, the primary pollutants of concern are CO and NO_x. Technical background on gas turbine emissions is given in Appendix B.6.5.

CO emissions can be easily characterized in the ASPEN simulation by specifying the fractional conversion of fuel gas CO in the gas turbine combustor to be less than 100 percent. Therefore, appropriate conversion rates were selected in the case studies to obtain desired estimates of CO emissions. CO emitted from the gas turbine represents incomplete combustion and an energy penalty on the combustor, because the heating value for conversion of CO to CO_2 is not realized. However, the energy penalty associated with the CO emission rates used in the cases studies in this work is typically negligible.

NO_x emissions were not characterized in the original DOE models. NO_x is comprised of NO and NO_2, and usually 95 percent of NO_x is in the form of NO when emitted at the stack. NO_x is obtained from thermal fixation of nitrogen in the combustor inlet air and from conversion of nitrogen-bearing compounds in the fuel. In IGCC systems, particularly those with hot gas cleanup, the most significant fuel-bound nitrogen species is ammonia. Therefore, reactions representing both thermal NO_x formation and fuel NO_x formation were added to the gas turbine combustor model. These reactions were written in a general form to accomodate any assumptions regarding the fraction of NO_x in the form of NO and the fractional conversion of combustor air nitrogen or ammonia to NO_x.

For thermal NO_x formation, the following equation is used:

$$N_2 + (2\text{-}x)\ O_2 \rightarrow 2x\ NO + 2\ (1\text{-}x)\ NO_2 \qquad \text{(A-19)}$$

where x is the molar fraction of NO_x that is in the form of NO. The fractional conversion rate of N_2 to NO_x is specified through a parameter in the combustor unit operation block in the ASPEN simulation.

For fuel NOx formation, the following equation is used:

$$2\ NH_3 + \left[\frac{3}{2} + xy + 2y(1\text{-}x)\right] O_2 \rightarrow (1 - y)N_2 + 2xy\ NO + 2y(1 - x)\ NO_2 + 3\ H_2O \qquad \text{(A-20)}$$

where y is the fractional conversion of ammonia to NO_x.

Cooling Air Circuitry

The turbine requires cooling air to keep the bulk metal temperatures of the blades and vanes, particularly in the first and second stages, sufficiently low to allow for long component life. See Appendix B.6.6 for more discussion. The DOE performance models of the IGCC systems contained provision for specifying cooling air flows. However, the models contained only two stages of unit operation blocks to represent the turbine. Furthermore, the model did not appear to account for pressure drops between the compressor air extraction point and the point at which the cooling air would be injected into the turbine gas stream. Therefore, a third stage was added to allow for better representation of the split of cooling air flows among the turbine stages and of the associated pressure drops. In addition, the total amount of compressor inlet air used for turbine blade and vane cooling was adjusted to 12 percent.

Model Validation

An effort was made to compare the results of the gas turbine performance model with published values for the performance of the MS7001F gas turbine: For these comparisons, the model was run using natural gas as a fuel. At a compressor inlet air temperature of 59°F, the model estimated that the gas turbine inlet air flow rate would be 919 lb/sec, the exhaust flow would be 938 lb/sec, the efficiency would be 34.68 percent, and that the power output would be 145 MW. The published values for the MS7001F are an inlet air flow of 981.7 lb/sec, and exhaust flow of 937.5 lb/sec, an efficiency of 34.57 percent, and a power output of 150 MW. While the model does not exactly reproduce the reported design values, it does estimate efficiency within 0.4 percent and plant output within about 3 percent. The model also is qualitatively consistent with the performance of gas turbines at varying ambient temperatures. As ambient temperature increases, both efficiency and the net power output decrease. However, no published data are currenlty available to compare with the model results at varying ambient temperature.

A.2 Background for IGCC Cost Models

This appendix discusses the basis for the development of cost models for three integrated gasification combined cycle (IGCC) systems. These systems include an oxygen-blown KRW-based system with cold gas cleanup, an air-blown KRW-based system with hot gas cleanup, and an air-blown Lurgi-based system with hot gas cleanup. These technologies are discussed in Chapter 3.

The cost models developed here are intended to be used for conceptual design studies for the purpose of evaluating generic features of process performance and cost. The modeling philosophy is discussed further in the next section. The cost models were developed based on published sources of cost information; these are discussed in Section A.2.2. The cost models are implemented with the IGCC performance models in the ASPEN modeling environment. Specifics related to model implementation are discussed in Section A.2.3. Many of the direct cost, operating cost, and auxiliary power requirement models were developed using regression analysis. The role of regression analysis is discussed in Section A.2.4.

The detailed documentation of the cost models is given in Appendices A.3 through A.8, including models for direct capital cost, total capital cost, fixed operating cost, variable operating cost, and annualized cost of electricity. The cost models are also summarized in Chapter 3.

A.2.1 Modeling Philosophy

There are a variety of approaches to developing cost estimates for process plants. These approaches differ in the level of detail with which costs are disaggregated into separate line items, as well as in the simplicity or complexity of analytic relationships used to estimate line item costs. The level of detail appropriate for the cost estimate depends on: (1) the state of technology development for the process of interest; and (2) the intended use of the cost estimates. The models developed here are intended to estimate the costs of innovative coal-to-electricity systems for the purpose of evaluating the comparative economics of alternative process configurations. The models are intended to be used only for preliminary or "study grade" estimates using representative (generic) plant designs and parameters.

In the electric utility and chemical process industries, there are generally accepted guidelines regarding the approach to developing cost estimates. EPRI (1986) has defined four types of cost estimates: simplified, preliminary, detailed, and finalized. The cost estimates developed here are best described as "preliminary." The differences between different types of cost estimates are briefly described below.

A *simplified* cost estimate is based on information about major stream flow rates and design parameters from a simple process flow diagram. The cost information used in a simplified estimate typically includes published cost curves or scaling relationships for generic process areas or for the plant as a whole. A simplified cost estimate may also be based on adjusting costs from similar published or in-house work on the basis of a single performance parameter. A simplified estimate is thus sensitive to only one (or a few) major performance parameter(s), such as the coal feed rate or the plant electrical output.

A *preliminary* cost estimate is based on a more disaggregated consideration of the costs of specific process areas and specific equipment items. A preliminary estimate also includes the use of ratio or scaling relationships to adjust costs for a variety of operating conditions. The preliminary estimate is sensitive to a larger number of performance parameters (perhaps a few dozen) than the simplified estimate.

Detailed and *finalized* cost estimates are generally developed only for site-specific projects that are intended for construction. For a large process plant, these types of estimates may cost millions of dollars to prepare. They are based on vendor quotations for specific equipment costs in response to specifications developed by an architect/engineering firm.

For the purposes of evaluating alternative technologies, and for research planning, preliminary cost estimates are the most appropriate. Preliminary cost estimates are sensitive to the performance and design parameters that are most influential in affecting costs. Thus, the goal of this study is to develop preliminary cost estimates for the three selected IGCC systems under study.

A major constraint on cost model development is the availability of data from which to develop cost versus performance relationships for specific process areas or for major equipment items. Limitations of cost data availability for the three selected IGCC systems are discussed in the next section. Data from published studies can be used to develop cost models for specific process areas using regression analysis. Regression analysis is used extensively for cost model development in this study (and elsewhere). An overview of the key concepts of regression analysis, and the philosophy of this study in applying regression analysis is given in Section A.2.4.

Alternatively, cost models for process areas consisting of only one major equipment item can be based on published equipment cost curves, either in place of or as a supplement to regression analysis. This approach was taken in several cases, such as for developing a model for zinc ferrite system direct capital cost. Such cases are noted in the text in Appendices A.3, A.4, and A.5.

A.2.2 Cost Data Availability

The primary constraint on cost model development is the availability of performance and cost data from which to develop correlations between process performance and equipment capital cost. A summary of performance and cost studies of IGCC systems and components that are used for cost model development is given in Table A-1. Limitations of these studies include disaggregation of capital costs to only about ten major process areas, differences in the battery limits of process areas from one study to another, differences in process technology assumed for a given plant section, and differences in the cost method employed.

A number of studies have been performed for DOE, including conventional designs with cold gas cleanup and advanced alternatives for hot gas cleanup. Studies of conventional systems include a series of performance and cost studies for Texaco, Lurgi, and KRW systems prepared by Bechtel for Argonne National Laboratory (Bechtel, 1983a; 1983b; 1983c), and a detailed study of Lurgi-based systems for synthetic fuel gas production by Foster Wheeler (Zahnstecher, 1984). A study of a KRW-based system developed by MW. Kellogg (Bostwick et al, 1981) was not used because it contained insufficient performance data and the cost estimates were inconsistent with more recent studies. Detailed performance and cost studies of hot gas cleanup systems include studies by General Electric (Cincotta, 1984; Corman, 1986) and an overview of sulfur recovery methods by Parsons (O'Hara, Chow, and Findley, 1987). A study by Gilbert/Commonwealth (Klett et al, 1987) was not used for cost model development because it presents only normalized (unitless) relative costs. A study by Westinghouse was specific to a phased addition at a particular site, and the sections of the report that describe the gasification section were deleted pending patent action (WEC, 1983). A study of a Texaco-based IGCC (UTC, 1983) contains little detail on the cost estimate. The available costs in the DOE studies do not use any consistent cost method or categorization of capital cost sections, and many are partially or not applicable to the cases of interest in this research.

Table A-1. Guide to Studies Used for Cost Model Development

Report No.	Process [a]	Contractor [b]	Sponsor [c]	Date
ANL/FE-83-15	Texaco	Bechtel Group	DOE	1983
ANL/FE-83-16	BGC/L	Bechtel Group	DOE	1983
ANL/FE-83-17	KRW	Bechtel Group	DOE	1983
AP-2207	Texaco & Lurgi	Ralph M. Parsons	EPRI	1982
AP-3084	Texaco	Fluor Engineers	EPRI	1983
AP-3129	Shell	Fluor Engineers	EPRI	1983
AP-3486	Texaco	Fluor Engineers	EPRI	1984
AP-3980	BGC/L	Ralph M. Parsons	EPRI	1985
AP-4018	KRW	Fluor Engineers	EPRI	1985
AP-4395	Texaco	Fluor Engineers	EPRI	1986
AP-4826	Texaco	Fluor Technology	EPRI	1986
AP-5950	Texaco	Bechtel Group	EPRI	1988
AP-6011	BGC/L	Bechtel et al.	EPRI	1988
DOE/ET/14928	Lurgi	General Electric	DOE	1986
DOE/FE/05081	Lurgi	Foster Wheeler	DOE	1984
DOE/MC/20315	Lurgi	General Electric	DOE	1984
DOE/MC/21097	Sulfur Recovery	Ralph M. Parsons	DOE	1987
GRI-86/0009	KRW & Lurgi	Kellogg Rust Synfuels	GRI	1985
GRI-87/0154	KRW	Fluor Technology	GRI	1986
GRI-87/0155	KRW	Fluor Technology	GRI	1986
GRI-87/0156	Lurgi	Fluor Technology	GRI	1986
GRI-87/0159	Lurgi	Fluor Technology	GRI	1986
GRI-87/0160	KRW	Fluor Technology	GRI	1987
GRI-87/0169	KRW	Fluor Technology	GRI	1988
GS-6160	Texaco	Fluor Daniel	EPRI	1988
GS-6161	Shell	NUSCo	EPRI	1988
GS-6176	Shell	Florida Power & Light	EPRI	1989
GS-6283	Shell	BG&E	EPRI	1989
GS-6318	Dow	Fluor Daniel	EPRI	1989

[a] BGC/L=British Gas Corporation/Lurgi; KRW = Kellogg Rust Westinghouse
[b] BG&E = Baltimore Gas and Electric; NUSCo = Northeast Utilities Services Company
[c] DOE = U.S. Department of Energy; EPRI = Electric Power Research Institute; GRI = Gas Research Institute.

EPRI has sponsored the most extensive publicly available performance and cost estimates of IGCC systems, including Texaco, Shell, KRW, British Gas Corporation/Lurgi, and Dow based systems. One contractor, Fluor Engineers, Inc., has prepared estimates for KRW (Fluor, 1985), Texaco (Fluor, 1983a; 1984; 1986), and Shell systems (Fluor, 1983b). Fluor Technology has prepared estimates of the performance and cost of alternative combined cycle systems integrated with Texaco gasification technology (Fluor Technology, 1986). Fluor Daniel has prepared a study of a Texaco system with Kraftwerk Union gas turbines (Fluor Daniel, 1988) and a study of a Dow-based IGCC system (Fluor Daniel, 1989). Bechtel Group, Inc. prepared an estimate of Texaco-based systems in a recent comparison of IGCC and conventional pulverized coal systems (Bechtel, 1988). Bechtel Group, Inc., in cooperation with others, also prepared a site-specific study of a BGC/L-based plant for Virginia Power (Bechtel et al., 1988). The Ralph M. Parsons Company has prepared studies of Texaco and BGC/L systems (Parsons, 1982; 1985). A recent set of studies focuses on site-specific applications of various IGCC technologies. In addition to the Bechtel study for Virginia Power, these include studies conducted with Baltimore Gas and Electric, Florida Power and Light, and Northeast Utilities Service Company concerning phased construction of Shell-based IGCC systems (BGE, 1989; FPL, 1989; NUSCo, 1988). These recent site-specific studies disaggregate costs based on a different definition of plant sections and plant section battery limits than the previous generic-site cost studies. Cost information is also available regarding the Cool Water demonstration plant (Cool Water, 1982; 1986; 1988).

All of the EPRI estimates present direct material costs for major process areas, generally including equipment and material installation costs, direct installation labor costs, sales tax, and indirect field costs and home office engineering, but may differ in the definition and battery limits of each process area. Process contingencies on a process area basis are estimated, and an overall project contingency is used. The total capital requirement also includes costs for royalties, spare parts inventory, organization and startup, working capital, allowance for funds during construction, and land. The amount of detail in the capital cost estimates varies from one study to another.

Figure A-1 illustrates how an IGCC plant is divided into process areas in most of the EPRI capital cost estimates. This figure is based on the KRW-based IGCC system estimate prepared by Fluor (1985). Costs are reported only for ten major process areas, which are common to most of the studies for EPRI. Each of these process areas contains several specific pieces of equipment, which may vary from one system to another. For example, the KRW system uses a dry coal feed system, whereas the Texaco system

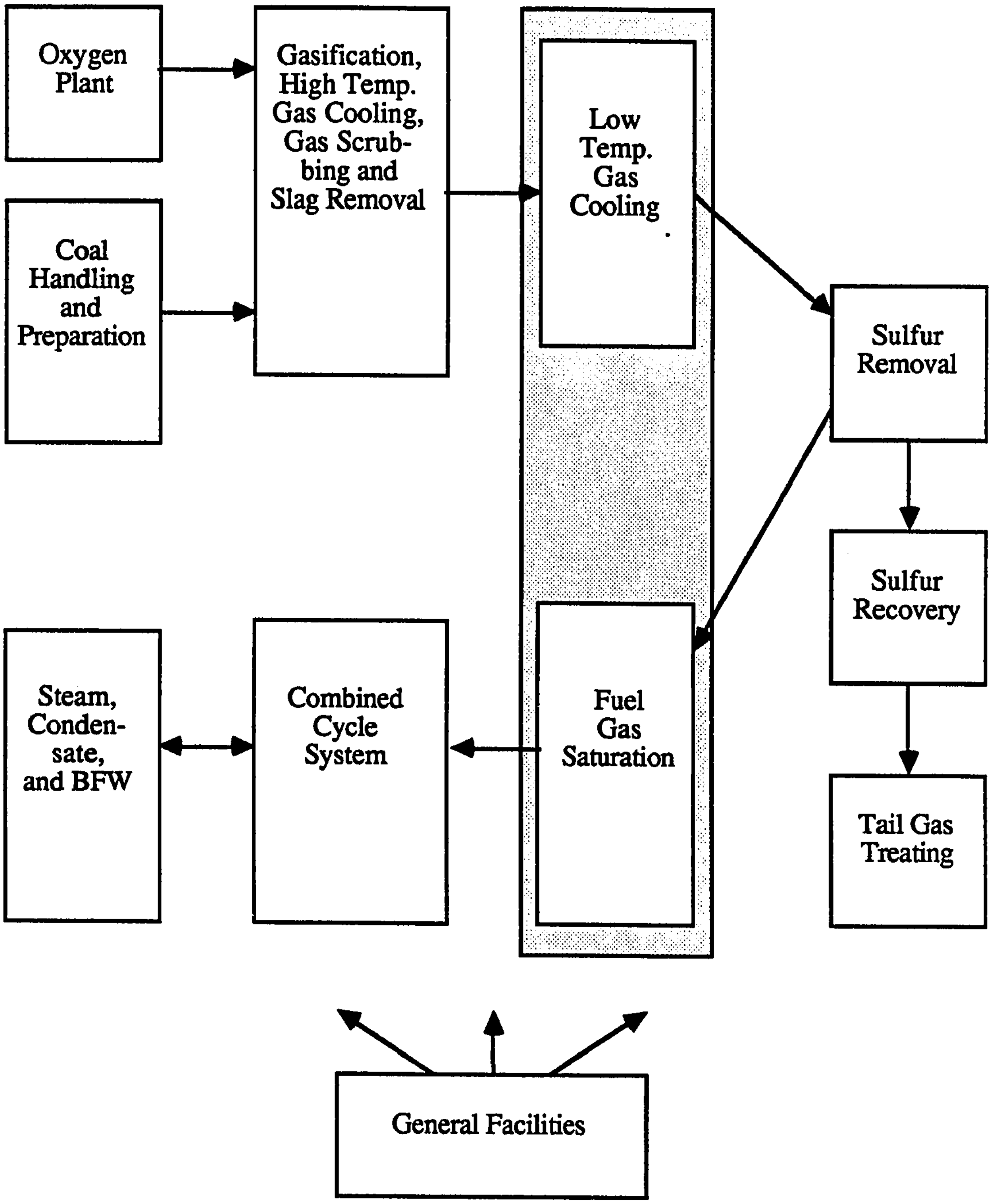

Figure A-1. Oxygen-blown KRW-based IGCC System Process Cost Areas uses a coal slurry feed system. Therefore, in many cases it is not possible to extract aggregated costs from one IGCC evaluation study and apply it to another system. Some reports contain more specifics; for example, the Bechtel report on a Texaco IGCC system includes a breakdown of general facilities costs into 20 line items.

Many of the cost estimates in the EPRI studies for specific equipment were developed in-house by EPRI's contractors or subcontractors. Manufacturers of some equipment were asked to submit quotes for a detailed commercial specification, while some quotes were obtained more informally by phone (e.g., Bechtel, 1988). General Electric provided Fluor with gas turbine combined cycle equipment costs. While Fluor considered about 350 separate items in their cost estimates of Texaco IGCC systems, much of the information is not included in the public report.

Several studies performed for the Gas Research Institute have been identified as useful for this research. These include evaluations of both KRW- and Lurgi-based systems. The studies of KRW-based systems include both coal-to-synthetic natural gas (SNG) facilities and IGCC plants, while the Lurgi-based studies are focused on coal-to-SNG facilities. Coal-to-SNG plants are similar to IGCC plants in the areas of coal feed, coal gasification, acid gas removal, and sulfur recovery. Kellogg Rust Synfuels, Inc. (KRSI) prepared the first detailed studies of coal-to-SNG systems for GRI, including both KRW and Lurgi gasification technology (Cover et al, 1985a; Cover et al, 1985b), and a comparison of how plant size affects the cost of both KRW and Lurgi-based systems (Cover et al, 1985c).

Fluor Technology has prepared all of the other studies of coal gasification systems for GRI. The studies of KRW-based coal-to-SNG plants include conversion of Western subbituminous coal to SNG (Smith, Hanny, and Smelser, 1986), conversion of Pittsburgh No. 8 coal to SNG (Smith and Smelser, 1987), and conversion of Pittsburgh No. 8 coal to SNG with reduced carbon conversion (Earley and Smelser, 1988a). Two studies, which focused on gas treating processes specific to coal-to-SNG systems, were based on the study of conversion of Western subbituminous coal to SNG, and contributed no additional information regarding gasification costs relevant to IGCC systems (Sandler and Smelser, 1987; Smelser et al, 1987). Three studies considered hot gas cleanup processes. These include performance and cost studies of: (1) a single train demonstration plant for either SNG production, electricity production, or both (Smelser, 1986a); (2) a commercial scale coal-to-SNG plant (Smith and Smelser, 1987); and (3) commercial scale coal-to-SNG and IGCC plants (Earley and Smelser, 1988b). The hot gas cleanup processes considered in these three studies include in-bed desulfurization in the gasifier, hot gas particulate removal using either sintered metal or ceramic candle filters, and hot gas desulfurization using the fixed bed zinc ferrite process.

The GRI-sponsored studies of Lurgi-based systems are based on conversion of lignite to SNG (Smelser, 1986b; 1986c). A summary of the GRI coal gasification technical analyses was recently published, which includes a brief summary of each of these studies (Smelser and Earley, 1988). All of the GRI studies are based on oxygen-blown systems. The coal-to-SNG plants are generally larger than IGCC plant sizes assumed in most studies, based on coal feed rate.

The capital cost model sections identified for each of the three selected IGCC systems are given in Table A-2. These cost model sections are based on a comparison of the design basis available for each design in the publicly available studies and the ASPEN performance simulations of each system.

While the readily available data forms a basis for developing a cost model, it is desirable to obtain more detail on the most costly and/or most risky components of an IGCC system. Of the major subsystems in the KRW-based IGCC cost estimates prepared by Fluor, the most costly subsystems are: (1) the combined cycle power generation system, (2) the gasifier, and (3) the oxidant feed system. These subsystems are typically the most expensive ones for any oxygen-blown IGCC system. Technical risk may also reside in the sulfur removal and recovery system, especially for advanced concepts such as in-situ or zinc ferrite systems. Furthermore, alternative zinc ferrite system designs, such as the moving bed system, may merit cost model development when a detailed design basis and cost data become available. These systems therefore represent priorities for data collection and model development.

The early stage of a data collection effort involves determining the appropriate level of detail for the cost model, and then identifying the key performance parameters that significantly influence costs, consistent with the chosen level of detail. *Because the readily available cost estimates are disaggregated only to major plant sections, initial efforts should be based on developing a model suited to the data availability constraints.* The cost models developed in this research consist of direct capital cost models for approximately ten major process areas for each IGCC technology, based on currently available performance and cost evaluations. Parameters of these models include key process flow rates whose values will be estimated from the ASPEN performance simulations. Variable and fixed costs are estimated based on fuel consumption, makeup material requirements, catalyst and

Table A-2. IGCC Capital Cost Model Sections

Area No.	Oxygen-Blown KRW with Cold Gas Cleanup	Air-Blown KRW with Hot Gas Cleanup	Air-Blown Lurgi with Hot Gas Cleanup
10	Oxygen Plant	Air Boost Compression	Air Boost Compression
20	Coal Handling	Coal Handling	Coal Handling
25	--	Limestone Handling	--
30	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasification, Coal Pressurization, Ash Depressurization
31	--	--	Coke Handling, Fines Agglomeration, Ash Removal
32	--	--	High Temperature Cyclones
35	--	Sulfation	--
40	Low Temperature Gas Cooling, Fuel Gas Saturation	--	--
50	Selexol Sulfur Removal	Zinc Ferrite Desulfurization	Zinc Ferrite Desulfurization
60	Claus Sulfur Recovery	--	Sulfuric Acid Plant
70	Tail Gas Treating	--	--
80	Steam, Condensate, Boiler Feed Water	Steam, Condensate, Boiler Feed Water	Steam, Condensate, Boiler Feed Water
85	Process Condensate Treatment	--	--
90	Combined Cycle	Combined Cycle	Combined Cycle
100	General Facilities	General Facilities	General Facilities

chemicals, other consumables, byproduct credits, maintenance materials, maintenance labor, and operating labor costs. The total revenue requirement for the plant is estimated using the commonly accepted EPRI costing method (EPRI, 1986).

A.2.3 Integration of Performance and Cost Models

The cost models must be compatible with the ASPEN IGCC flowsheet simulations. The performance models must include the key parameters that are required to determine capital and operating costs. Generally, the performance models estimate all the parameters that are required for the cost models with a few exceptions. However, a few process areas (such as cooling water makeup treatment chemical requirements) are not modeled in the IGCC performance models, some operating requirements are not modeled. However, these affect the annual costs of the system. Therefore, in the cases where the ASPEN models contain insufficient information for the cost models, regression models representing operating requirements are developed for the purpose of estimating annual costs. Table A-3 shows a comparison of parameters available in the ASPEN performance model of the KRW IGCC system with cold gas cleanup and parameter requirements for the capital cost model. Tables A-4 and A-5 provide comparable information for the KRW and Lurgi IGCC systems with hot gas cleanup.

ASPEN contains features that are potentially useful for estimating the costs of some process technologies. The ASPEN costing system consists of two main components: the cost estimation system (CES) and the economic evaluation system (EES). The former is used to estimate installed equipment costs and to track utility use within the plant. The latter is used to estimate capital investment, operating costs, and measures of profitability. The discussion in this section draws heavily on reports by Scientific Design Company (Schwint, 1985; 1986).

A.2.3.1 The ASPEN Cost Estimation System

The CES consists of a number of cost models, many of which correspond closely with unit operation blocks (e.g., pumps) and a number of others which are more general (e.g., vessels) which can be sized based either on a stream or a specific unit operation block (e.g., FLASH2). Each cost block contains a correlation for purchased equipment cost which is a function of one or more size parameters. The user may override the correlation with one of the form:

Table A-3. Major Plant Cost Sections, ASPEN Flowsheet Performance Sections, and Key Performance Parameters Affecting Cost for KRW IGCC With Cold Gas Cleanup

Area No.	Cost Section	ASPEN Flowsheet Section	Key Parameters
10	Oxygen Plant	N/A	Oxygen Flow Rate to Gasifier
20	Coal Handling	N/A	Coal Feed to Gasifier
30	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasifier Solids Separation	Coal Feed to Gasifier Syngas Flow Rate Gasifier Temperature, and Pressure, Coal Properties, Ash and Particulate Flow
40	Low Temperature Gas Cooling, Fuel Gas Saturation	Gas Processing Steam Cycle	Fuel Gas Flow Rate, Gas Temperature and Pressure
50	Selexol Sulfur Removal	Gas Processing	Fuel Gas Flow Rate, Hydrogen Sulfide Removal Efficiency
60	Claus Sulfur Recovery	Claus Plant	Claus Sulfur Production Rate
70	Tail Gas Treating	Beavon-Stretford	Beavon-Stretford Sulfur Production Rate
80	Steam, Condensate, Boiler Feed Water	Steam Cycle	Raw Water Input, Polished Water Flow Rate
85	Process Condensate Treatment	Solids Separation Effluent Water Primary Treatment	Scrubber Blowdown
90	Combined Cycle	Gas Turbine, Steam Cycle	Gas Turbine Model, Steam Flow Rate to Steam Turbine, Steam Turbine Electrical Output
100	General Facilities	N/A	Based on Other Costs

Table A-4. Major Plant Cost Sections, ASPEN Flowsheet Performance Sections, and Key Performance Parameters Affecting Cost for KRW IGCC with Hot Gas Cleanup

Area No.	Cost Section	ASPEN Flowsheet Section	Key Parameters
10	Boost Air Compression	Gas Turbine	Extracted Air Flow Rate
20	Coal Handling	N/A	Coal Feed to Gasifier
25	Limestone Handling	N/A	Limestone Feed to Gasifier
30	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasifier	Coal Feed to Gasifier
35	Sulfation	Gasifier	Ash and Spent Limestone Flow Rate from Gasifier
50	Zinc Ferrite Desulfurization	Zinc Ferrite	Sulfur Flow Rate in Syngas, Syngas Volume Flow Rate, Syngas Pressure
80	Steam, Condensate, Boiler Feed Water	Steam Cycle	Raw Water Input Polished Water Flow Rate
90	Combined Cycle	Gas Turbine	Gas Turbine Make and Model
		Steam Cycle	Steam Flow Rate to Steam Turbine, Steam Turbine Electrical Output
100	General Facilities	N/A	Based on Other Costs

Table A-5. Major Plant Cost Sections, ASPEN Flowsheet Performance Sections, and Key Performance Parameters Affecting Cost for Lurgi IGCC With Hot Gas Cleanup

Area No.	Cost Section	ASPEN Flowsheet Section	Key Parameters
10	Boost Air Compression	Gas Turbine	Extracted Air Flow Rate
20	Coal Handling	N/A	Coal Feed to Gasifier
30	Gasification, Coal Pressurization, Ash Depressurization	Gasifier	Coal Feed to Gasifier
31	Coke Handling, Fines Agglomeration, Ash Removal	Gasifier	Coal Feed to Gasifier
32	High Temperature Cyclones	N/A	Syngas Flow Rate, Syngas Pressure
50	Zinc Ferrite Desulfurization	Zinc Ferrite	Sulfur Flow Rate in Syngas, Syngas Volume Flow Rate, Syngas Pressure
60	Sulfuric Acid Plant	Zinc Ferrite	Off-Gas Flow Rate, Sulfur Dioxide Concentration, Off-Gas Temperature
80	Steam, Condensate, Boiler Feed Water	Steam Cycle	Raw Water Input Polished Water Flow Rate
90	Combined Cycle	Gas Turbine	Gas Turbine Make and Model
		Steam Cycle	Steam Flow Rate to Steam Turbine, Steam Turbine Electrical Output
100	General Facilities	N/A	Based on Other Costs

$$C = B\left(\frac{A_{cap}}{B_{cap}}\right)^n \tag{A-21}$$

where:

	C	=	Estimated cost at actual capacity
	B	=	Base cost at base capacity
	B_{cap}	=	Base capacity
	A_{cap}	=	Actual capacity
	n	=	Scaling exponent

and where capacity is a sizing parameter specified in the ASPEN User's Manual (MIT, 1987). The user-correlation form is not, in general, the form used by the ASPEN cost models; for example, the pump cost correlation is based on both the volumetric flow and the pressure head while the user correlation must be based only on the volumetric flow. Furthermore, a number of the cost models include several correlations to represent the breakdown of costs for complex systems (e.g., motor-driven pumps). While the user correlation may be useful for rough capital cost estimates, and while Equation (A-21) is often used by engineers to scale equipment costs, it lacks generality in terms of functional form. Furthermore, the user has no discretion over the input variable for the correlation.

Some of the major equipment in an IGCC system does not correspond directly to any of the existing ASPEN cost blocks. For example, while in principle a gas turbine could be costed by summing costs for a compressor, combustor vessel, turbine, and other components, in practice this would lead to unreliable results. A single cost block which models the cost of a gas turbine system is therefore more appropriate. Similarly, a heat recovery steam generator could be modeled as a collection of heat exchangers, but is more appropriately modeled as an integrated system. Although a gasifier could be modeled using one of the vessel cost models, the specialized nature of gasifiers implies that these cost correlations will not produce meaningful results. Therefore, the ASPEN CES is not an appropriate means for estimating IGCC capital costs.

Use of the ASPEN CES also results in an uneven level of detail (e.g., small pumps and large reactor vessels), and constrains the nature of the capital cost correlations that can be used. The lack of appropriate process area models (e.g., the gas turbine, HRSG, gasifier, and oxygen plant) indicates that either significant development work within the CES would be needed, or that an alternative approach should be developed.

While in principle existing ASPEN cost models could be modified for application to IGCC systems, such an effort is well beyond the scope of this research. As an alternative, Fortran blocks will be developed as needed to incorporate features essential to the capital cost model. This option is discussed below.

A.2.3.2 The ASPEN Economic Evaluation System

The Economic Evaluation System (EES) is used to determine operating costs, including the costs of imported utilities and raw materials. The EES can also be used to estimate the capital investment required to produce utilities on-site. In the case of an IGCC system, most or all of the utilities, except water, would be produced by the plant itself, except during startup. Utilities include water, steam, and electricity. Raw materials include both inputs and byproducts (e.g., coal, lube oil, Selexol solvent, sulfur).

The EES also calculates capital investment. For this purpose, the plant can be divided into cost sections. Each plant section can contain any number of cost blocks. In principle, if only one cost block contains a correlation, and the other cost blocks are specified as having zero capital cost, then the plant section cost would be based solely on the single cost block. The zero cost blocks might be used only to track utility use. Such an approach would minimize the number of blocks that would have to be created or modified, and could result in costs based on process areas and not specific equipment. However, limitations regarding the functional form and input variable selection for cost correlations would remain, and computation time would likely be excessive.

ASPEN uses an extensive cost indexing system to convert costs from one year to another for a variety of cost items (e.g., plant, equipment, labor, commodity, raw material, and fuel). Material and labor installation cost factors can be specified at either the block, section, or plant level to estimate installed costs based on purchased costs. There is provision to estimate costs for a variety of specific general facility items, including storage tanks, service buildings, site development, and land, plus keywords to specify indirect costs and additional depreciable or non-depreciable costs. There are also process and project contingency factor keywords. It appears that these last two can only be specified at the plant level. While many of these features are useful in calculating general facility costs and installation costs, the applicability of the EES depends on whether an IGCC cost model can be integrated within ASPEN.

Finally, ASPEN can be used to calculate the profitability of a plant. As a matter of convenience, it would be desirable to develop a cost summary similar in nature to the performance summary developed for the METC IGCC flowsheets. Such a summary would include annualized costs based on the EPRI costing method.

A.2.3.3 Cost Model Implementation via Fortran Blocks

There are three general approaches to developing a cost model of IGCC systems using the ASPEN performance models. They are:

1) Modify ASPEN to fully integrate cost correlations and cost models in a generalizable fashion;
2) Add user cost models and Fortran blocks to augment the existing CES and EES to facilitate calculation of special equipment costs and summary report generation; or
3) Do not use CES or EES. Instead, develop new Fortran block subroutines to access key flowsheet parameters, calculate costs based on correlations developed specifically for IGCC systems, incorporate uncertainty factors and error terms, calculate annualized costs according to EPRI guidelines, and write a summary report.

The first two approaches would utilize, to varying degrees, the current capabilities of the CES and EES. The first option, however, is beyond the scope of the current research to develop a cost model for IGCC systems. The second option is also less desirable since it would lead to potentially awkward or inefficient solutions.

The approach adopted here is to develop IGCC cost models independently of ASPEN, and to then use Fortran blocks to sample the key ASPEN performance variables needed to calculate the capital and annual costs. This approach requires the minimum resources to implement the cost models, and allows more effort to be devoted to the development and analysis of cost models. The cost models will be implemented as separate subroutines, to be linked with the ASPEN flowsheets via Fortran blocks.

A.2.4 Role of Regression Analysis

Regression analysis is used to help understand the interrelationships among a given set of variables. The use of regression analysis here is oriented toward developing useful and reasonable relationships between process area costs and key performance parameters. The emphasis is not on the use of extensive formal statistical tests but rather on the practical application of regression analysis for cost model development. Thus, some statistical tests, along with engineering judgments and the availability of data, are used to guide the selection of parameters, the representation of relationships in the regression models, and validation of the models. The "goodness" of the regression models are indicated witn common summary statistics, graphical comparison of the model predictions with the actual data, and evaluation of the appropriateness of the model relationships with *a priori* engineering expectations.

This section will briefly discuss issues related to developing and interpreting the regression models. The issues related to developing the regression models include developing a data set for analysis, selecting parameters for inclusion in the model, and validating the model. Specific issues related to the development and use of the models in this study are then discussed.

A.2.4.1 Overview of Multivariate Linear Least Squares Regression Analysis

The discussion in this section draws on Ang and Tang (1975), Chatterjee and Price (1977), DeGroot (1986), Dillon and Goldstein (1984), Edwards (1976), Montgomery and Peck (1982), and Weisberg (1985). An overview of key concepts is presented; details of multivariate regression can be obtained elsewhere in many texts such as the ones cited here.

In general, regression analysis involves describing the mean and variance of a random variable, Y, as a function of the value of another variable, X, or a set of variables $\mathbf{X}=(X_1, X_2, X_3,..., X_k)$. The variables in the vector $\mathbf{X}$ may take on specific values $\mathbf{x}=\mathbf{x_i}(x_{1,i}, x_{2,i}, x_{3,i},...,x_{k,i})$. For each value $\mathbf{x_i}$ in an actual data set, there is a corresponding value y_i. We use the notation $E(Y|\mathbf{X}=\mathbf{x_i})$ to indicate the mean, or expected value, of Y associated with a specific vector of values $\mathbf{x_i}$ of the variables $\mathbf{X}$. The notation $Var(Y|\mathbf{X})$ represents the conditional variance of Y on $\mathbf{X}$. If we expect that the value of Y can be estimated from a weighted linear combination of the k variables in $\mathbf{X}$, and if the conditional variance of Y is independent of the specific values $\mathbf{x_i}$ of $\mathbf{X}$, then:

$$E(Y|\mathbf{X}=\mathbf{x}) = a + b_1X_1 + b_2X_2 + b_3X_3 + \cdots + b_kX_k \qquad \text{(A-22)}$$

$$Var(Y|\mathbf{X}) = \sigma^2 = \text{constant} \qquad \text{(A-23)}$$

The parameters in the linear equation are estimated based on a limited number, n, of observed pairs of $(\mathbf{x_i},y_i)$, using multi-variable linear regression with constant variance. The linear regression model is written as:

$$E(Y'|\mathbf{X}=\mathbf{x_i}) = a + b_1X_1 + b_2X_2 + b_3X_3 + \cdots + b_kX_k \qquad \text{(A-24)}$$

or as:

$$Y' = a + b_1X_1 + b_2X_2 + b_3X_3 + \cdots + b_kX_k + \varepsilon \qquad \text{(A-25)}$$

The linear fit is usually obtained by selecting the values of a and b_i to minimize the sum of the square of the errors between $E(Y'|\mathbf{X}=\mathbf{x_i})$ from Equation (A-24) and the values of Y from actual data, y_i. Equation (A-25) differs from Equation (A-24) in that the model is used to predict the conditional random values of Y', rather than the conditional expected value of Y'. Equation (A-25) includes an error term, ε, that represents the variance in Y that is unexplained by the model. Thus, for a specific data point x_i, there is a corresponding data value y_i, a conditional mean value $E(Y'|X=x_i)$, and a conditional random distribution for Y'. Using the method of least squares, as documented in any standard text, we obtain estimates for the coefficients of the regression model. It is important to recall that the coefficients of the model, a and b_i, known as the partial regression coefficients, and the values of $E(Y'|X=x_i)$ or the parameters of the conditional distribution for Y' calculated using the model, are only estimates of the respective "true" population values of the parameters α and β_i and the "true" population of the values of Y associated with each value x_i.

Common statistical measures of the adequacy of the regression model in describing the data set (X,Y) include the standard error of the estimate, the coefficient of determination, the t-test for significance of each partial regression coefficient, and the F-test for the significance of the regression model and coefficient of determination. Confidence intervals, in addition to significance tests, can also be used. Proper application of these statistics requires the existence of certain properties in the data set (X,Y) and in the regression model. Several of these key assumptions are:

- random sample of n paired values (X,Y) (e.g., values of X are not preselected or screened)
- X and Y are multivariate normal
- for each value of x, there is an associated normal population of Y
- for each value of x, the variance of Y is constant
- no error in the measurement of X
- residual errors are not autocorrelated
- residual errors are normally distributed
- residual errors have constant variance

While these assumptions are often only approximately satisfied when developing regression models, the use of statistical evaluation methods based on these assumptions may provide some insight to guide the development of the model, even if a strict interpretation of the results is not correct. Therefore, blind application of significance tests to accept or reject parameters may not be appropriate. The most important consideration in selecting variables for use in a model, and for selecting the functional form of the model, is the analyst's knowledge of the substantive area under study and of each of the variables. The analyst will generally have expectations regarding the sign and magnitude of the coefficient for each variable, as well as which variables should be most significant in predicting the dependent variable.

The use of statistical tests is thus viewed here as an aid to, but not as a substitute for, the judgment of the analyst regarding the relationships among the variables. For example, it is common to test the significance of a model parameter by determining whether it is possible to reject a hypothesis that its coefficient is equal to zero. However, in many practical regression situations, it is known, based on theory or experience, that the coefficient must be greater than zero and, therefore, such a significance test is not particularly relevant. The potential inability to reject the hypothesis that a coefficient is zero in a regression model may be more an artifact of a small number of data points than due to a lack of relationship between Y and the predictive variable of concern.

Statistical tests are useful in identifying the independent variables which are relatively more important in predicting Y than others for the available data. For example, one can examine a correlation matrix of X and Y to determine which variables X_i are most highly correlated with Y. These variables are logical candidates for inclusion in the regression model. However, if a potential predictive variable X_i is also highly correlated with another variable X_j, then the inclusion of both may not significantly improve the model and may lead to counter-intuitive results in terms of the sign or magnitude of the coefficient for one of the variables. In such cases, one of the variables would be excluded from the model. Statistical tests can be used to identify independent variables that have only a weak predictive power. These variables would also typically be excluded from the model. A few of the statistical measures used to evaluate regression models will be discussed here, with an indication of how they are used in this study.

The issues of statistical tests and model validation are closely linked. Statistical tests are used to determine the adequacy of the model in representing a known data set. To the extent that the model is used only to interpolate information from within the data set, checking the adequacy of the model is the same as model validation. A regression model can be used for prediction beyond the range of the original data set only if there is some basis in prior experience, industry practice, or physical theory for the relationships between variables. If the form of the regression model is not based on theoretical or expert judgment about the relationship between the dependent and independent variables, the model should not be used for extrapolation. The user is cautioned that the primary purpose of the models developed in this study is for interpolation within the range of data values used to develop the models. Furthermore, the user is cautioned that the models are intended for application with very specific systems. Limitations on the ranges of predictive variables and discussion of the design basis for process areas are presented as part of cost model documentation in later sections of this appendix.

In using multiple regression models, it is easy to inadvertently extrapolate beyond the original domain for X, because that domain is jointly defined by the pairing of the values of each independent variable used to generate the model. Therefore, range checks on each independent variable separately will not guarantee the avoidance of "hidden" extrapolation. However, because the regression models are developed with some engineering basis for the relationship between variables, hidden extrapolation may be acceptable, and individual range checks on the independent variables will be used as a practical convenience.

Standard Error

The standard error of the estimate is the standard deviation of the residual errors ε for Y'. The standard error is a measure of the variability in Y that is not captured by the model. If the functional form of the regression model is "correct", this variability can be attributed to factors that are not quantified in the database and therefore cannot be investigated quantitatively. If the functional form of the regression model is not appropriate, then some portion of the standard error may be associated with an incorrect choice of the model, rather than unexplainable variability in the data set. Therefore, it is often useful to compare alternative functional forms of the model in terms of the standard error.

The standard error is estimated based on the residual sum of squares and the degrees of freedom of the residuals. The residual sum of squares is the sum of the squares of the difference between the values of $E(Y'|X=x_i)$ estimated by the model in Equation (A-24) and the values y_i from the data. The degrees of freedom of the regression model are the number of variables, k. The degrees of freedom of the residuals are the number of data points less the number of partial regression coefficients, including the intercept term. Thus, the standard error is given by:

$$s = \sqrt{\frac{\sum_{i=1}^{n} [E(Y'|X=x_i) - y_i]^2}{n-k-1}} \qquad \text{(A-26)}$$

This is an unbiased estimate of the standard deviation of the error. The error is assumed to be normally distributed with a mean of zero. In practice, this assumption may be difficult to verify, particularly for a small number of observed data points. Typical methods for evaluating the normality of the error include plotting the residuals against the fitted values $E(Y'|X=x_i)$, or plotting the errors on normal probability paper. A normality test may also be based on a one-sample Kolmogorov-Smirnov test (e.g., see DeGroot, 1984). In this test, the estimated cumulative probability distribution (cdf) for the errors is compared to a cdf based on the standard normal distribution. The maximum difference between the values of the sample and normal cdf's, adjusted for sample size, is the basis for estimating the test statistic. If the test statistic is larger than a specified value, based on the acceptable significance level for the test, then the hypothesis that the errors are normally distributed is rejected.

The estimate of the standard error is dependent on the actual data as well as the number of data points. As the quantity (n-k-1) becomes small, the estimate of the standard error will tend to increase. The standard error can be used to place a confidence interval on the values of Y' using Equation (A-24) or to generate conditional random values of Y' using Equation (A-25) and a probabilistic modeling capability. In the application of the regression models developed in this work, the standard error is used as a basis for generating conditional random values of Y'.

<u>Coefficient of Determination</u>

The most commonly used measure of the adequacy with which a regression model fits the data is the coefficient of determination, R^2, which is defined as:

$$R^2 = 1 - \frac{\sum_{i=1}^{n} \left(y_i - E(Y'|X=x_i)\right)^2}{\sum_{i=1}^{n} \left(y_i - E(Y')\right)^2} \qquad \text{(A-27)}$$

The numerator of the fractional term is the sum of the square of the residual errors between the actual data and the predicted conditional expected values of Y' from Equation (A-24). The denominator is the sum of the square of the differences between the actual data and the sample mean. The value of the coefficient of determination is interpreted as the proportion of the total variance in Y which is explained by the regression model, and it varies from 0 to 1, with values near 1 typically considered to represent "good" fits. The coefficient of determination is the square of the multiple correlation coefficient, R, between Y and the regression model. The multiple correlation coefficient is a measure of the degree of linear relationship between the dependent variable Y' and the linear combination of predictive variables.

The coefficient of determination is not a sufficient measure of the goodness of the model. At a minimum, evaluation of a regression model should include consideration of how reasonably the functional form and values of the coefficients represent the expected relationships between variables, the significance level of the coefficients and the regression model as a whole, and a graphical comparison of the model results with the actual data. The coefficient of determination may be highly influenced by extreme data points. If those data points are removed, the correlation coefficient may be drastically altered. The addition of a new data point may lead to a large change in the value of the coefficient of determination. Also, if the range of the predictive variables is reduced or increased, the correlation coefficient may change considerably.

Statistical Significance of the Model

It may be appropriate to consider a significance test for the correlation coefficient. A significance test based on the t-statistic can be used for this purpose to test the hypothesis that the correlation is not significantly different from zero. The hypothesis that a parameter is equal to zero is known as the null hypothesis. The likelihood that a parameter is significantly different from the null hypothesis is determined using a test statistic, such as the t-test. The value of the test statistic computed from the data is then compared to the value of the statistic estimated for the significance level of the test. It is common to use significance levels of 0.05 or 0.01 as the basis for comparison. If the probability of a obtaining a value of the test statistic is less than the significance level (e.g., 5 percent or 1 percent), then the null hypothesis is rejected as being sufficiently improbable that it is regarded as false.

The null hypothesis for the correlation coefficient is a hypothesis that the correlation is zero. A correlation of zero implies that the regression model is not useful, and that the best predictor for the value of Y is the mean of Y. Instead of doing a significance test, it is also possible to use a transformation of the correlation coefficient for use in developing a confidence interval for the correlation (Edwards, 1976). However, statistical tests on the correlation coefficient are related to statistical tests on the coefficient of determination. Furthermore, a test of the null hypothesis for the coefficient of determination is implicitly a test of the null hypothesis that the regression coefficients for the predictive parameters are all zero (Edwards, 1976; Dillon and Goldstein, 1984). This hypothesis is commonly tested using the F test statistic. Thus, an F test implicitly is a test of the null hypothesis for the coefficient of determination as well as for all of the partial regression coefficients simultaneously .

The F test involves first computing the F-ratio of the regression model, which is related to the coefficient of determination as follows:

$$F = \frac{\left(\frac{R^2}{k}\right)}{\left(\frac{1 - R^2}{n - k - 1}\right)} \tag{A-28}$$

As the coefficient of determination becomes large, the value of the F-ratio increases. The value of the F-ratio is then compared to the value of the F-distribution (published in many texts) for a selected significance level based on the degrees of freedom of the numerator (k) and denominator (n-k-1) of the F-ratio. Therefore, the F-test is influenced by both the number of data points and the number of predictive parameters included in the model. If the F-ratio is larger than the selected value of the F-distribution, then it is possible to reject the hypothesis that all the regression coefficients are equal to zero. However, rejection of this hypothesis does not imply that all of the regression coefficients are significantly different from zero; it only implies that at least one coefficient is significantly different from zero. Furthermore, even if the regression model is statistically significant, it may not necessarily be the best model of the data or even a theoretically valid model of the data. In this study, the F-ratio is compared to a significance level of 0.001 as the basis for rejecting the null hypothesis. In cases where the null hypothesis cannot be rejected at this significance level, but in which there is an engineering basis for preserving the regression model nonetheless, the significance level at which the model can be rejected is indicated.

To test the significance of individual regression coefficients, a commonly used technique is a t-test. For each regression coefficient, most computer regression packages will report the results of a t-test of the hypothesis that the individual regression coefficients are significantly different from zero. If a regression coefficient is not significantly different from zero, it can be deleted from the model with usually little effect on the residual error. In addition, the standard error for each coefficient is generally reported, which permits the evaluation of confidence intervals for the coefficients, using the t-distribution.

The regression models in this report have been developed using Statworks™, a statistics package for the Apple Macintosh™ computer. This package reports the statistics discussed here, and also facilitates use of the Kolmogorov-Smirnov test for testing the normality of the errors.

A.2.4.2 Application of Regression Analysis to Model Development

In general, the regression models developed here (and detailed in Appendix A) for process perfomrance and cost have high coefficients of determination and meet the F-test of significance at a significance level below 0.001. These results are not unexpected, because the development of the models is based on prior engineering knowledge of the primary relationships between performance, design, and cost. In this section, issues specifically related to the development of the regression models in this research are discussed. These issues relate to the number of observations available in each model data set, the use of transformation of variables to develop nonlinear models using linear regression, the selection of predictive parameters, the collection of data, and the reporting of results.

Number of Observations

The number of data points used to develop the regression model has an important effect on variable selection and interpretation of model results. As the number of data points becomes small, the number of independent variables that can be used may become constrained. It is often possible to obtain a model with a high coefficient of determination by selecting a large number of independent parameters; however, such a model may contain counter-intuitive relationships, or relationships that violate principles of engineering. This often occurs when the range of a predictive variable is small, when other important predictive variables have not been included in the model, or when there is correlation or collinearity between predictive variables. It is often appropriate to include only a small set of independent parameters that are expected to be fundamentally important and robust as more data are gathered, rather to include all possible variables for which data are currently available. To select the most important parameters, one may begin by including all possible predictive variables in the model. Those variables with regression coefficients that fail the t-test for significance are then deleted to yield a new model with fewer predictive variables. The deletion or inclusion of a variable may be tempered by judgment regarding relationships that must be included in the model, assuming that the coefficients of the particular variable are of the correct sign and magnitude.

For small numbers of data, the estimates for the standard error, and the significance levels for the F-ratio, will tend to increase, because the degrees of freedom are reduced. Therefore, confidence intervals on the regression coefficients and the estimate for Y will usually be larger than when more data are available. As more data become available, the regression models can be redeveloped. While the specific values of the regression coefficients would likely change, they would be expected to remain within the confidence intervals, unless the new data are from a different sample population than the original data.

In this case, the original regression model is not an appropriate representation of the new data. It is important, therefore, to ascertain if the basis for the new data is the same as for the older data (e.g., same design for process area equipment, same battery limits for the process area). In some cases, it may not be appropriate to add the new data without also including other predictive variables to capture the differences in the basis for the new and old data.

Transformation of Variables

While linear regression analysis has been used for all the regression model developed in this report, in many cases variable transformations have been used because the relationship between the dependent and predictive variables is non-linear. For example, the simplest cost model involves exponential scaling of a performance parameter, X, (such as a flowrate) to estimate cost, Y, as follows:

$$Y = aX^b \tag{A-29}$$

This functional form is standard in the chemical process industry, and cost capacity exponents for standard process plants are published in various sources (e.g., Peters and Timmerhaus, 1980; Ulrich, 1984; Humphreys and Wellman; 1987). The exponential scaling rule can be converted to linear form using the natural logarithm to transform the variables. A general assumption for the functional form of the cost models used here is a multi-variate extention of Equation (A–29) to k predictive performance variables:

$$Y = aX_1^{b_1}X_2^{b_2}\cdots X_k^{b_k} \tag{A-30}$$

This model represents the expected exponential scaling relationship between key process flowrates or design parameters and cost. In most cases, the exponent is expected to be less than one, representing the "economy of scale" of building larger units compared to smaller units. Typically, the exponent of one of the parameters will be much larger than for the other parameters. This result is expected, for example, when the flow rate of one material stream is expected to have a major influence in cost, while other parameters, such as temperature, may have only a secondary effect. The model in Equation (A-30) can be transformed to linear form using the natural logarithm:

$$\ln(Y) = \ln(a) + b_1\ln(X_1) + b_2\ln(X_2) + \cdots + b_k\ln(X_k) \tag{A-31}$$

A linear regression is then developed based on the transformed variables. The transformation of variables affects the interpretation of distribution of the errors. If the errors for Equation (A-31) are normally distributed, which is the underlying assumption for the statistical tests discussed in the previous section, then the errors for Equation (A-30) will be lognormally distributed. The statistical tests are applied to the transformed model of Equation (A-31). These cases are noted in the text.

Recall that the probability density function (pdf) for the normal distribution is given by:

$$f(x) = \frac{1}{\sigma\sqrt{2\pi}} \exp\left(-\frac{[x-\mu]^2}{2\sigma^2}\right); \quad -\infty \le x \le \infty \qquad \text{(A-32)}$$

where μ is the mean and σ is the standard deviation. If y is lognormally distributed, then ln(y) is normally distributed. The pdf for the lognormal distribution is given by:

$$f(y) = \frac{1}{\phi y\sqrt{2\pi}} \exp\left(-\frac{[\ln(y)-\xi]^2}{2\phi^2}\right); \quad 0 \le y \le \infty \qquad \text{(A-33)}$$

The parameters of the lognormal distribution are ξ and ϕ. These parameters correspond directly to the mean and standard deviation of the normal distribution for ln(y). The mean and variance of the lognormal distribution are given by:

$$\mu_y = \exp\left(\xi + \frac{\phi^2}{2}\right) \qquad \text{(A-34)}$$

$$\sigma_y^2 = \omega(\omega-1)\exp(2\xi) \qquad \text{(A-35)}$$

where,

$$\omega = \exp\left(\phi^2\right)$$

Using these relationships, the parameters of the lognormal distribution of errors for the nonlinear regression models can be estimated from the parameters of the normal distribution for the errors of the linearized model. Therefore, the statistical model based on the functional form in Equation (A-30) is given by:

$$Y' = aX_1^{b_1}X_2^{b_2}\cdots X_k^{b_k}\varepsilon \qquad \text{(A-36)}$$

where the error term is multiplicative and lognormal, not additive and normal as with the linear model in Equation (A-25). The mean of $\ln(\varepsilon)$ is zero and the standard deviation is the standard error of the estimate for the linearized model. These parameters for $\ln(\varepsilon)$ are used to estimate the mean and standard deviation for the lognormal distribution of ε using the relationships shown in Equations (A-33), (A-34), and (A-35). The median of the lognormal error term in Equation (A-36) will always be 1. The mean of the error term will typically be a value close to, but larger than, 1 and the standard deviation will typically be less than 1. The parameters that are reported for lognormal error terms for the appropriate models described in Appendix A are the mean and standard deviation given by Equations (A-34) and (A-35).

Two-Step Regressions

In many cases, the relationship between cost and performance parameters is expected to be nonlinear, as described by Equation (A-36). However, the cost is also directly proportional to the number of trains of equipment for a given process area. To capture both the nonlinear relationships between performance and cost and the linear relationship between the number of trains and cost, a two-step approach to developing the regression models may be required. The primary reason for the two-step approach is because it is not possible to specify that the exponent of the number of trains must be equal to one when developing the nonlinear model using linear regression packages. In the first step, a linearized regression of cost and performance parameters as just described is developed on the basis of a single train of equipment. In the second step, the predicted values from the nonlinear model for a single train are combined with information about the number of trains to predict the total cost of the process area. Thus, the final regression model from this process contains predictive variables for both performance and the number of total and operating trains.

The first step in the process involves estimating the coefficient and exponents of a model of the form of Equation (A-30) on the basis of a single train of equipment. The values of Y estimated in this fashion are then multiplied by the corresponding total number of trains to form a new predictive variable. This predictive variable is then used in a simple linear regression model. The first regression yields a model of the form:

$$Y^1 = a\left(\frac{X_1}{N_o}\right)^{b_1}\left(\frac{X_2}{N_o}\right)^{b_2}\cdots\left(\frac{X_k}{N_o}\right)^{b_k} \qquad \text{(A–37)}$$

Note that this is a general functional form; in some cases, the predictive performance parameters (such as temperature or pressure) do not depend on the number of trains, and therefore would not be divided by the number of operating trains. The estimated values of cost from Equation (A-37) represent the cost per operating train. However, we are ultimately interested in the total cost for the process area. Therefore, we calculate a new predictive variable which is the estimated cost for all operating and spare trains:

$$X^{(2)} = N_T\, Y^1 \qquad \text{(A-38)}$$

We then use this new variable as the basis for a simple linear regression of the form:

$$Y' = a^{(2)} + b^{(2)}X^{(2)} + \varepsilon \tag{A-39}$$

Typically, the value of $b^{(2)}$ from this model is close to 1.0. The value of $a^{(2)}$ may occasionally be small enough (or statistically insignificant) to exclude from the model by estimating the regression without a constant. Note that the error term here is in the linear space. If the errors conform to a hypothesis of normality, then the error can be represented as normally distributed with a mean of zero. Based on Equations (A-37), (A-38), and (A-39), we can write the final regression model as:

$$Y' = a^{(2)} + b^{(2)}\left[N_T\, a\left(\frac{X_1}{N_o}\right)^{b_1}\left(\frac{X_2}{N_o}\right)^{b_2}\cdots\left(\frac{X_k}{N_o}\right)^{b_k}\right] + \varepsilon \tag{A-40}$$

where the term within the square brackets is treated as a single variable in the simple linear regression. Thus, the first regression is essentially a method for grouping a number of performance parameters into a single aggregate predictive term, while the second regression permits the addition of the linear relationship between cost and the number of trains of equipment. This approach permits the calculation of model statistics based on total, rather than per train, process area costs, which are the ultimate measures of interest.

Selection of Predictive Variables

Direct capital cost regression models for each IGCC plant section, and in some cases estimates of annual consumable material and auxiliary material requirements, have been developed based on an analysis of approximately 30 detailed performance and cost studies of IGCC and coal-to-SNG (synthetic natural gas) systems, as discussed previously. These models have been developed based on analysis of plant section direct costs and key plant section performance parameters. In each regression model, the parameters selected for inclusion in the model and the analytic relationships between model inputs and outputs were based on engineering judgments, statistical analysis, and data availability. These regression models relate the total direct cost (which includes delivered equipment cost, installation labor, and installation materials) to the statistically most significant performance parameters influencing cost. These parameters are typically mass flow rates, although in some cases parameters such as removal efficiency, pressure, or temperature were found to be statistically significant. In cases where parameters that are expected to be important were not found to be statistically significant, the variation in these parameters often is small for the available data samples (e.g., most KRW gasifier designs are at about 450 psig and 1850 ^{o}F), or the variation in these parameters is highly correlated with variations in the statistically most significant parameter (e.g., the syngas output from the KRW gasification section is highly correlated with the coal feed rate). Similarly, some parameters that are expected to be important in influencing cost may yield counter-intuitive results in the regression models (e.g., cost inversely proportional to mass flow rate). This, too, occurs when two candidates for predictive parameters are highly correlated.

Collecting Data

For the IGCC cost models, performance and cost data were collected into separate data bases for each plant section, based on similarity of plant section definitions. Only direct equipment costs were collected. Direct costs include equipment, material, and labor costs associated with installing plant equipment. Because indirect costs are treated differently in different studies (e.g., EPRI vs. GRI), these were not included in the cost databases. All direct costs were adjusted to a common year using the Chemical Engineering plant cost index (January 1989 = 351.5). Because the studies varied in the amount of detail for each plant section, only a few performance parameters may be reported in common among studies for a given plant section. This limits the number of parameters that are candidates for regression analysis.

Reporting Results

For each plant section in the IGCC systems, the direct capital cost model is reported in Appendix A along with the error of the regression, the coefficient of determination, the number of data points used in developing the regression, and the range of values over which the regression was developed. The error term is typically expressed as a normal distribution with a mean of zero and a standard deviation estimated from the difference between the direct costs available in the literature and the direct costs estimated from the regression model. In cases where a non-linear variable transformation was used, the error is reported as a lognormal distribution. The error term provides a measure of the variance of the direct cost estimate. In principle, the variance would be zero if the model accounted for all the parameters that influence costs and if the model were of an appropriate functional form. However, because the models are simplified and include only one or a few parameters, not all of the variation in cost is captured. The variance represents differences in plant location, design, or performance parameters that are not included in the cost model.

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A.3 Capital Cost of a KRW-Based System with Cold Gas Cleanup

Figure A-2 shows an oxygen-blown KRW IGCC power plant with cold gas cleanup. The system shown in the figure is based on the design presented in a study for EPRI (Fluor, 1985) and the configuration of the ASPEN flowsheet of a KRW-based IGCC system with cold gas cleanup (Stone, 1985). In the gasification section, coal is partially oxidized in a reaction with oxygen and steam to produce a gas containing CO and H_2. Oxygen for the gasification reaction is provided by an air separation plant, and steam is provided from the plant steam cycle. The hot fuel gas leaving the gasifier is cooled by generating saturated steam in high temperature heat exchangers as part of the IGCC power plant steam cycle. Particulates in the fuel gas are removed by cyclones, which recycle fines to the gasifier, and by a particulate scrubber. The sulfur in the fuel gas, primarily H_2S, is removed in the acid gas removal section using the Selexol process, and the off-gas from this section, containing concentrated H_2S, is sent to a Claus plant for sulfur recovery. Prior to entering the acid gas removal section, the coal syngas must be cooled to approximately 100 ^oF. The clean fuel gas is reheated (not shown in the figure) and then combusted in gas turbines, which generate electric power. The hot exhaust gas from the gas turbines is used to superheat saturated steam from the gasification section and to reheat steam for the steam turbine, which also generates electricity. A portion of the electrical output from the generators must be used to power equipment in the plant, most notably the air separation plant. The remaining electricity is exported for sale. Overall, the plant consumes coal, air, and water, and produces ash, sulfur, flue gas, wastewater, and electricity. Although many individual components of IGCC systems are commercially proven in other applications, there is only limited experience with some specific components.

The IGCC plant is divided into 13 sections for the purpose of estimating total direct capital costs. These sections include the following:

1. Coal handling (Area 20)
2. Oxidant feed (Area 10)
3. Gasification (Area 30)
4. Gas cooling (Area 40)
5. Acid gas removal (Area 50)
6. Sulfur recovery (Area 60)

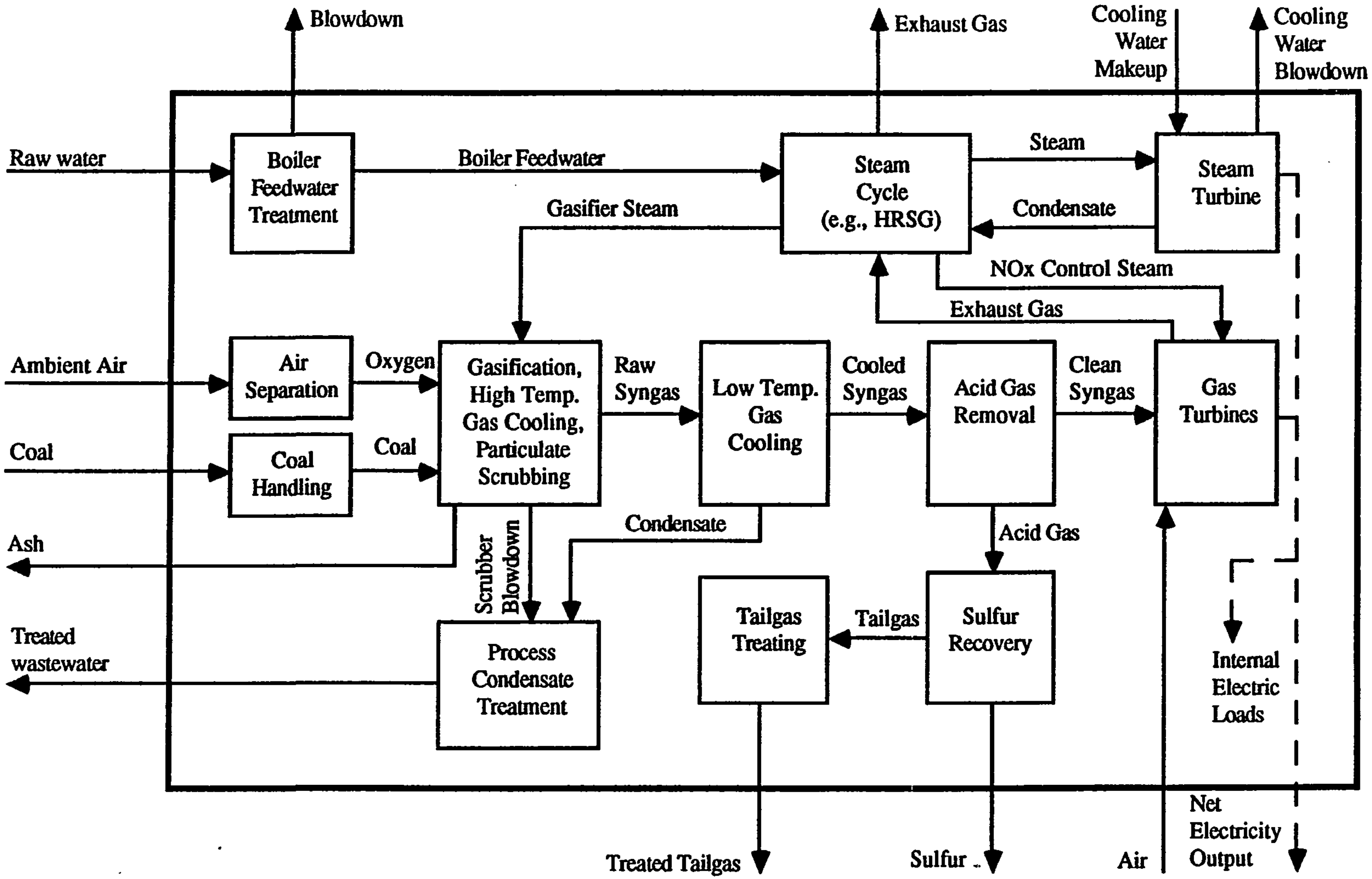

Figure A-2. Schematic of Oxygen-Blown KRW-based IGCC System with Cold Gas Cleanup

7. Tail gas treating (Area 70)
8. Boiler feedwater system (Area 80)
9. Process condensate treatment (Area 85)
10. Gas turbine (Area 91)
11. Heat recovery steam generators (HRSG) (Area 92)
12. Steam turbine (Area 93)
13. General facilities (Area 100)

The direct cost correlations for each plant section are described in the following sections of this report. While some of the process area models may be applicable to a variety of IGCC or coal-to-SNG systems, the models are intended for the specific purpose of estimating the direct cost of the KRW-based IGCC system shown in Figure A-2. The reader is cautioned that the models should not be used to estimate costs for systems with a design basis that differs from the assumptions discussed here. Nor, except when noted, should the models be extrapolated. The most salient features of the design basis for each process area are discussed in this report. However, the purpose here is not to recapitulate each detail of the process area design basis, but rather to document the development of the cost models. Therefore, the reader may wish to read this report in conjunction with some of the performance and cost studies cited here to obtain more detail about specific process areas. Perhaps the single most useful reference is the EPRI-sponsored study by Fluor (1985).

A.3.1 Coal Handling Section

The coal handling section for a KRW-based IGCC system must deliver dry pulverized coal to the coal surge bin in the gasification section. The design basis for the coal handling system assumed here is specific to KRW-based systems; therefore, the coal handling cost model should not be applied to other types of gasification systems. However, the KRW-based coal handling is generic to both IGCC and coal-to-SNG systems; therefore, data from conceptual design studies of both types of systems were used to develop the direct capital cost model. These studies include: Bechtel and WE, 1983; Fluor, 1985; Cover et al, 1985a; Smith et al, 1986; Smith and Smelser, 1987; and Earley and Smelser, 1988a. The first two of these studies are for IGCC systems, while the others are for coal-to-SNG systems. An additional study by Earley and Smelser (1988b) develops costs for a KRW-based IGCC system. This study uses the same coal handling design and cost estimate as Earley and Smelser (1988a).

The coal types represented in the data set include Illinois No. 6, Pittsburgh No. 8, Wyodak subbituminous, Texas lignite, and North Dakota lignite. The design basis for the coal handling section is similar across the KRW-based design studies. All equipment in the coal handling section is commercially available. The equipment includes bottom dump railroad car unloading hoppers, vibrating feeders, conveyors, belt scale, magnetic separator, sampling system, double boom stacker, bucket wheel reclaimer, surge bins, hammer mill, vibrating fluid bed dryer, circulating gas blower, baghouse, and dust suppression system. There is typically one train of equipment for coal receiving, and multiple trains for coal crushing and drying. However, detailed information on the size or number of trains of process area equipment is limited, particularly for the coal-to-SNG studies. In addition, some studies report only a total direct cost for coal receiving, crushing, and drying.

The coal throughput determines the sizing of most of the process area equipment. Therefore, this is the primary predictive parameter for direct cost. While other parameters, such as coal moisture content, also may affect cost, they are only secondary in importance. Only 15 data points for direct cost and coal mass throughput are available from the design studies of KRW-based systems. There are fewer than 15 data points for potential secondary predictive parameters, except for coal moisture content. However, there is a high correlation between coal moisture content and mass flow rate for this particular set of data. Because of these limitations in the available data, and because it is desirable to develop the regression model based on all of the available cost sample points, a single variable regression equation based on coal feed rate to the gasifier was developed. Both linear and nonlinear functional forms were investigated. A linear model yielded marginally better summary statistics (e.g., coefficient of determination, F-ratio, t-statistics for individual parameters) than a nonlinear exponential scaling model. About half of the data points are drawn from one study in which the cost of coal handling appears to scale linearly with coal mass flow rate. Therefore, a linear model for the direct cost of the coal handling process area as a function of the coal mass flow delivered to the gasifier is assumed:

$$DC_{CH} = 3{,}300 + 4.09\ m_{cf,G,i} \qquad (R^2 = 0.859;\ n = 15) \qquad \text{(A-41)}$$

where,

$$4{,}700 \leq m_{cf,G,i} \leq 23{,}000 \text{ tons/day}$$

The range of values for the coal mass flow indicates the limits within which the regression model should be applied. The model should not be used to predict coal handling direct cost for coal mass flow rates significantly outside this range. The standard error of the

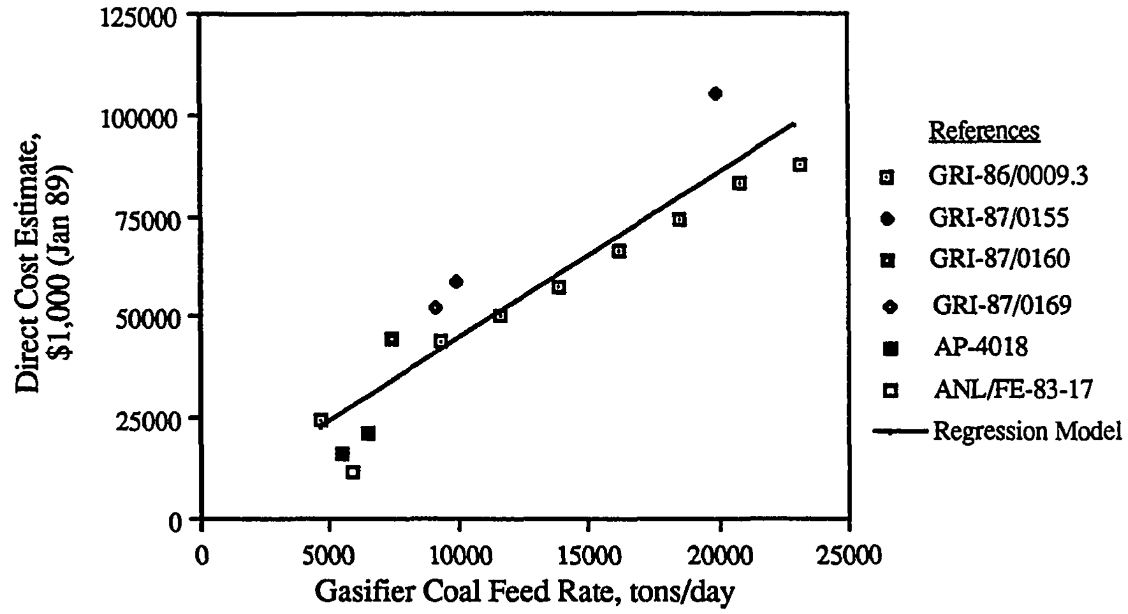

Figure A-3. Predicted and Actual Direct Costs for the Coal Handling Process Area

regression model is $10 million. Based on only these 15 data points, it is not possible to reject the hypothesis that the error is normally distributed at a significance level of 0.05 using a one-sample Kolmogorov-Smirnov test. The coal handling section cost data and regression model are shown graphically in Figure A-3. From the graph, it is apparent that the errors may not be normally distributed. Therefore, summary statistics for the regression model should be interpreted with care. However, the regression model does capture the key relationship between coal mass flow rate and coal handling system direct capital cost.

A.3.2 Oxidant Feed

The oxidant feed section is applicable to all oxygen-blown gasification systems. Performance and cost data for 31 oxygen plants were taken from 14 studies of oxygen-blown IGCC systems, all prepared for EPRI. These plants all include electric motor-driven compressors. Data from coal-to-SNG systems were not included because many of these use steam-driven, rather than motor-driven, compressors. Electric motor-driven systems offer advantages in terms of plant operation, although steam-driven systems may be more energy efficient. These plants produced between 625 and 11,350 lbmole/hr of oxygen per train. A typical plant consists of two parallel operating trains with no spare trains. Each train includes an air compression system, air-separation unit and an oxygen compression system. For more detail on the oxygen plant design, see Fluor (1985). The oxygen plants produce an oxidant feed to the gasifier containing typically 95 to 98 percent oxygen on a volume basis. It is possible to recover argon as a saleable byproduct from high purity oxygen plants (99.5 percent oxygen or greater) (BOC Cryoplants, 1987); however, the available data are not for the oxygen purity levels and plant designs required to do this. The oxygen plants represented in the database are considered commercially available.

The direct cost of oxygen plants is expected to depend mostly on the oxygen feed rate to the gasifiers, because the size and cost of compressors and the air separation systems are proportional to this flow rate. The oxygen purity of the oxidant feed stream is expected to affect the cost of the air separation system. As oxygen purity increases, it is expected that the cost of the oxygen plant will increase because the size of equipment in the air separation plant (e.g., high pressure column) increases (BOC Cryoplants, 1987). The ambient temperature determines the volume flow rate of air entering the inlet air compressor; as ambient temperature increases, the volume flow rate increases for a given mass flow, thereby requiring an increased compressor size.

A number of regression models were considered in which alternative combinations of predictive parameters and functional forms were assumed. These regressions were based on nonlinear variable transformations using the natural logarithm, as discussed in Chapter 2. A single-variate regression of cost and oxygen flow rate, using an exponential scaling formulation (see Equation (1) for an example), not unexpectedly yielded excellent results ($R^2 = 0.90$). The scaling exponent in this case was 0.9. The addition of terms for ambient temperature and oxidant purity yielded a marginal improvement in the summary statistics for the model. From an engineering viewpoint, the inclusion of these additional predictive terms significantly improves the utility of the model, allowing costs to be sensitive to both primary and secondary factors. Therefore, the following multi-variate regression is assumed for the oxidant feed process area direct capital cost:

$$DC_{OF} = 14.35 \frac{N_{T,OF}\, T_a^{0.067}}{(1 - \eta_{Ox})^{0.073}} \left(\frac{M_{O,G,i}}{N_{O,OF}}\right)^{0.852} \qquad (R^2 = 0.936;\ n = 31) \qquad \text{(A-42)}$$

where,

$$20 \leq T_a \leq 95$$

$$625 \leq \left(\frac{M_{O,G,i}}{N_{O,OF}}\right) \leq 11{,}350 \text{ lbmole/hr}$$

$$0.95 \leq \eta_{Ox} \leq 0.98$$

The robustness of the exponential scaling relationship between oxygen flow rate and direct capital cost is indicated by the similarity of the exponent for oxygen flow rate in the single and multi-variable regression models. In the single variable model previously described, the exponent was 0.9, while for the multivariate model above it is 0.85. The limits for each parameter indicated above represent the ranges for which the regression model is valid. While to obtain accurate results these ranges should not be violated, it is not a severe violation to exceed the range for the oxygen flow rate per train, particularly on the high side, because the model reasonably captures the expected relationship between oxygen flow rate and cost. An alternative to extrapolating the model for oxygen flow rate per train, however, is to alter the number of trains so that the flow rate per train is within the limits given above. The ambient temperature and oxygen purity parameters should not be extrapolated.

Regression of a linearized model (see Chapter 2) including a term for the number of trains was possible for this process area because the estimated exponent for the total number of trains was approximately 1.0. This result is expected because the cost of the oxidant feed plant should be a linear function of the number of trains. A one-sample Kolmogorov-Smirnov test indicated that the residual error of the linearized model is reasonably similar to a normal distribution. In the nonlinear form reported here, the error is therefore lognormally distributed, as discussed in Chapter 2. The mean of the lognormal error is 1.012 and the standard deviation is 0.155. The 90 percent probability range for this distribution is 0.78 to 1.29, implying a 90 percent confidence limit from 78 to 129 percent of the predicted values from the model.

A graphical representation of the model is shown in Figure A-4. Figure A-4 shows a comparison of the direct costs estimated by the regression model versus the costs available in the engineering studies. The dashed line in the figure represents the ideal case where the model would predict the study costs with no error. In the cases where a point lies above the reference line, the model over-predicts costs, and in the cases where a point lies below the reference line, the model under-predicts costs. The largest error in prediction occurs for two cases at the upper end of the cost range. These two cases represent designs with several trains (four in one case, five in the other) and costs based on older studies. If these two cases are excluded from the estimate of the model error, the standard deviation of the residual errors is reduced from \$10.8 million to \$8.7 million.

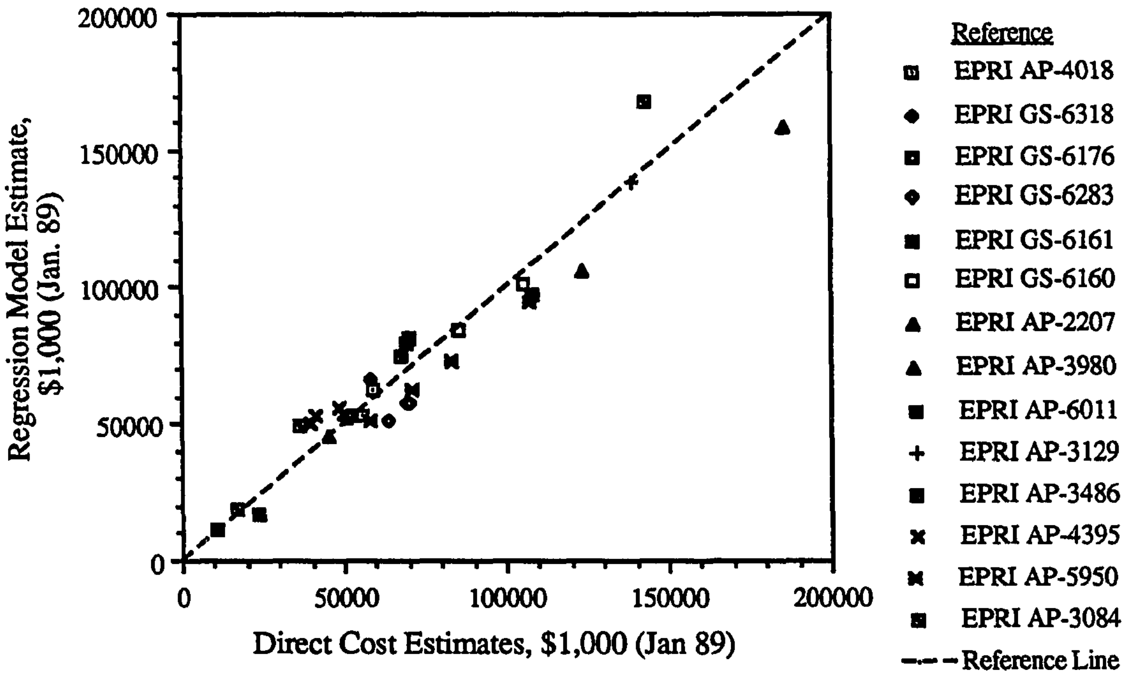

Figure A-4. Predicted vs. Report Direct Costs for the Oxidant Feed Section

A.3.3 Gasification, Ash Removal, High Temperature Gas Cooling, and Particulate Removal

For a KRW-based system, the gasification section includes equipment associated with coal pressurization, gasification, char recycle, syngas recycle, ash removal, high temperature gas cooling, and particulate removal from the syngas.

A.3.3.1 Technology Overview

The KRW gasifier is a pressurized fluidized bed, agglomerating dry ash design. The gasifier vessel is a refractory lined carbon steel pressure vessel. The gasifier has a number of different "zones," including low velocity ash cooling and removal, moderate velocity ash separation and gasification, and high velocity combustion and devolatilization. In addition, a freeboard disengaging zone is used to reduce the amount of char entrained in the outgoing syngas. The diameter of the gasifier is largest in the freeboard zone, in order to reduce syngas velocity and entrainment of char. Char leaving the gasifier is removed from the syngas by cyclones and recycled to the gasifier to improve overall carbon conversion efficiency (Smith et al, 1986).

No commercial scale KRW gasification system has yet been built. Kellogg Rust Synfuels, Inc. (KRSI) has operated a 24 ton coal per day process development unit (PDU) test facility in Waltz Mill, PA since 1975. By contrast, typical performance and cost studies of commercial scale gasifiers assume an average of about 1,000 tons per day of coal feed, which implies a scale up of a factor of 40. Furthermore, the test facility is operated at a maximum pressure of 230 psig, compared to the typically assumed value of 450 psig in design studies. The PDU has higher heat losses, higher fines elutriation, higher recycle gas flow rates, higher ash annulus gas velocities, and a lower bed height than most design studies. Furthermore, the carbon conversion rates and oxygen-to-coal and steam-to-coal ratios assumed in most design studies may not be justifiable based on PDU experience, because the PDU has operated at significantly higher steam-to-oxygen ratios than commonly assumed (Shinnar, Avidan, and Weng, 1988).

The basic processes involved in gasification are chemical, thermal, and hydrodynamic. Many of the chemical processes do not depend on scale because they take place at a particle level. However, hydrodyamic processes are generally scale-dependent, and influence the thermal history of particles and gases in the gasifier. The main areas of design that have the least commercial experience are the combustion jet zone and the ash separation zone. Analytic models of the jetting and ash separation zone have been developed by KRW based on a variety of tests at different scales, ranging from four inch to ten foot diameter for cold flow facilities and four inch to 24 inch diameter for hot flow facilities. KRSI claims an excellent correlation between the analytic models and the observed test results (Smith et al, 1986).

Perhaps the most significant scale-up uncertainty is in the jet combustion zone. It is important that the jet surface area and the solid recirculation rate near the jet be sufficient to allow for dissipation of the heat from combustion, otherwise agglomeration, clinkering and sintering of bed material will occur. Commercial designs may require the use of multiple jets, rather than a single jet, for better distribution of heat. However, this alternative may introduce problems if the jet velocities are not uniform. Also, multiple jets may interact to form stagnant regions between jets. KRSI recommends more extensive testing using semi-circular and circular models of multiple jets before designing a commercial reactor with multiple jets.

Gasifier performance for a specific coal is predicted by M.W. Kellogg based on analytic models and empirically-derived data. Experimental data are required to determine reaction rates and the influence of contained mineral matter on coal reactivity. The caking properties of the coal are also important, as is the ash fusion temperature. M.W. Kellogg has devised a number of bench-scale tests that are used to determine the empirical data needed for the analytic models. The combination of bench scale testing and mathematical modeling is reported to yield a good predictive capability for gasifier performance.

However, the predictive capability is limited by uncertainties in free board temperature, bed density, bed carbon content, and bed height (Floyd and Agrawal, 1989).

Six performance and cost studies of KRW-based systems were used to develop the gasification section cost correlation. These include two studies of IGCC systems (Bechtel and WE, 1983; Fluor, 1985) and four studies of coal-to-SNG systems (Cover et al, 1985a; Smith et al, 1986; Smith and Smelser, 1987; Earley and Smelser, 1988a). An additional earlier study (Bostwick et al, 1981) was reviewed for possible inclusion in the cost model database, but was excluded because the cost estimate was unusually low compared to the other data and because information on the performance of the system was limited.

In all six studies, the gasification section definitions are similar. The coal-to-SNG systems are the same as the IGCC systems in the areas of coal pressurization, gasification, ash removal, and particulate scrubbing. In fact, it appears that the physical dimensions of the gasifier are similar across all studies. Typical dimensions are an overall height of about 100 to 115 ft, a maximum outer diameter of about 14 feet, and a minimum outer diameter of about 5.5 ft. All systems use a coal surge bin, coal pressurization lockhopper, coal feed lockhopper, and rotary feed valve to deliver pressurized coal to the gasifier. All systems also use an ash receiving lockhopper and an ash depressurization lock hopper for ash removal. All systems use pneumatic transport of coal from the rotary feed valve to the gasifier, and pneumatic cooling and separation of ash. All systems have a recycle gas compressor and motor. Particulate recovery appears to be the same for the coal-to-SNG as for the IGCC cases, although relatively little detail is provided on this part of the plant section. A typical particulate recovery system includes a venturi scrubber and a gas scrubber. Dirty wash water is filtered in a filtration system, and the filter cake is sent to disposal. Blowdown from the wash water and sour condensate from knock-out drums are sent to the process condensate treatment section of the plant.

There appear to be some differences in the area of high temperature gas cooling and char recycle. For example, the number of high temperature heat exchangers appears to vary from one to two in the coal-to-SNG systems. With the exception of the most recent KRW-based IGCC study (Fluor, 1985), all the systems produced high pressure superheated steam from high temperature syngas cooling. The IGCC systems include a fuel gas reheater that is not used in the coal-to-SNG systems. In general, the coal-to-SNG systems appear to use only one cyclone for char recycle, compared to two cyclones in the IGCC cases.

A data base of performance and cost information for a total of 15 different plant sizes from the six references was developed. Of these 15 cases, seven are a variation in size from one case (Cover et al, 1985a) and one is a variation in size for another case (Smith et al, 1986); detailed performance data are not available for these scaled cases. For all 15 cases, the data base includes the number of spare and operating gasifier trains, number of particulate scrubbing units, gasifier operating temperature, gasifier operating pressure, coal feed rate to the gasifier, oxidant feed rate to the gasifier, and total direct capital cost. For seven of the cases, more extensive data are available. These data include steam feed rate to the gasifier, boiler feed water to the gasification section, raw water make up required for particulate scrubbing, outlet syngas from the gasification section, ash disposal, high pressure steam production from gas cooling, sour water output to process condensate treatment, and particulate cake to disposal. The 15 cases in the database include Illinois No. 6, Pittsburgh No. 8, Texas lignite, North Dakota lignite, and Wyodak subbituminous coal as feedstocks, with gasifier coal feed moisture contents ranging from about 1 to 23 percent.

A.3.3.2 Factors Affecting Cost

The cost of the gasification section is expected to depend on factors such as the coal feed rate, oxygen feed rate, steam feed rate, syngas flow rate, high pressure steam flow rate from the heat recovery system, ash disposal flow rate, raw water makeup for the particulate scrubbing unit, sour water flow rate to process condensate treatment, operating temperatures and operating pressures. While data on all of these parameters is available for some of the cost studies, data are not available for all the cost studies, such as for sour water flow rates. In some cases, such as gasifier operating temperature and pressure, there is little variation from one study to another. In other cases, such as with coal feed rate and syngas flow rate, there is a high correlation between parameters, so that inclusion of both in a regression equation does not significantly improve the "goodness-of-fit" and may yield unexpected results (e.g., cost inversely proportional to syngas flow rate).

A.3.3.3 Model Form

Prior to investigating the development of regression models, an attempt was made to develop a detailed cost model for the gasification section based on estimating the equipment cost of each major equipment item within the process area. This approach requires detailed information regarding the design basis of each equipment item and the performance parameters that determine equipment size. For example, the cost of a gasifier vessel would be estimated using the shell weight of the gasifier vessel, which is calculated based on the dimensions of the gasifier. The dimensions of the gasifier would be determined from detailed information regarding the ash separation zone, the coal combustion jet zone, the fluidized bed reaction zone, and the transport disengaging zone. The shell weight is multiplied by a unit material cost, and the costs of refractory lining, manholes, and installation would then developed. Because the gasifier is a vertical pressure vessel, the installation costs must account for the difficulty of erecting the large vessel at the site. However, this approach proved unsatisfactory for several reasons. The required detailed design information for the KRW gasifier is not reported in the cost studies. In addition, detailed design information is not reported for other components of the gasification process area. Finally, because the direct costs are reported only for the process area as a whole, there would be no direct way to validate the cost of each individual equipment item. Therefore, a regression model approach was used. The advantage of the regression model is that, if it is not extrapolated, it will yield cost estimates consistent with the published studies.

Several alternative regression models were investigated. These models included single variable regressions on parameters such as coal feed rate (on as-received, dried, or moisture- and ash-free bases) or syngas flow rate, and multivariate regressions based on various combinations of two or more predictive parameters such as coal feed rate, syngas flow rate, oxidant feed rate, ash removal rate, and number of trains. From these analyses, a regression model of gasification section costs was developed (using "Cricket Statworks" on a Macintosh computer) based on three performance parameters, the number of operating trains, and the number of total trains. This model was developed using the two-step approach described in Chapter 2. The performance parameters selected are coal feed rate (moisture- and ash-free basis), oxidant feed rate, and ash removal rate. This combination of parameters yielded reasonable results for the scaling of gasification section costs to mass flow rates. Of these three performance parameters, the model is least sensitive to the ash removal rate, as is expected. The coal feed rate directly affects the size of the coal pressurization and gasifier systems, and the combination of coal feed rate and oxidant feed rate influences the syngas flow rate, which in turn is a primary sizing parameter for high temperature gas cooling and the particulate scrubbing system. The regression model is given as:

$$DG_G = 27{,}700 + 0.0184\, N_{T,G} \left(\frac{m_{cm,G,i}}{N_{O,G}}\right)^{0.867} \left(\frac{m_{Ox,G,i}}{N_{O,G}}\right)^{0.987} \left(\frac{m_{ash,G,o}}{N_{O,G}}\right)^{0.083}$$

$$(R^2 = 0.948;\ n = 15) \qquad \text{(A-43)}$$

where,

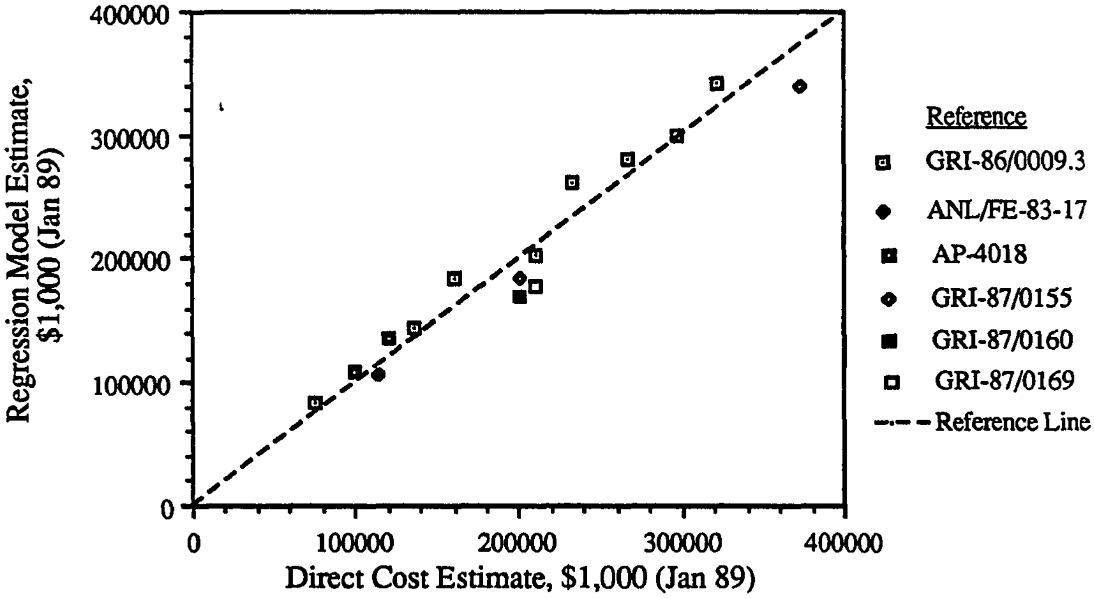

Figure A-5. Predicted vs. Actual Direct Costs for Gasification Section.

$$800 \leq \left(\frac{m_{cm,G,i}}{N_{O,G}}\right) \leq 1{,}250 \text{ tons/day}$$

$$1{,}450 \leq \left(\frac{m_{Ox,G,i}}{N_{O,G}}\right) \leq 2{,}000 \text{ lbmole/hr}$$

$$50 \leq \left(\frac{m_{ash,G,o}}{N_{O,G}}\right) \leq 200 \text{ tons/day}$$

The model should not be extrapolated for values of the predictive parameters outside the ranges given above. These ranges indicate the allowable size for each gasification train. The standard error of the estimate is $20.5 million. A graph of the cost data and the regression model is shown in Figure A-5. The model should only be used to estimate the cost of oxygen-blown KRW gasification systems operating at 465 psia and between 1,575 and 1,875 ^{o}F.

The number of trains used in the gasification section can be estimated from the following correlations:

$$N_{T,G} = \mathrm{INT}\left(0.556 + 1.49\times10^{-3} m_{cm,G,i}\right) \qquad (R^2 = 0.944;\ n = 15) \qquad \text{(A-44)}$$

$$N_{O,G} = \mathrm{INT}\left(0.557 + 9.10\times10^{-4} m_{cm,G,i}\right) \qquad (R^2 = 0.936;\ n = 15) \qquad \text{(A-45)}$$

where INT means the nearest integer value. The number of spare trains can be estimated from the difference between the number of total and number of operating trains.

A.3.4 Low Temperature Gas Cooling and Fuel Gas Saturation

In IGCC systems featuring "cold gas cleanup," the syngas is cooled to about 100 ^{o}F before entering the acid gas removal plant section. Additionally, in many IGCC designs, moisture is added to the fuel gas in a fuel gas saturator to reduce NO_X formation during syngas combustion in the gas turbine. The low temperature gas cooling section consists primarily of a series of shell and tube heat exchangers. The fuel gas saturator is a vertical column with sieve trays in which fuel gas is contacted countercurrently with hot water flowing downward. Data for this particular plant section design was available from four studies for nine different sizes. In all four studies, the number of trains was two, with no spare. These studies were based on KRW, Texaco, and Dow IGCC systems (Fluor, 1985a; Fluor, 1986; Bechtel, 1988; Fluor Daniel, 1989). Although all "cold gas" IGCC systems have a fuel gas cooling process area, not all IGCC system designs are based on fuel gas moisturization. Alternatively, many are based on direct steam injection in the gas turbine.

The performance parameters that were common to all nine cases are the outlet syngas flow rate, the syngas moisture content, and the temperature and pressure of the outlet syngas. Of these, the syngas moisture content was not a statistical significant parameter in the regression models, nor is it expected to be the most important determinant of cost. The syngas flow rate generally affects the size of all the equipment in the plant section and for this reason is a major determinant of process area capital cost. The syngas outlet temperature affects the sizing of heat exchangers within the system. The syngas outlet pressure affects the selection of pressure vessel wall thickness. A regression model obtained from a two-step process (see Chapter 2) based on syngas outlet temperature, pressure, and flow rate, and the number of operating and total trains, is given as:

$$DC_{LT} = 8.78\times10^{-21}\, N_{T,LT}\, T_{syn,LT,o}^{2.38}\, P_{syn,LT,o}^{6.06} \left(\frac{M_{syn,LT,o}}{N_{O,LT}}\right)^{0.79} \quad (R^2 = 0.966;\ n = 9) \quad \text{(A-46)}$$

where,

$85 \leq T_{LT,o} \leq 120^{o}F$

$385 \leq P_{LT,o} \leq 435$ psia

$16{,}000 \leq \left(\frac{M_{syn,LT,o}}{N_{O,LT}}\right) \leq 37{,}200$ lbmole/hr

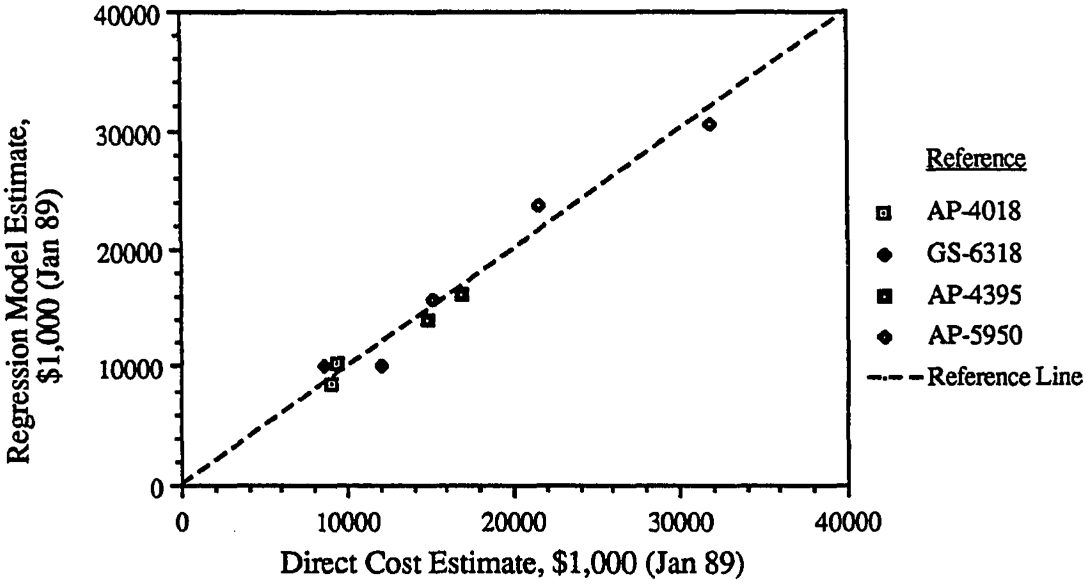

Figure A-6. Predicted versus Reported Direct Costs for the Gas Cooling Process Area.

The model should not be extrapolated for temperature or pressure. Extrapolation of the model for syngas flow rate per train is possible, because the scaling relationship between syngas flow rate and cost is quite reasonable. However, such extrapolation is not recommended, and should be confined to values not significantly different (e.g., within 50 percent) from the given range. The preferred alternative to extrapolation is to alter the number of trains so that the syngas mass flow rate per train is within the limits given above.

The standard error of the estimate is $1.5 million. The error has a mean of zero and is approximately normally distributed, based on a Kolmogorov-Smirnov test. A comparison of the regression model and cost study direct cost estimates is shown in Figure A-6. The figure indicates that there is excellent agreement between the cost estimated by the regression model and reported in the design studies over a range of costs which vary by a factor of three.

A.3.5 Sulfur Removal

A number of different sulfur removal and recovery systems have been studied in IGCC and coal-to-SNG plant designs. The most common configuration is the Selexol process for sulfur removal from the raw syngas, a two-stage Claus plant for recovery of elemental sulfur, and the Shell Claus off-gas treating (SCOT) process for treatment of the tailgas from the Claus plant. However, a number of alternative designs have also been considered. These include integration of the Selexol and SCOT processes in the

LONGSCOT design, as well as the use of alternative processes including the Dow GAS/SPEC MDEA and Selectox processes. The design basis assumed here is a Selexol unit for sulfur removal, a two-stage Claus plant for sulfur recovery, and either a SCOT or a Beavon-Stretford unit for Claus plant tail gas treatment. In this section, the development of a cost model for the Selexol process is discussed.

The proprietary Selexol process selectively removes hydrogen sulfide from the raw syngas. Typically, about 95 percent of the hydrogen sulfide is removed through counter-current contact of the syngas with Selexol solvent. The Selexol process also removes approximately 15 percent of the carbon dioxide in the flue gas. The composition of the acid gas stream which is sent from the Selexol unit to a sulfur recovery plant is typically over 50 percent carbon dioxide (Bechtel , 1983a; Bechtel, 1988; Cover et al, 1985a, 1985b; Fluor, 1983a, 1983b, 1984, 1985; Parsons, 1982). The studies cited here include both IGCC and coal-to-SNG systems based on a variety of gasifiers, including KRW, Texaco, and Shell designs. From these studies, 28 individual data points were developed. Thus, the database for the Selexol cost model represents a variety of coal gas compositions.

From the available performance and cost information for the Selexol process applied to gasification systems, a database containing total direct cost, syngas inlet flow rate, syngas composition (e.g., carbon dioxide, hydrogen sulfide, carbonyl sulfide, water vapor), removal efficiency of syngas components, acid gas flow rate and composition, and syngas temperature and pressure was developed. The inlet crude syngas temperatures for these data ranged from 95 to 120 ^{o}F and the inlet pressures ranged from 315 to 557 psia.

The inlet syngas is contacted countercurrently in a packed bed with Selexol solvent. For a more detailed discussion of this process area, the reader is referred to any of the design studies used as a basis for cost model development, and in particular Fluor (1985). The reactions occurring in the absorber reduce the temperature of the syngas. The treated syngas flows through a knock-out drum to remove solvent mist and is then heated in a heat exchanger by the incoming fuel gas. The cost of the Selexol section includes the acid gas absorber, syngas knock-out drum, syngas heat exchanger, flash drum, lean solvent cooler, mechanical refrigeration unit, lean/rich solvent heat exchanger, solvent regenerator, regenerator air-cooled overhead condenser, acid gas knock-out drum, regenerator reboiler, and pumps and expanders associated with the Selexol process.

The absorption of hydrogen sulfide by the solvent is influenced by the liquid to gas molar ratio in the absorption tower, the partial pressure of the hydrogen sulfide in the syngas, the contact temperature, the number of absorption stages or trays in the tower, and the amount of residual hydrogen sulfide left in the regenerated solvent (EPA, 1983). The absorption tower must be sized based on the syngas volume flow rate and the number of trays required for contacting solvent with the syngas. The solvent circulation rate depends on both the syngas molar flow rate and the desired removal efficiency for hydrogen sulfide. As the removal efficiency is increased, the solvent circulation rate must be increased (EPA, 1983). The solvent circulation rate affects the cost of most of the process equipment in the Selexol process. However, data for the circulation rate are not reported in the design studies. Therefore, to a first order approximation, the cost of the Selexol process is assumed to depend on the syngas flow rate for the syngas temperature and pressure range of the database. The hydrogen sulfide removal efficiency is expected to have a secondary effect on cost, because it also influences the solvent circulation rate. Other parameters such as syngas temperature or the concentration of hydrogen sulfide in the syngas may also have secondary effects on the process area cost.

Several alternative regression model formulations were attempted based on syngas flow rate, temperature, pressure, hydrogen sulfide concentration, and the removal efficiency for hydrogen sulfide. The cost of the Selexol process was found to depend primarily on the syngas flow rate entering the acid gas absorber. The cost is also influenced to a much smaller degree by the hydrogen sulfide removal efficiency. Other parameters had less significant or statistically insignificant effects in explaining the cost of the system. Therefore, these additional parameters were excluded from the model. Thus, the cost correlation for the Selexol process is:

$$DC_S = \frac{0.420\ N_{T,S}}{(1 - \eta_{HS})^{0.059}} \left(\frac{M_{syn,S,i}}{N_{O,S}}\right)^{0.980} \qquad (R^2 = 0.909;\ n = 28) \qquad \text{(A-47)}$$

where,

$$2{,}000 \le \left(\frac{M_{syn,G,i}}{N_{O,S}}\right) \le 67{,}300 \text{ lbmole/hr}$$

$$0.835 \le \eta_{HS} \le 0.997$$

The range for the syngas molar flow rate per train indicates the size range for a single train. Because the scaling exponent for the syngas flow rate term is within the range typically expected for chemical process plants, extrapolation above this range may yield satisfactory results. However, the range for syngas molar flow per train is actually quite large, implying that extrapolation is unlikely in practice. Moreover, the preferred alternative to extrapolation is to adjust the number of trains so that the molar flow rate per train is within

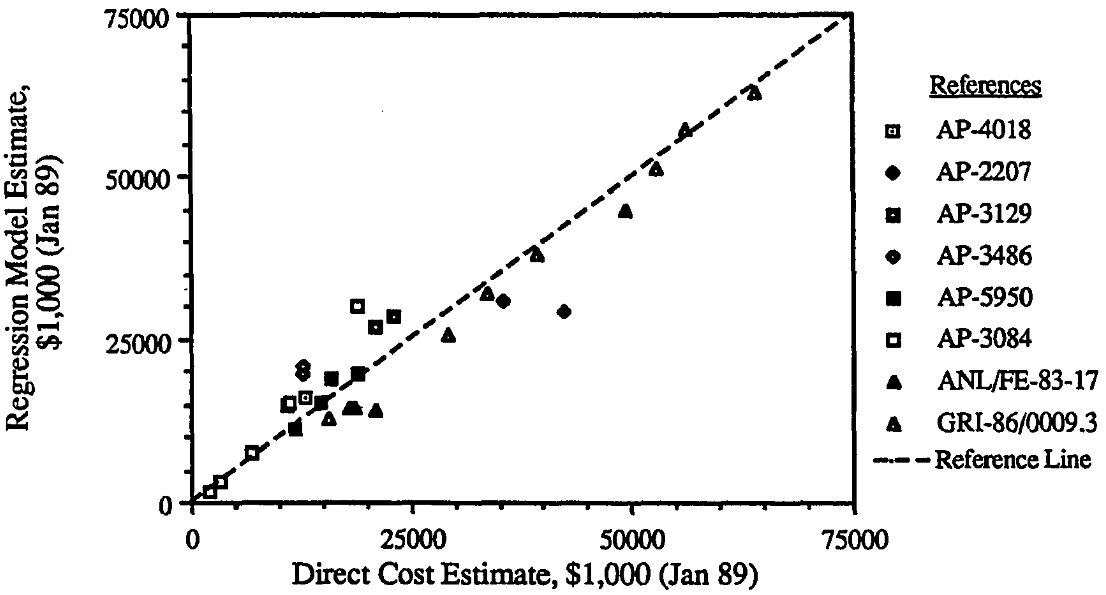

Figure A-7. Predicted versus Reported Costs for the Selexol Process Area.

the given range. The range for the hydrogen sulfide removal efficiency should not be extrapolated. Based on a two-step regression as described in Chapter 2, the standard error of the estimate is $5.1 million. A graph comparing the regression model estimates of direct cost with the costs reported in the literature is given in Figure A-7.

A.3.6 Sulfur Recovery

In most IGCC cost studies, sulfur recovery is assumed to be achieved using a Claus plant to produce elemental sulfur. This section presents an overview of the design features of a Claus plant in the IGCC process environment. For additional detail see Fluor (1985) or any of the other detailed design studies of IGCC or coal-to-SNG systems used to develop this process area cost model.

The inlet stream to the Claus plant is the acid gas from the sulfur removal section. In this study, only data for Claus plants that process the acid gas from a Selexol unit are considered. The acid gas typically contains primarily carbon dioxide and hydrogen sulfide. In order to produce elemental sulfur, a 2:1 ratio of hydrogen sulfide and sulfur dioxide is required. Therefore, a portion of the incoming acid gas is combustion in a two-stage sulfur furnace. The furnace temperature is high enough in the first stage (typically 2,500 ^{o}F) to destroy any ammonia in the acid gas. Intermediate pressure steam (e.g., 350 psia) is generated from the waste heat produced in the sulfur furnace, cooling the feed gas to the Claus converters to about 600 ^{o}F. Further cooling to 350 ^{o}F occurs in a sulfur condenser, generating low pressure steam (e.g., 55 psia). Sulfur flows to a gravity sump, and is kept molten by condensing low pressure steam that flows through coils in the bottom of the sump.

Some of the furnace gas is used to heat the feed gas from the first condenser to approximately 450 ^{O}F prior to entering the sulfur converter, where hydrogen sulfide and sulfur dioxide react in the presence of a catalyst (e.g., Kaiser S-501) to produce elemental sulfur and water. This reaction is exothermic, and the outlet temperature of the gas is approximately 630 ^{O}F. The conversion rate is limited by thermal equilibrium. Gaseous sulfur is recovered in a second condenser. The cooling may be accomplished by heating water for fuel gas saturation. The feed gas then is mixed with remaining combustion gases and then enters the second converter. A third condenser, in which water for fuel gas saturation may be heated, is used for final sulfur recovery. The effluent gas from the Claus plant then passes through a coalescer and then on to tail gas treatment.

A direct cost correlation was developed for two-stage Claus plants based on data from a number of gasification plant studies. A number of data points are not included in this correlation because they represent either three-stage Claus plants or two-stage Claus plants with tail gas incineration and no tail gas treatment, with the incinerator costs included in the direct cost. The cost of a Claus plant is known to scale primarily with the recovered sulfur mass flow rate capacity using the standard exponential scaling model of Equation (1) with an exponent of approximately 0.6 (e.g., EPA, 1983b). It appears that this scaling rule may have been the basis for developing the cost estimates of Claus plants used in the design studies, because an excellent goodness-of-fit was found for a single variable regression based on sulfur recovered. The scaling exponent that was obtained in the single variate analysis was 0.668. The regression model was further developed to represent the number of operating and spare trains for each data point in the database. The resulting cost equation is:

$$DC_c = 6.28\, N_{T,C}\left(\frac{m_{s,C,o}}{N_{O,C}}\right)^{0.668} \qquad (R^2 = 0.994;\ n = 21) \qquad \text{(A-48)}$$

where,

$$695 \leq \left(\frac{m_{s,C,o}}{N_{O,C}}\right) \leq 18{,}100 \text{ lbmole/hr}$$

The standard error of the estimate is $235,000, and this is probably primarily attributable to differences in site location. The regression model is shown graphically in Figure A-8.

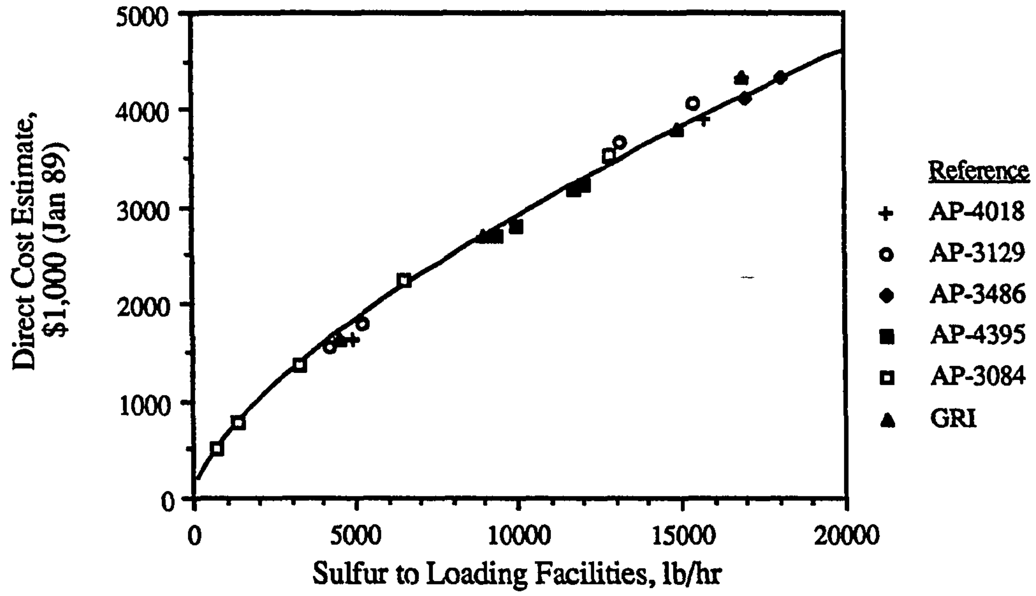

Figure A-8. Direct Cost of the Claus Plant Process Area.

As indicated above, the capacity of a single train varies by a factor of over 20. Typically, one or two operating trains and one spare train are used, each with equal capacity. Because there was a prior expectation that the cost of the Claus plant should be modeled using an exponential scaling relationship based on recovered sulfur capacity, with a coefficient near 0.6, this model can be extrapolated at the high end of the range. However, as with all other models, it is recommended that the number of trains be selected so that extrapolation is not required.

A.3.7 Tail Gas Treating

Cost models for two tail gas treatment technologies are developed in this section. The Shell Claus Off-gas Treating (SCOT) process is commonly assumed in the studies performed for EPRI. The Beavon-Stretford process is assumed in the ASPEN flowsheet for the oxygen-blown KRW-based IGCC system with cold gas cleanup.

A.3.7.1 SCOT Process

The Shell Claus Off-gas Treating (SCOT) process removes hydrogen sulfide from atmospheric pressure gases, such as the tail gas from a Claus plant, in a manner similar to the Selexol process, and the concentrated acid gas stream from the SCOT process is then sent to the Claus plant for sulfur recovery. The process is proprietary, and limited performance data regarding this process are available in the detailed cost studies. The process is available commercially.

The tail gas from the Claus plant typically contains a number of sulfur species, including hydrogen sulfide, sulfur dioxide, carbonyl sulfide, and elemental sulfur. However, the SCOT solvent is not suitable for removing sulfur species other than hydrogen sulfide. Therefore, the tail gas is reacted in a hydrogenation reactor in the presence of a cobalt-molybdenum catalyst to convert all sulfur species to hydrogen sulfide. Hydrogen and carbon monoxide are required for these reactions, and are supplied by mixing flash gas from the acid removal unit with the Claus plant tail gas. The flash gas is partially combusted in a reducing gas generator prior to entering the hydrogenation reactor, which raises the temperature of the inlet gases to the appropriate reaction temperature. The outlet gas from the hydrogenation reactor is cooled in a waste heat boiler, where intermediate pressure steam (e.g., 100 psia) is generated. The gas is further cooled by direct contact with water in a quench water cooling tower.

The cooled gas containing hydrogen sulfide then enters an absorber vessel where almost all of the hydrogen sulfide and a portion of the carbon dioxide is absorbed by countercurrent contact with an amine solution. The treated tail gas goes to the gas turbine air intake, and contains approximately 300 ppm of total sulfur.

Like the Selexol process, the hydrogen sulfide-rich amine solution is pumped through a lean/rich solvent heat exchanger, and then enters the regenerator vessel, where acid gases are stripped from the solution by indirect application of heat. This heat is supplied by condensing steam in the regenerator reboiler. The lean solvent leaves the regenerator and is pumped through the lean/rich solvent heat exchanger to the absorber vessel. The acid gas is cooled by cooling water in an overhead condenser, and flows through a knock-out drum for removal of any amine solution prior to recycle to the Claus plant sulfur furnace.

The cost of the SCOT process is expected to vary directly with performance parameters such as the tail gas flow rate, flash gas flow rates, and amine solution flow rates. However, these data are not available from the cost studies because of the proprietary nature of the process. The SCOT process is used in conjunction with the Claus process to achieve approximately 99.7 to 99.9 percent recovery of the sulfur removed from the syngas. The Claus plant removal efficiencies vary within a relatively narrow range of about 91 to 95 percent. This implies that the total sulfur recovered to the sulfur loading facilities could serve as a proxy for the unavailable detailed performance information about the SCOT process.

In analyzing the relationship between the cost of the SCOT process and the total amount of sulfur sent to the loading facilities, two groupings of data were identified. These include a set of ten cost estimates based on four studies performed for EPRI in 1983 to 1986, referred to here as "Case 1", and a set of ten cost estimates based on five studies performed for EPRI or GRI in 1987 and 1988, referred to as "Case 2". The Case 2 cost estimates are significantly higher than the Case 1. Furthermore, the Case 1 cost estimates correlate almost exactly with the sulfur flow rate to the loading facilities, indicating that perhaps a simple scaling equation was used in developing these costs. The Case 2 estimates are more scattered, indicating perhaps that more detailed cost estimates, based on several performance parameters, were used or that the studies were performed with different assumptions or data. The latter may be a likely cause of the variation in costs, because three separate contractors prepared the cost estimates.

The cost correlations for Case 1 and Case 2 are summarized below, and shown graphically in Figure A-9:

Case 1: $DC_{SC} = 2.64\ m_{s,S,o}^{0.75}$ $(R^2 = 0.988;\ n = 10)$ (A-49)

where,

$4{,}900 \leq m_{s,S,o} \leq 30{,}800$ lb/hr

and,

Case 2: $DC_{SC} = 1{,}700 + m_{s,S,o}$ $(R^2 = 0.865;\ n = 10)$ (A-50)

where,

$3{,}700 \leq m_{s,S,o} \leq 16{,}600$ lb/hr

In Case 1, the standard error is a low \$60,000, while in Case 2 the standard error is \$700,000. The number of trains assumed in the cost studies is one for all cases except for one study (AP-3129).

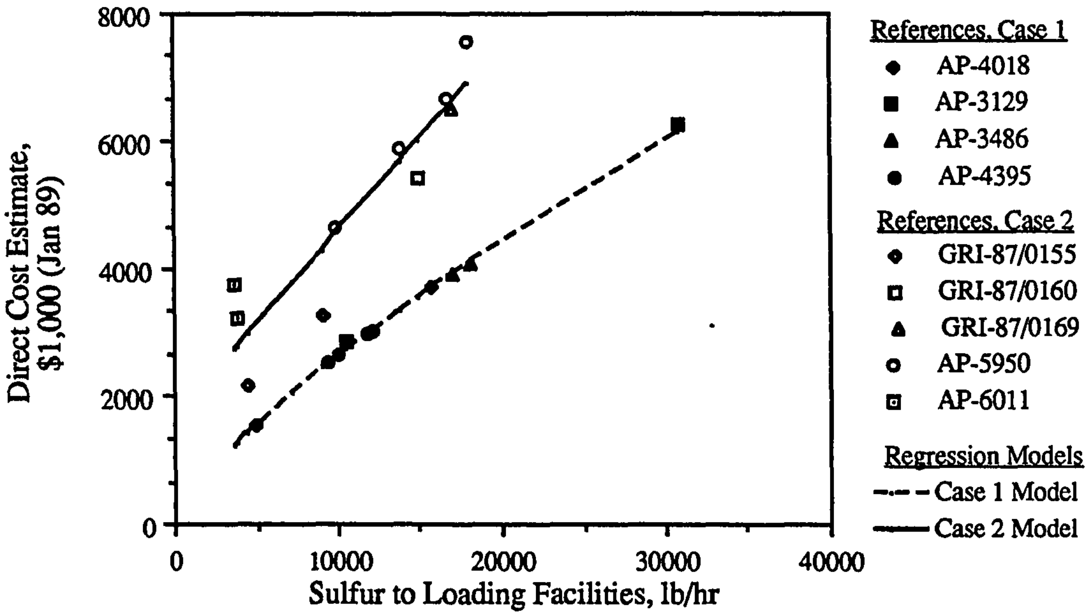

Figure A-9. Direct Cost of the SCOT Process Area.

The Case 2 correlation is recommended for use in the cost models because it represents the most recent cost studies. Because this model is based on a proxy variable, rather than on performance variables that directly affect the sizing of SCOT process equipment, this model should not be extrapolated.

A.3.7.2 Beavon-Stretford Process

In this section, an overview of the performance and design of the Beavon-Stretford process is presented as background information for the development of a regression cost model. See Fluor (1983a) or Fluor (1983b) for a more detailed discussion of this process. The Beavon-Stretford process is a modification of the Stretford process, which is designed to remove hydrogen sulfide from atmospheric pressure gas streams and convert it to elemental sulfur. However, the Stretford process is not appropriate for handling effluent gases containing sulfur dioxide, carbonyl sulfide, or elemental sulfur. Therefore, a Beavon unit is used to catalytically reduce or hydrolize these species to hydrogen sulfide in the presence of a cobalt molybdate catalyst.

Because hydrogen is required for the reactions occurring in the Beavon unit, flash gas from the acid gas removal section is used as a feed stream. The flash gas is partially combusted in a reducing gas generator, mixed with the Claus plant tail gas, and the total gas stream then enters the Beavon hydrogenation reactor. The hot gas from the reactor is cooled in a waste heat boiler where intermediate pressure (e.g., 100 psia) steam is generated. The gas stream is further cooled in the desuperheater section of a thermally integrated desuperheater/absorber vessel. The cooling of the gas stream is accomplished by heat transfer with cooling water, which is recirculated through an air-cooled heat exchanger. The gas stream then enters the absorber portion of the vessel, where over 99 percent of the hydrogen sulfide is removed by contact with a Stretford solution containing sodium carbonate. The treated gas is vented to the atmosphere.

The Stretford solution flows to a soaker/oxidizer, where anthraquinone disulfonic acid (ADA) is used to oxidize the reduced vanadate in the Stretford solution. The ADA is regenerated by air sparging, which also provides a medium for sulfur flotation. The sulfur overflows into a froth tank, and the underflow from the oxidizer/soaker is pumped to a Stretford solution cooling tower and then to a filtrate tank.

The sulfur from the froth tank is pumped to a primary centrifuge, where the wet sulfur cake product is reslurried and sent to a second centrifuge, after which the sulfur is again reslurried. The slurry is then pumped through an ejector mixer, where the sulfur is melted and separated in a separator vessel. The sulfur goes to a sump.

The process is considered commercially available. The capital cost of a Beavon-Stretford unit is expected to vary with the volume flow rate of the input gas streams and with the mass flow rate of the sulfur produced. Data from two EPRI-sponsored studies were used to develop a regression cost model (Fluor, 1983a; 1983b). An additional two studies were reviewed for inclusion in the database, but information regarding key process parameters (e.g., recovered sulfur flow rate) was not reported. The two EPRI studies report limited performance and cost data for nine different Beavon-Stretford unit sizes. For example, there is incomplete information about inlet gas streams flow rates. Because of the limited availability of performance data, a regression analysis based only on the sulfur produced by the Beavon Stretford process was developed. However, this regression yielded an excellent fit to the data:

$$DC_{BS} = 57.5 + 66.2\, N_{T,BS} \left(\frac{m_{s,BS,o}}{N_{O,BS}}\right)^{0.645} \qquad (R^2 = 0.998;\ n = 7) \qquad \text{(A-51)}$$

where,

$$75 \leq m_{s,BS,o} \leq 1{,}200 \text{ lb/hr}$$

The high coefficient of determination indicated for this model implies either that an exponential cost model is an excellent predictor of the costs of Beavon-Stretford units, or that the costs developed in the EPRI studies were based on a simple scaling model as an approximation. Therefore, it is not immediately clear if this model merely represents an

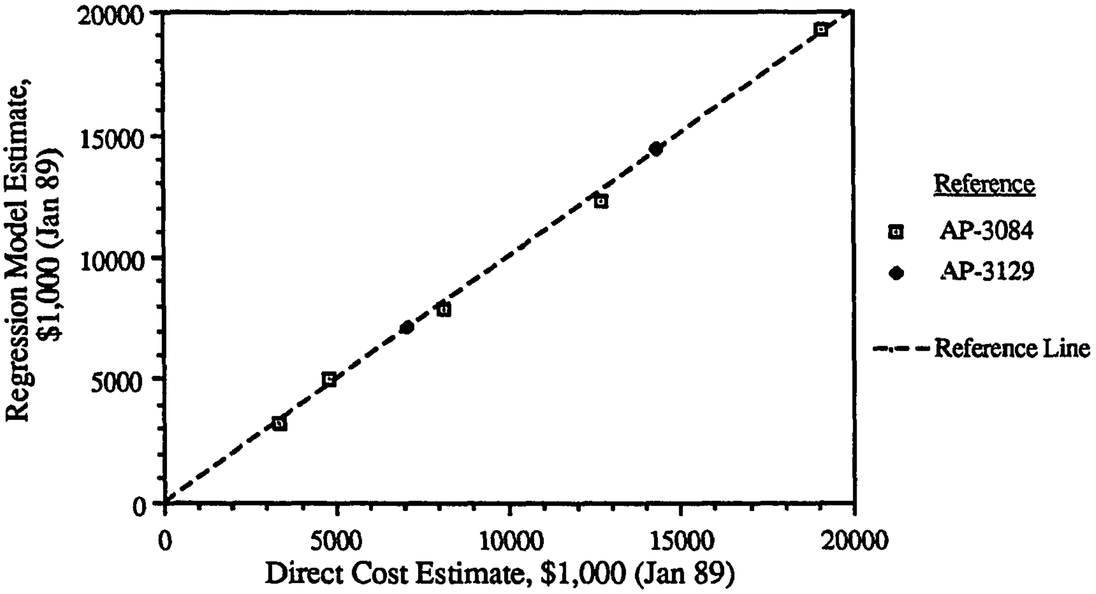

Figure A-10. Predicted versus Reported Direct Costs for the Beavon-Stretford Process Area.

accepted industry practice for developing preliminary cost estimates, or if it accurately reflects the cost of Beavon-Stretford units. The standard error of the estimate is $260,000, and the errors have an approximately normal distribution with a mean of zero. Typically, two operating and one spare train are assumed. Figure A-10 shows a comparison of the regression model estimate and the cost study estimates. Although the regression model is an excellent fit to the data, it is recommended that the number of trains be adjusted so that the recovered sulfur flow rate per train does not exceed the limits given above. As a default, the number of operating and total trains for this process area is assumed to be the same as for the Claus plant process area.

A.3.8 Boiler Feedwater System

The boiler feedwater system consists of equipment for handling raw water and polished water in the steam cycle. This equipment includes a water dimineralization unit for raw water, a dimineralized water storage tank, a condensate surge tank for storage of both dimineralized raw water and steam turbine condensate water, a condensate polishing unit, and a blowdown flash drum. The major streams in this process section are the raw water inlet and the polished water outlet. Data on the cost of the boiler feedwater section and the flow rates of the raw water and polished water streams is available from five studies for 14 plant sizes. These studies include Texaco-based, Shell-based, and KRW-based IGCC systems (Fluor, 1983a; 1983b; 1984; 1985; 1986). Because all of these studies were developed by the same contractor using a consistent approach, they provide an excellent basis for developing a cost model. The boiler feedwater section is generic to the steam cycle.

The cost of the boiler feedwater section is expected to depend on both the raw water flow rate through the dimineralization unit and the polished water flow rate through the polishing unit. The polished water flow rate includes primarily both the raw water and the steam turbine condensate. The steam cycle condensate is typically larger than the raw water flow rate. A two-variable regression model of the boiler feed water system cost as a function of the raw water and polished water flow rates was found to yield good results:

$$DC_{BF} = 0.145\, m_{rw}^{0.307} m_{pw}^{0.435} \qquad (R^2 = 0.991;\ n = 14) \qquad \text{(A-52)}$$

where,

$24{,}000 \leq m_{rw} \leq 614{,}000$ lb/hr

$234{,}000 \leq m_{pw} \leq 3{,}880{,}000$ lb/hr

For this model, a nonlinear variable transformation was used. The error of the linearized model is approximated by a normal distribution. Therefore, the error of the nonlinear model shown above is represented by a lognormal distribution. The median of the errors is 1.0, with a mean of 1.002 and a standard deviation of 0.063. The 90 percent probability range for the error is approximately 0.9 to 1.1, implying a 90 percent confidence band of 90 to 110 percent of the nominal cost estimate.

Typically only one train of equipment is used in this section, and all the equipment is commercially available. A comparison of the regression model cost estimates and the direct cost estimates from the detailed cost studies is shown in Figure A-11. This model should not be extrapolated beyond the range of the predictive variables as indicated above. However, because the cost of the boiler feed water section is a very small portion of the total direct cost for a typical IGCC plant, the effect of any errors introduced by modest extrapolations may be acceptable for some purposes.

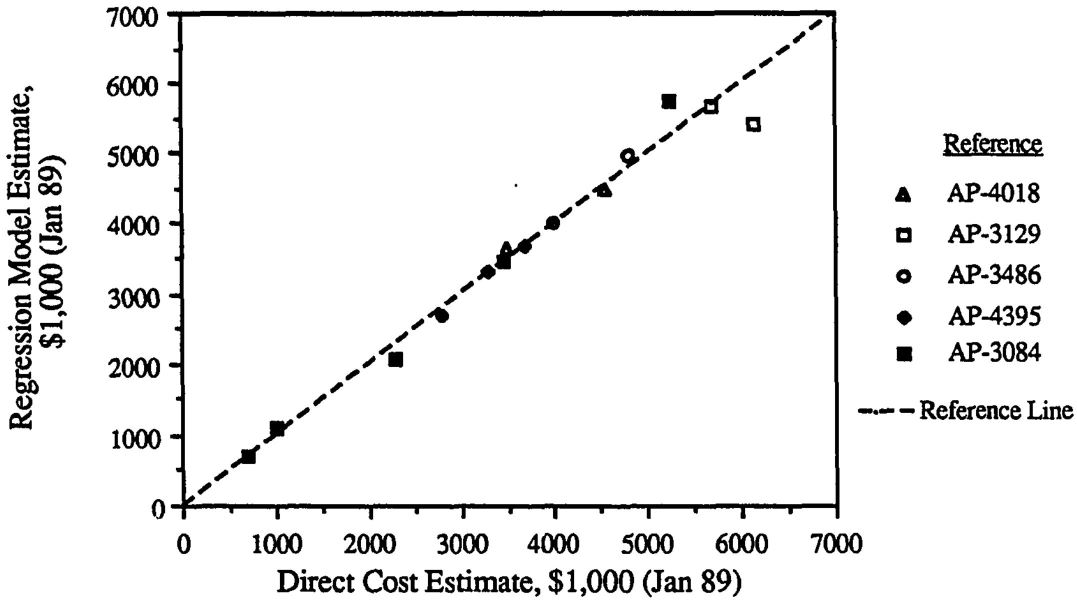

Figure A-11. Predicted versus Reported Direct Costs for the Boiler Feedwater Process Area.

A.3.9 Process Condensate Treatment

The process condensate treatment section is used to treat blowdown from the particulate scrubber and process condensate from gas cooling (Fluor, 1983b; 1985). These streams contain ammonia, carbon dioxide, and hydrogen sulfide, and the scrubber blowdown also has a high chlorides content. The blowdown and condensate stream are treated in separate strippers. The overhead vapor streams from both strippers are cooled in air-cooled heat exchangers and then they flow through knock-out drums prior to feed to the Claus plant sulfur furnace. The stripped bottoms product from the blowdown water stripper is cooled by the incoming process condensate water and then sent to a water treatment plant for biological treatment prior to flow to the cooling tower. The bottoms from the process condensate water stripper is sent as make up to the gas scrubbing unit.

Because the treated process condensate is used as make-up to the gas scrubbing unit, and because blowdown from the gas scrubbing unit is the larger of the flow streams entering the process condensate treatment section, it is expected that process condensate treatment direct cost will depend primarily on the scrubber blowdown flow rate. Because only two cost studies were identified with similar designs and sufficient detail for regression analysis, a single variate regression analysis was used:

$$DC_{PC} = 2{,}583 + 0.00213\ m_{sbd} \qquad (R^2 = 0.998;\ n = 4) \qquad \text{(A-53)}$$

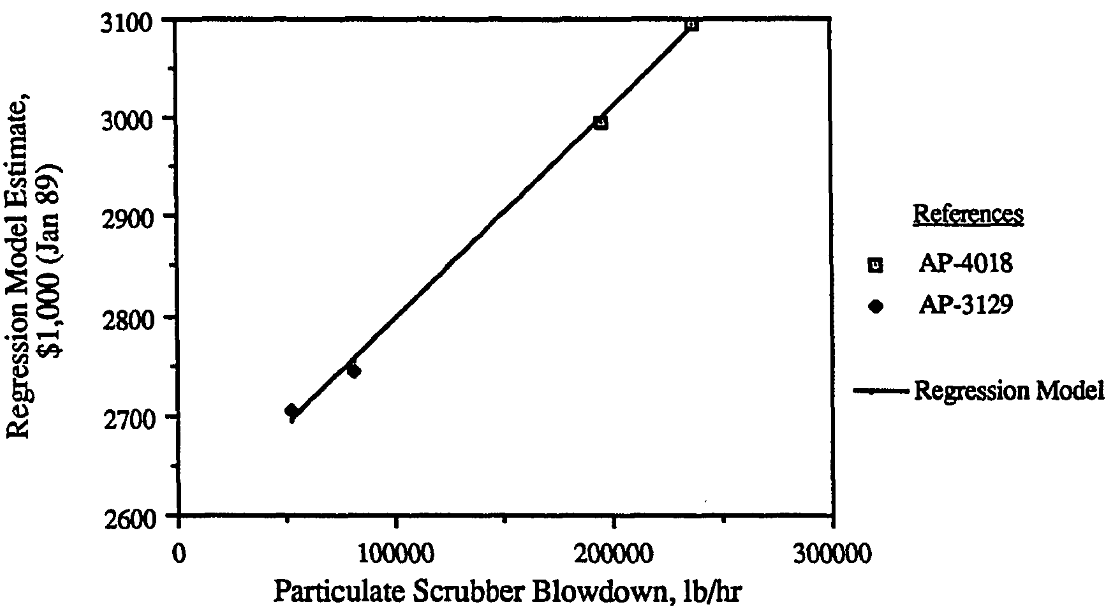

Figure A-12. Direct Cost of the Process Condensate Process Area.

where,

$$52{,}000 \leq m_{sbd} \leq 237{,}000 \text{ lb/hr}$$

The standard error of the estimate is $11,000. Typically only one operating train is used, with no spare. The equipment is commercially available. However, the designs depend on the chemical composition of the coal feedstock. Detail regarding the design of these systems is not available from the cost studies. The regression model is shown graphically in Figure A-12.

A.3.10 Gas Turbine

The most commonly assumed gas turbine in IGCC performance and cost studies is the General Electric (GE) model MS7001F, also referred to as the "Frame 7F". This gas turbine is designed for a turbine inlet temperature of 2,300 ^{o}F and has a power output of about 125 to 150 MW. By contrast, typical gas turbines have firing temperatures in the range of 2,020 to 2,150 ^{o}F. The thermal efficiency of gas turbines increases as the firing temperature increases. The higher firing temperature is the result of advances in turbine blade manufacturing. The Frame 7F turbine blades are manufactured using a process known as "directional solidification," which has been used on smaller aircraft engine turbine blades. Because of improvements in molding technology, the process can now be applied to the larger turbine blades of the Frame 7F. Further advances in manufacturing techniques may lead to the capability to cast turbine blades as single crystals with no grain boundaries, permitting an additional 50 to 150 ^{O}F increase in firing temperature (Smock, 1989).

The first Frame 7F has completed factory tests at General Electric and has been delivered to a Virginia Power site in Chesterfield, VA as part of a combined-cycle power plant. General Electric has rated this machine at 150 MW with a heat rate of 9,880 Btu/kWh.

There are a number of design factors that affect the cost of a gas turbine in an IGCC process environment. For example, the firing of medium-BTU coal gas, as opposed to high-BTU natural gas, requires modification of the fuel nozzles and gas manifold in the gas turbine (BGE, 1989). Some additional concerns associated with firing coal gas are discussed by Cincotta (1984). The presence of contaminants in the syngas may affect gas turbine maintenance and long term performance. Liquid droplets may cause uneven combustion or may burn in the turbine first-stage nozzles, causing damage. Solids can deposit on fuel nozzles or cause erosion in the hot gas path of the gas turbine (e.g., combustor, turbine). Alkali materials that deposit on hot gas path parts cause corrosion. It is expected that, at fuel gas temperatures less than 1,000 ^{O}F, that alkali material is essentially condensed on any particulate matter in the raw syngas, and that the alkali removal efficiency is approximately the same as the particle removal efficiency. For sufficiently high particle removal efficiencies, erosion is not expected to be a problem. Corrosion is not expected to be any worse than for distillate oil firing. Deposition of particles is expected to be within the allowance of reasonable maintenance schedules. The design for an advanced high firing temperature gas turbine employs advanced air film cooling which could be affected by the ash content of combustion products.

Another design issue is the gas turbine fuel inlet temperature. A study by Fluor (Earley and Smelser, 1988) assumes that hot desulfurized syngas from an advanced hot gas cleanup process is fed directly to the gas turbine at 1,200 ^{O}F. The Fluor study indicates that General Electric expects that a fuel system capable of a 1,200 ^{O}F fuel inlet temperature could be developed by 1994. The maximum fuel temperature test to date has been at 1,000 ^{O}F. An earlier study with hot gas cleanup included a hot gas cooler to reduce the gas temperature to 1,000 ^{O}F (Corman, 1986). For the KRW system with cold gas cleanup, the coal gas temperature is within the limits of current technology. However, the gas turbine costs developed here should not be used in conjunction with IGCC systems featuring hot gas cleanup without some adjustments to account for the uncertainty in using a higher fuel inlet temperature.

Unfortunately, there is currently a lack of reported data from which to develop a detailed gas turbine cost model that is explicitly sensitive to the type of factors discussed above. In preliminary cost estimates, the typical approach to accounting for these uncertainties in performance, or for the possibility of increased capital cost due to design modifications, is through process contingency factors. The approach taken here is to use the available cost data for the GE Frame 7F to develop a cost estimate for a single gas turbine. In the use of this cost estimate for actual case studies in a later task, judgments about the uncertainty in cost, and about the likelihood of cost increases for applications with coal gases, will be encoded using either process contingency factors or the new probabilistic modeling capability developed for ASPEN, as discussed in Chapter 2.

Although cost estimates of the GE Frame 7F are available in a number of IGCC cost studies, recent cost estimates are significantly higher than older estimates. However, the more recent estimates are expected to be more reliable, because the Frame 7F was at or near commercialization at the time of the recent studies. In four recent site-specific IGCC studies performed for EPRI (BGE, 1989; Fluor Daniel, 1988, 1989; FPL, 1989), the cost of the Frame 7F in the first phases of a phased IGCC construction schedule ranged from \$30.8 to 33.6 million, with an average of \$32.0 million (Jan 89). This cost excludes equipment associated with combined cycle systems, which are discussed in the following two sections. In two other studies (JCP&L, 1989; NUSCo, 1988), the cost of the Frame 7F for application in natural gas-fired combined cycle plants was estimated at \$28.3 and \$26.8 million, respectively. The higher estimate of \$32.0 million per unit is consistent with the expectation that the cost of the gas turbine modified to fire medium-BTU coal gas will be higher than for the standard natural gas-fired unit. This high estimate will be used in the cost model:

$$DC_{GT} = 32{,}000\ N_{T,GT} \tag{A-54}$$

A competitor to the GE Frame 7F is under development by Mitsubishi Heavy Industries and Westinghouse Electric. The prototype model 501F is expected to achieve a rating of 148.8 MW and a turbine inlet temperature of 2,300 ^oF. This model is expected to be commercially available in 1992 (GTW, 1989). No cost data are currently available for this model; however, competition between the Frame 7F and the 501F could result in similar prices for both machines.

A Kraftwerk Union (KWU) gas turbine, model 84.2, was analyzed in an EPRI study (Fluor Daniel, 1988). This is a commercially available, moderate firing temperature machine that is rated at approximately 100 MW with a cost of about \$24.2 million per unit.

The combustor features a low-NO_X design, and does not require water injection when operated on natural gas.

A.3.11 Heat Recovery Steam Generator

The heat recovery steam generator (HRSG) is a set of heat exchangers in which heat is removed from the gas turbine exhaust gas to generate steam. Typically, steam is generated at two or three different pressures, and associated with the HRSG is one steam drum for each steam pressure level. High pressure superheated steam is generated for use in the steam turbine, and typically the exhaust from the steam turbine first stage is reheated. The input streams to the HRSG section include the gas turbine exhaust and boiler feedwater to the deaerator. The major output stream is the high pressure steam to the steam turbine. Several parts of the HRSG must be sized to accommodate the high pressure steam flow, including the superheater, reheater, high pressure steam drum, high pressure evaporator, and the economizers.

Most studies of IGCC systems aggregate the cost of the HRSG units with the cost of the gas turbine and the steam turbine. Only four studies were identified in which the cost of the HRSG units were reported as a separate line item. A study of Texaco and British Gas/Lurgi IGCC systems includes performance and cost estimates for several sizes of HRSGs used in combination with reheat steam turbines (Parsons, 1982). These HRSG units include two steam pressure levels, and are used in conjunction with a conventional gas turbine. The high pressure steam varies from 650 psia to 1520 psia for these HRSGs. The exhaust gas flow rate and temperature indicate that the gas turbine is a GE Frame 7E or equivalent. A study by Bechtel and WE (1983c) for a KRW-based system included an HRSG design with three pressure levels using a large 130 MW gas turbine with a high exhaust gas temperature. A study of Texaco-based IGCC systems included performance and cost estimates for reheat steam turbines and HRSGs with two pressure levels (Fluor Technology, 1986). A recent study of Dow-based IGCC systems includes performance and cost estimates for two-pressure level reheat HRSGs applied in conjunction with large advanced gas turbines (Fluor Daniel, 1989).

A detailed approach to estimating the cost of HRSGs is reported by Foster-Pegg (1986). This approach requires detailed performance and design information for each heat exchanger in the HRSG. The necessary design values were not reported in the performance and cost studies, nor was sufficient detail about performance available to develop such a model. Furthermore, the level of detail in the Foster-Pegg model is not justifiable for the applications envisioned for this model for several reasons. The technical and cost growth risks of IGCC systems reside primarily in process areas such as gasification, gas cleanup, and advanced gas turbine designs. The HRSG is a conventional, commercially available component. Therefore, the priorities for cost model development should be with the more innovative systems. Secondly, in comparative studies of IGCC systems, the cost of HRSGs will be similar, and will not be a factor in distinguishing one system from another. Instead, differences in the gasification process area, gas cleanup, and byproduct recovery, as examples, are expected to be important in distinguishing alternative systems. Third, the purpose of this model is not to develop detailed, final estimates of site-specific costs for a particular project, but to develop preliminary cost estimates for the purpose of research planning. Therefore, there is not a need for a highly detailed cost model for this particular process area.

The cost of the HRSG is expected to depend on factors such as the high pressure steam flow rate to the steam turbine, the pressure of the steam, the gas turbine exhaust gas volume flow rate, the number of steam drums, and, to a lesser extent, the boiler feed water or saturated steam flowrates in each of the heat exchangers in the HRSG. A variety of regression models were investigated to represent these potential predictive parameters. However, because only 10 data points are included in the database, only a limited number of predictive parameters can be reasonably included in the model, based on statistical considerations. Furthermore, some parameters that are expected to be important in determining HRSG cost, such as the gas turbine exhaust flow rate, are not statistically important for this data set. When the gas turbine exhaust flow rate, high pressure inlet steam flow rate to the steam turbine, and the steam pressure are included in a regression model, the exponent for exhaust flow rate is small and is not statistically significant. The exhaust gas flow rate is not an influential predictive parameter because the cost studies are based primarily on either GE Frame 7E or 7F gas turbines; therefore, there was not a large range of variation for the exhaust gas flow rate. A simple regression model based only on the high pressure steam flow rate to the steam turbine yielded a high coefficient of determination. A multivariate regression based on the high pressure steam flow to the steam turbine and the pressure of the steam yielded satisfactory results:

$$DC_{HR} = -5{,}364 + 7.24\times10^{-3}\, N_{T,HR}\, P_{hps,HR,o}^{1.526} \left(\frac{m_{hps,HR,o}}{N_{O,HR}}\right)^{0.242}$$

$$(R^2 = 0.966;\ n = 10) \qquad \text{(A-55)}$$

where,

$650 \leq P_{hps,HR,o} \leq 1{,}545$ psia

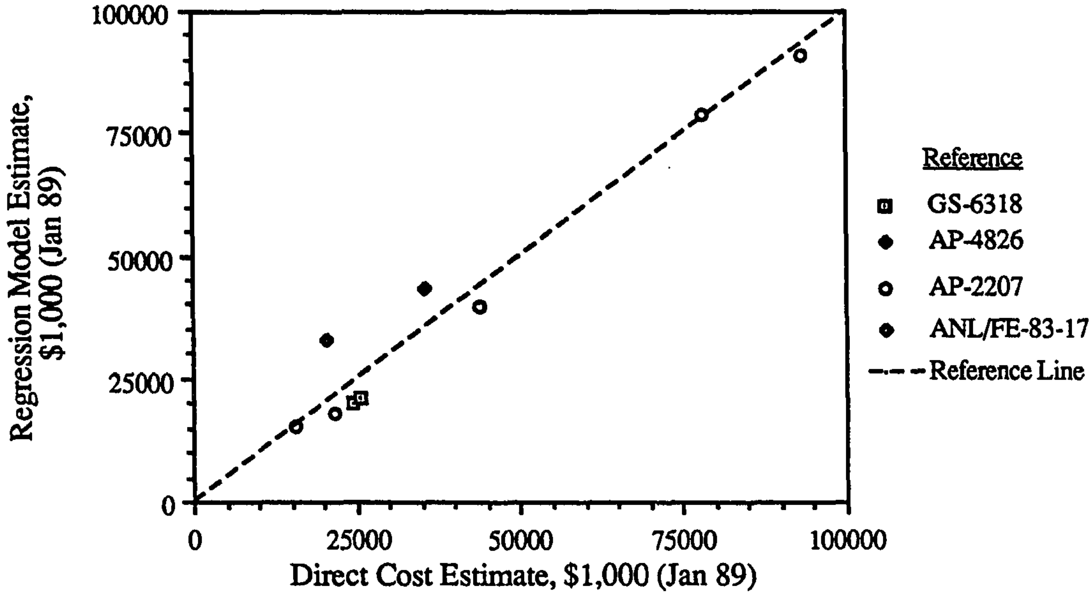

Figure A-13. Predicted versus Reported Direct Costs for the Heat Recovery Steam Generator.

$$66{,}000 \leq \frac{m_{hps,HR,o}}{N_{O,HR}} \leq 640{,}000 \text{ lb/hr}$$

The standard error of the estimate is \$6.0 million. In all the cost studies one HRSG is assumed for each gas turbine, with no spares. A comparison of the regression model and the cost estimates is shown graphically in Figure A-13. The regression model should not be extrapolated beyond the ranges for the predictive parameters indicated above.

A.3.12 Steam Turbine

A typical steam turbine for an IGCC plant consists of high-pressure, intermediate-pressure, and low-pressure turbine stages, a generator, and an exhaust steam condenser. The high pressure stage receives high pressure superheated steam from the HRSG. The outlet steam from the high pressure stage returns to the HRSG for reheat, after which it enters the intermediate pressure stage. The outlet from the intermediate pressure stage goes to the low pressure stage.

The cost of a steam turbine is expected to depend on the mass flow rate of steam through the system, the pressures in each stage, and the generator output, among other factors. Nine cost estimates for the steam turbine were available from four studies. A single-variate regression based on the generator output was found to yield reasonable results:

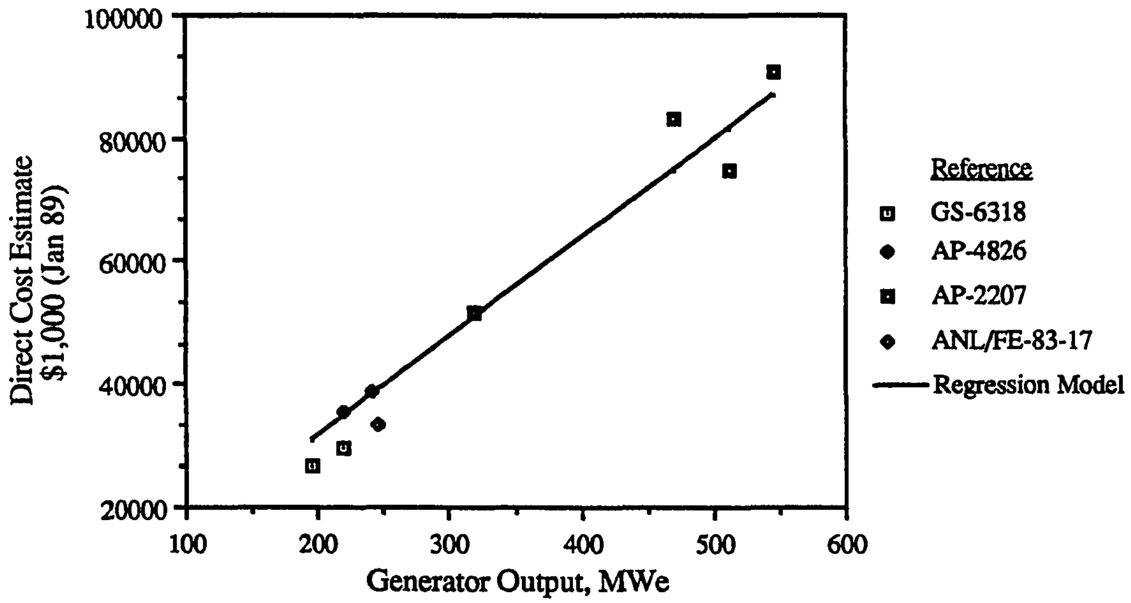

Figure A-14. Direct Capital Cost of the Steam Turbine Process Area.

$$DC_{ST} = 158.7\ MW_{ST} \qquad (R^2 = 0.958;\ n = 9) \qquad \text{(A-56)}$$

where,

$$200 \leq MW_{ST} \leq 550$$

The standard error of the estimate is $5.5 million. Only one steam turbine is used in most IGCC designs. A graphical representation of the regression model is shown in Figure A-14.

A.3.13 General Facilities

General facilities include:

Cooling water systems
- Process heat rejection
- Condensation of exhaust steam
- Cooling of mechanical equipment

Plant and instrument air
- Motor driven compressors, dryer packages

Potable and utility water
- Motor driven pumps, in-line chlorination
- Storage tanks, supply drum

Fuel system
- Fuel oil for startup, 3 30,000 bbl tanks, two motor driven pumps and suction heaters

Nitrogen system

Liquid N_2 storage sphere, air fin heaters

Effluent water treating

Storm and oily water
Utility wastewater
Cooling tower blowdown

Flare system

Elevated flare stacks, pilot flame LPG tank, pumps, and vaporizor, condensate separator,

Fire water system

Fire water loop, motor driven jockey pump, hydrants, monitors, fire water pumps, fire water storage tank

Interconnecting piping

Major process and utility piping between battery limits of plant units.

Buildings

Substations
Control houses
Operators shelters
Administration
Laboratory
Cafeteria
Change house and guard house
Fire house
First aid
Maintenance
Warehouse

Railroad facilities, roads, and lighting

Unloading coal, loading liquid sulfur
15,000 lineal ft of track, 3 switches, 1 bumper
6 RR crossings. Heated liquid sulfur storage tanks. Security and plant service roads, fencing, parking, street lighting.

Computer control system

Control centers, consoles, panels, control and measurement devices, computer, interfaces, graphic displays, logging and reporting, and data highway system

Electrical system

captive transformers
high voltage electric power distribution from switchyard to high voltage substations
high voltage substations, step-down

transformers

These equipment items are typically represented in preliminary, study-grade cost estimates as a percentage of the direct costs. Most studies assume that general facilities are approximately 15 percent of direct costs. The cost of general facilities can be estimated as:

$$GF = f_{GF} \sum_{i=1}^{12} DC_i \qquad \text{(A-57)}$$

where,

$f_{GF} = 0.15$

A.4 Capital Cost of a KRW-based System with Hot Gas Cleanup

A schematic of an air-blown KRW-based IGCC system with hot gas cleanup is shown in Figure A-15. The schematic represents process elements based on design and cost studies prepared for GRI (Smelser, 1986a; Earley and Smelser, 1988b) and the configuration assumed in the ASPEN simulation model (Craig, 1988). The primary features of a hot gas cleanup design, compared to the IGCC system with cold gas cleanup presented in Figure A-15, are: (1) elimination of an oxygen plant; (2) in-situ desulfurization with limestone or dolomite; (3) external (e.g., not in the gasifier) desulfurization using a high temperature removal process; (4) reduced requirement for syngas cooling prior to desulfurization; (5) elimination of sulfur recovery and tail gas treating; and (6) addition of a circulating fluidized bed boiler for sulfation of spent limestone (to produce an environmentally acceptable waste) and conversion of carbon remaining in the ash. In the GRI studies, recycle syngas is cooled for the purposes of pneumatic transport of coal and ash, and for cooling of the raw syngas through mixing. The sour condensate from the raw gas cooler knock-out drum must be treated prior to discharge. However, the ASPEN flowsheet contains no provision for estimating process condensate streams which require treatment. Instead, the design basis for the ASPEN flowsheet includes the use of water quench, rather than heat exchange, for high temperature syngas cooling. Therefore, there is no knock-out drum for process condensate removal.

Many of the direct cost models developed in Section A.3 are applicable to an air-blown KRW-based system with hot gas cleanup. These cost models include:

1. Coal Handling (Area 20);
2. Boiler Feedwater System (Area 80);
3. Gas Turbine (Area 91);
4. Heat Recovery Steam Generators (Area 92);
5. Steam Turbine (Area 93);
6. General Facilities (Area 100).

The reader is referred to the discussions of these systems in Section A.3. In this section, the direct capital cost models required in addition to those above are presented. These cost models include:

1. Limestone Handling (Area 25);
2. Oxidant Feed (Area 10);
3. Gasification (Area 30);
4. Zinc Ferrite Desulfurization (Area 50);
5. Sulfation (Area 35).

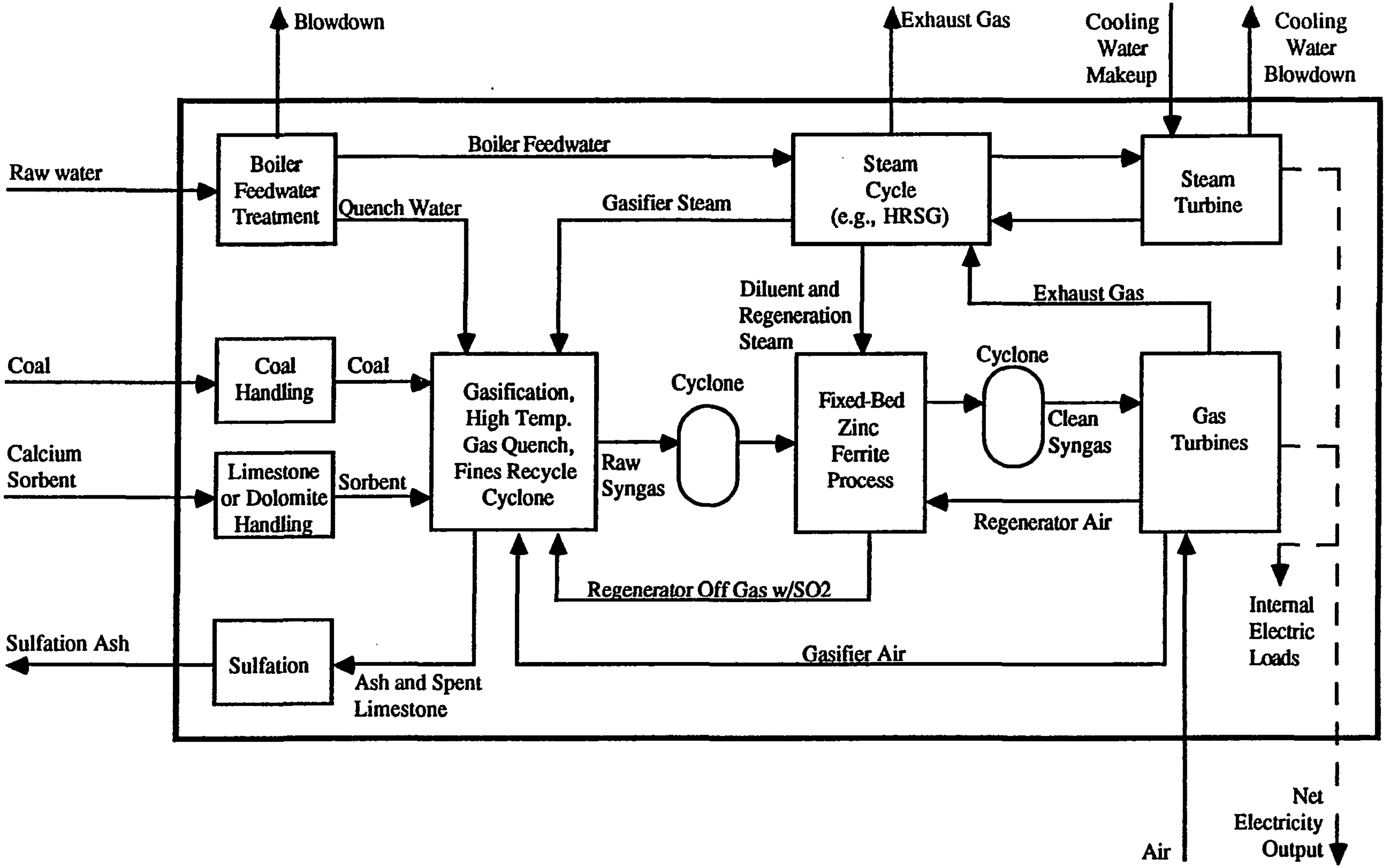

Figure A-15. Simplified Schematic of Air-Blown KRW IGCC System with Hot Gas Cleanup

The additional direct cost models required for a hot gas cleanup system are presented in the following subsections. The cost correlations presented in this chapter are generally more tentative, with a higher degree of uncertainty, than the cost curves presented in the previous chapter. This is due partly to the limited amount of data available for this advanced technology system, and to the risks of cost growth associated with new process technology.

In this analysis, it is assumed that there is no need for cooling of the fuel gas exiting the desulfurization unit prior to entering the gas turbine. Furthermore, no allowances are made for either steam injection or fuel gas moisturization, because these are not included in the ASPEN flowsheet simulation. Instead, in the ASPEN simulation, raw gas cooling takes place by adding water to the hot syngas, thereby moisturizing the syngas and reducing NO_x emissions in the gas turbine. One design study proposes the use of catalytic NO_x reduction in the gas turbine exhaust gas, presumably using selective catalytic reduction (SCR) (Earley and Smelser, 1988b). This option is not considered in this study.

The discussion of each process area will provide an overview of the design basis for each section. Earley and Smelser (1988b) provide additional detail on the limestone handling, gasification, zinc ferrite desulfurization, and sulfation process areas. Corman (1986) provides a discussion of air extraction from the gas turbine for oxidant feed to air-blown gasifiers.

A.4.1 Limestone Handling

Limestone or dolomite are used as reagents for in-bed removal of sulfur compounds in the gasifier reactor. Limestone handling facilities are required to receive limestone, store it, and deliver it to the coal surge hoppers of the gasifier trains. A typical limestone handling design for a commercial scale system includes: an unloading hopper to receive minus 2 inch limestone from a unit train; a conveyor to a storage silo with seven day capacity; a conveyor from the storage silo to an impact crusher to produce minus 12 mesh limestone; two pulverized limestone storage conveyors with a combined capacity of 24 hours; pneumatic transport to the coal/limestone surge bin in the gasification section; and a dust suppression system. Three cost estimates are available for limestone handling systems associated with hot gas cleanup IGCC systems (Smelser, 1986a; Smith and Smelser, 1987; Earley and Smelser, 1988b). From these data a simple linear regression model of direct cost as a function of limestone feed rate was developed:

$$DC_L = 1{,}160 + 0.026\, m_{L,G,i} \qquad (R^2 = 0.907;\ n = 3) \qquad \text{(A-58)}$$

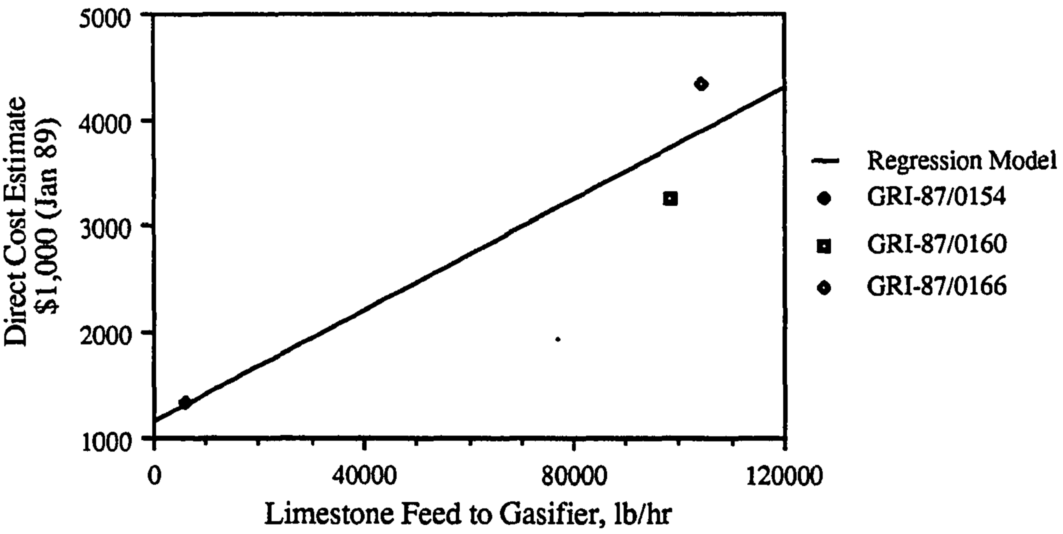

Figure A-16. Direct Cost of Limestone Handling Section.

The standard error of the estimate for this model is $655,000. A graph of the model is shown in Figure A-16. Because of the small number of data points, the statistical significance of the model is questionable. The significance level for the F-ratio of this model is approximately 0.2, compared to far less than 0.001 for the models developed for the KRW-based IGCC system with cold gas cleanup. Thus, it is not possible to reject the hypothesis that the regression model is statistically insignificant (i.e. that the regression coefficient of the predictive term is zero) based on commonly used significance levels of 0.01 or 0.05. However, it is clear from an engineering standpoint that the cost of the limestone handling system does increase as the limestone throughput capacity increases. Therefore, the inability to reject the hypothesis that the model is statistically insignificant is the result of a small number of data points, rather than due to a lack of relationship between cost and limestone flow rate. The low F-ratio for this model implies a wide confidence interval for the regression coefficient of the limestone flow rate.

Although the coefficients of the regression model are not known with certainty, the cost of the limestone handling system is small compared to other parts of the system. Therefore, uncertainty about the estimated costs for the limestone handling system will be a very small contributor to uncertainty in the total cost of the IGCC system. Furthermore, the technology in the limestone handling system is commercially available, which implies that the limestone handling process area will not be a source of cost growth or technical failure of an IGCC system.

A.4.2 Oxidant Feed

For air-blown gasification systems, gasifier air is extracted from the gas turbine compressor and is further boosted in pressure using a boost compressor. For this analysis, it will be assumed that the only change compared to an oxygen blown system is the removal of the oxygen plant and addition of an air boost compressor and air precooler for each gas turbine.

Air extraction reduces the amount of air that enters the gas turbine combustor. However, in air-blown gasification systems, the fuel gas contains a substantial amount of nitrogen. Therefore, coal gases from air-blown systems have a substantially lower heating value than coal gases from oxygen-blown systems. The reduced mass flow rate of air to the gas turbine combustor due to air extraction is thus offset (more or less) by the increased mass flow rate of the low-BTU coal gas. Furthermore, in order to maintain the design mass flow rate of exhaust gas through the turbine, the air mass flow rate into the compressor can be restricted through the use of variable position inlet guide vanes (IGV) (NUSCo, 1988). For example, the exhaust gas flow rate for the GE Frame 7F gas turbine is approximately 920 lb/sec when operating at an ambient temperature of 59 ^{o}F. A study of a Shell-based phased IGCC plant indicates a compressor inlet air flow of 888 lb/sec when firing natural gas, which is reduced to 814 lb/sec when co-firing medium-BTU coal gas and distillate oil at the same ambient temperature (VEPCo, 1989). The use of IGVs allows flexibility in gas turbine operation for a variety of fuels. Therefore, it is not expected that the gas turbine will require substantial redesign due to air extraction.[1]

In the METC ASPEN simulation of an air-blown KRW-based IGCC system, the boost air compressor is assumed to be a single stage device with an isentropic efficiency of 88 percent (Craig, 1988). The outlet pressure from the GE Frame 7F compressor is approximately 198 psia, assuming a pressure ratio of 13.5. A typical pressure ratio for the air boost compressor is approximately 2 or 3. The ASPEN model calculates the shaft work required for the compressor. This information can be used to estimate the direct cost of the compressor and the electric motor drive for the compressor.

The ASPEN simulation also assumes precooling of the air extracted from the gas turbine prior to boost compression. The cooling occurs in two heat exchangers, one to heat

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[1] Modifications may be required to handle the lower heating value fuel gas generated by air-blown systems, as compared to oxygen-blown systems.

low pressure feed water and the other to heat high pressure feedwater. Based on the heat duty, inlet and outlet temperatures, and an assumed universal heat transfer coefficient, the required heat transfer area can be estimated, and this information can be used to estimate the direct cost of the precoolers.

Several sources of compressor cost data were analyzed. These include a conceptual design study for air-blown Lurgi-based systems with hot gas cleanup (Corman, 1986), the ASPEN Cost Estimating Manual (Schwint, 1986), Chemical Engineering magazine (Hall, Matley, and McNaughton, 1982) and chemical engineering texts (Peters and Timmerhaus, 1980; Ulrich, 1984). Based on this review, the compressor shaft work input was selected as the sizing parameter. The Corman study includes budgetary cost estimates for the boost air compressor and the electrical power requirements for these units. From this information, a simple cost correlation was developed based on the compressor shaft work requirement:

$$DC_{CM} = -98 + 26\, N_{T,CM}\left(\frac{W_{s,CM}}{N_{O,CM}}\right)^{0.485} \qquad (R^2 = 0.819;\ n = 6) \qquad \text{(A-59)}$$

where,

$$1{,}900 \le \left(\frac{W_{s,CM}}{N_{O,CM}}\right) \le 8{,}200\ \text{kW}$$

The standard error of the estimate is \$230,000. The model is shown graphically in Figure A-17. The F-test of significance for this model indicates that it is possible to reject the hypothesis that the regression coefficient of the predictive term is equal to zero at a 0.01 significance level. Thus, we may conclude that the model is statistically significant. From previous engineering experience, it is possible that this model may under-predict the cost of large compressors. For example, the scaling exponent for compressors from Ulrich (1984) is approximately 0.94, while Peters and Timmerhaus (1980) report approximately 0.7, and Hall et al (1982) shows approximately 0.94. However, unlike these other data sources, the estimate based on data from Corman (1986) is specific for this application. Furthermore, the cost of the boost air compressor is a relatively small component of the total IGCC system direct cost, so the effects of any errors in cost are not expected to be significant. The model should not be extrapolated.

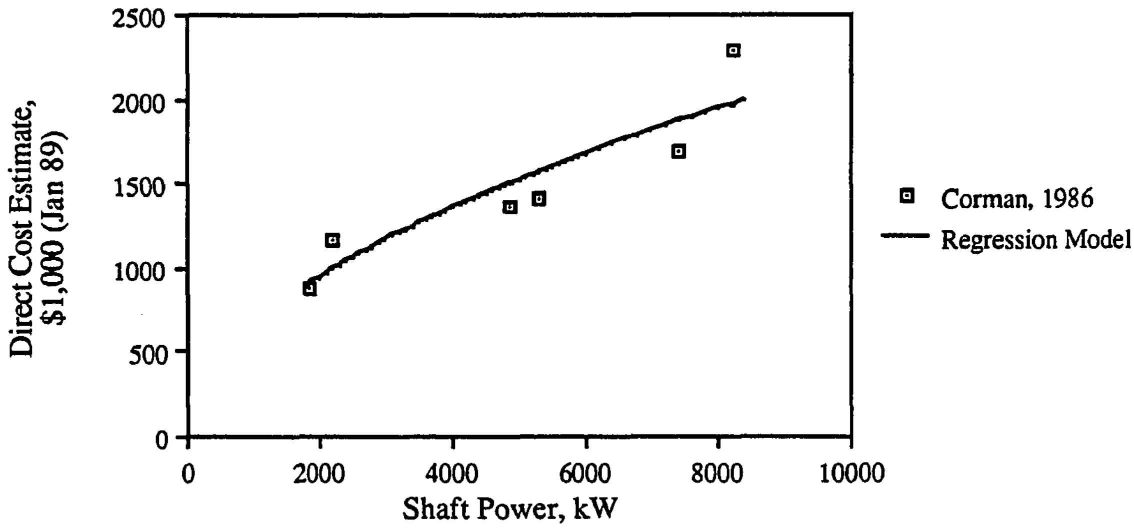

Figure A-17. Direct Cost of Air Boost Compressor.

The cost of the boost compressor precoolers was developed based on a cost curve for fixed tube sheet shell and tube heat exchangers in Ulrich (1984), adjusted for compatibility with cost estimates of precoolers in Corman (1986). The precooler is used to reduce the temperature of the air extracted from the gas turbine compressor prior to entering the boost compressor. The extraction air is cooled by heating boiler feed water. The heat exchanger cost depends on the heat transfer area, which in turn depends on the heat flow, universal heat transfer coefficient, and temperature difference as given by:

$$A = \frac{q}{U \Delta T_{LM}} \tag{A-60}$$

For heat transfer between air and water, a heat transfer coefficient of 40 Btu/(ft^2-hr-^{o}F) was assumed based on Ulrich (1984).

The cost curve for fixed sheet shell and tube heat exchangers in Ulrich was approximated by a cubic polynomial. The heat transfer area for the six different size precoolers used in the Corman study was estimated using the assumed heat transfer coefficient and the reported values of air flow, water flow, and air temperatures. The data reported in the cost study was used to determine the heat flux in the heat exchanger and the log mean temperature difference. The estimated equipment purchase cost from the cost curve in Ulrich was then adjusted to reproduce, as closely as possible, the cost estimates in the Corman study. This was done using a regression analysis of the purchased equipment cost estimated from the Ulrich curve and the reported direct cost estimate for each of the six precoolers in the cost study. The resulting cost curve is shown in Figure A-18, which is a graph of the total direct cost of a single heat exchanger versus the heat transfer area. A

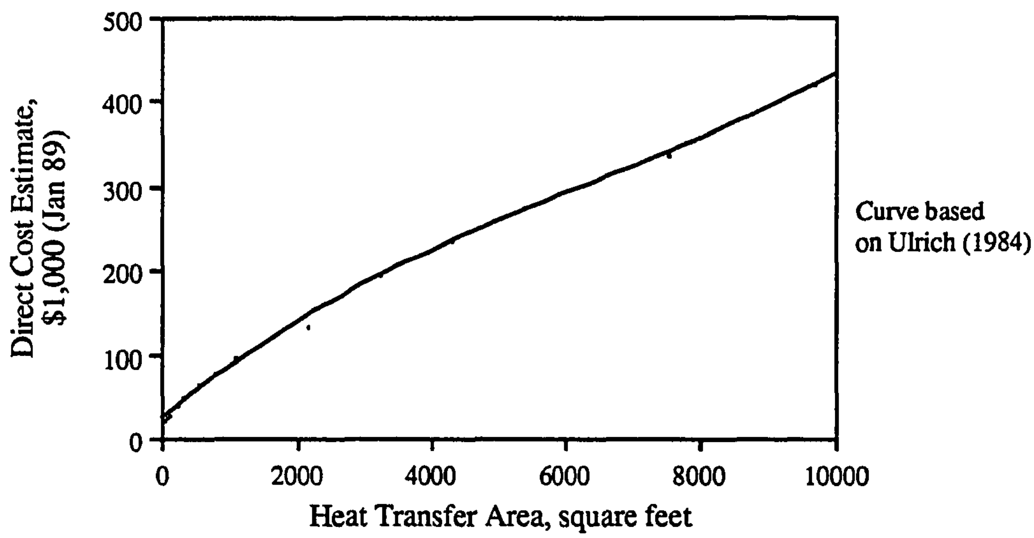

Figure A-18. Direct Cost of Precooler Heat Exchanger.

comparison of the direct cost estimate and the reported cost estimates for the six precooler cases is given in Figure A-19. The cost correlation is:

$$DC_{PC} = (24 + 0.067A - 5.5x10^{-6}A^2 + 2.8x10^{-10}A^3)\ N_{T,PC} \qquad \text{(A-61)}$$
$$(R^2 = 0.972;\ n = 6)$$

where,

A = Heat transfer area per precooler, ft^2

The standard error of the estimate is \$30,000. The estimate of goodness of fit is based on the data shown in A-19. The cost of the precoolers is on the order of \$0.5 million for a 500 MW system. This is not a significant percentage of the total plant capital cost.

The direct cost for the oxidant feed section for an air-blown system is the sum of the boost compressor and precooler costs, and is given by:

$$DC_{OF} = DC_{CM} + DC_{PC} \qquad \text{(A-62)}$$

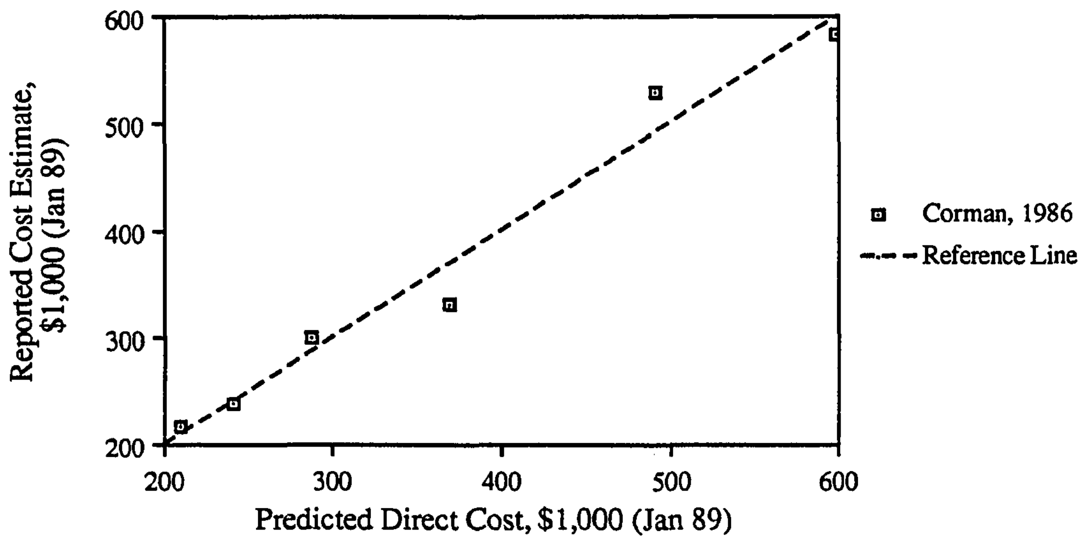

Figure A-19. Comparison of Reported vs. Predicted Precooler Costs.

A.4.3 Gasification

In the currently available design studies for KRW-based systems with hot gas cleanup (Smelser, 1986a; Smith and Smelser, 1987; Earley and Smelser, 1988b), the gasifier is oxygen-blown with in-bed desulfurization using limestone. Although the amount of syngas cooling is reduced compared to cold gas cleanup designs, high temperature syngas cooling cannot be eliminated. The gasifier syngas exit temperature in the detailed studies is 1,850 ^{o}F, which must be cooled below 1,200 ^{o}F prior to entering the hot gas desulfurization system.

The equipment assumed for the gasification section in the hot gas design studies includes the following:

- coal and limestone pressurization (surge bin, pressurization lockhopper, feed lockhopper)
- gasifier
- solids (ash and spent sorbent) removal
- raw syngas recycling and cyclone
- primary heat exchanger
- particulate filter (sintered metal candle filter or ceramic candle filter)
- gas recycle cooler
- secondary gas cooler
- cooling gas compressor
- recycle gas condenser

• recycle gas compressor
• knock-out drum
• steam drum.

Sulfur removal during gasification using calcium-based sorbents has been demonstrated in the KRW process development unit (PDU). The sulfur absorption rate depends on the molar calcium-to-sulfur ratio (Ca/S) of the sorbent, the type of sorbent material used, and the sulfur content of the coal, among other factors. For a given sorbent, the sulfur removal increases as the Ca/S ratio is increased; however, insufficient data are reported in the literature reviewed to date to develop a regression relationship between sulfur removal and Ca/S ratio. A few different types of calcium-based sorbents have been used in testing, including high calcium, magnesium, and dolomitic limestones. The dolomitic limestones have shown the most consistent sulfur removal performance. The sulfur removal efficiency is limited by equilibrium constraints, and the equilibrium sulfur removal efficiency increases as the coal sulfur content increases (Schmidt, Sadhukhan, and Lin, 1989).

The nominal expected in-bed desulfurization for a high sulfur coal is 90 percent (Haldipur et al, 1988). For dolomitic limestone, this removal rate is commonly assumed to occur at a Ca/S ratio of 1.8 (e.g., Haldipur et al, 1988; Earley and Smelser, 1988). However, the actual Ca/S ratio required to achieve 90 percent sulfur removal for a given coal and sorbent also depends on the residence time in the gasifier. Based on a graph presented by Haldipur et al (1988) for the PDU using a 4.5 percent sulfur coal, 90 percent sulfur removal may be achievable with dolomitic limestone at a Ca/S ratio as low as 1.4. In contrast, for a high calcium limestone, a minimum Ca/S ratio is approximately 2.4, and a conservative assumption would be a ratio of 4. Additional data are reviewed in Appendix B.

There is general agreement that calcium-based sorbents catalyze the gasification reactions, increasing the reactivity of eastern coals. For example, the reactivity of Pittsburgh No. 8 coal appears to triple with the use of a limestone sorbent (Floyd and Agrawal, 1989). This implies that carbon conversion efficiencies may be higher with in-bed desulfurization. A sorbent may also reduce the caking tendency of bituminous coals, thereby allowing the gasifier to operate at higher temperatures, which also would tend to increase carbon conversion (Shinnar, Avidan, and Weng, 1988). Based on PDU tests with limestone, the design operating temperature for the Appalachian clean coal technology demonstration project featuring a KRW gasifier using a high sulfur eastern coal was raised from 1,850 ^{o}F to 1,900 ^{o}F (Banchik, Buckman, and Rath, 1988).

There is less consensus on the effect of calcium sorbents on the environmental performance of the gasifier. For example, a conceptual design study (Earley and Smelser, 1988) reports that ammonia production is less with in-bed desulfurization than without, particularly for high calcium limestone. But an environmental study by Radian (Scheffel and Skinner, 1988) based on testing of the PDU does not indicate any reduction in ammonia production with dolomite compared to gasification without a sorbent. For the KRW system with cold gas cleanup, ammonia yield from the gasifier is not an air pollution concern because the ammonia is almost completely removed by wet scrubbing. However, for the hot gas cleanup system, the ammonia yield will affect NO_x emissions in the gas turbine combustor (e.g., Cincotta, 1984). Earley and Smelser (1988) also report that pilot plant data indicate reduced production of methane with sorbents. The increased gasification rate resulting from use of a calcium sorbent is reported to increase fines consumption in the gasifier and reduce fines elutriation (Banchik, Buckman, and Rath, 1988).

To accommodate in-bed desulfurization, the gasifier vessel may be slightly increased in size compared to no in-bed desulfurization. For example, one study assumed a gasifier size of 101 feet overall length and 14 feet maximum outside diameter without in-bed desulfurization, and 115 feet overall length and 14 feet maximum outside diameter with in-bed desulfurization (Smith and Smelser, 1987). The fluidized bed height for the in-bed desulfurization case is approximately 4 feet higher, due to increased bed volume. The limestone addition results in high levels of ash in the bed and higher bed densities than the conventional gasifier. The higher bed density permits a slightly higher superficial velocity (1.72 ft/s vs. 1.6 ft/s) in the freeboard (uppermost) zone of the gasifier.

Because only three cost estimates for an oxygen-blown coal gasification system with in-bed desulfurization and hot gas cleanup are currently available, the number of variables that can be included in a regression model is limited. The gasifier coal feed rate was identified as the single most important parameter affecting cost. All three data points are based on a Pittsburgh No. 8 coal feedstock. Two alternative simple models of the coal gasification cost were developed:

<u>Case</u> 1: $DC_G = 12{,}200 + 0.205\, m_{cm,G,i}$ $\quad (R^2 = 0.999;\ n = 3)$ (A-63)

<u>Case</u> 2: $DC_G = 16.5\, (m_{cm,G,i})^{0.68}$ $\quad (R^2 = 0.999;\ n = 3)$ (A-64)

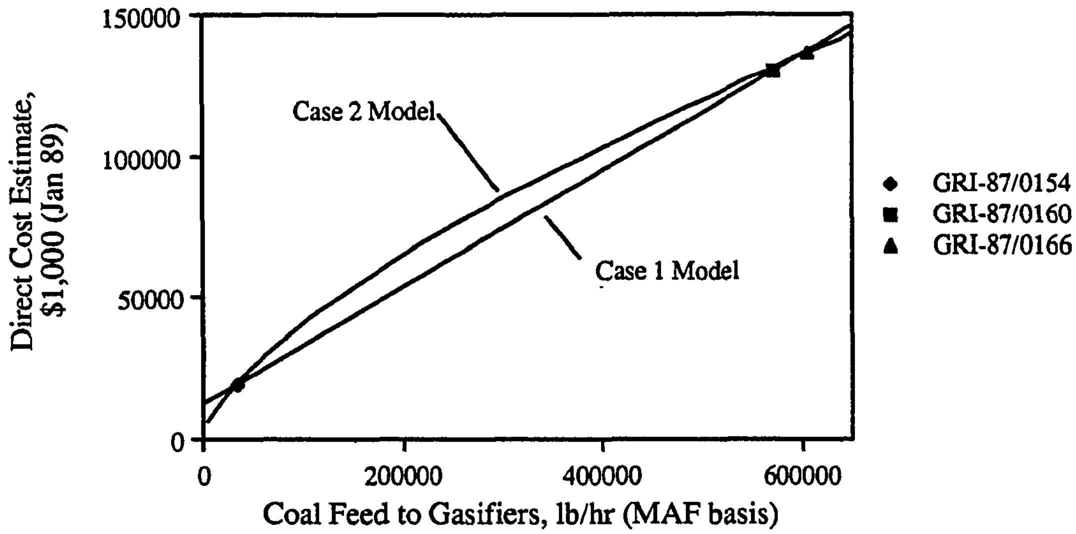

Figure A-20. Direct Cost of Oxygen-Blown Gasification Section for High Temperature Gas Cleanup IGCC System.

The Case 1 model is a simple linear fit to the data, and the Case 2 model is a traditional "capacity-exponent" model commonly assumed in chemical engineering cost estimation. A graph of the three cost estimate data points and the two alternative cost models is given in Figure A-20. The two cost models produce similar results over the range of coal feed rates shown in the figure, with the Case 2 model slightly conservative compared to the Case 1 model for coal feed rates below about 600,000 lb/hr on a moisture and ash free basis. The Case 2 model implies that there is an economy of scale to building larger plants. The standard error of these models is negligible. The models are statistically significant at the 0.01 significance level, based on the F-test.

Equations (A-63) and (A-64) are valid only for oxygen-blown systems. In order to handle the larger oxidant and syngas volume flow rates, the cost of an air blown system would be higher than a comparable oxygen-blown system. In an air-blown system, the syngas flow rate is substantially increased. The components of the gasification section that are affected by significant changes in the syngas volume flow rate include the cyclones, heat exchangers, and particulate filters. The gasifier vessel dimensions are also affected. The coal feed and ash removal systems would be unaffected.

A correction factor for the cost of an air-blown gasification system compared to an oxygen-blown gasification system can be estimated based on a few assumptions. Comparing results from the two KRW ASPEN simulations (oxygen-blown with cold gas cleanup vs. air-blown with hot gas cleanup), the syngas flow rate for the air-blown system is larger than for the oxygen-blown system by a factor of 2.5. If approximately one-quarter of the total direct cost of the gasification section is affected by increases in the syngas flow rate, and if the relationship between cost and syngas flow rate can be modeled using a capacity-exponent formulation, then it is reasonable to assume, based on the Case 2 model presented as Equation (A-64), that the scaling exponent is approximately 0.7. If the syngas flow rate increases by a factor of 2.5 for an air-blown system, then the direct cost of an air-blown gasification section is approximately 20 percent greater than for an oxygen-blown system.

There is an additional discrepancy between the ASPEN flowsheet design and the available cost data. The flowsheet assumes gas cooling by quenching the high temperature raw gas with a sufficient quantity of boiler feedwater. The cost data includes high temperature gas cooling using heat exchangers to generate steam. Therefore, the capital cost for this process area would tend to be over-estimated for a given syngas flow rate. Because cost estimates are not directly available for an air-blown system with raw gas quench, some judgment is required to adjust the costs accordingly.

Some basis for the adjustment is available from the studies for Texaco-based IGCC systems, in which alternative gas cooling systems were considered (Fluor, 1984; Fluor, 1986). In place of a radiative syngas cooler and a convective syngas cooler, a case is considered in which the raw gas enters a quench chamber, where water is introduced. From the Texaco study, it appears that the high temperature gas cooling equipment is a significant part of the total cost. In the Texaco system, the raw syngas has a temperature of 2,400 to 2,600 ^{o}F, compared to 1,850 ^{o}F typically assumed for KRW-based systems. The gas is cooled to 450 ^{o}F in the Texaco design. For the KRW-based system with hot gas cleanup, cooling is required only to about 1,100 ^{o}F. Therefore, the proportional decrease in cost of the quench vs. radiative plus convective design for the Texaco system is probably higher than for a KRW system with hot gas cleanup. The quench system has a direct cost lower than the radiative plus convective system by a factor of 0.43. We might then assume that for the KRW system, the cost of a quench system is on the order of a factor of 0.75 of the cost of a system with syngas heat exchangers.

For initial analyses, this implies that uncertainty in the cost of the gasification process area due to process configuration can be represented by two multiplicative factors. The first factor adjusts the costs of a system with syngas coolers to the cost of a system with raw gas quench, and is initially assumed to be 0.75. The second factor adjusts the costs of an oxygen-blown system to the cost of an air-blown system, based on the difference in volumetric gas flow rates. This factor is initially assumed to be 1.2. These uncertainties can be resolved by developing a more detailed cost model based on the design assumed in the ASPEN simulation. Such a model is not possible at this time because there is no available cost estimate of an air-blown system with raw gas quench with which to develop a design basis and validate the model. Thus, the preliminary cost model assumed for the air-blown KRW-based system with raw gas quench is:

$$DC_G = (0.75)(1.2)\left[16.5\,(m_{cm,G,i})^{0.68}\right] \tag{A-65}$$

Clearly, this is an approximate, study-grade estimate. Additional detail on the costs of specific equipment in the gasification section would be required to refine the correction factors or to develop a new model.

A.4.4 Desulfurization

Three studies of KRW-based systems with zinc ferrite desulfurization have been prepared for GRI (Smelser, 1986a; Smith and Smelser, 1987; Earley and Smelser, 1988b). These studies are based on a fixed-bed design, in which a minimum of two reactor vessels are required. In the fixed-bed design, one vessel operates in the sulfur absorption mode, while the other vessel operates in the sorbent regeneration mode. In the absorption mode, hydrogen sulfide and carbonyl sulfide react with zinc ferrite, and are thus removed from the syngas stream. In the regeneration mode, air and steam are reacted with the spent sorbent, resulting in an off-gas containing primarily water vapor with a concentration of about 5 percent sulfur dioxide. This off-gas is recycled to the gasifier in the KRW-based systems.

The design basis for the zinc ferrite system was significantly changed in the latest of the three studies. The change was in the amount of sulfur loading capacity in the sorbent at which regeneration of the sorbent is required. In the two earlier studies, it was assumed that a nearly theoretical level of over 30 weight percent sulfur in the spent sorbent could be achieved. However, based on recent tests the design basis was updated to a sulfur loading of only 10 percent, substantially increasing the sorbent requirement. As a result, the number of reactor vessels per gasifier train was increased from two (one absorbing and one regenerating at any given time) to four (two absorbing and two regenerating). The likely reason for the increase in the number of vessels (although not stated explicitly in the report) is a constraint on the maximum vessel diameter which is economically transportable (limited by railroad clearances) for shop fabricated pressure vessels, as discussed by Kasper (1988).

Cost estimates for fixed bed zinc ferrite systems were developed as part of analyses of fixed bed gasification IGCC systems (Corman, 1986). However, insufficient definition of the design basis, and aggregation of desulfurization and sulfur recovery costs, precludes use of these estimates for development of cost models. For example, design parameters such as the absorption cycle time and the maximum assumed sulfur loading in the sorbent just prior to regeneration are not given.

A study of hot gas desulfurization processes was performed for METC (Klett et al, 1986). However, in this report, all cost estimates were normalized, with no indication of how the normalized costs convert into absolute dollar costs.

Because a fixed-bed zinc ferrite design is assumed in the cost studies, it is adopted as the basis for a preliminary cost model. Because so few cost estimates of this technology are available, and because the zinc ferrite plant section consists primarily of vertical pressure vessels, a preliminary estimate was developed based on published cost curves for vertical pressure vessels. These costs were used with the KRW-based system studies to estimate costs for sorbent handling, piping, valving, and the control system. The cost model is based on equipment cost curves in Ulrich (1984), a design basis for fixed bed zinc ferrite systems developed by Kasper (1988), and cost and design data from the most recent IGCC study with zinc ferrite hot gas cleanup (Earley and Smelser, 1988b). In addition, a comparison of the cost model prediction and the cost estimate from another hot gas cleanup gasification system study (Smith and Smelser, 1987) was made to verify the model.

Cost curves for vertical pressure vessels from Ulrich (1984) were used to develop an analytic expression for pressure vessel cost as a function of vessel internal diameter, length, and operating pressure. Assumptions in the design basis governing the maximum allowable syngas velocity through the vessels determine the minimum vessel diameter. The maximum vessel diameter is assumed to be limited to the largest diameter that can be shop fabricated and rail transported, which is 12.5 feet. A length-to-diameter ratio of 4 is assumed for simplicity, representing a reasonable economical aspect ratio (Kasper, 1988).

The number of operating absorber vessels is assumed to be at least as great as the number of operating gasifiers, but may be larger by a factor of two, four, etc, depending on limitations imposed by maximum gas velocity or large sorbent requirements. The cost of auxiliary equipment, such as sorbent handling, piping, valving, and control systems, was estimated by difference between the pressure vessel cost estimate and the total direct cost estimate in the most recent IGCC cost study with hot gas desulfurization (Earley and Smelser, 1988b). This study includes five operating and one spare gasification train. Each train includes a gasifier; associated raw gas handling equipment, and one zinc ferrite process train. Each zinc ferrite process train consists of four vertical reactor vessels, of which two are in an absorption mode and two are in a regeneration mode at any given time. The plant has a total of 24 zinc ferrite vessels, of which ten are in operation during full plant load.

The raw syngas flow rate into the zinc ferrite section in the Earley and Smelser (1988b) study is estimated to be 61,200 lbmole/hr, on a plant basis. Ten zinc ferrite absorbers are operated to remove sulfur from this syngas, with an inlet temperature of 1,125 ^{o}F and inlet pressure of 453 psia. Therefore, the volume flow rate per absorber is approximately 64 ft^3/sec, assuming that the syngas is an ideal gas.

The absorption cycle is assumed in the study to be 168 hours. The sulfur flow rate in the syngas is 1,734 lb/hr of sulfur, which is in the form of hydrogen sulfide and carbonyl sulfide. Therefore, 291,400 lb of sulfur must be absorbed on a plant basis every 168 hours.

To estimate the amount of sorbent required to absorb this amount of sulfur, a simple relationship was developed based on the stoichiometry of the absorption reactions. The main absorption reactions are:

$$6\ ZnFe_2O_4 + CO + H_2 \leftrightarrow 6\ ZnO + 4\ Fe_3O_4 + CO_2 + H_2O \qquad \text{(A-66)}$$

$$ZnO + H_2S \leftrightarrow ZnS + H_2O \qquad \text{(A-67)}$$

$$Fe_3O_4 + 3\ H_2S + H_2 \leftrightarrow 3\ FeS + 4\ H_2O \qquad \text{(A-68)}$$

and it is assumed that all carbonyl sulfide reacts to form hydrogen sulfide according to:

$$COS + H_2O \leftrightarrow H_2S + CO_2 \qquad \text{(A-69)}$$

The overall absorption reaction is then:

$$6\ ZnFe_2O_4 + 18\ H_2S + 5\ H_2 + CO \leftrightarrow 6\ ZnS + 12\ FeS + 23\ H_2O + CO_2 \qquad \text{(A-70)}$$

If 6 moles of zinc ferrite, weighing 1,447 pounds, react completely, then 576 pounds of sulfur are absorbed and the spent sorbent (consisting of ZnS and FeS) weighs 1,638.5 pounds, assuming no inert materials in the sorbent. From these relationships it is apparent that the maximum theoretical sulfur loading is 35 percent, if all the sorbent reacts. However, test results and expectations for commercial scale design are for less than theoretical loading, with long term loading as low as 10 percent (Kasper, 1988). Let the fresh sorbent charge to the operating absorber vessels be S_C, the weight fraction of sulfur loading in the spent sorbent be L_S, the absorption time be t_a, and the weight of the spent sorbent at the end of the absorption cycle be S_S, then:

$$M_{s,ZF,i} = (M_{HS,ZF,i} + M_{COS,ZF,i}) t_a \tag{A-71}$$

$$m_{s,ZF,i} = M_{s,ZF,i} \left(\frac{32 \text{ lb S}}{\text{lbmole S}} \right) \tag{A-72}$$

$$S_s = S_c + 192 \left(\frac{M_{s,ZF,i}}{18} \right) \tag{A-73}$$

$$L_s = \frac{m_{s,ZF,i}}{S_s} \tag{A-74}$$

substituting Equations (A-72) and (A-73) into (A-74) and simplifying, the following result is obtained:

$$S_c = \left(\frac{32 - 10\frac{2}{3} L_s}{L_s} \right) M_{s,ZF,i} \tag{A-75}$$

On a mass, rather than molar basis, Equation (A-75) becomes:

$$S_c = \left(\frac{1 - \frac{1}{3} L_s}{L_s} \right) m_{s,ZF,i} \tag{A-76}$$

Equation (A-75) or (A-76) can be used to determine the initial charge of fresh sorbent required for the operating absorber vessels as a function of the sulfur capture in the absorbers and the long term sorbent loading expected. As a check on the equations, for a maximum theoretical sorbent loading of 35.1 percent and a sulfur capture of 576 pounds, the estimated sorbent charge required is 1,447 pounds, which is the result previously obtained.

At a 10 percent weight loading of sulfur on the spent zinc ferrite sorbent and a sorbent capture of 291,400 pounds per cycle, the total sorbent requirement for the example IGCC plant is estimated to be equivalent to 2.8 million pounds of fresh sorbent charged to the ten on-line absorbers.

The volume of the sorbent charge in each reactor vessel is given by the following equation:

$$V_c = \frac{S_c}{D_b} \tag{A-77}$$

At a sorbent bulk density, D_b, of 82 lb/ft^3 (Kasper, 1988), the total sorbent volume is estimated to be 34,350 ft^3, or 3,435 ft^3 for each on-line reactor vessel. For simplicity, this is assumed to be the vessel volume. The design basis developed by METC indicates that the reactor vessels are only slightly larger in volume than the sorbent bed volume.

The minimum diameter is constrained to keep the superficial gas velocity below 2 ft/sec (Kasper, 1988). In ASPEN, the gas volume flow rate can be obtained directly. However, as an approximation for model use outside of ASPEN, the raw syngas is assumed to behave like an ideal gas. The following equation can be used to estimate the minimum acceptable diameter for each zinc ferrite reactor vessel:

$$d_{min} = 1.128\left(\frac{M_{syn,ZF,i}\left(T_{ZF,i} + 460\right)10.73}{V_s\, P_{ZF,i}\, N_{O,ZF}\, 3{,}600}\right)^{\frac{1}{2}} \tag{A-78}$$

Kasper (1988) recommends a vessel length-to-diameter ratio of two to four, which is a range of the most economical aspect ratios. Furthermore, Kasper recommends that the vessel diameter should not exceed 12.5 feet, which is an upper limit for the size of rail-transportable shop-fabricated vessels. If a length-to-diameter ratio of 4 is assumed, then the reactor vessel inner diameter is given by:

$$d = \left(\frac{V_c}{\pi}\right)^{\frac{1}{3}} \tag{A-79}$$

If the estimated vessel diameter is smaller than the minimum diameter, the number of absorber vessels should be decreased. If the estimated diameter is larger than 12.5 feet, the number of absorber vessels should be increased.

Using the above relations for the example IGCC case, the dimensions of each vessel are estimated to be 10.3 feet inner diameter and 41.2 feet minimum length. The estimated diameter is larger than the required minimum diameter of 6.4 feet, so no adjustment in the number of vessels is required.

To estimate the cost of a range of reactor vessel sizes for the zinc ferrite system, the cost curves in Ulrich (1984) have been approximated as analytic expressions using regression analysis. The cost of the pressure vessel depends on the operating pressure, diameter, and length. The diameter and length are determined based on the sorbent requirements, which in turn depends on an assumption regarding sulfur loading, as indicated in the previous discussion. The cost of the auxiliaries is assumed to scale with the cost of the reactor vessels; for simplicity it is assumed that the cost of the auxiliaries is a simple multiplier of the cost of the pressure vessels. This assumption can be refined if more data become available.

The method for estimating pressure vessel cost is first to estimate the purchased cost of a vertical vessel for operation at low pressures, then to adjust the cost for higher pressures, and finally to adjust the cost to include installation costs. Eight curves for the purchased cost of a low pressure vertical vessel are given in Ulrich (1984, p 307), one for each of eight diameters as a function of vessel height. These curves were approximated using linear regression models, of the form:

$$y = a + bH \tag{A-80}$$

The dependent variable in these estimates was the purchased cost of the vessel, and the independent variable was the vessel height. The coefficient of determination of these estimates ranged from 0.988 to 0.999. The parameters of Equation (A-80) are a function of the vessel diameter. A regression analysis of the slope and intercept of all eight linear cost approximations using a third degree polynomial yielded a single analytical expression for pressure vessel cost as a function of diameter and height, in units of feet. This expression for a single vessel (in thousands of June 1982 dollars) is:

$$\begin{aligned} PC_{ZF,v} = {} & \left(4.1 - 4.35d + 0.958d^2 - 0.0391d^3\right) + \\ & \left(0.23 + 0.166d - 0.018d^2 + 7.87\times10^{-4}d^3\right)H \end{aligned} \tag{A-81}$$

The cost estimate from Equation (A-81) is adjusted to January 1989 dollars using a multiplier of 1.122, based on the *Chemical Engineering* plant cost index. To adjust the cost for increased pressure and installation costs, the following correction factor was developed based on curves presented in Ulrich (1984, p. 308):

$$f_{p,i} = 2.5 + 0.093\left(P_{syn,ZF,i}\right)^{0.64} \tag{A-82}$$

The direct cost of the zinc ferrite plant section also includes piping, valves, and a control system. These auxiliary items were estimated by difference between the vessel cost and the total direct cost estimate presented in Earley and Smelser (1988b). As already discussed, 24 reactor vessels are assumed in the detailed cost study. The total cost for pressure vessels is estimated to be \$10.6 million, in January 1989 dollars. This compares with a reported total direct cost of \$14.1 million, implying that the cost of auxiliaries is \$3.5 million, or 32 percent of the vessel costs.

An attempt was made to verify the cost model against other cost estimates. However, limited design data is available for the study by Corman (1986) and the exact value of the weight percent sulfur loading used in the Smith and Smelser (1987) study is not know, except that it is above 30 percent (Earley and Smelser, 1988b). The upper limit is 35 percent, as already discussed. The cost model was applied to try to reproduce the results of the Smith and Smelser (1987) study.

For the Smith and Smelser study, it is assumed that a sulfur loading of 32 percent was used. The syngas flow rate and sulfur flow rate for each vessel are given, which are very close to the values used in the Earley and Smelser (1988b) study, as is the assumed cycle time of one week. There are five operating vessels and a total of twelve vessels, associated with five operating and one spare gasifier. For each gasifier, there are two zinc ferrite vessels, allowing one to operate in the absorption mode and the other in the regeneration mode. The cost estimated from the equations presented above is \$4.7 million in June 1985 dollars, including the multiplier of 32 percent for auxiliaries. The cost estimate in the GRI study is reported with only one significant figure as \$4 million, also in June 1985 dollars. These costs agree fairly well, and also indicate the sensitivity of the zinc ferrite equipment capital costs to the long term attainable sulfur loading in the sorbent.

The direct cost model for the zinc ferrite section can be summarized in the following equation:

$$DC_{ZF} = 1.477\ f_{p,i}[A(d) + B(d)\ H]N_{T,ZF} \tag{A-83}$$

where

$f_{p,i}$	=	from Equation (A-82);
$A(d)$	=	$4.1 - 4.35d + 0.958d^2 - 0.0391d^3$;
$B(d)$	=	$0.23 + 0.116d - 0.018d^2 + 7.87\times10^{-4}d^3$.

As previously discussed, the vessel diameter should not exceed 12.5 feet, and the minimum diameter for which this model is valid is 1.6 feet. The factor 1.477 in Equation (A-83) includes the multiplier of 1.324 representing the cost of auxiliary equipment. It also adjusts the cost year from mid 1982 to January 1989. The vessel dimensions are determined based on the sorbent requirement per vessel, which depends upon the sulfur flow rate in the syngas, the absorption cycle time, and the long term sulfur loading capacity of the sorbent. The number of vessels should be selected so that the superficial velocity of the syngas is below a threshold value (e.g., 2 ft/sec), and so that the diameter of the vessels is small enough (e.g., less than 12.5 feet) to be shop fabricated and rail transported to the site.

A.4.5 Sulfation

For the air-blown KRW-based system with in-bed desulfurization, the spent sorbent is a waste stream. This sorbent contains calcium sulfide (CaS) which is not acceptable for landfilling. The purpose of the sulfation unit is to oxidize the calcium sulfide to the more stable calcium sulfate ($CaSO_4$), which is then suitable for disposal. The spent sorbent is contained in the gasifier ash waste stream, which also includes unconverted carbon. Therefore, a sulfation unit can recover the heating value of unconverted carbon. A circulating fluidized bed combustor has been proposed for the sulfation application. The heat released from the sulfation reaction and combustion of unconverted carbon is used to generate steam. However, the current ASPEN simulation does not include a model of this system; therefore, the recoverable chemical energy in the ash and spent limestone leaving the gasifier are not included in the plant energy balance.

The sulfation unit is an additional emission source. Fluidized bed combustors generally have more uniform flame temperatures than conventional types of combustion systems; therefore, the NO_x emissions from these systems are comparatively low. The sulfur emissions from the sulfation unit depend on the sulfur content of the feed stream, which is expected to be low. The SO_2 and NO_x emission rates are expected to be 0.01 lb SO_2/MMBtu and 0.15 lb NO_x/MMBtu, respectively. These are well below the NSPS limits for steam generators fired with bituminous coal. Particulate emissions are controlled by a fabric filter to less than 0.03 lb/MMBtu (Earley and Smelser, 1988).

There is uncertainty regarding the effectiveness of the sulfation unit in converting calcium sulfide to calcium sulfate. A test of ash and spent limestone in a research reactor yielded only 25 to 35 percent conversion of the calcium sulfide at solids residence times from one to two hours. A proprietary additive has been formulated to improve the conversion rate, and it is expected to result in 85 percent conversion of the calcium sulfide. However, no information is currently published on the amount required or the cost of the additive. Furthermore, there is currently no assessment of the effect of an incomplete conversion rate or of the additive on the environmental characteristics of the sulfation unit waste stream.

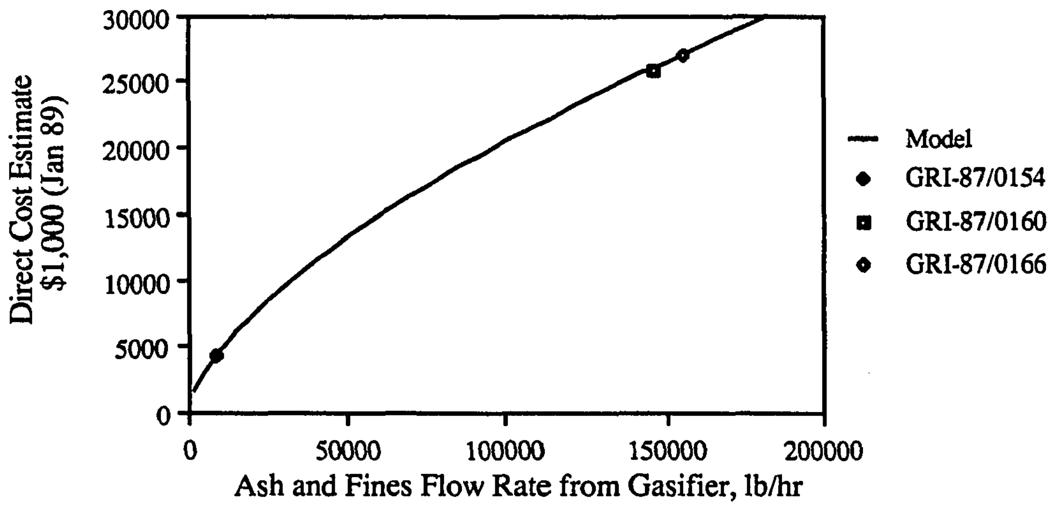

Figure A-21. Direct Cost of the Sulfation Unit.

A simple cost curve for the sulfation section was developed based on the three available KRW-based gasification system studies with hot gas cleanup. This model is:

$$DC_{SF} = 13.0\ (m_{ash} + m_{fines})^{0.639} \qquad (R^2 = 1.00;\ n = 3) \qquad \text{(A-84)}$$

The standard error of the model is negligibly small. The F-ratio of this model is significant at the 0.001 significance level. A graph of this model is shown in Figure A-21.

A.5 Capital Cost of a Lurgi-based System with Hot Gas Cleanup

The plant design assumed as the basis for cost model development is illustrated in Figure A-22. This design is based on an ASPEN flowsheet developed by the U.S. Department of Energy (DOE) simulating an air-blown Lurgi gasifier IGCC system with hot gas cleanup (Klara, Rastogi, and Craig, 1988), and on a design study of simplified IGCC plants (Corman, 1986). The design study by Corman assumes the fixed bed zinc ferrite process for high temperature and high pressure fuel gas desulfurization. The ASPEN flowsheet was originally based on a moving bed zinc ferrite desulfurization design. However, this is a proprietary design of the General Electric Company, and insufficient cost data were available for cost model development. Therefore, the fixed bed zinc process has been assumed for hot gas desulfurization, and the ASPEN performance model was modified by the author accordingly to represent this process. Performance and cost data for the fixed bed zinc ferrite process, as well as technical judgments about uncertainties (see Appendix B) are more readily available. Therefore, the cost model of the fixed bed zinc ferrite process developed in Section A.4 is also applicable to the Lurgi-based IGCC system.

In the air-blown Lurgi system with hot gas cleanup, coal gasification is accomplished in a fixed-bed, dry ash Lurgi gasifier. The Lurgi dry-ash gasifier, which was developed in the 1930's, features a counter-current flow of coal past the steam and oxidant feed streams. This results in a widely varying temperature profile through the reactor, and also in a low syngas exit temperature compared to other gasifiers (e.g., KRW). The Lurgi gasifier has only a limited ability to handle coal fines, which tend to become entrained in the exiting syngas or deposited in the exiting tars (which are a characteristic product of this design). Because entrained fines represent a loss of energy efficiency (due to unconverted carbon), typical designs for Lurgi systems involve sizing of the coal input to the gasifier. The typical coal feed is between 1/4 and 2 inches in size (SFA, 1983). Alternatively, a portion of the coal may be fed to the gasifier in the form of fines (coal particles less than 1/4 inch in size). Provisions for cyclones to collect the fines that pass through the gasifier, and the agglomerating and recycling of fines to the gasifier, are necessary to improve carbon conversion efficiency. This option is assumed in the available cost study, and is therefore adopted as a basis for the cost model.

Because the Lurgi gasifier has a relatively low syngas outlet temperature (typically near 1,000 ^{o}F), high temperature syngas cooling is not required prior to hot gas cleanup.

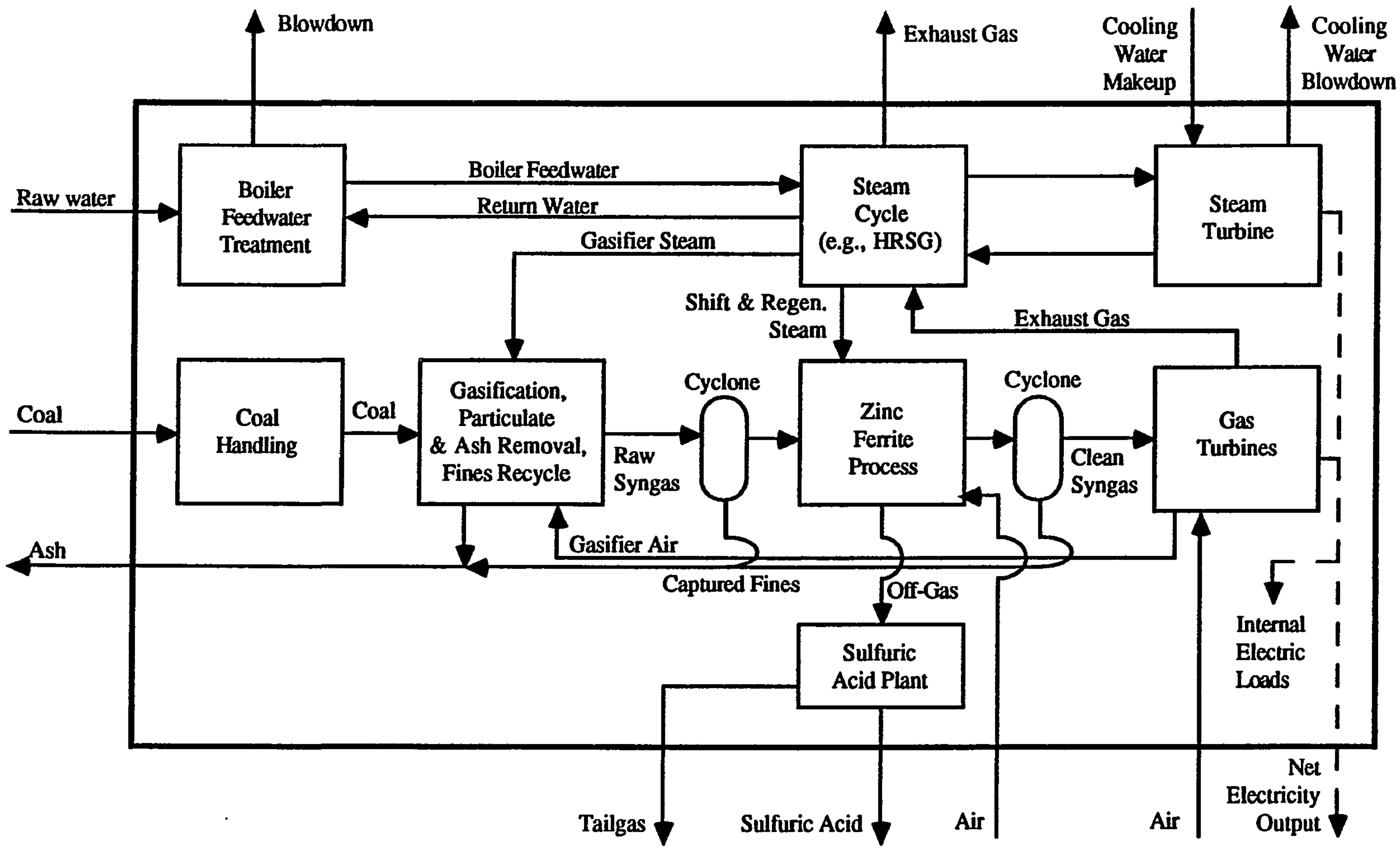

Figure A-22. Schematic of Air-Blown Dry-Ash Lurgi Gasifier IGCC System with Hot Gas Cleanup

The plant design assumed here includes high temperature particulate removal using high efficiency cyclones, and high temperature desulfurization using the zinc ferrite process, as assumed in a recent study (Corman, 1986). The clean syngas is then fed directly to GE Frame 7F gas turbines. The exhaust from the gas turbines passes through heat recovery steam generators, which supply steam for the steam turbine, the zinc ferrite process, and the gasifier. Bleed air from the gas turbine is used as the oxidant in the gasifiers. The off-gas from the zinc ferrite process is delivered to a sulfuric acid plant for byproduct recovery.

In the fixed bed zinc ferrite design, regeneration occurs in cycles, using steam as a diluent and producing an off-gas stream containing primarily water vapor and nitrogen. Air required for the regeneration reactions is supplied by the air booster compressor. The sulfur dioxide concentration is typically below five percent (Corman, 1986). While sulfuric acid plants have processed off-gases containing as little as three percent sulfur dioxide in the copper smelting industry (EPA, 1981), the costs increase significantly as the off-gas volume flow rate increases.

Many of the direct cost models developed in Sections A.3 and A.4 are applicable to air-blown Lurgi-based systems with hot gas cleanup. From Section A.3, the applicable direct cost models are:

1. Boiler Feedwater System (Area 80);
2. Gas Turbine (Area 91);
3. Heat Recovery Steam Generator (Area 92);
4. Steam Turbine (Area 93);
5. General Facilities (Area 100).

From Section A.4, the applicable direct cost models are:

1. Air Boost Compression (Area 10);
2. Zinc Ferrite Desulfurization (Area 50).

In this section, additional direct cost models are presented for the following plant sections:

1. Coal Handling (Area 20);
2. Gasification (Area 30);
3. Coke, Fines, and Ash Handling (Area 31);
4. High Temperature Cyclones (Area 32);
5. Sulfuric Acid Plant (Area 60).

The direct cost models for each of these additional plant sections are presented below. These models are intended only for the purpose of estimating the cost of an air-blown Lurgi-based IGCC system with hot gas cleanup.

A.5.1 Coal Handling

Lurgi gasifiers are constrained in the amount of coal fines that can be included in the gasifier coal feed, due to entrainment of fines in the exit gas and possible instabilities in the gasifier bed (SFA, 1983; Corman, 1986). To accommodate these limitations, which apply to both dry-ash and slagging Lurgi gasifiers, four general approaches have been used in different studies for utilizing coal in Lurgi-based plants. These include: (1) separation and return of all or a portion of the fines to the coal supplier (e.g., Cover et al, 1985b; Smelser, 1986b and 1986c); (2) separation of sized coal for use in Lurgi gasifiers and fines for use in other types of gasifiers, or for combustion in conventional boilers (e.g., Cover et al, 1985b; Smelser, 1986b and 1986c); (3) separation of a portion of the fines for injection directly into the gasifier bed via "tuyeres" (e.g., Bechtel, 1983a); and (4) agglomeration of separated coal fines into briquettes for feed to the gasifier coal surge bin (e.g., Parsons, 1985; Bechtel et al, 1988). All systems require coal receiving, storage, and reclaim facilities. All systems typically include some type of screening to separate coal fines (less than 1/4 inch) from the sized coal (typically 2 x 1/4 inch). The systems vary in the size of the received coal, with many studies assuming that the coal is two inches or smaller when received. In a few studies, larger sized coal is first crushed to the minus two inch size recommended for use in Lurgi gasifiers (Bechtel, 1983; Zahnstecher, 1984).

The assumption in a recent General Electric study (Corman, 1986) is that the coal is received as minus two inch size, and that all of the coal, including fines, are fed directly to the gasifier. In this study, only the fines that are captured from the raw syngas in a primary cyclone are agglomerated into briquettes, using bentonite (a type of clay) as a binder. In other studies that include fines agglomeration, the fines in the coal feed are agglomerated prior to entering the gasifier. The coal assumed in the General Electric study is an Illinois No. 6, which is a caking coal. It is reported that up to 30 percent fines by weight in the coal feed could be handled in the Lurgi gasifier if the coal is a caking type and if the small fines (less than 1/8 inch) met certain size criteria, which are not reported. Corman reports that at most 5 percent of the coal feed is expected to carry-over in the form of fines recovered in the primary cyclone.

The Corman study scaled cost estimates for the coal handling section from a previous study performed by General Electric (Cincotta, 1984). In the Cincotta study, it is assumed that sized coal (2 x 1/4 inch) is delivered by a dedicated unit train operating between the power plant and a supplier of sized coal. The coal handling system includes an unloading hopper for railroad bottom dump cars, vibrating feeders for the hopper cones, a dual train of belt conveyors between the vibrating feeders and the sampling system, and a belt conveyor from the sampling system to the stacker. Two 100 percent coal reclaiming systems are included. Coal is reclaimed by gravity feed from the active storage pile to underground reclaim hoppers. A vibrating feeder supplies a reclaim belt conveyor to a series of belt conveyors that distribute the coal to the gasifier surge bins using traveling trippers.

Corman (1986) reports that the coal handling section costs can be estimated based on the coal feed mass flow rate (lb/hr basis) to the gasifiers and the following relationship:

$$DC_{CH} = 4.3\text{x}10^{-4} \left(m_{cf,G,i}\right)^{0.85} \tag{A-85}$$

The reported costs of this coal handling system are high compared to the cost reported for coal handling for the KRW-based systems (e.g., for a 420,000 lb/hr system, the KRW coal handling cost is estimated to be $24 million compared to $26 million for the Lurgi system). This is unreasonable because, in addition to coal receiving, storage, and reclaiming, the KRW-based coal handling section also includes systems for coal pulverization and coal drying. Furthermore, the General Electric estimate is high compared to the estimates prepared for the Gas Research Institute, which include crushing of minus six inch coal to minus two inches, plus fines separation. However, because no detail on equipment costs is given in any of the studies, it is not possible to determine which study is "correct".

For the purpose of comparing Lurgi and KRW systems, it is important that the relative costs be reasonable. Therefore, a cost correlation based on the Gas Research Institute estimates was developed. This provides consistency with the data used in developing the coal handling cost curve for the KRW-based systems, which included data from four studies prepared for GRI. The fact that that GRI estimates include allowances for coal crushing, fines separation, and fines export can be considered as conservative in cost for a system intended to receive sized coal and deliver all of the coal, including fines, to the gasifiers. Such a coal handling system may be required, however, if the gasifiers are unable to perform as predicted with large fines loading in the feed coal. Furthermore, such a coal handling system allows receipt of less expensive coal, such as minus 6 inch, rather than requiring that the coal be sized prior to delivery to the plant.

An additional data point for the coal handling systems was obtained from a study by Bechtel (Bechtel, 1983a) for a plant size better suited for IGCC systems. Using all

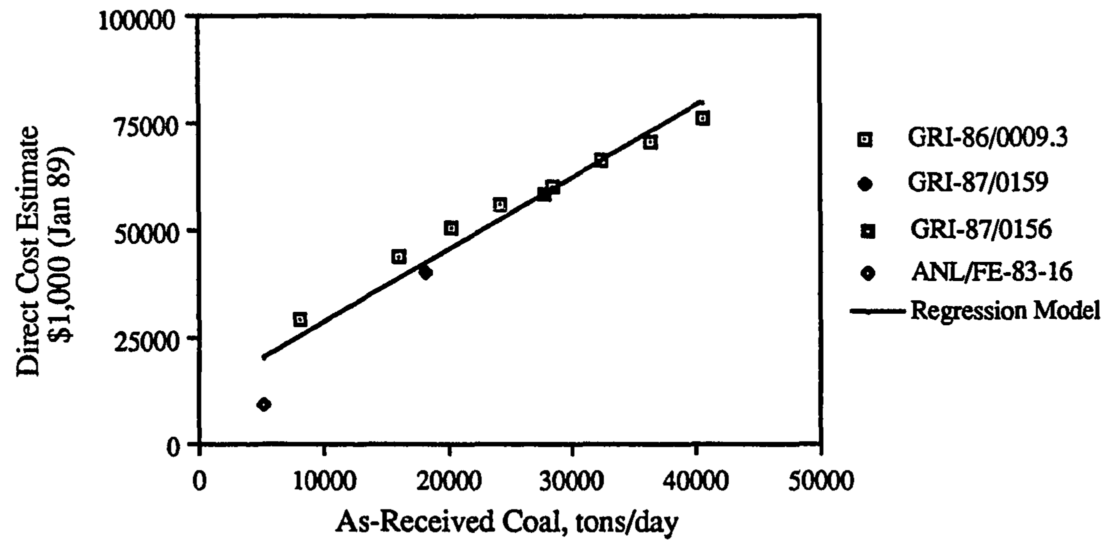

Figure A-23. Direct Cost of Coal Handling for a Lurgi-based IGCC System.

available data, a simple relationship between the cost of the coal handling section and the coal feed rate to the gasifier was developed:

$$DC_{CH} = 11{,}200 + 1.70\ m_{cf,CH,i} \qquad (R^2 = 0.943;\ n = 11) \quad \text{(A-86)}$$

where,

$$5{,}000 \leq m_{cf,CH,i} \leq 36{,}000 \text{ tons/day}$$

The standard error of the estimate is $5 million. Unlike the cost equation developed for KRW-based systems, this one is based on the coal feed into the coal handling system, rather than into the gasifier. However, in the case where all of the coal, including fines, are fed to the gasifier, the as-received mass flow and the gasifier feed mass flow are equivalent. A comparison of the cost data and the regression model is shown graphically in Figure A-23. The data used to develop the model in Equation (A-86) include lignite and Illinois No. 6 coals.

A.5.2 Gasification

The only currently available detailed design study of a commercial-size air-blown Lurgi-based IGCC system with hot gas cleanup is Corman (1986). Other studies of similar Lurgi-based systems have been done, but either no details are given on the development of performance and cost estimates (e.g., Craig and Koch, 1988), or too little information is reported from which to develop a cost model (e.g., Klett et al, 1987).

For the KRW-based systems, the gasification section costs include the gasifier pressure vessels, coal feed hoppers, ash removal hoppers, particulate removal systems, and other equipment. In the Corman study, costs are reported separately for three component parts of the gasification system. These are: (1) the gasifier pressure vessels, coal feed hopper, and ash removal hoppers; (2) coke handling, fines recycle and agglomeration, and ash removal from the ash lockhopper; and (3) particulate removal from the syngas using high temperature, high pressure cyclones. The costs for these gasification subsections are treated separately here.

A.5.2.1 Gasifier and Associated Subsystems

From Corman (1986) and Zahnstecher (1984), the cost per train of a gasifier subsystem can be estimated. The gasifier subsystem includes a coal surge bin, the gasifier pressure vessel, and an ash lockhopper. The coal surge bin is centrally mounted on top of the gasifier pressure vessel. Coal is charged to the surge bin from the coal handling system, and the coal flows from the surge bin to a rotating distributor plate in the gasifier. The gasifier vessel includes a jacket for generation of steam. Ash is removed from the gasifier as a dry solid to a centrally located ash lockhopper located at the bottom of the gasifier.

The typical gasifier design assumed as a basis in cost studies is a Lurgi Mark IV system. This is a commercial offering for oxygen-blown systems. It has a nominal diameter of 12.7 feet, which is at the upper limit of rail-shippable vessel size. The cost per unit of these systems is approximately the same in several cost estimates. However, the capacity of each gasifier, and therefore the total number of gasifiers required for a given plant, depends on the coal throughput that can be handled per vessel. For an Illinois No. 6 coal, the coal throughput estimated by Lurgi, as reported by Corman, is 300 lb/hr-ft^2 of grate area for a system operating at 300 psia, and 500 lb/hr-ft^2 of grate area for a system operating at 600 psia. The grate area can be approximated based on the working diameter of the vessel, which can be assumed to be 12.7 feet. Thus, the coal throughput per gasifier operating on Illinois No. 6 coal may be estimated as follows, assuming a simple linear model:

$$CTP = 127\left(100 + \frac{2}{3} P_{syn,G,o}\right) \qquad \text{(A-87)}$$

Corman states that the throughput values estimated by Lurgi may be optimistic for air-blown systems. Therefore, in actual model applications, the coal throughput estimated by Equation (A-87) might be viewed as an upper bound. (Expert judgments for Lurgi gasifier coal throughput are discussed in Appendix B). This equation is only applicable for the

Mark IV gasifier operating with Illinois No. 6 coal and gasifier pressures ranging from 300 to 600 psia. The coal throughput per gasifier is in units of lb/hr.

The cost of the coal surge bin, gasifier vessel, and ash lockhopper is approximately \$7.7 million (Jan 89) per train based on three estimates for the Mark IV design presented in Corman (p III-3). The cost of the gasifier system from Zahnstecher (1984) is approximately 7.85 million (Jan 89) per train; however, this is for a conventional oxygen-blown system with cold gas cleanup, including a raw gas wash cooler to remove tars and oils which would condense on downstream equipment. Because gas cooling is not required in hot gas cleanup systems, condensation of tars and oils is not expected to pose a problem; therefore no raw gas wash cooling is required. Furthermore, the heating value of the syngas is increased by the presence of tars and oils in the vapor phase, compared to the syngas in a cold gas system in which tars and oils are removed. Therefore, the cost estimate from Corman will be used. The number of operating trains of gasifiers is determined based on the coal throughput per gasifier. The number of total trains is determined based on the number of operating trains and the availability of the gasifier, which is 87 percent (Zahnstecher, 1984). Thus:

$$N_{O,G} = INT\left(\frac{m_{cf,G,i}}{CTP}\right) \quad \text{(A-88)}$$

$$N_{T,G} = INT\left(\frac{N_{O,G}}{AV}\right) \quad \text{(A-89)}$$

where AV = 0.87. The function INT indicates that the number of trains should be rounded to the nearest integer. The direct cost equation for the gasification section will be presented after the discussion of coke, fines, and ash handling and particulate removal using cyclones.

A.5.2.2 Coke, Fines, and Ash Handling

The gasifier train costs do not include coke handling, fines briquetting and recycle, or ash removal. Coke is required as a startup fuel for Lurgi gasifiers. A typical coke handling system would include a receiving hopper for coke delivered by truck, and a means for reclaiming the coke and delivering it to the gasifier during startup. A similar system is required for handling bentonite, which is used as a binder for fines agglomeration. The fines agglomeration section includes fines surge bins, binder storage, mixers, roll briquettors, and conveyors. Briquetted fines are delivered to the gasifier coal surge bin for feed back to the gasifier. No detail on binder requirements are given in the Corman study. Ash removal includes facilities to removal ash from the gasifier ash lockhopper and convey

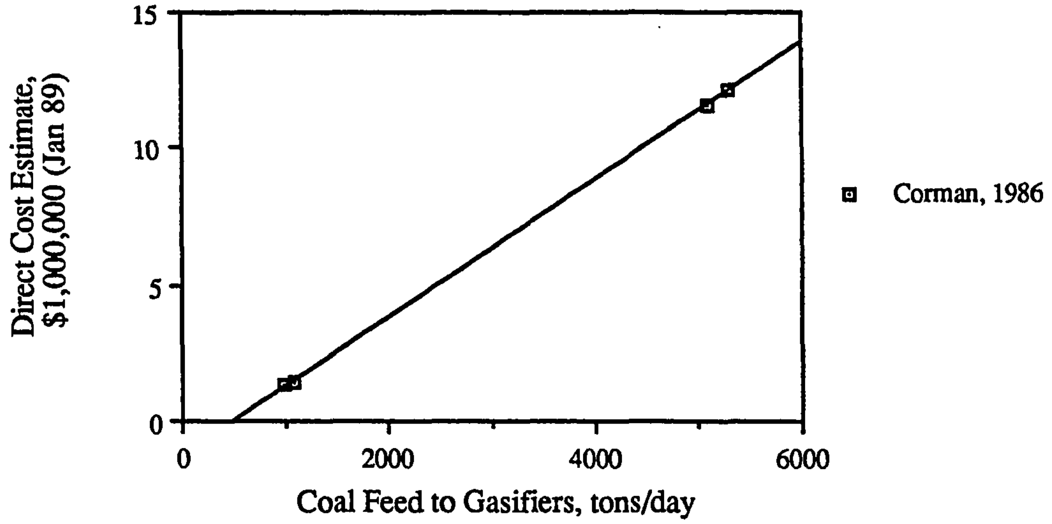

Figure A-24. Direct Cost of Lurgi Gasification Subsystems: Coke Handling, Fines Agglomeration, and Ash Removal.

it for disposal (e.g., to trucks). The costs of these systems are included in the Corman study, but are not detailed. They are estimated here by difference between the total direct cost reported for the gasification section of the plant and the total direct cost of the gasifier vessels and associated equipment. The cost of the coke handling system is expected to be proportional to the size of the gasifier, as measured by the coal feed rate. The cost of the agglomeration system is expected to be proportional to the fines recycle rate (which, for the Corman study, is proportional to the coal flow rate). The cost of the ash removal system is expected to be proportional to the ash content of the coal and the coal feed rate to the gasifier. A simple linear model of the cost of the coke, agglomeration, and ash removal subsystems as a function of plant coal feed rate was developed:

$$DC_{SS} = -1{,}200 + 2.5\, m_{cf,G,i} \qquad (R^2 = 1.00;\ n = 4) \tag{A-90}$$

where,

$$1{,}000 \leq m_{cf,G,i} \leq 5{,}200 \text{ tons/day}$$

The standard error of this estimate is negligible. A graphical representation of the sum of the costs of these subsystems and the coal feed rate to the gasifiers is given in Figure A-24. This model is applicable only to Illinois No. 6 coal.

A.5.2.3 High Temperature Particulate Removal

In the Corman study, particulate removal occurs without gas cooling and in two stages. Primary cyclones, located between the gasifier outlet and the zinc ferrite inlet, are used to remove approximately 98 percent of the particulates in the raw syngas. Secondary cyclones, located between the zinc ferrite absorber outlet and the gas turbine inlet, are used for removing any entrained sorbent material from the zinc ferrite process as well as a significant portion of the remaining particulates not removed in the primary cyclone. The cost estimate is based on single-stage cyclones offered by General Electric, which consist of a stainless steel cyclone enclosed in a refractory lined carbon steel pressure vessel (Corman, 1986). The inner diameter of the cyclones ranges from 50 to 60 inches in the six cases considered in the study.

Each cyclone unit consists of a cyclone, solids lockhopper, solids ball valve, and support structural steel. The cost of each unit is expected to depend on the syngas volume flow rate, the syngas pressure, and to a lesser extent, the solids throughput. A multi-variate regression of the six cost estimates for cyclone systems was developed based on the syngas volume flow rate through each cyclone, and the operating pressure of the gasifier (which was assumed as the design pressure for the cyclones). For a particulate removal system in which both primary and secondary cyclones are used, the volume flow rate of the syngas through each cyclone is equal to the total syngas volume flow rate divided by one-half the number of cyclones. The size of the primary and secondary cyclones is assumed to be the same in this analysis, because the syngas volume flow rate changes only slightly between the desulfurization system inlet and outlet. The direct cost model is:

$$DC_{CY} = 0.98\, N_{T,CY} \left(P_{G,o}\right)^{0.28} \left(\frac{2\, V_{syn,G,o}}{N_{O,CY}}\right)^{0.43} \qquad (R^2 = 0.997;\ n = 6) \qquad \text{(A-91)}$$

where,

$$260 \le P_{G,o} \le 630 \text{ psia, and}$$

$$3{,}900 \le \left(\frac{2\, V_{syn,G,o}}{N_{O,CY}}\right) \le 6{,}000 \text{ acfm.}$$

The number of operating cyclones should be twice the number of operating gasifiers, and the number of total cyclone^r should be twice the number of total gasifiers to ensure that there are two stages of cyclones per gasification train. The regression models should not be extrapolated beyond the ranges shown above. A comparison of the direct cost estimated from the multi-variate regression of Equation (A-91) and the direct costs reported in Corman (1986) are shown graphically in Figure A-25.

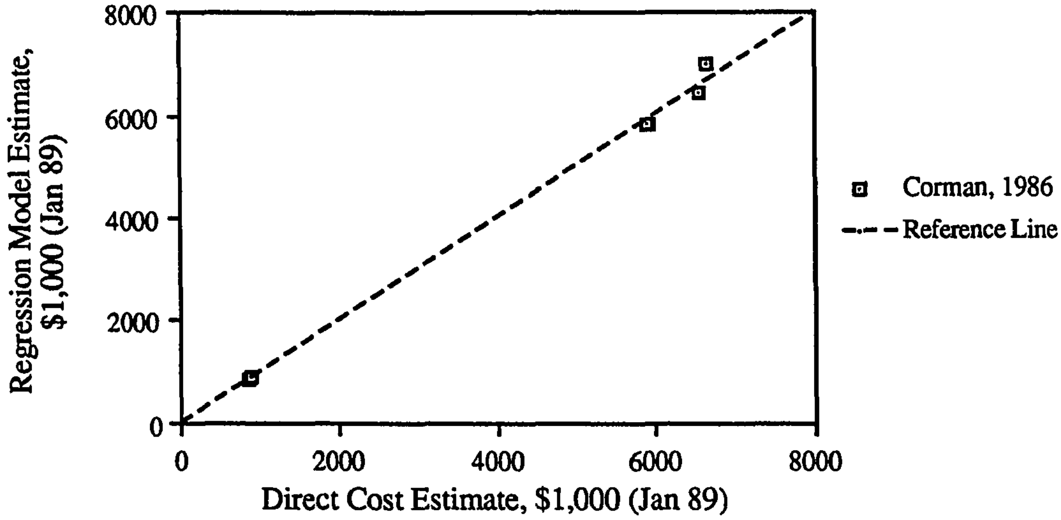

Figure A-25. Predicted vs. Actual Direct Cost for High Temperature Cyclones.

The total direct cost for the Lurgi gasification section is given by the sum of its components:

$$DC_G = 7{,}700\left(\frac{f_{p,i}}{7.14}\right)N_{T,G} + DC_{ss} + DC_{CY} \tag{A-92}$$

The term $f_{p,i}$ is the pressure correction factor presented in Equation (A-82). This factor is assumed to adjust the cost of the gasifiers for different pressure levels. The factor is divided by 7.14, which is the value of the factor at a pressure of 450 psia. This is the design pressure which is the basis for the cost estimate. Because the pressure range for the coal throughput estimate is 300 to 600 psia, the pressure levels for the gasifier should not exceed this range.

A.5.3 Sulfuric Acid Plant

The zinc ferrite desulfurization system generates an off-gas stream containing sulfur dioxide which must be processed to remove or recover the sulfur. The proposed method for recovering the sulfur is a sulfuric acid recovery plant (Corman, 1986; Cincotta, 1984; O'Hara et al, 1987). For a fixed bed zinc ferrite process, the sulfur dioxide concentration in the off-gas may be 1 to 4 percent on a volume basis. For a moving-bed zinc ferrite design, in which off-gas is recycled as a diluent for the exothermic reactions in the regenerator, the concentration of sulfur dioxide is higher, on the order of 10 percent. The ASPEN simulation is based on a moving bed zinc ferrite process.

A detailed performance and cost model of sulfuric acid plant was developed by Frey (1987). The model is sensitive to a number of factors affecting the cost of a sulfuric acid plant, including the volume flow rate of the off-gas stream, the amount of combustible gases in the off-gas, the amount of sulfur dioxide in the off-gas, and the inlet temperature of the off-gas. The sulfuric acid plant model is based on a detailed design and cost estimate developed for DOE by Monsanto under subcontract to Science Management Corporation (SMC, 1983). The plant is an interpass absorption contact design with off-gas conditioning.

The detailed sulfuric acid plant model was run for 100 cases with variation in the off-gas molar flow rate, sulfur dioxide concentration, moisture concentration, and off-gas temperature. A variant of Monte Carlo analysis employing Latin hypercube sampling was used to select combinations of parameters for the 100 cases. From the resulting cost estimates, regression analysis was used to develop a simple expression for the cost of the sulfuric acid plant as a function of key performance parameters. The cost of the sulfuric acid plant is most sensitive to the total flow rate of the gas streams throughout the system. The total off-gas molar flow rate, concentration of sulfur dioxide, and temperature were identified as parameters with significant influence on cost. The simplified cost model is:

$$DC_{SA} = 52 + 18.2\, M_{SA,i}^{0.67}\, f_{SO_2}^{0.12}\, T_{SA,i}^{0.12} \qquad (R^2{=}0.956;\ n = 100) \qquad \text{(A-93)}$$

where,

$$1{,}480 \leq M_{SA,i} \leq 70{,}000 \text{ lbmole/hr;}$$

$$0.01 \leq f_{SO_2} \leq 0.13\text{; and}$$

$$500 \leq T_{SA,i} \leq 1{,}500\ ^{\circ}\text{F.}$$

The model should not be extrapolated beyond the ranges of the predictive variables shown above. The standard error of the estimate is $1.4 million. The plant does not include provision for off-gas particulate removal.

A.6 Total Capital Cost Model

A framework for estimating the total capital requirement (TCR) for an IGCC power plant is presented in this section. The method is based on the EPRI Technical Assessment Guide (TAG) (1986), and is presented in the following section. A review of process contingency factors commonly assumed in the gasification systems cost studies is presented in Section A.6.2. Section A.6.3 discusses how uncertainty in the capital costs can be represented more rigorously using probabilistic methods.

A.6.1 Estimating the Total Capital Requirement

Table A-6 lists the items included in the total capital requirement (TCR) for an IGCC system. The total direct cost (TDC) is the summation of the plant section direct costs (presented in previous chapters) and the general facilities cost. This is given by:

$$\mathrm{TDC} = \left(\sum_{i=1}^{n} \mathrm{DC}_i + \mathrm{GF} \right) \left(\frac{I_{PCI}}{351.5} \right) \tag{A-94}$$

The total direct cost is referenced to January 1989 dollars, using the *Chemical Engineering* Plant Cost Index. The cost can be adjusted for other years using the appropriate value of the cost index in Equation (A-94).

Indirect construction costs include workers benefits, supervision and administrative labor, purchased and rented construction equipment, and construction facilities, which may include temporary buildings, roads, utilities, railroad, and minimal recreation facilities for the workers. From an analysis of the cost estimate for a KRW-IGCC plant located in Chicago, the indirect construction costs are approximately 25 percent of the total direct cost (Fluor, 1985). For the Lurgi-based system, the only indirect costs reported are for field supervision labor (Corman, 1986). Therefore, for this analysis it is assumed that the indirect construction cost is given by:

$$C_{ICC} = f_{ICC}\,\mathrm{TDC} \tag{A-95}$$

where, as a nominal (default) value,

$$f_{ICC} = 0.25.$$

The cost of sales tax is specific to the state where the power plant is to be constructed. A common assumption is the Illinois sales tax of six percent, applicable to all material costs. Material costs comprise typically 80 percent of the total direct cost and 10

Table A-6. Items Included in the Total Capital Requirement

Total Plant Cost (TPC)
- Summation of Direct Costs for All Process Areas
- General Facilities
- Sales Tax
- Environmental Permits
- Indirect Construction Costs
- Engineering and Home Office Overhead and Fees
- Process Contingency
- Project Contingency

Total Plant Investment (TPI)
- TPC
- Allowance for Funds Used During Construction (AFUDC)

Total Capital Requirement (TCR)
- TPI
- Prepaid Royalties
- Spare Parts Inventory
- Preproduction (or startup) Costs
- Inventory Capital (fuel storage, etc.)
- Initial Chemicals and Catalyst Charges
- Land

Basis: EPRI (1986)

percent of the indirect costs, based on data in Fluor (1985). Therefore, the sales tax can be estimated as:

$$C_{tax} = r_{tax}\left(0.8\ TDC + 0.1\ C_{ICC}\right) \tag{A-96}$$

where,

$$r_{tax} = 0.06\text{, as a default value.}$$

The engineering and home office costs include the costs associated with: (1) engineering, design, and procurement labor; (2) office expenses; (3) licensor costs for basic process engineering; (4) office burdens, benefits, and overhead costs; (5) fees or profit to the architect/engineer. EPRI recommends that a value of 7 to 15 percent of the total direct cost, indirect construction cost, and sales tax be used. For this analysis, a value of 15 percent is assumed. The engineering and home office cost is thus given by:

$$C_{EHO} = f_{EHO}\left(TDC + C_{ICC} + C_{tax}\right) \tag{A-97}$$

where,

$f_{EHO} = 0.15$ as a default value.

An item not commonly quantified in capital cost estimates is the cost of obtaining environmental permits required for the power plant. An allowance for permitting costs is included here, with a default value of one million dollars. This represents the costs associated with obtaining the services of a consultant who provides the various services associated with permits, including estimates of emissions and discharges of gaseous, liquid, and solid wastes; dispersion modeling of air emissions; and preparation of permit applications. The permitting cost assumed here is a rough, order-of-magnitude estimate only. The cost of environmental permits is thus given by:

$$C_{EP} = 1{,}000 \tag{A-98}$$

The total indirect cost (TIC) is the sum of the indirect construction costs, engineering and home office costs, sales tax, and environmental permitting costs:

$$TIC = C_{ICC} + C_{tax} + C_{EHO} + C_{EP} \tag{A-99}$$

A major cost item for advanced technology plants is the process contingency. As discussed in Chapter 2, the process contingency is used in deterministic cost estimates to quantify the expected increase in the capital cost of an advanced technology due to uncertainty in performance and cost for the specific design application. In the EPRI cost method, the process contingency is estimated based on separate consideration of contingencies for each process section. The contingency is expressed as a multiplier of the sum of the direct and indirect capital costs for each plant section. Recommended ranges of process contingency factors are shown in Table A-7. The process contingency decreases as the commercial experience with a process area increases. For example, in a fully commercialized process, which has been used in similar applications, the process contingency may be zero. For a new concept early in the development stage, the process contingency may be over 40 percent of the process area cost. Experience has shown that cost estimates for innovative technologies early in the development phase tend to be low by a factor of two or more compared to the cost of the first commercial-size demonatration plant (EPRI, 1986; Merrow, Phillips, and Myers, 1981). However, the cost for subsequent plants tends to decrease, which is known as the "learning curve" effect. Process contingencies employed for innovative technologies are intended to represent the expected costs of a commercialized (e.g., fifth of a kind) plant (EPRI, 1986). The process contingency for each major plant section is estimated as follows:

Table A-7. Process Contingency Factors Recommended by EPRI[a]

State of Technology Development	Percentage of Process Area Cost
New concept with limited data	≥ 40
Concept with bench scale data available	30 to 70
Small pilot plant data (e.g., 1 MW) available	20 to 35
A full-size module has been operated (e.g., 20-100 MW)	5 to 20
The process is used commercially	0 to 10

[a] Cost estimates using these contingency factors are intended to represent the cost of commercialized (e.g., fifth of a kind) process plants. Source: EPRI (1986)

$$C_{PC,i} = f_{PC,i}\left[DC_i + TIC\left(\frac{DC_i}{TDC}\right)\right] \quad \text{(A-100)}$$

where,

$f_{PC,i}$ = process contingency for plant section i.

Equation (A-100) includes a term which prorates the total indirect costs to each plant section based on the ratio of the plant section direct cost to the plant total direct cost. This approach was used in Fluor (1985). The total process contingency allowance for the plant is given by the sum of process contingencies for each plant section:

$$C_{PC} = \sum_{i=1}^{n} C_{PC,i} \quad \text{(A-101)}$$

The process contingencies for each plant section are discussed in Section A.6.2.

In contrast to the process contingency, the *project* contingency is used in deterministic cost estimates to represent the expected increase in the capital cost estimate that would result from a more detailed estimate for a specific project at a particular site. EPRI defines four levels of cost estimates, based on the type of information used to develop the estimate. These are listed, with brief explanatory notes, in Table A-8. The type of estimates developed in this work are best classified as "preliminary." The estimates are based on the costs of major equipment, and are taken from studies which present process diagrams for major plant sections typically including 10 or 20 equipment items per section. In the EPRI study of KRW IGCC systems (Fluor, 1985), a project contingency

Table A-8. Project Contingency Factors Recommended by EPRI[a]

Type of Estimate	Design Information	Percentage of Direct Cost
Simplified	General site, process flow diagram	30 to 50
Preliminary	Major equipment, preliminary piping and instrumentation diagrams	15 to 30
Detailed	Complete process design, site-specific, engineering design in progress, construction contract and schedule.	10 to 20
Finalized	Complete engineering of process plant	5 to 10

[a] Expressed as a percentage of the total of total direct, total indirect, and process contingency.
Source: EPRI (1986)

of 15 percent was assumed, which is at the low end of the recommended range for preliminary estimates. Based on the overlap between recommended project contingencies for preliminary and "detailed" cost estimates, it is assumed here that the project contingency has a typical value of 20 percent. This value is also consistent with the range of recommended values for preliminary estimates. Furthermore, an upward adjustment in the contingency factor is consistent with the findings of a number of studies by the Rand Corporation, which indicate that contingency factors typically are under-estimated for advanced technology process plants (e.g., Milanese, 1987). Reasons for under-estimates often include: a lack of a organized record comparing the cost estimates of past projects with actual costs (which would provide a basis for correcting future cost estimates); the mis-application of a standard contingency factor (e.g., 10 percent) regardless of the type of estimate; and deliberate under-estimation of costs to ensure further consideration of the project by management. The project contingency is given by:

$$C_{PJ} = f_{PJ}\left(TDC + TIC + C_{PC}\right) \qquad \text{(A-102)}$$

The total plant cost (TPC) is the sum of the total direct cost, total indirect cost, process contingency, and project contingency:

$$TPC = TDC + TIC + C_{PC} + C_{PJ} \qquad \text{(A-103)}$$

The total plant cost is an "instantaneous" estimate; i.e., it is estimated as if the entire plant were constructed at a single instant, thereby disregarding the time value of money and the time required for construction.

The total plant investment, as shown in Table A-6, includes the total plant cost plus an allowance for funds used during construction (AFUDC), also referred to as "interest during construction." If the expenditure for the total plant cost is spread uniformly over the construction period, measured in years, then the total plant investment is given by the following:

$$TPI = AF \times TPC \tag{A-104}$$

where,

$$AF = \frac{Z^N - 1}{N\,(Z - 1)} \tag{A-105}$$

$$Z = \frac{(1+i)}{(1+e_a)} \tag{A-106}$$

and,

i = interest cost for spent funds,

e_a = annual escalation rate for plant equipment (e.g., inflation rate)

The total process capital cost should be based on the date at which construction begins. The total plant investment is expressed in the same year dollars as the total process capital.

As shown in Table A-6, the total capital requirement includes the total plant investment plus several other items. Prepaid royalties are fees paid to the owners of proprietary process technology designs, and are typically estimated as a fraction of the total plant investment if specific data are not available. The spare parts inventory is also estimated as a fraction of the total plant investment. Preproduction costs include one month of fixed operating costs, one month of variable operating costs (excluding fuel and byproduct credit) based on full plant capacity, one-quarter of the full capacity fuel cost for one month, and two percent of the total plant investment. The operating costs and fuel costs are developed in Section A.7. The total preproduction cost is given by the following:

$$PPC = PP_{FC} + PP_{OC} + PP_{Fuel} + 0.02\,TPI \tag{A-107}$$

The preproduction costs for fixed operating costs, variable operating costs, and fuel are presented in Section A.7.3 in Equations (A–169), (A-170), and (A-171), respectively.

Inventory capital includes the costs of fuels and other consumables which are inventoried prior to plant startup. For a baseload power plant, this includes 60 days of fuel and consumable inventories based on 100 percent of plant capacity. Examples of consumables include make-up catalyst and chemicals (e.g., for a Claus sulfur plant or Selexol process) and water treatment chemicals. The fuel and consumable requirements are developed in Section 6.0 for each of the three IGCC power plant systems. The inventory capital is given by the following:

$$IC = \sum_{i=1}^{n} IC_i \tag{A-108}$$

The inventory capital includes costs for coal, boiler feed water dimineralizer and treatment chemicals, water polishing chemicals, cooling water treatment chemicals, plant and instrument air adsorbent, liquified petroleum gas for a flare, and fuel oil. For a KRW-based system with cold gas cleanup, the inventory capital also includes Selexol solvent, Claus plant catalyst, and either SCOT or Beavon-Stretford catalyst and chemicals. For a KRW-based system with hot gas cleanup, the inventory capital also includes limestone and zinc ferrite sorbent. For a Lurgi-based system with hot gas cleanup, the inventory capital also includes limestone, coke, bentonite, and zinc ferrite sorbent. The annual operating requirements for these processes are discussed in Section A.7. A detailed expression for the inventory capital is developed and presented in Section A.7.3, Equation (A-108).

The initial catalyst and chemicals charge is distinct from the inventory capital, and includes the cost of catalyst or chemicals that are contained in process equipment. Examples of initial catalyst or chemicals are the required catalyst inventory for a Claus plant or zinc ferrite process or the initial charge of solvent for the Selexol process.

The initial catalysts and chemicals required for a KRW-based system with cold gas cleanup include fuel oil for gas turbine startup, Selexol solvent, Claus plant catalyst, and SCOT catalyst and chemicals. If a Beavon-Stretford unit is used in place of a SCOT unit, then Beavon catalyst and the Stretford chemical are required. Data from a number of cost studies have been analyzed to determine the initial requirement for catalyst and chemicals as a function of key performance parameters.

The initial requirement for Selexol solvent is expected to depend primarily on the mass flow of hydrogen sulfide, the primary sulfur species in raw syngas, and on the

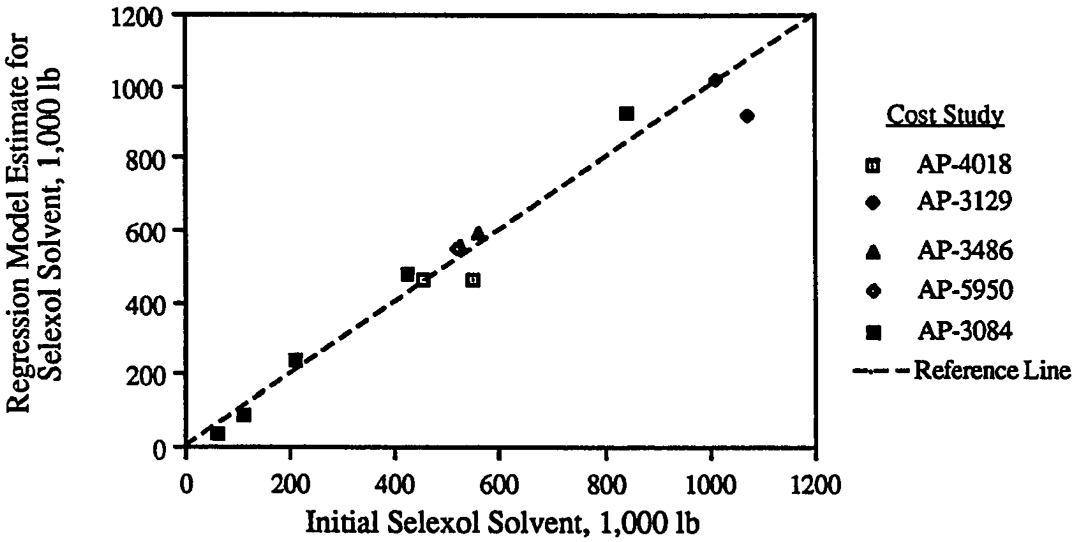

Figure A-26. Initial Solvent Requirement for the Selexol Process.

concentration of the hydrogen sulfide. A multivariate regression yielded the following result for the initial solvent requirement, expressed in pounds:

$$CHEM_{i,S} = -25{,}200 + 16.6\left(\frac{M_{HS,S,i}^{0.935}}{f_{HS}^{1.04}}\right) \quad (R^2 = 0.959;\ n = 12) \qquad \text{(A-109)}$$

where,

$$50 \le M_{HS,S,i} \le 900 \text{ lbmole/hr},$$

$$0.004 \le f_{HS} \le 0.012.$$

The standard error of the estimate is 68,000 lb of solvent. The solvent requirement estimated from the regression model is compared to the reported solvent requirement in Figure A-26.

The initial catalyst requirement for two-stage Claus plants was found to depend on the recovered sulfur mass flow rate. The initial catalyst requirement, in tons, is given by:

$$CAT_{i,C} = 5.03\times10^{-3}\ m_{s,C,o} \qquad (R^2 = 0.959;\ n = 12) \qquad \text{(A-110)}$$

where,

$$1{,}000 \le m_{s,C,o} \le 30{,}800 \text{ lb/hr}.$$

The standard error of the estimate is 4.1 tons. The regression model is shown graphically in Figure A-27.

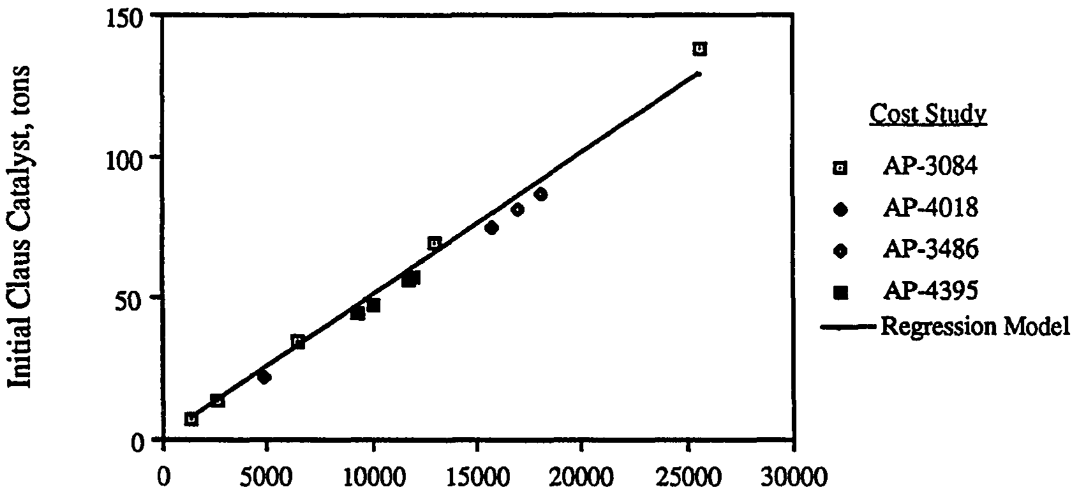

Figure A-27. Initial Catalyst Requirement for Two-Stage Claus Plant.

Because the SCOT process is proprietary, no detailed data are reported regarding the performance of this process. The recovered sulfur mass flow rate is the only performance parameter that is widely reported in the cost studies that is relevant to the SCOT process. Therefore, single-variate regression analyses based on this parameter were used to estimate both the initial SCOT catalyst and SCOT chemical requirements. Data from three of the studies (Fluor, 1984; Fluor, 1985; Fluor, 1986) satisfied a simple straight-line approximation for catalyst or chemical requirement as a function of the recovered sulfur flow rate. Data from an earlier study (Fluor, 1983b) appear inconsistent with the three more recent studies, and were not included in the regression analysis. The resulting equations for the initial SCOT catalyst and chemical are:

$$CAT_{i,SC} = 19.3 + 0.161\ m_{s,C,o} \qquad (R^2 = 1.00;\ n = 7) \qquad \text{(A-111)}$$

$$CHEM_{i,SC} = 1{,}270 + 10.7\ m_{s,C,o} \qquad (R^2 = 1.00;\ n = 7) \qquad \text{(A-112)}$$

where,

$$5{,}000 \le m_{s,C,o} \le 18{,}000$$

The catalyst requirement is in units of cubic feet, and the initial chemical requirement is in units of pounds. The standard errors of the estimates are negligibly small The regression equations are shown graphically in Figures A-28 and A-29 for the initial catalyst and chemical requirement, respectively. The regression equations reproduce almost exactly the reported requirements for most of the data shown, but over-estimate the requirements for the three data points not included in the regression model.

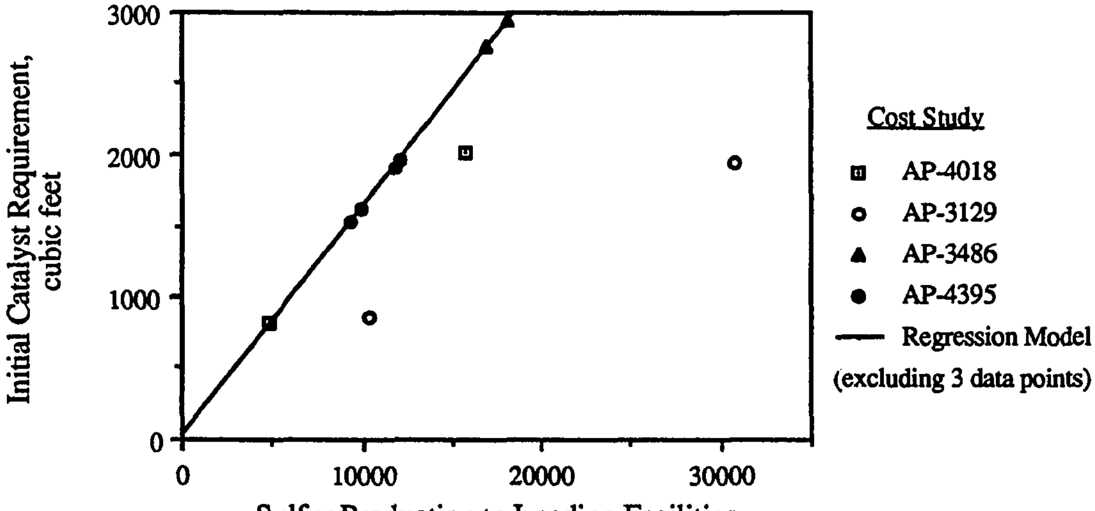

Figure A-28. Initial Catalyst Requirement for the SCOT Process.

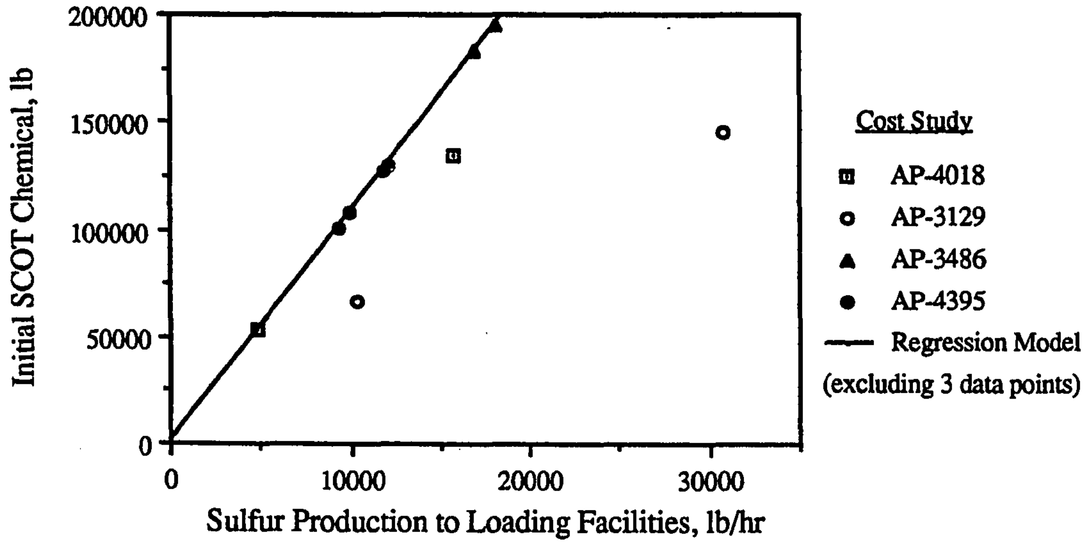

Figure A-29. Initial Chemical Requirement for the SCOT Process.

As an alternative to the SCOT process, some cost studies specify the Beavon-Stretford process for Claus plant tail gas treatment. The Beavon-Stretford process requires a catalyst for the Beavon unit and a special chemical for the Stretford unit. The initial catalyst and chemical requirements for the Beavon-Stretford process were estimated from the values reported in Fluor (1983a), which includes data for a range of plant sizes. From these data, a simple linear relationship of catalyst and chemical requirements as a function of the sulfur recovered in the Beavon-Stretford unit was identified. In the case of the

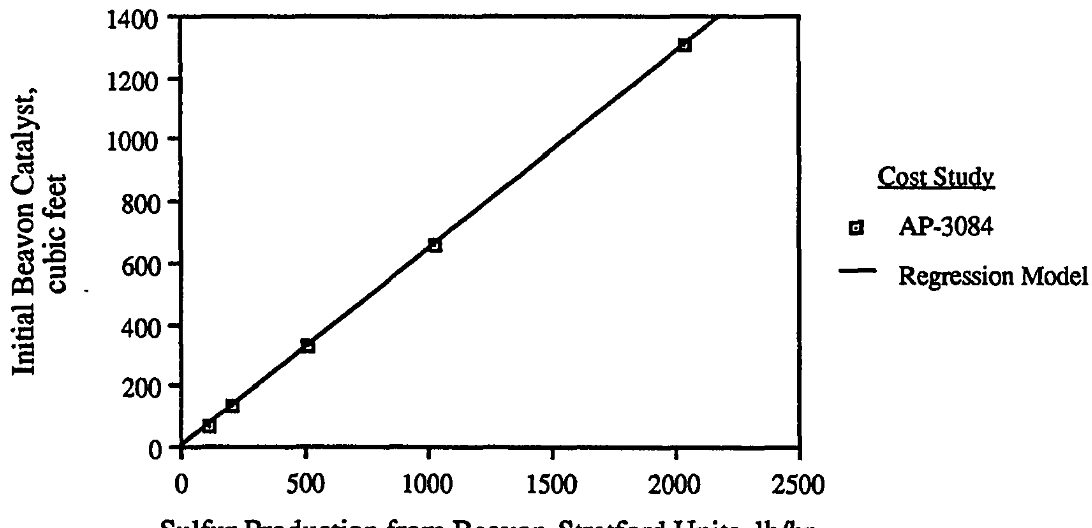

Figure A-30. Initial Catalyst Requirement for the Beavon-Stretford Process.

Beavon catalyst, the mass requirement as a function of sulfur flow rate can be estimated. In the case of the Stretford chemicals, the mass requirement is not given. However, the cost of the initial Stretford chemicals as a function of the recovered sulfur flow rate was developed. The resulting regression models for the initial catalyst requirement, in cubic feet, and the initial chemical requirement, in dollars, are:

$$CAT_{i,BS} = -1.3 + 0.641\ m_{s,BS,o} \qquad (R^2 = 1.00;\ n = 5) \qquad \text{(A-113)}$$

$$C_{i,BS,Chem} = 85.8\ m_{s,BS,o} \qquad (R^2 = 1.00;\ n = 5) \qquad \text{(A-114)}$$

where,

$$100 \le m_{s,BS,o} \le 2{,}100 \text{ lb/hr.}$$

The standard error of the estimate for these models is negligible. The regression models for the initial catalyst and initial chemical requirements are shown graphically in Figures A-30 and A-31, respectively.

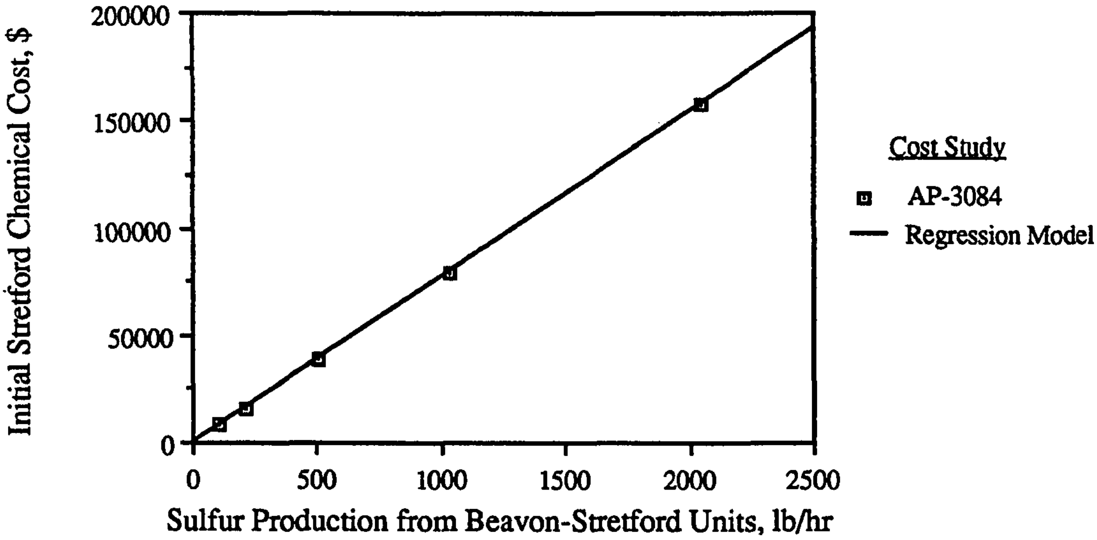

Figure A-31. Initial Stretford Chemical Cost for the Beavon-Stretford Process.

Fuel oil is required as a startup and auxiliary fuel for the gas turbines. For a nominal 550 MW power plant, 80,000 barrels of fuel oil are required initially. As an approximation, this amount may be scaled with the size of the plant. The estimated initial fuel oil requirement in barrels is therefore given by:

$$FO_i = 80{,}000\left(\frac{MW}{550}\right) \quad \text{(A-115)}$$

In addition to the initial catalysts and chemicals summarized above, chemicals are also required for treatment of the makeup cooling water. These chemicals include sulfuric acid (93 percent purity), corrosion inhibitor, and surfactant. However, the initial inventory of these chemicals is small, and in most cost estimates the cost of these is approximately 1,000 dollars (e.g., Fluor, 1985). Therefore, an allowance of 1,000 dollars for the initial charge of these chemicals is given by:

$$C_{CH,i,CW} = 1 \quad \text{(A-116)}$$

The total cost of initial catalysts and chemicals for a KRW-based system with cold gas cleanup is then given by:

$$C_{IC\&C} = UC_{FO}FO_i + C_{CH,i,CW} + \sum_j UC_j CHEM_{i,j} + \sum_k UC_k CAT_{i,k} \quad \text{(A-117)}$$

The subscript j refers to the name of the chemical and the subscript k refers to the name of the catalyst. The unit costs of chemicals and catalysts are discussed in Section A.7.

For a KRW-based system with hot gas cleanup, the initial catalyst and chemical requirement includes zinc ferrite sorbent, fuel oil, and cooling water treatment chemicals. The required initial zinc ferrite sorbent charge for the on-line desulfurization units is discussed in Section A.4.4 and is presented in Equation (A-76). The total amount of sorbent required is given by:

$$CHEM_{i,ZF} = S_c \left(\frac{N_{T,ZF}}{N_{O,ZF}} \right) \qquad \text{(A-118)}$$

The fuel oil and cooling water treatment chemical requirements are estimated in the same manner as for the KRW-based system with hot gas cleanup.

For a Lurgi-based system with hot gas cleanup, coke and bentonite are required in addition to zinc ferrite sorbent, fuel oil, and cooling water treatment chemicals. Coke is required as a gasifier startup fuel. For a nominal 180 MW power plant, the initial coke requirement is 200 tons, based on Bechtel et al (1988). Therefore, the initial coke requirement may be estimated by:

$$CHEM_{i,coke} = 200 \left(\frac{MW}{180} \right) \qquad \text{(A-119)}$$

Bentonite is used as a binder for fines agglomeration, based on the design specified in Corman (1986). However, the amount of binder required is not given. From Parsons (1985), the ratio of bentonite to coal fines is estimated to be 0.035 on a weight basis. The initial charge is equivalent to one week of consumption at full load. Therefore, the initial charge of bentonite in pounds is estimated by:

$$CHEM_{i,Bent} = 5.88\ m_{fines} \qquad \text{(A-120)}$$

The requirements for the other initial chemicals and catalysts are estimated using the equations developed for the KRW-based systems. The total cost of initial catalysts and chemicals are then estimated by multiplying the total requirements by the appropriate unit costs and summing, as in Equation (A-95).

The cost of land is estimated based on the total land required for an IGCC power plant and the cost per acre of land. Plant site requirements have been estimated by Bechtel Group (1988) based on analysis of data in 25 studies of gasification systems. The data are presented graphically in the Bechtel Group study. With the exception of two outlier points for phased systems, regression analysis was used to develop an analytic expression for the

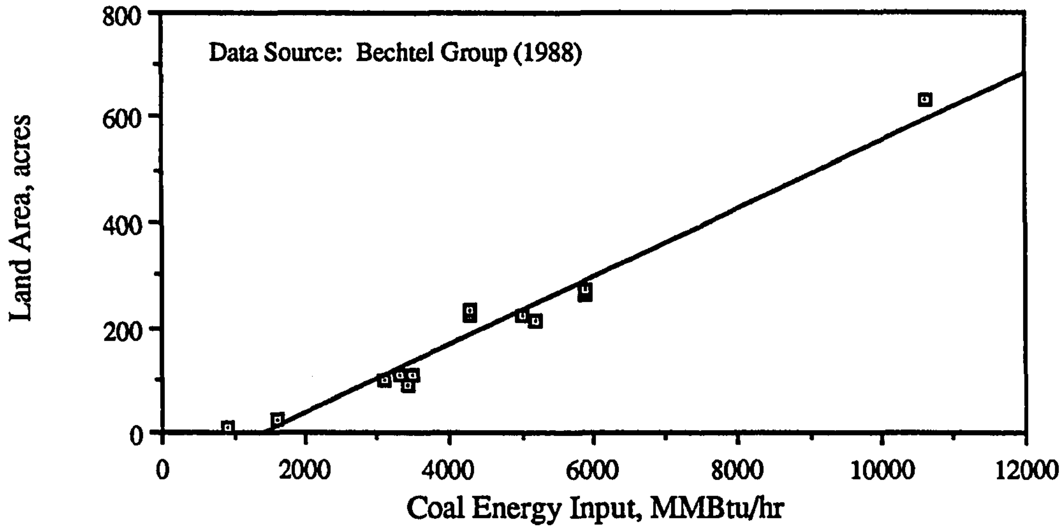

Figure A-32. Land Requirements for IGCC Plants.

land requirement as a function of the plant coal energy input. The land requirement can be estimated from the following correlation:

$$A_L = -93 + 0.065\, Q_{coal} \qquad (R^2 = 0.962;\ n = 13) \qquad \text{(A-121)}$$

where,

$$900 \leq Q_{coal} \leq 11{,}000 \text{ MMBtu/hr}$$

The standard error of the estimate is 32 acres. The data and the regression equation are shown graphically in Figure A-32.

Finally then, the total capital requirement (TCR) for an IGCC system is given by:

$$TCR = 1.01\ TPI + PPC + IC + C_{IC\&C} + A_L UC_L \qquad \text{(A-122)}$$

The first term includes an allowance of 0.5 percent of the total process investment for prepaid royalties and 0.5 percent for spare parts inventory. The unit cost of land is assumed to be $6,500 per acre, based on EPRI (1986).

It is common to express the total capital requirement on a normalized ($/kW) basis. For IGCC plants, it is important to specify the ambient temperature for which the normalized cost is reported, because the gas turbine power output is a function oi ambient temperature.

Table A-9. Reported Process Contingency Factors for Selected IGCC Systems[a]

Area Description	Area No.	KRW, Cold Gas Cleanup	KRW, Hot Gas Cleanup	Lurgi, Hot Gas Cleanup
Oxidant Feed	10	0	<5	<5
Coal Handling	20	0 - 12.5	12.5	0
Limestone Handling	25	--	0	--
Gasification	30	20 - 37.5	37.5	5
Sulfation	35	--	35 - 60	--
Low Temp. Gas Cooling	40	0	--	--
Selexol	50	0 - 12.5	--	--
Zinc Ferrite	50	--	40	5.8
Claus Plant	60	0 - 8.75	--	--
Sulfuric Acid Plant	60	--	--	5.9
SCOT	70	0 - 8.75	--	--
Beavon-Stretford	70	10	--	--
Boiler Feed Water System	80	0	0	0
Process Condensate Treatment	85	10 - 50	--	--
Combined Cycle System	90	2.7	2.7	2.7

References: Corman, 1986; Earley and Smelser, 1988a; Earley and Smelser, 1988b; Fluor, 1985; Smelser, 1986; Smith, Hanny, and Smelser, 1986; Smith and Smelser, 1987

[a] These contingency factors are intended to be representative of commercial, fifth-of-a-kind plants.

A.6.2 Process Contingencies

Reported process contingencies for the major plant sections of the three IGCC systems considered in this study are summarized in Table A-9. These contingency factors are based on a review of several studies (Corman, 1986; Earley and Smelser, 1988a; Earley and Smelser, 1988b; Fluor, 1985; Smelser, 1986; Smith, Hanny, and Smelser, 1986; Smith and Smelser, 1987). For many commercially available systems, the process contingencies are zero. Contingency factors tend to be largest for those process areas which have not been built on a commercial scale, such as the gasification, sulfation, and zinc ferrite sections. As noted earlier, the process contingency is to represent the expected cost increases for a mature, "fifth-of-a-kind" plant. However, there is considerable variability in the contingencies used in different studies. The studies prepared for GRI tend to have the largest contingencies, while the studies of Lurgi-based systems with hot gas cleanup prepared by General Electric had the lowest overall process contingency factors of any of the studies examined.

The contingency factors developed in the GRI studies are based on a disaggregated consideration of several factors that pose a risk of increased process area costs. These factors are summarized in Table A-10, which is taken from GRI (1983). The process development allowance, which is equivalent to the process contingency used in EPRI studies, is based on an average (expert judgment) "score" for four attributes of a technology. These attributes are state of development, availability of experimental data, untested assumptions used in the design, and operability and control difficulty. Recommended scores for each of these attributes are indicated in Table A-10. Contingency factors estimated in this manner (i.e., based on a disaggregated consideration of several factors) are likely to be more accurate than contingency factors estimated as a single multiplier, as in the EPRI studies. The GRI estimates also tend to be significantly higher than the EPRI estimates. Although Lurgi gasifiers are commercially available, the contingency factors reported by Corman for the Lurgi-based system appear to be unrealistically low, particularly for the hot gas cleanup system, resulting in a downward bias of the capital cost estimates.

Because contingency factors for new processes tend to be significantly underestimated (Milanese, 1987), it is reasonable to use the largest contingency factors reported in the literature in first-pass cost analyses. However, the traditional approach to estimating contingency factors is inadequate for properly quantifying the risk of cost increases in new process technology. An alternative approach, using probabilistic analysis, is described below.

Table A-10. Process Development Allowances Recommended by GRI[a]

State of Development	Experimental Data	Assumptions	Operability and Control	Value (Percent of Direct Cost)
Conceptual	Phenomena identified	Critical assumptions for key processes or equipment	Not defined and not demonstrated	90 to 100
Laboratory	Laboratory only	Critical assumptions for supporting equipment or processes	Poorly defined and not demonstrated	70 to 90
PDU	PDU Level	Assumptions for supporting equipment or processes not demonstrated (and provided by others)	Preliminary definition, demonstration in development stage	50 to 70
Pilot plant	Pilot Plant Level	Assumptions for equipment or processes demonstrated in significantly different applications	Fair definition, demonstration may not be at design conditions	30 to 50
Demonstration plant	Demonstration plant level	Assumptions for equipment or processes demonstrated in similar but not identical applications	Good definition demonstrated in similar but not identical process	10 to 30
Commercial plant	Complete, demonstrated, commercial plant level	No assumptions: commercially demonstrated for the design case	Well defined, and commercially demonstrated for the design case	0 to 10

[a] The process development allowance is a weighted sum of the values assigned for state of development, experimental data, assumptions, and operability and control. The PDA may be estimated based on a simple average of these values. The PDA is intended to accountfor expected cost increases for a commercial (mature) plant.
SOURCE: GRI (1983)

A.6.3 Representing Uncertainty in Capital Cost Estimates

The concept of a contingency factor is based on adjusting the capital cost estimate such that the probability of a cost over-run is below some acceptable threshold value (e.g., 50 percent). A low contingency factor for a new technology therefore implies a high probability of cost overrun. Therefore, contingency factors should be developed based on a probabilistic analysis of the capital cost of a process plant. A probabilistic analysis facilitates the disaggregated and quantitative consideration of uncertainties for specific parameters in an engineering model. Engineers are better able to make judgments about specific parameters than about an aggregate contingency factor. In fact, the GRI approach to estimating contingency costs involves disaggregation of the contingency factor into specific components, about which judgments are more easily made.

In a probabilistic modeling approach, any parameter in an engineering model may be treated as uncertain and quantified using probability distributions instead of point estimates. Uncertainties are particularly important for advanced concepts in an early phase of development, for which significant scale-up (or possibly redesign) is required for commercial application.

An important application of the cost models developed here is the development of probabilistic estimates of IGCC system cost based on a disaggregated consideration of uncertainties in the performance and cost parameters of the engineering models. Although some of these uncertainties are indicated in the open literature, in many cases probability distributions for key performance and cost parameters must be estimated in consultation with process developers or other experts. The development of probabilistic performance and cost estimates for IGCC systems is addressed in other parts of the work described here.

A.7 OPERATING COST MODEL

IGCC plant operating costs are also estimated using the method presented in the EPRI Technical Assessment Guide (EPRI, 1986). The items included in the operating costs are listed in Table A-11. These can be divided into fixed and variable costs. Fixed costs are independent of the plant capacity factor or plant load. Variable costs, which include fuel, consumables, ash disposal, and byproduct credit, are directly proportional to the amount of energy produced by the plant. Fixed operating costs are discussed in Section A.7.1. Variable operating costs are discussed in Section A.7.2. Preproduction costs, which are required for the capital cost estimate, are summarized in Section A.7.3. Because the ASPEN simulations include little or no detail on the power consumption of several plant sections, regression models for power consumption of selected plant sections are given in Section A.7.4. In Section A.7.5, a discussion of how uncertainty may be incorporated into the operating cost model is presented.

A.7.1 Fixed Operating Costs

Fixed operating costs include operating labor, maintenance labor and materials, and overhead costs associated with administrative and support labor. The operating labor cost is based on an estimate of the number of personnel hours required to operate the plant multiplied by an average labor rate. It is common to assume that four shifts per day are required for plant operation, allowing two hours overlap for transition between shifts. Furthermore, an allowance for personnel on sick leave or vacation can be incorporated into the "shift factor." A shift factor of 4.75 is assumed as a default in this study, based on Bechtel (1988).

The number of operators required per shift for major plant sections are indicated for the three IGCC systems in Table A-12. These estimates are based on data reported in several studies (Bechtel, 1983b; Bechtel, 1988; Bechtel et al, 1988; Corman, 1986; and Fluor, 1985). In some cases, the numbers are taken directly from the studies, while in other cases the numbers represent a judgment based on a review of several data sources. For the gasification and combined cycle areas, an assumption was made that the number of operating personnel are directly proportional to the number of operating trains for the process area.

Table A-11. Items Included in the Operating Cost Model

Fixed Operating Costs
- Operating Labor
- Maintenance Labor
- Maintenance Materials
- Administrative and Support Labor

Variable Operating Costs
- Consumables
  - Ex: Raw water, chemicals, catalyst, limestone
- Ash Disposal
- Fuel
- Byproduct Credit

Basis: EPRI, 1986.

Table A-12. Operating Labor Requirements for Selected IGCC Systems[a]

		Number of Operators per Shift		
Description	Area No.	KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Oxidant Feed	10	5	--	--
Coal Handling	20	5	4	3
Limestone Handling	25	--	1	--
Gasification	30	2 $N_{O,G}$	2 $N_{O,G}$	2 $N_{O,G}$
Sulfation	35	--	1	--
Gas Cooling	40	--	--	--
Selexol	50	2	--	--
Zinc Ferrite	50	--	5	5
Claus Plant	60	2	--	--
Sulfuric Acid Plant	60	--	--	2
Tail Gas	70	1	--	--
Boiler Feed Water	80	1	1	1
Process Cond. Treat.	85	1	--	--
Combined Cycle	90	4 $N_{O,GT}$ + 1	4 $N_{O,GT}$ + 1	4 $N_{O,GT}$ + 1
General Facilities	100	3	3	3

[a] Based on a nominal 500 MW plant.
Sources: Bechtel (1983b); Bechtel (1988); Bechtel et al (1988); Corman (1986); Fluor (1985).

The total operating labor cost is estimated by summing the number of plant operators per shift for all process areas, applying the shift factor, and applying the average labor rate as follows:

$$OC_L = ALR\left(\frac{2{,}080 \text{ hours}}{\text{year}}\right) SF \sum_i O_i \tag{A-123}$$

where,

ALR = Average labor rate (default is \$19.70/hr based on EPRI, 1986),
SF = Shift factor (default is 4.75, based on Bechtel, 1988), and
O_i = Number of operators in process area i, from Table A-12.

The cost for maintenance material and labor for new technologies is typically estimated as a percentage of the installed capital cost for each process section. Recommended maintenance cost factors for the three IGCC systems are presented in Table A-13. These maintenance cost factors are based on information reported in two cost studies (Fluor, 1985; Corman, 1986). The total maintenance cost for the plant is given by:

$$OC_M = \sum_i f_{M,i} \left(DC_i + IDC_i + C_{PC,i} + C_{PJ,i}\right) \tag{A-124}$$

where,

$$IDC_i = TIC\left(\frac{DC_i}{TDC}\right) \tag{A-125}$$

$$C_{PJ,i} = C_{PJ}\left(\frac{DC_i}{TDC}\right) \tag{A-126}$$

and,

$f_{M,i}$ - Maintenance cost factor for plant section i (from Table A-13).

The maintenance cost can be divided into materials and labor components by assuming that 60 percent of the maintenance cost is associated with maintenance materials and the remainder is associated with maintenance labor (EPRI, 1986). Thus:

$$OC_{MM} = 0.60\ OC_M \tag{A-127}$$

$$OC_{ML} = 0.40\ OC_M \tag{A-128}$$

The administrative and support labor cost is assumed to be 30 percent of the operating and maintenance labor cost:

$$OC_{AS} = 0.30\ (OC_L + OC_{ML}) \tag{A-129}$$

Table A-13. Annual Maintenance Cost Factors for Selected IGCC Systems

		Maintenance Cost Factor[a]		
Description	Area No.	KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Oxidant Feed	10	2.0	2.0	2.0
Coal Handling	20	3.0	3.0	3.0
Limestone Handling	25	--	3.0	--
Gasification	30	4.5	4.5	4.5
Coke, Fines, and Ash	31	--	--	3.0
Cyclones	32	--	--	1.5
Sulfation	35	--	4.0	--
Gas Cooling	40	3.0	--	--
Selexol	50	2.0	--	--
Zinc Ferrite	50	--	4.5	4.5
Claus Plant	60	2.0	--	--
Sulfuric Acid Plant	60	--	--	2.0
Tail Gas	70	2.0	--	--
Boiler Feed Water	80	1.5	1.5	1.5
Process Cond. Treat.	85	2.0	--	--
Combined Cycle	90	1.5	1.5	1.5
General Facilities	100	1.5	1.5	1.5

[a] Annual maintenance cost as a percent of plant section direct, indirect, and contingency costs, as presented in Equation (81).
Sources: Corman (1986); Fluor (1985).

The total fixed operating cost is the sum of the operating labor, maintenance, and administrative and support labor costs:

$$FOC = OC_L + OC_M + OC_{AS} \quad (A\text{-}130)$$

A.7.2 Variable Operating Cost

Variable operating cost includes fuel, consumables, ash disposal, and byproduct credits. Default unit costs of fuels, consumables, ash disposal, and byproduct credits are given in Table A-14. In the following sections, the total requirements for fuel and

Table A-14. Unit Costs of Fuel, Consumables, Ash Disposal, and Byproducts

Item	Units	Cost	Basis	Reference	1/89[a]
Fuel					
Illinois No. 6 Coal	$/MMBtu	1.55	1/85	EPRI, 1986	1.61
Gulf Lignite	$/MMBtu	1.50	1/85	EPRI, 1986	1.64
Wyoming Subbituminous	$/MMBtu	1.85	1/85	EPRI, 1986	1.99
Consumables					
Sulfuric Acid (93%)	$/ton	89.40	1/87	BGE, 1989	110
NaOH (50%)	$/ton	175	1/87	BGE, 1989	220
Na2 HPO4	$/lb	0.55	1/87	BGE, 1989	0.70
Hydrazine	$/lb	2.50	1/87	BGE, 1989	3.20
Morpholine	$/lb	1.02	1/87	BGE, 1989	1.30
Lime	$/ton	65	1/85	EPRI, 1986	80
Soda Ash	$/ton	125	1/87	BGE, 1989	160
Corrosion Inhibitor	$/lb	1.55	1/83	Fluor, 1985	1.90
Surfactant	$/lb	1.05	1/83	Fluor, 1985	1.25
Chlorine	$/ton	200	1/87	BGE, 1989	250
Biocide	$/lb	3.00	1/83	Fluor, 1985	3.60
Selexol Solvent	$/lb	1.50	1/83	Fluor, 1985	1.80
Claus Catalyst	$/ton	365	1/83	Fluor, 1985	440
Sulfuric Acid Catalyst	$/liter	1.70	4/88	Higgins, 1988	1.90
SCOT Catalyst	$/ft^3	192	1/83	Fluor, 1985	230
SCOT Chemicals	$/lb	0.30	1/83	Fluor, 1985	0.36
Beavon-Stretford Catalyst	$/ft^3	154	6/81	Fluor, 1983a	170
Beavon-Stretford Chemicals	$/(lb/hr sulfur)[b]	154	6/81	Fluor, 1983a	170
Zinc Ferrite Sorbent	$/lb	3.00	6/88	Kasper, 1988	3.30
Plant Air Adsorbent	$/lb	2.30	1/83	Fluor, 1985	2.80
Flare - LPG	$/bbl	9.24	1/87	BGE, 1989	11.70
Wastewater Chemicals	$/gpm ww	695	1/83	Fluor, 1985	840
Fuel Oil	$/bbl	42	1/83	Fluor, 1985	42
Water	$/1,000 gal	0.60	1/85	EPRI, 1986	0.73
Limestone	$/ton	15	1/85	EPRI, 1986	18
Bentonite	$/lb	0.024	6/84	Parsons, 1985	0.029
Coke	$/lb	0.0175	6/84	Parsons, 1985	0.021
Ash Disposal					
Ash Disposal	$/ton	8	1/85	EPRI, 1986	10
Sludge	$/ton (dry)	9.25	1/85	EPRI, 1986	11.30
Byproduct Credits					
Sulfur	$/ton	125	1/89	Frey, 1989	125
Sulfuric Acid	$/ton	40	1/89	Frey, 1989	40

[a] Prices except for coal adjusted to 1/89 basis using the industrial chemicals producer prices index in *Chemical Engineering* magazine. The values of this index are: 6/81 = 369.4; 1/83 = 339.9; 6/84 = 345.5; 1/85 = 337.7; 1/87 = 323.9; 4/88 = 367.6; 6/88 = 372.8; 1/89 = 411.25. Coal prices adjusted based on interpolation of data in EPRI (1986), Table B.4-2.

[b] At 100% capacity factor.

consumable materials are developed, followed by a discussion of ash disposal and byproduct quantities. The material requirements are then used in conjunction with the unit costs in Table A-14 to estimate the total variable cost.

A.7.2.1 Fuel Consumption

Fuel consumption for the IGCC power plants is estimated by the ASPEN simulation models on a mass flow rate basis. The total annual fuel consumption, in units of million Btu, is then determined as follows:

$$\hat{Q}_{coal} = 8{,}760\ c_f\ m_{cf,CH,i}\ HV_{coal} \tag{A-131}$$

A.7.2.2 Feed Water Treating Consumables

Consumption of chemicals required for water treatment are a significant portion of the operating cost of an IGCC plant. The water streams that require treatment include boiler feed water and cooling water. In the boiler feed water system, raw water is treated and mixed with steam condensate. The combined stream is then chemically polished. Raw water used in the cooling water system must be treated to avoid fouling and corrosion in the cooling water system.

All of the IGCC systems require consumables for treatment of boiler feed water and steam cycle condensate. The chemicals required for raw boiler feed water dimineralization include sulfuric acid and sodium hydroxide. The chemicals required for raw boiler feed water treatment include sodium phosphate, hydrazine, and morpholine. The required quantity of these chemicals is proportional to the raw water intake rate. The chemicals required for polishing the boiler feed water and steam cycle condensate include sulfuric acid and sodium hydroxide. The required quantity of the polishing chemicals is proportional to the flow rate of water through the polishing unit, which is a primarily a combination of the raw water feed rate and the flow rate of condensate from the steam turbine, with a minor contribution (e.g., less than 5 percent) from condensate in other process units.

The required quantity of each steam cycle water treatment chemical has been estimated as a function of the raw or polished water flow rate based on detailed data available in a number of cost studies (Fluor, 1983a; Fluor, 1983b; Fluor, 1984; Fluor, 1985; Fluor, 1986). For each chemical and application, the regression model is presented and summarized below. Each regression model is shown graphically in the indicated figures.

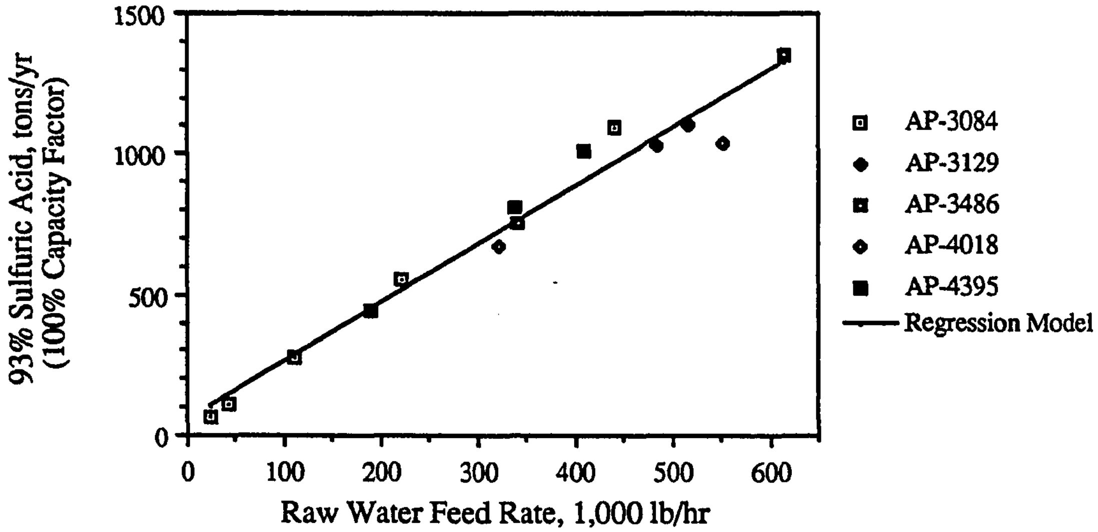

Figure A-33. Sulfuric Acid Requirement for Boiler Feed Water Demineralization.

Sulfuric Acid for Raw Water Dimineralization

$$m_{sa,BF,i} = c_f(47.0 + 2.09\times10^{-3}\ m_{rw}) \qquad (R^2 = 0.969;\ n = 14) \qquad \text{(A-132)}$$

where,

$$24{,}200 \le m_{rw} \le 613{,}000\ \text{lb/hr.}$$

The sulfuric acid flow rate is expressed in units of tons per year. The standard error of the estimate is 70 tons/year. The regression model is shown graphically in Figure A-33.

Sodium Hydroxide for Raw Water Dimineralization

$$m_{sh,BF,i} = c_f(9.5 + 4.20\times10^{-4} m_{rw}) \qquad (R^2 = 0.969;\ n = 14) \qquad \text{(A-133)}$$

where,

$$24{,}200 \le m_{rw} \le 613{,}000\ \text{lb/hr.}$$

The sodium hydroxide flow rate is expressed in tons per year. The standard error of the estimate is 15 tons/year. The regression model is shown graphically in Figure A-34.

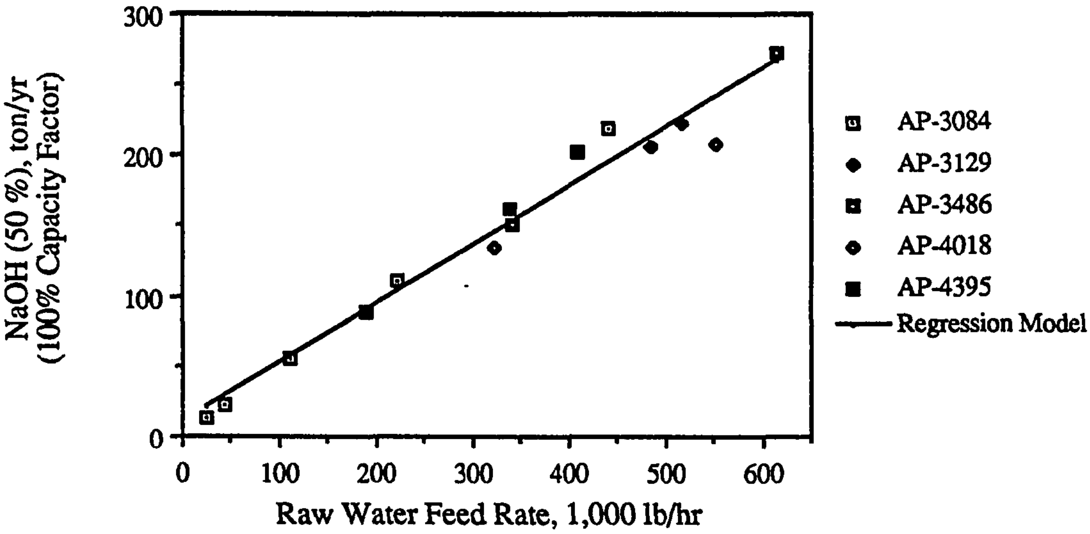

Figure A-34. Sodium Hydroxide Requirement for Boiler Feed Water Demineralization.

Sodium Phosphate for Raw Water Treating

$$m_{sp,BF,i} = c_f(115 + 3.61\times10^{-3} m_{rw}) \qquad (R^2 = 0.962;\ n = 14) \qquad \text{(A-134)}$$

where,

$$24{,}200 \le m_{rw} \le 613{,}000 \text{ lb/hr.}$$

The sodium phosphate flow rate is expressed in units of lb/yr. The standard error of the estimate is 140 lb/yr. The regression model is shown graphically in Figure A-35.

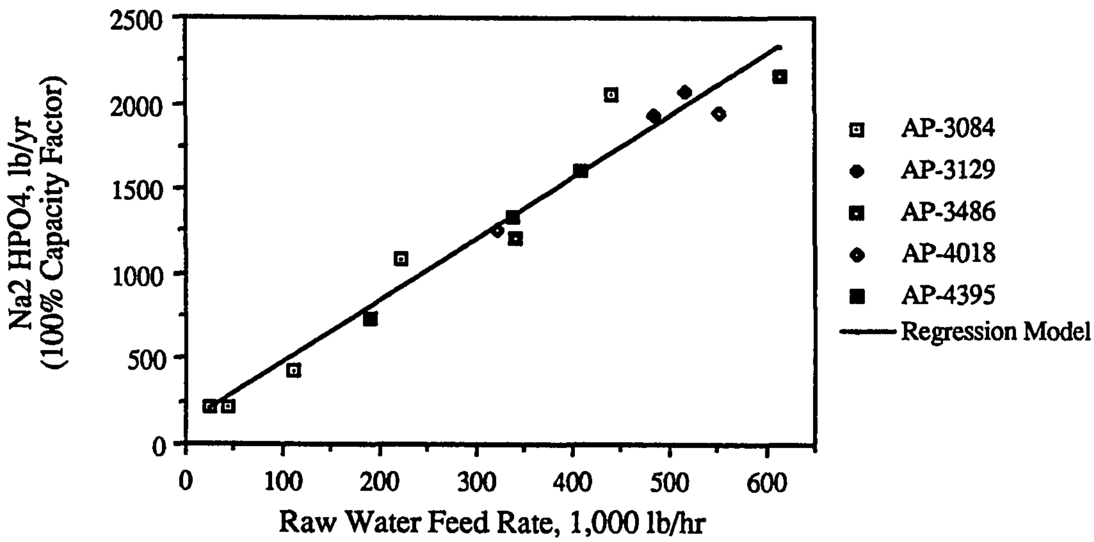

Figure A-35. Sodium Phosphate Requirement for Boiler Feed Water Treating.

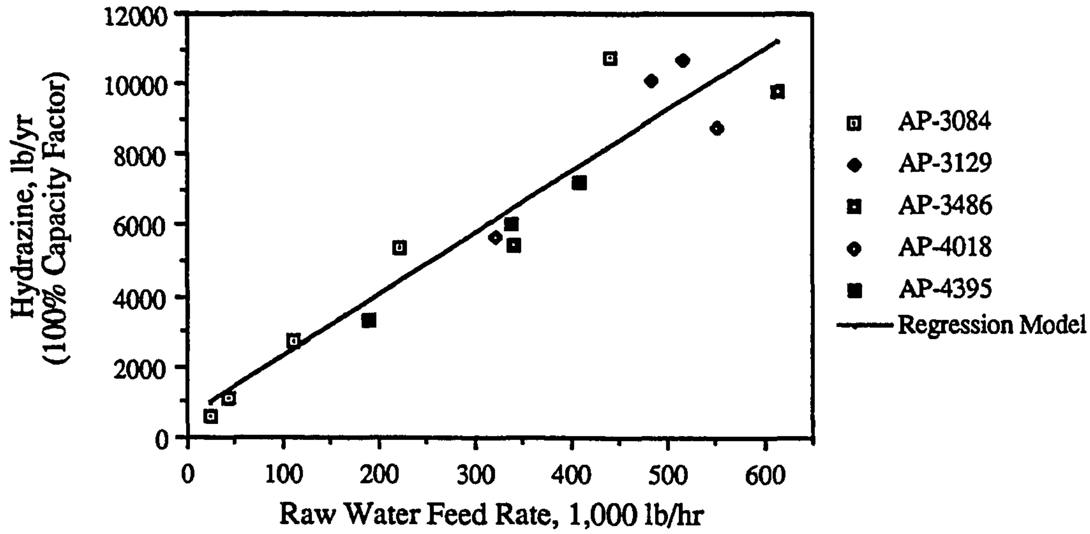

Figure A-36. Hydrazine Requirement for Boiler Feed Water Treating.

Hydrazine for Raw Water Treating

$$m_{hy,BF,i} = c_f(529 + 0.0174\ m_{rw}) \qquad (R^2 = 0.898;\ n = 14) \tag{A-135}$$

where,

$$24{,}200 \leq m_{rw} \leq 613{,}000 \text{ lb/hr.}$$

The hydrazine flow rate is expressed in units of lb/yr. The standard error of the estimate is 1,200 lb/yr. The regression model is shown graphically in Figure A-36.

Morpholine for Raw Water Treating

$$m_{mo,BF,i} = c_f(420 + 0.0163\ m_{rw}) \qquad (R^2 = 0.965;\ n = 14) \tag{A-136}$$

where,

$$24{,}200 \leq m_{rw} \leq 613{,}000 \text{ lb/hr.}$$

The morpholine flow rate is expressed in units of lb/yr. The standard error of the estimate is 610 lb/yr. The regression model is shown graphically in Figure A-37.

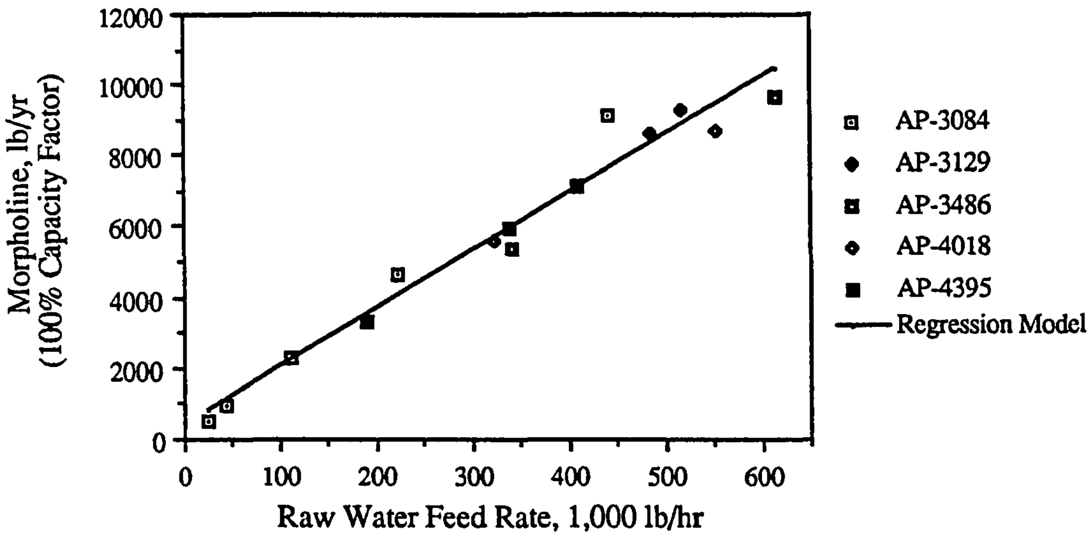

Figure A-37. Morpholine Requirement for Boiler Feed Water Treating.

Sulfuric Acid for Condensate Polishing

$$m_{sa,BFP,i} = c_f(15 + 5.4\times10^{-5}\ m_{pw}) \qquad (R^2 = 0.992;\ n = 7) \qquad \text{(A-137)}$$

where,

$$1{,}200{,}000 \leq m_{pw} \leq 2{,}200{,}000\ \text{lb/hr}.$$

The sulfuric acid flow rate is expressed in units of tons per year. The standard error of the estimate is 2 tons/year. The regression model is shown graphically in Figure A-38. As indicated in the figure, seven data points were excluded from the regression analysis. These data points are from the earliest of the five studies, and are inconsistent with the sulfuric acid requirement reported in the more recent studies. Furthermore, the data that were used in the regression model include the estimates developed specifically for a KRW-based system (Fluor, 1985).

Sodium Hydroxide for Condensate Polishing

$$m_{sh,BFP,i} = c_f(30 + 1.07\times10^{-4}\ m_{pw}) \qquad (R^2 = 0.991;\ n = 7) \qquad \text{(A-138)}$$

where,

$$1{,}200{,}000 \leq m_{pw} \leq 2{,}200{,}000\ \text{lb/hr}.$$

The sodium hydroxide flow rate is expressed in units of tons per year. The standard error of the estimate is 4 tons/year. The regression model is shown graphically in Figure A-39. As indicated in the figure, seven data points were excluded from the regression analysis, for the same reasons noted above for the sulfuric acid requirement regression model.

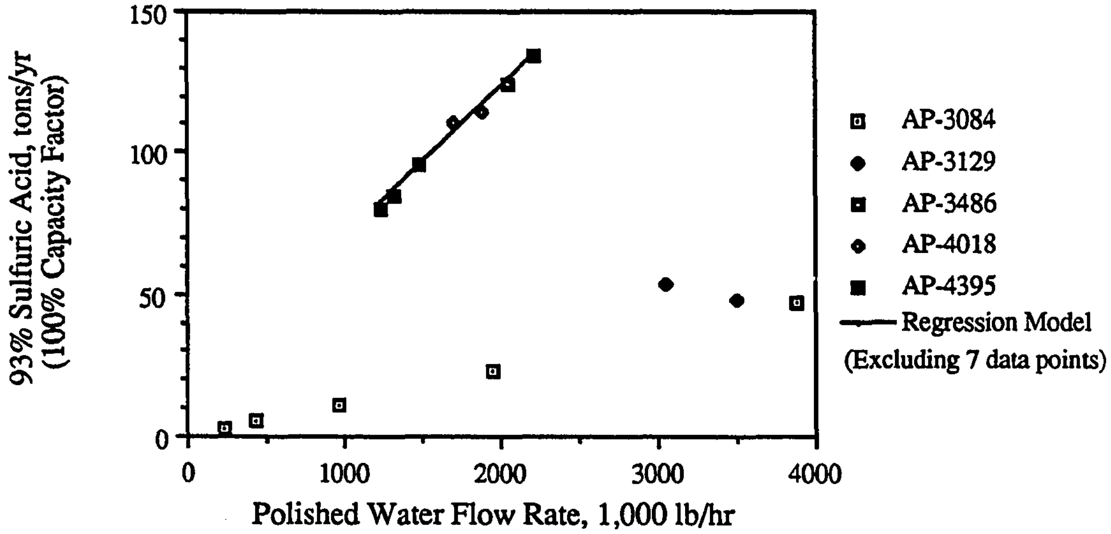

Figure A-38. Sulfuric Acid Requirement for Polishing.

Chemical treatment is also required for the cooling water system. From the same five studies used to development the boiler feed water chemical treatment requirements, the cooling water treatment requirements were estimated. The amount of each chemical required was found to be directly proportional to the makeup cooling water flow rate. The required chemicals include lime, soda ash, sulfuric acid, a corrosion inhibitor, a surfactant, chlorine, and a biocide. The regression models for the annual required flow rate of each of these treatment chemicals is presented below. Each regression model is represented graphically in the indicated figures.

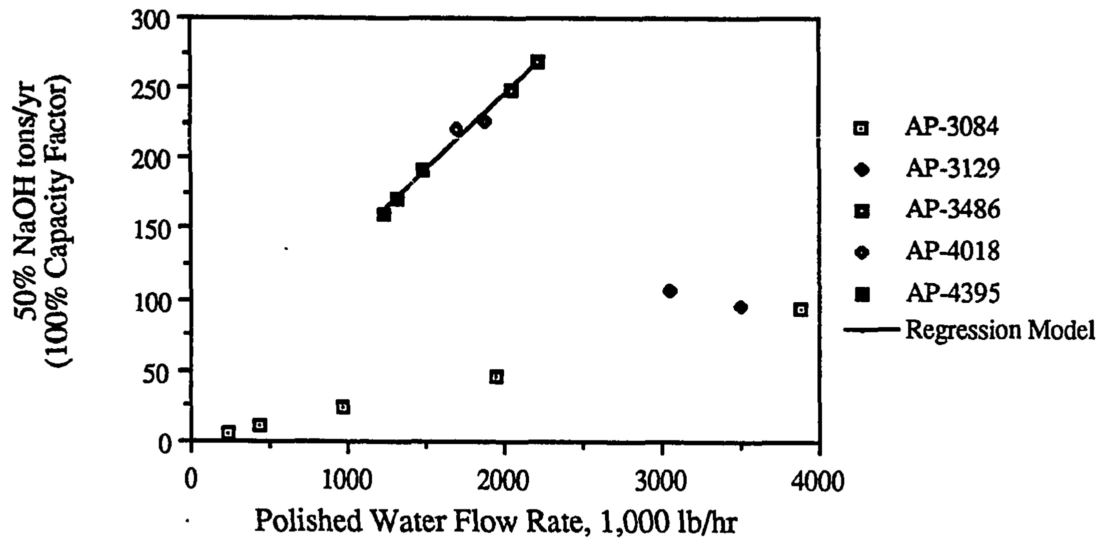

Figure A-39. Sodium Hydroxide Requirement for Polishing.

Lime for Cooling Water Treating

$$m_{lime,CW,i} = c_f(-26 + 0.143\ m_{cw,i}) \qquad (R^2 = 0.991;\ n = 12) \qquad \text{(A-139)}$$

where,

$$415 \leq m_{cw,i} \leq 7{,}700 \text{ gal/min.}$$

The lime flow rate is expressed in units of tons per year. The standard error of the estimate is 27 tons/year. The regression model is shown graphically in Figure A-40. As indicated in the figure, two data points, which are outliers, were excluded from the regression analysis.

Soda Ash for Cooling Water Treating

$$m_{so,CW,i} = 0.154\ c_f\ m_{cw,i} \qquad (R^2 = 0.989;\ n = 14) \qquad \text{(A-140)}$$

where,

$$415 \leq m_{cw,i} \leq 8{,}300 \text{ gal/min.}$$

The soda ash flow rate is expressed in units of tons per year. The standard error of the estimate is 47 tons/year. The regression model is shown graphically in Figure A-41.

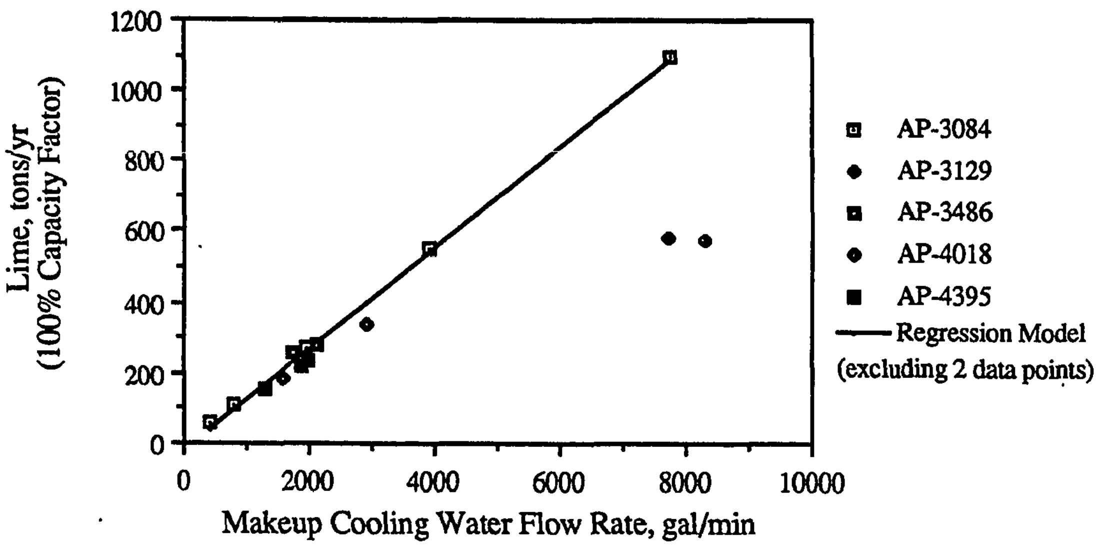

Figure A-40. Lime Requirement for Makeup Cooling Water Treating.

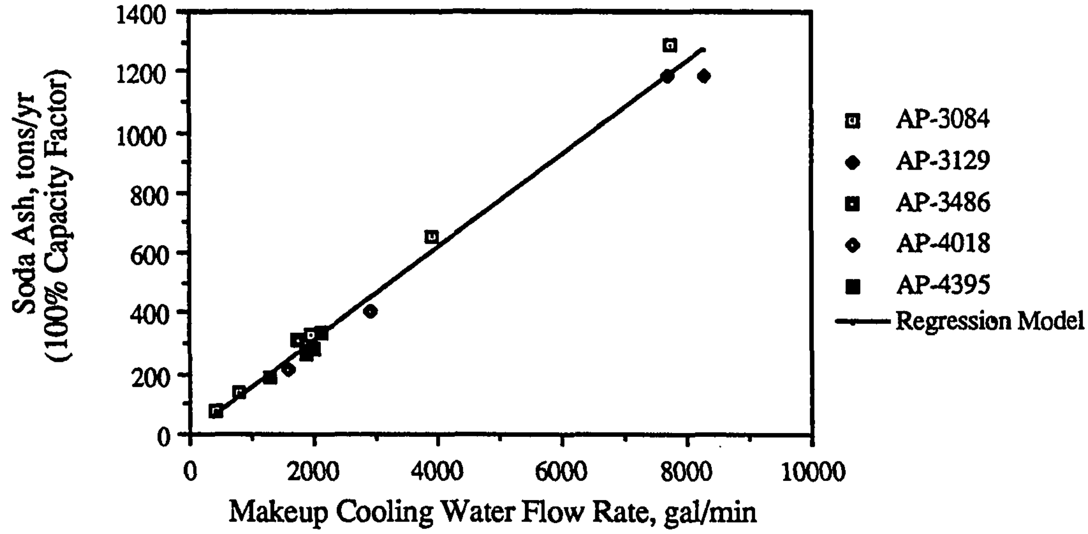

Figure A-41. Soda Ash Requirement for Makeup Cooling Water Treating.

Sulfuric Acid for Cooling Water Treating

$$m_{sa,CW,i} = c_f(6 + 0.147\ m_{cw,i}) \qquad (R^2 = 0.984;\ n = 14) \qquad \text{(A-141)}$$

where,

$$415 \le m_{cw,i} \le 8{,}300 \text{ gal/min.}$$

The sulfuric acid flow rate is expressed in units of tons per year. The standard error of the estimate is 53 tons/year. The regression model is shown graphically in Figure A-42.

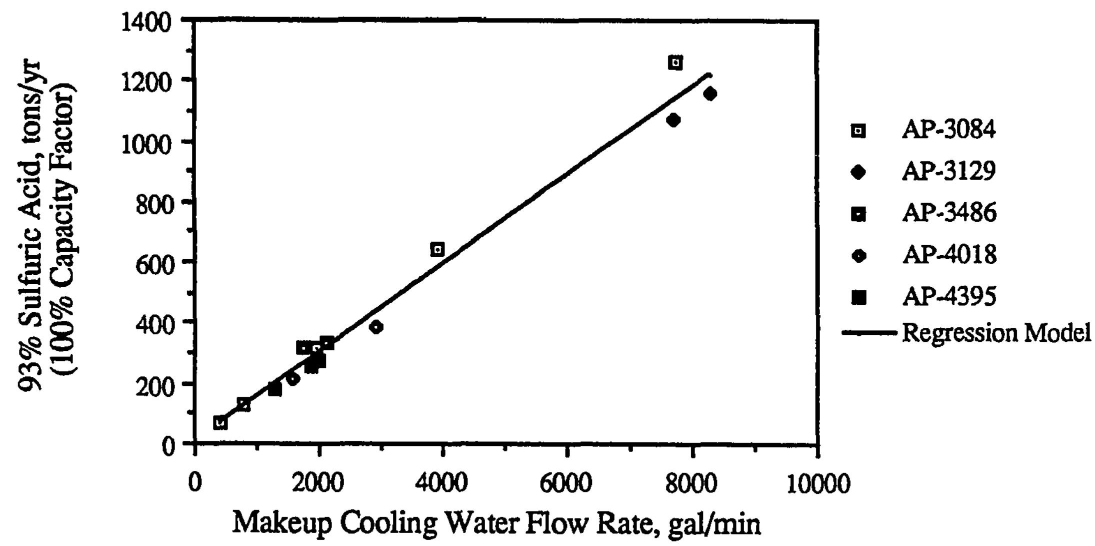

Figure A-42. Sulfuric Acid Requirement for Cooling Water Treating.

Corrosion Inhibitor for Cooling Water Treating

$$m_{ci,CW,i} = c_f(-7{,}280 + 28.8\ m_{cw,i}) \qquad (R^2 = 0.964;\ n = 12) \qquad \text{(A-142)}$$

where,

$$415 \le m_{cw,i} \le 7{,}700 \text{ gal/min.}$$

The corrosion inhibitor flow rate is expressed in units of pounds per year. The standard error of the estimate is 11,100 lb/yr. The regression model is shown graphically in Figure A-43. As indicated in the figure, two data points, which are outliers, were excluded from the regression analysis.

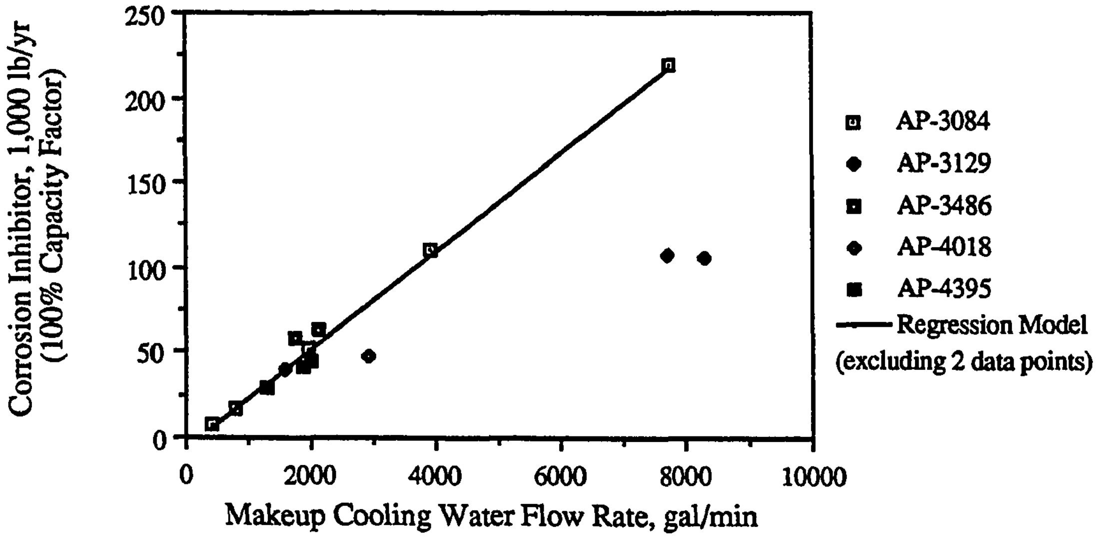

Figure A-43. Corrosion Inhibitor Requirement for Cooling Water Treating.

Surfactant for Cooling Water Treating

In all of the cost studies used to develop the cooling water treatment chemical requirements, the mass flow rate of surfactant is equal to the mass flow rate of corrosion inhibitor. Therefore, the surfactant requirement, in units of lb/yr, is given by:

$$m_{su,CW,i} = m_{ci,CW,i} \qquad \text{(A-143)}$$

Chlorine for Cooling Water Treating

$$m_{Cl,CW,i} = c_f(4.1 + 3.74\times10^{-3}\ m_{cw,i}) \qquad (R^2 = 0.920;\ n = 14) \qquad \text{(A-144)}$$

where,

$$415 \le m_{cw,i} \le 8{,}300 \text{ gal/min.}$$

The chlorine flow rate is expressed in units of tons per year. The standard error of the estimate is 3.1 tons/year. The regression model is shown graphically in Figure A-44.

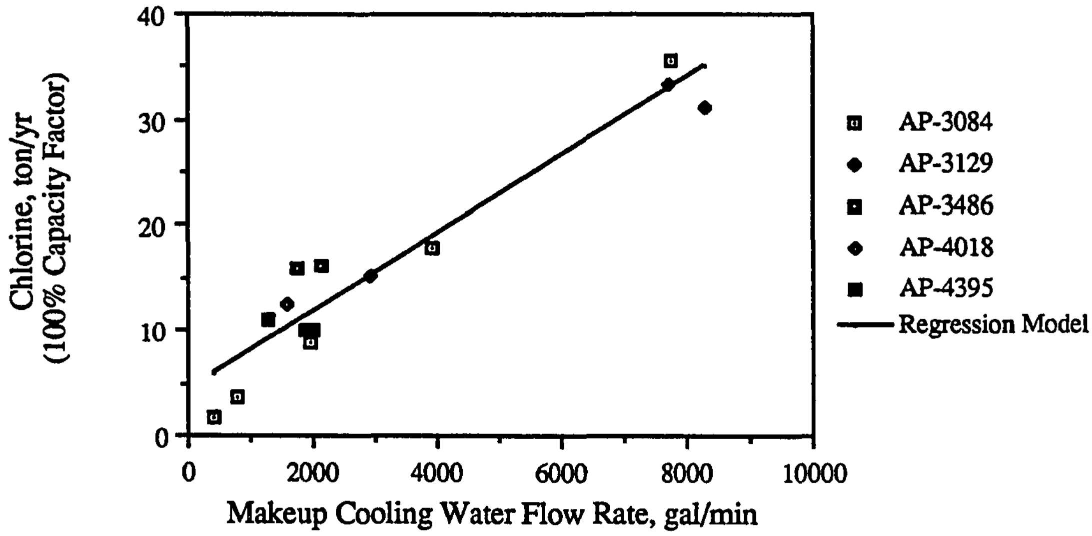

Figure A-44. Chlorine Requirement for Cooling Water Treating.

Biocide for Cooling Water Treating

$$m_{bio,CW,i} = c_f(4{,}800 + 4.12\ m_{cw,i}) \qquad (R^2 = 0.922;\ n = 14) \qquad \text{(A-145)}$$

where,

$$415 \le m_{cw,i} \le 8{,}300 \text{ gal/min.}$$

The biocide flow rate is expressed in units of pounds per year. The standard error of the estimate is 3,360 pounds/year. The regression model is shown graphically in Figure A-45.

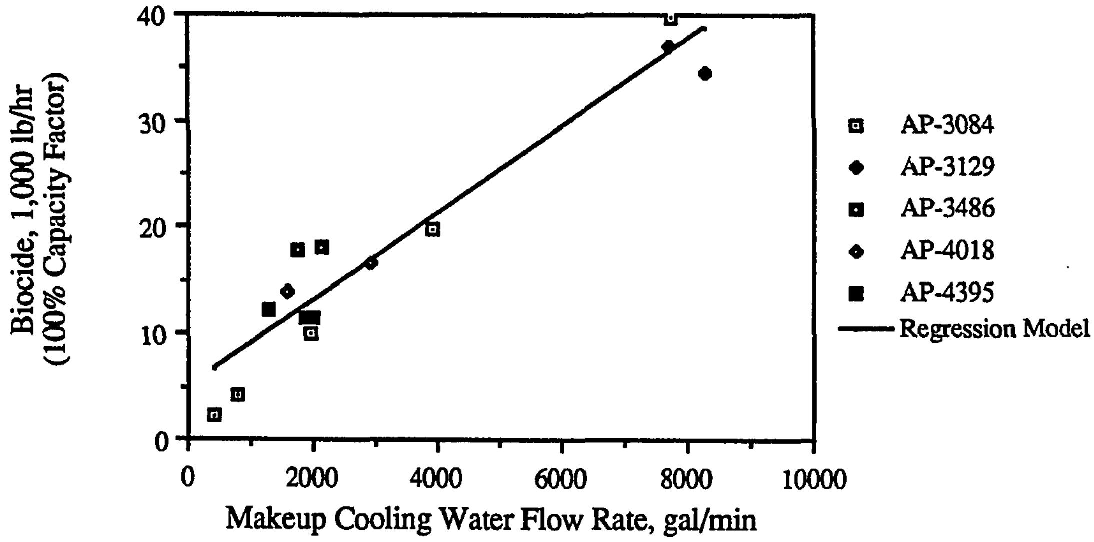

Figure A-45. Biocide Requirement for Cooling Water Treating.

A.7.2.3 Sulfur Removal and Recovery

Makeup chemicals or catalysts are required for the sulfur removal and recovery systems in all IGCC designs. For cold gas cleanup systems, the makeup requirements include Selexol solvent, Claus plant catalyst, and either SCOT or Beavon-Stretford catalyst and chemicals. For the hot gas cleanup system with off-gas recycle, the only requirement is for makeup zinc ferrite sorbent. For a hot gas cleanup system with sulfuric acid recovery, makeup sulfuric acid catalyst is also required. The operating material requirements for these systems are summarized below.

Makeup Selexol Solvent

$$m_{ss,S,i} = c_f(-350 + 1.58\ M_{syn,S,i}) \qquad (R^2 = 0.989;\ n = 11) \qquad \text{(A-146)}$$

where,

$$4{,}000 \leq M_{syn,S,i} \leq 74{,}500 \text{ lbmole/hr.}$$

The makeup Selexol solvent flow rate is expressed in units of pounds per year. The standard error of the estimate is 4,400 pounds/year. The regression model is shown graphically in Figure A-46.

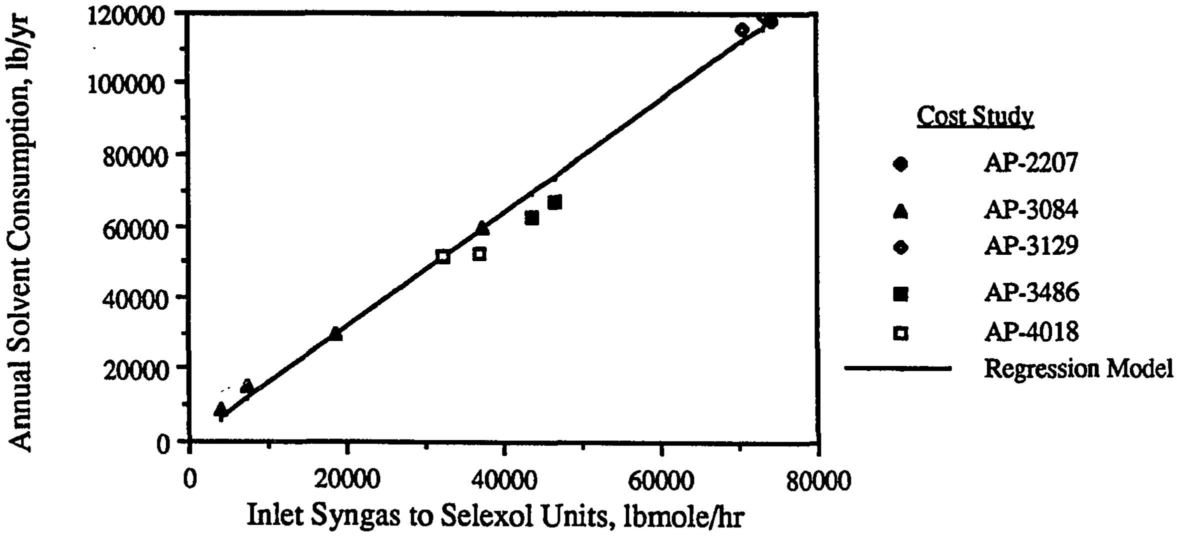

Figure A-46. Annual Solvent Requirement for the Selexol Process.

Makeup Claus Plant Catalyst

$$m_{cat,C,i} = 9.61\times10^{-4}\ c_f\ m_{s,C,o} \qquad (R^2 = 0.843;\ n = 13) \qquad \text{(A-147)}$$

where,

$$1{,}000 \leq m_{s,C,o} \leq 26{,}000 \text{ lb/hr.}$$

The makeup Claus plant catalyst requirement is expressed in units of tons per year. The standard error of the estimate is 3 tons/year. The regression model is shown graphically in Figure A-47.

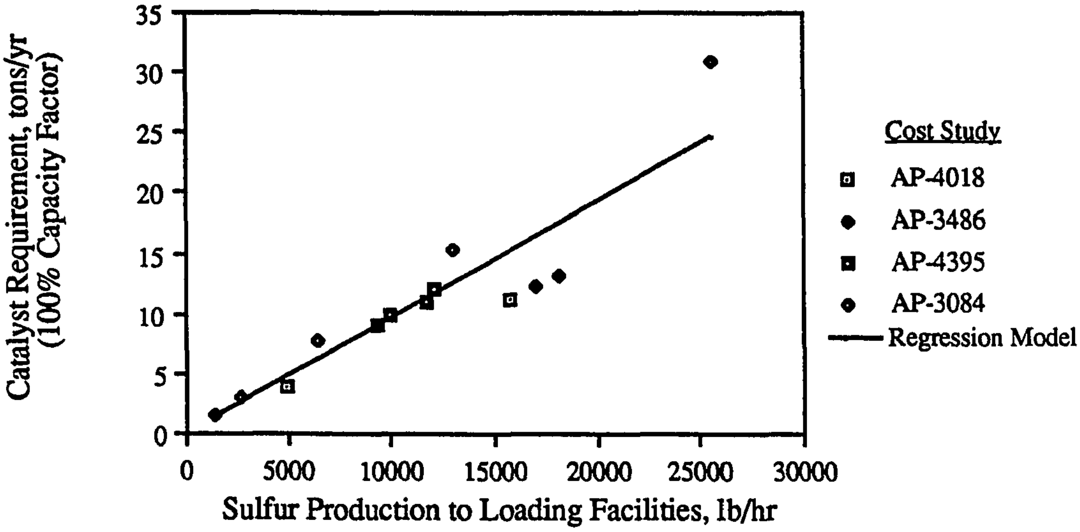

Figure A-47. Annual Catalyst Requirement for Two-Stage Claus Plant.

Makeup SCOT Catalyst

$$m_{cat,SC,i} = 0.0822\ c_f\ m_{s,C,o} \qquad (R^2 = 0.993;\ n = 7) \qquad \text{(A-148)}$$

where,

$$5{,}000 \leq m_{s,C,o} \leq 18{,}000 \text{ lb/hr.}$$

The makeup SCOT catalyst requirement is expressed in units of cubic feet per year. The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-48. As indicated in the figure, three outlier data points were excluded from the regression analysis. Two of these points are from the least recent study. The regression model over-estimates the outlier points, and therefore it is conservative for these outlier cases.

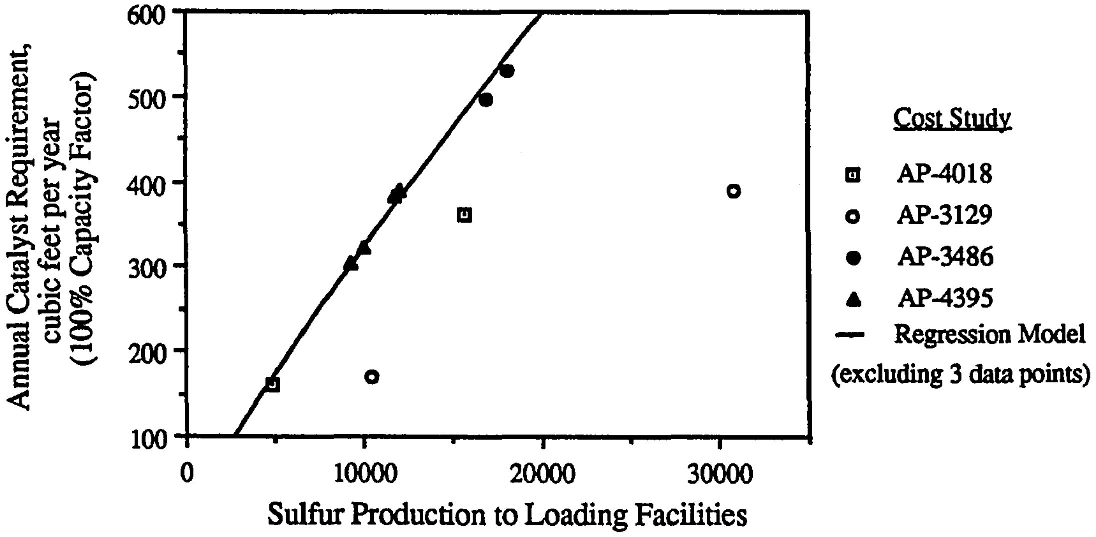

Figure A-48. Annual Catalyst Requirement for the SCOT Process.

Makeup SCOT Chemicals

$$m_{chem,SC,i} = c_f(-270 + 1.10\ m_{s,C,o}) \qquad (R^2 = 0.883;\ n = 10) \qquad \text{(A-149)}$$

where,

$$5{,}000 \leq m_{s,C,o} \leq 30{,}800 \text{ lb/h}$$

The makeup SCOT chemical requirement is expressed in units of pounds per year. The standard error of the estimate is 3,000 lb/yr. The regression model is shown graphically in Figure A-49.

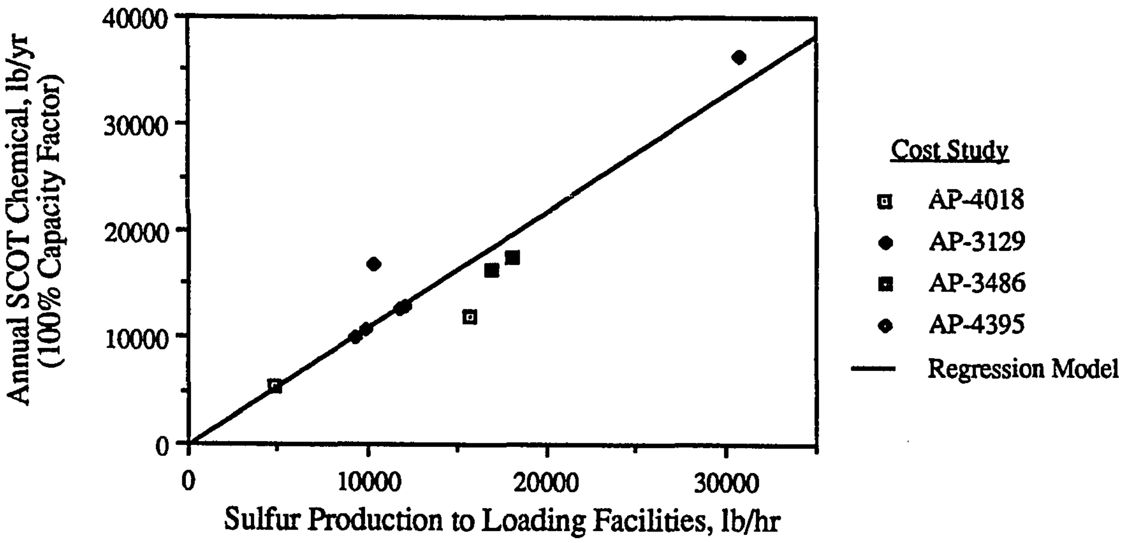

Figure A-49. Annual SCOT Chemical Requirement.

Makeup Beavon-Stretford Catalyst

$$m_{cat,BS,i} = 0.0856\ c_f\ m_{s,BS,o} \qquad (R^2 = 1.00;\ n = 5) \qquad \text{(A-150)}$$

where,

$$100 \le m_{s,BS,o} \le 2{,}000 \text{ lb/hr}$$

The makeup Beavon-Stretford catalyst requirement is expressed in units of cubic feet per year. The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-50. Two outlier data points were excluded from the analysis, as indicated in the figure. These points, both from the same study (Fluor, 1983b), appear inconsistent with the more extensive set of data from the other study (Fluor, 1983a).

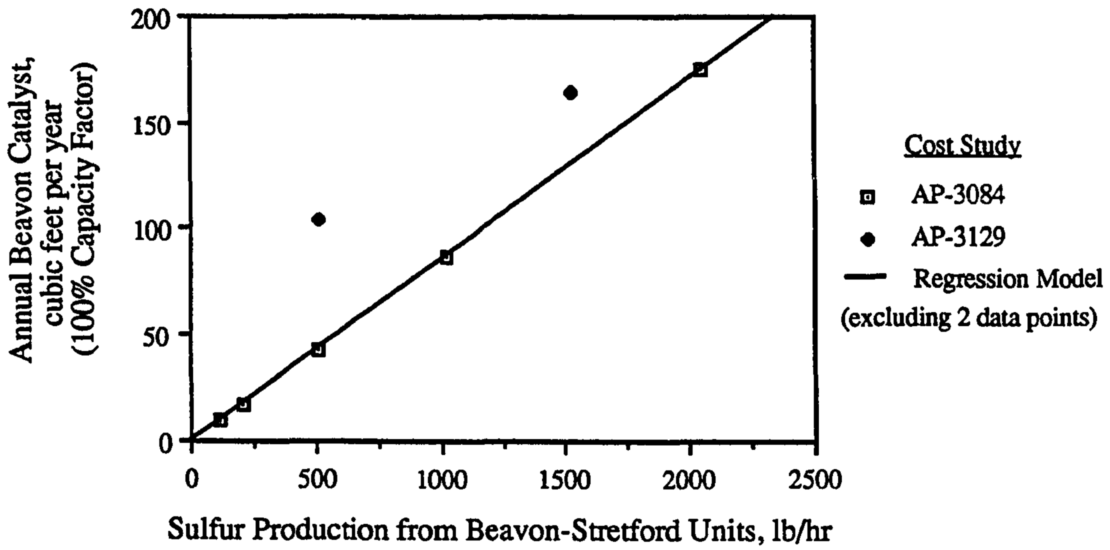

Figure A-50. Annual Catalyst Requirement for the Beavon-Stretford Process.

Makeup Beavon-Stretford Chemicals

$$C_{chem,BS,i} = 170\ c_f\ m_{s,BS,o} \qquad (R^2 = 1.00;\ n = 5) \qquad \text{(A-151)}$$

where,

$$100 \le m_{s,BS,o} \le 2{,}000 \text{ lb/hr}$$

Unlike the other consumable chemicals and catalysts, data are not available regarding the makeup mass flow rate for the Stretford chemicals. However, data are available regarding the cost of the Stretford chemicals. The regression model shown in Equation (A-151) is the cost of the Stretford chemicals, in dollars, as a function of the sulfur recovered in the Beavon-Stretford process. The standard error of the model is negligible. The model is shown graphically in Figure A-51.

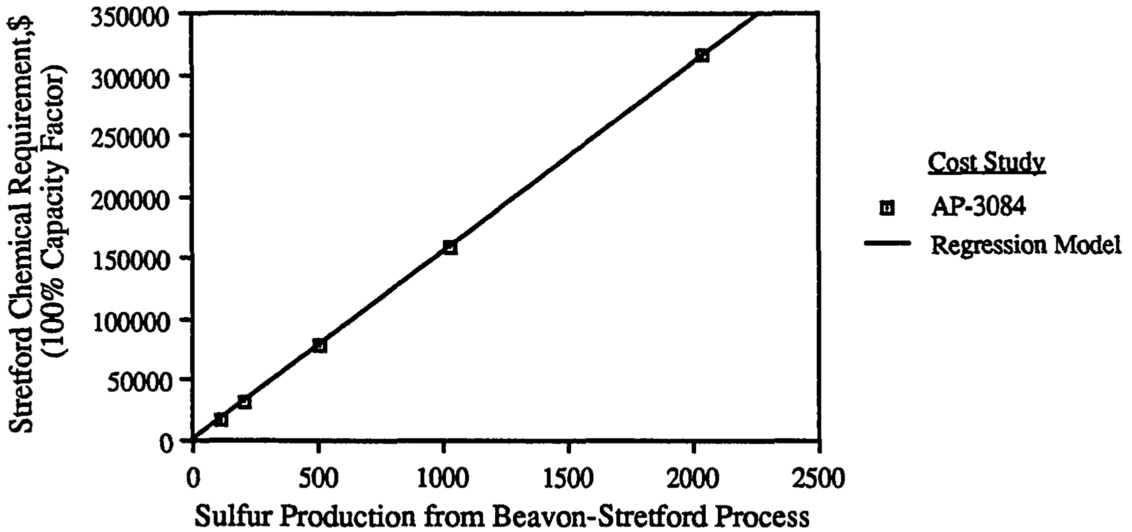

Figure A-51. Annual Chemical Cost for the Beavon-Stretford Process.

Makeup Zinc Ferrite Sorbent

Makeup zinc ferrite sorbent is required to replace sorbent that disintegrates into fines and that may be entrained with the syngas or cause a significant increase in the pressure drop across the reactor. A replacement schedule of 20 percent of the sorbent in the reactors is suggested every 80 cycles (Kasper, 1988). Spent sorbent is assumed to be returned to the manufacturer for reprocessing (Banchik, Buckman, and Rath, 1988). The annual required replacement rate is therefore given by:

$$m_{zf,ZF,i} = \frac{8{,}760\ c_f}{2\ t_a}\left(\frac{f_{att}\ S_c}{80}\right) \tag{A-152}$$

where, as a default, it is assumed that $f_{att} = 0.20$.

Makeup Sulfuric Acid Plant Catalyst

The sulfuric acid plant, which is assumed for sulfur recovery in the Lurgi-based system, requires makeup catalyst. The number of liters of makeup catalyst depends on the design capacity of the sulfuric acid plant. Based on data from Monsanto (Higgins, 1988), the annual makeup catalyst, in liters, is approximated as:

$$m_{cat,SA,i} = 1.1\ M_{SA,i} \tag{A-153}$$

A.7.2.4 Other Consumables

Other consumables that are required for all systems are gas turbine startup fuel, plant and instrument air adsorbent, wastewater treatment chemicals, water, and liquified petroleum gas for flares. For the KRW-based system with hot gas cleanup, limestone is required for in-bed desulfurization. For the Lurgi-based system, coke is required as a gasifier startup fuel, and bentonite is required as a binder for fines agglomeration.

The consumables required for all systems are estimated based on a simple scaling relationship with plant size, on the premise that the requirement of these items scales directly with the plant capacity. The requirements for these items are based on data reported in Fluor (1985), for a nominal 550 MW plant.

The annual required gas turbine startup fuel oil, in barrels, is assumed to be:

$$m_{fo,GT,i} = 48{,}000\ c_f \left(\frac{MW}{550}\right) \qquad \text{(A-154)}$$

The annual plant and instrument air adsorbent, in pounds, is assumed to be:

$$m_{ads,GF,i} = 3{,}600\ c_f \left(\frac{MW}{550}\right) \qquad \text{(A-155)}$$

The wastewater chemical requirement is given in dollar, not mass, terms. The reported cost of wastewater chemicals is given in Table A-14 as an equivalent annual cost, at 100 percent capacity factor, based on the wastewater flow rate. The cost of wastewater chemicals is therefore given by:

$$C_{ww} = c_f\, UC_{ww}\, m_{ww} \qquad \text{(A-156)}$$

The plant water consumption is given by the total of the raw water consumption for the steam cycle and the cooling water makeup water. The total is:

$$m_{water} = 8{,}760\ c_f\ m_{rw} \qquad \text{(A-157)}$$

The consumption of liquified petroleum gas for maintaining the plant flare is assumed to be:

$$m_{LPG} = 4{,}200\ c_f \left(\frac{MW}{550}\right) \qquad \text{(A-158)}$$

The limestone requirement for in-bed desulfurization depends on the sulfur content of the coal and the calcium-to-sulfur molar ratio required to achieve the design in-bed sulfur removal efficiency. For 90 percent sulfur removal, a molar ratio of 1.8 is commonly assumed (Smelser, 1986; Smith and Smelser, 1987; Earley and Smelser, 1988). Therefore, the annual limestone requirement, in pounds, is given by:

$$m_{limestone} = 3.13 \left(\frac{R_{Ca/S}}{f_{CaCO_3}} \right) m_{coal}\, f_s \tag{A-159}$$

where,

$$m_{coal} = 8{,}760\ c_f\, m_{cf,CH,i} \tag{A-160}$$

and,

$R_{Ca/S}$ = molar ratio of calcium to sulfur, with a default value of 1.8.

f_{CaCO_3} = weight percent calcium carbonate in limestone. Default is 0.95.

The coke requirement for Lurgi-based systems is assumed to be directly proportional to the plant energy output, based on data in Bechtel et al (1988):

$$m_{coke} = 610\ c_f \left(\frac{MW}{180} \right) \tag{A-161}$$

The required quantity of coke is expressed in tons.

The bentonite requirement is assumed to be directly proportional to the fines flow rate through the gasifier. The bentonite requirement for coal fines agglomeration is reported as 26 tons per day to agglomerate 600 tons per day of coal fines (Bechtel et al, 1988). Therefore, the annual bentonite requirement, in pounds, is assumed to be:

$$m_{bent} = 0.043\ (8{,}760)\ c_f\, m_{fine} \tag{A-162}$$

A.7.2.5 Ash Disposal and Byproducts

The annual ash disposal requirement includes bottom ash from the gasifiers and ash recovered from the syngas in cyclones or scrubbers. In the case of the systems with hot gas cleanup, all ash is in dry form. In the case of the KRW-based system with cold gas cleanup, a portion of the ash is disposed of as filter cake from the syngas particulate scrubbing unit, which contains typically about 60 percent dry ash (e.g., Fluor, 1985). The total ash to disposal may be approximately estimated as the total ash content in the coal:

$$m_{ash,o} = m_{coal}\, f_{ash} \tag{A-163}$$

The amount of sulfur recovered in the Claus plant (and Beavon-Stretford unit, if applicable) for the KRW-based system with cold gas cleanup, or the amount of sulfuric acid recovered for the Lurgi-based system, is the basis for estimating the byproduct credit. These flows are estimated in the ASPEN simulations.

A.7.2.6 Total Variable Operating Costs

To estimate the total variable operating cost, the annual material requirements appropriate to the given IGCC system must be multiplied by their respective unit costs. In two cases (Beavon-Stretford chemicals and wastewater treatment chemicals) the unit costs are based on a process flow rate (sulfur recovered in the Beavon-Stretford unit and the wastewater discharge rate) because the material requirements of the consumables themselves are not reported. The variable operating costs should be reported in the same categories as presented in Table A-11 (i.e., fuel, consumables, ash disposal, and byproduct credit). The major components of the total variable operating cost, expressed in dollars, are:

$$OC_{fuel} = m_{coal}\, UC_{coal} \quad \text{(A-164)}$$

$$OC_{cons} = \sum_{cons} m_i\, UC_i \quad \text{(A-165)}$$

$$OC_{ash} = m_{ash,o}\, UC_{ash} \quad \text{(A-166)}$$

$$OC_{byp} = (1 - f_{bm})\, m_{byp}\, UC_{byp} \quad \text{(A-167)}$$

where,

f_{bm} = fraction of byproduct revenue required for marketing and shipping costs. The default is assumed to be 10 percent, or $f_{bm} = 0.10$.

The total variable cost is then:

$$VOC = OC_{fuel} + OC_{cons} + OC_{ash} - OC_{byp} \quad \text{(A-168)}$$

A.7.3 Preproduction Costs and Inventory Capital

In Section A.6, preproduction costs are summarized in Equation (A-107). These costs include one month of fixed operating costs; one month of variable operating costs excluding fuel and byproduct credit; one-quarter of the fuel requirement for one month of operation at full load; and an additional allowance based on a percentage of the total plant investment. Note that the operating costs are expressed in units of dollars, while the capital costs are expressed in units of 1,000 dollars. Therefore, a conversion factor is required when estimating preproduction and inventory capital costs based on the operating cost equations presented earlier. One month of fixed operating cost (in thousands of dollars) is given by:

$$PP_{FC} = \frac{1}{12}\left(\frac{FOC}{1{,}000}\right) \tag{A-169}$$

Similarly, one month of variable operating costs, excluding fuel and byproduct credits, can be estimated as follows:

$$PP_{OC} = \left(\frac{0.083}{c_f}\right)\left(\frac{OC_{cons} + OC_{ash}}{1{,}000}\right) \tag{A-170}$$

The preproduction fuel cost is given by:

$$PP_{Fuel} = \left(\frac{0.021}{c_f}\right)\left(\frac{OC_{fuel}}{1{,}000}\right) \tag{A-171}$$

The preproduction capital costs are then fully specified by substituting Equations (A-169), (A-170), and (A-171) into Equation (A-107).

The inventory capital costs include 60 days of fuel and consumable inventories, based on full plant load. These costs can be estimated from the variable operating costs as follows:

$$IC = \left(\frac{0.164}{c_f}\right)\left(\frac{OC_{fuel} + OC_{cons}}{1{,}000}\right) \tag{A-172}$$

Equation (A-172) can be substituted for Equation (A-108) in Section A.6.

A.7.4 Auxiliary Electrical Requirements

A number of process areas consume significant amounts of electric power, thereby reducing the saleable electrical output of the power plant. When reporting costs on a normalized basis (e.g., $/kW or mills/kWh), it is important to use an accurate estimate of the net electrical production available for sale. For many process areas, the ASPEN performance models do not estimate the internal electrical loads. Therefore, simple regression models of power consumption versus key flow rates have been developed for major process sections. These models provide more accurate estimates of the plant electrical requirements than are currently available in the ASPEN IGCC simulation models.

For each major plant section for which power consumption data are available, a regression model was developed. These models are presented below.

KRW Coal Handling

$$W_{e,CH} = 378 + 0.533\ m_{cf,CH,i} \qquad (R^2 = 0.918;\ n = 6) \qquad \text{(A-173)}$$

where,

$$5{,}900 \le m_{cf,CH,i} \le 27{,}000 \text{ tons/day}$$

The standard error of the estimate is 1,580 kW. The regression model is shown graphically in Figure A-52.

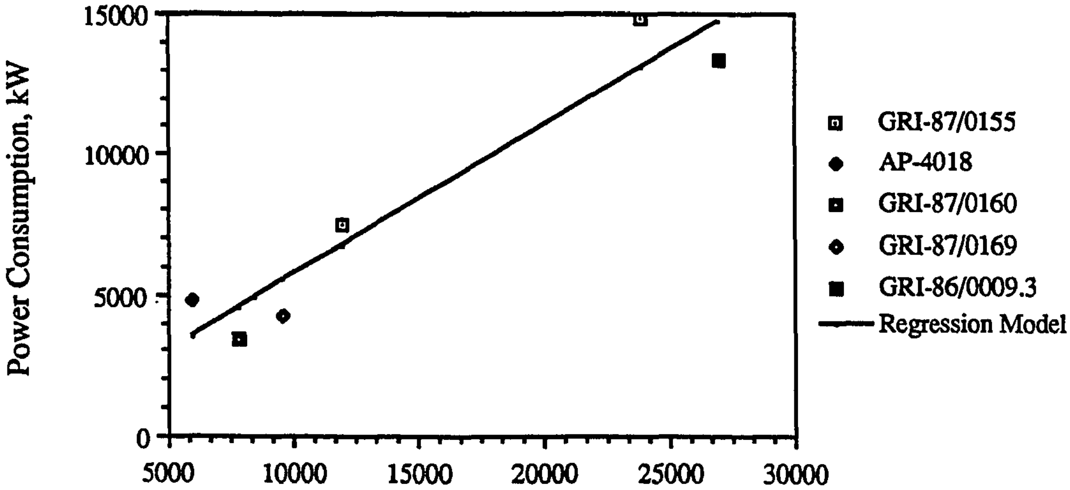

Figure A-52. Power Requirement for KRW Coal Handling.

Oxygen Plant

$$W_{e,OF} = 6.09\ M_{O,G,i} \qquad (R^2 = 0.967;\ n = 27) \qquad \text{(A-174)}$$

where

$$1{,}250 \le M_{O,G,i} \le 22{,}700 \text{ lbmole/hr.}$$

The standard error of the estimate is 6,600 kW. The regression model is shown graphically in Figure A-53. A factor to adjust the power consumption for changes in ambient temperature was developed based on analysis of data from Fluor (1985). This factor is:

$$f_{e,OF} = 0.9466 + 3.73\times10^{-3}T_a + 9.019\times10^{-6}T_a^2 \qquad \text{(A-175)}$$

This factor has a value of 1.0 at an ambient temperature of 59 °F, and it is valid only for temperatures between 20 °F and 95 °F. The modified expression for oxidant feed power consumption is therefore given by:

$$W_{e,OF} = 6.09\left(0.9466 + 3.73\times10^{-3}T_a + 9.019\times10^{-6}T_a^2\right) M_{O,G,i} \quad \text{(A-176)}$$

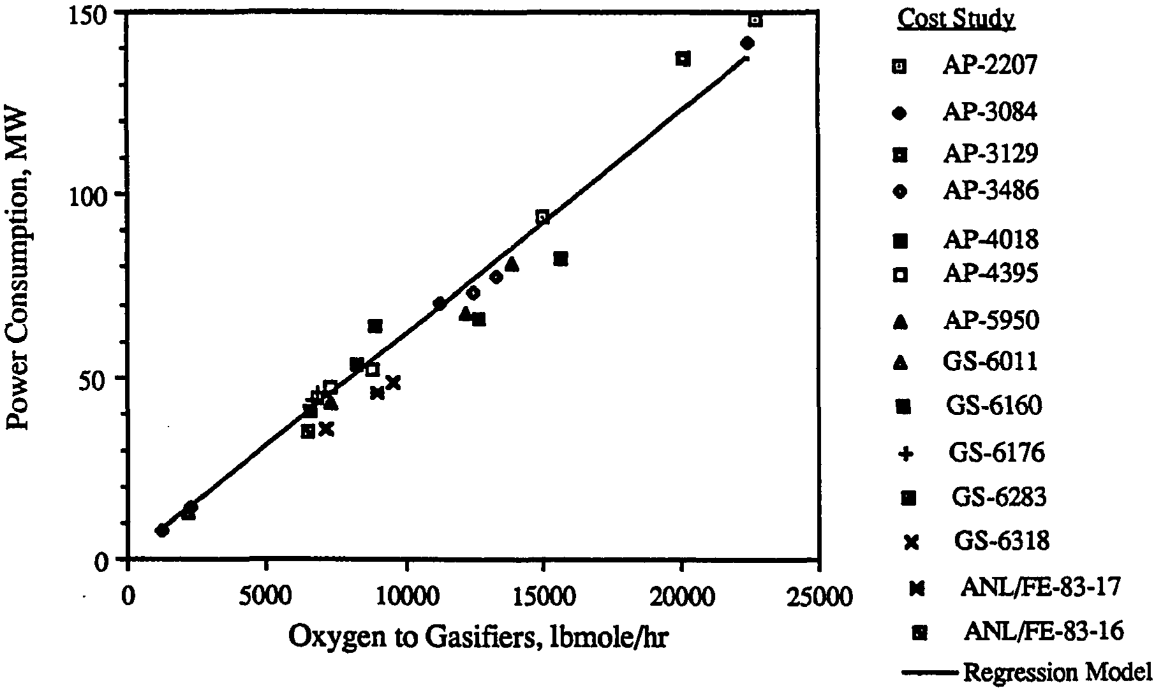

Figure A-53. Power Requirement for Air Separation Plant.

KRW Gasification (Cold Gas Cleanup)

$$W_{e,G} = 0.0476\, M_{syn,G,o} \qquad (R^2 = 0.8948;\ n = 7) \quad \text{(A-177)}$$

where,

$$35{,}400 \leq M_{syn,G,o} \leq 120{,}600 \text{ lbmole/hr.}$$

The standard error of the estimate is 520 kW. The regression model is shown graphically in Figure A-54.

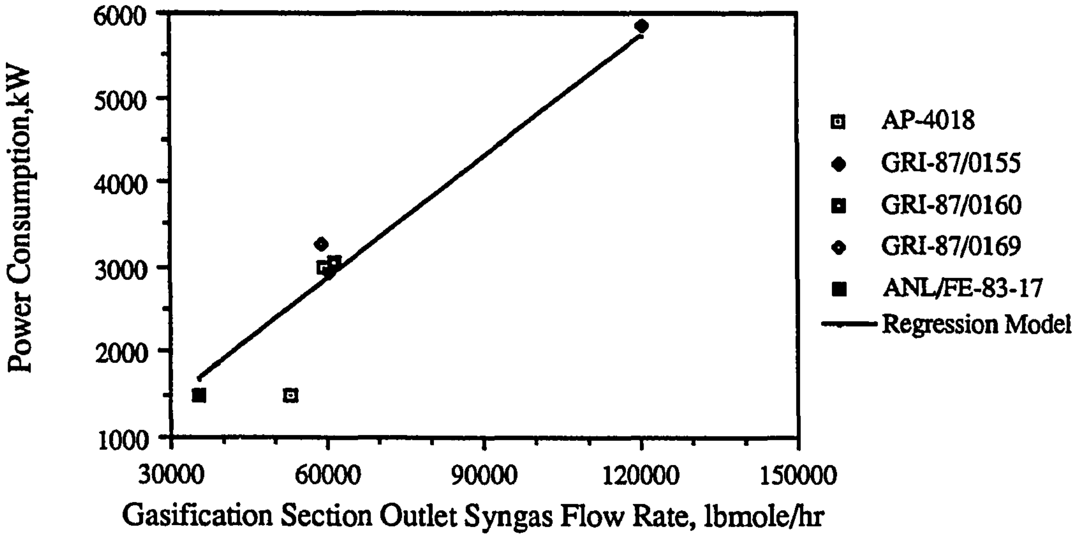

Figure A-54. Power Requirement for Gasification Section (KRW-based System with Cold Gas Cleanup).

Low Temperature Gas Cooling

$$W_{e,LT} = -5{,}600 + 0.0108\ M_{syn,LT,o} + 14.19\ P_{syn,LT,o} \quad (R^2 = 0.890;\ n = 10) \quad \text{(A-178)}$$

where,

$$32{,}300 \leq M_{syn,LT,o} \leq 74{,}400 \text{ lbmole/hr}$$

$$373 \leq P_{syn,LT,o} \leq 463 \text{ psia.}$$

For this plant section, a multivariate regression was found to yield significantly better results than a single-variate analysis. The standard error of the estimate is 240 kW. A comparison of reported and predicted power consumption is given in Figure A-55.

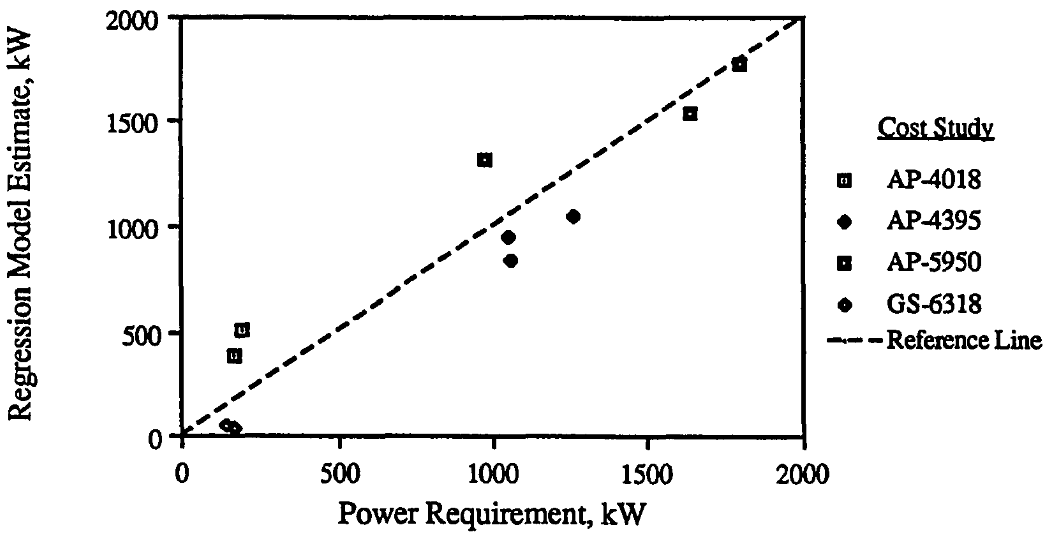

Figure A-55. Power Requirement for Low Temperature Fuel Gas Cooling Section.

Selexol Process

$$W_{e,S} = 348 + 0.478\left(M_{syn,S,i}\right)^{0.839} \qquad (R^2 = 0.881;\ n = 18) \qquad \text{(A-179)}$$

where,

$$4{,}000 \le M_{syn,S,i} \le 74{,}500 \text{ lbmole/hr.}$$

The standard error of the estimate is 550 kW. The regression model is shown graphically in Figure A-56.

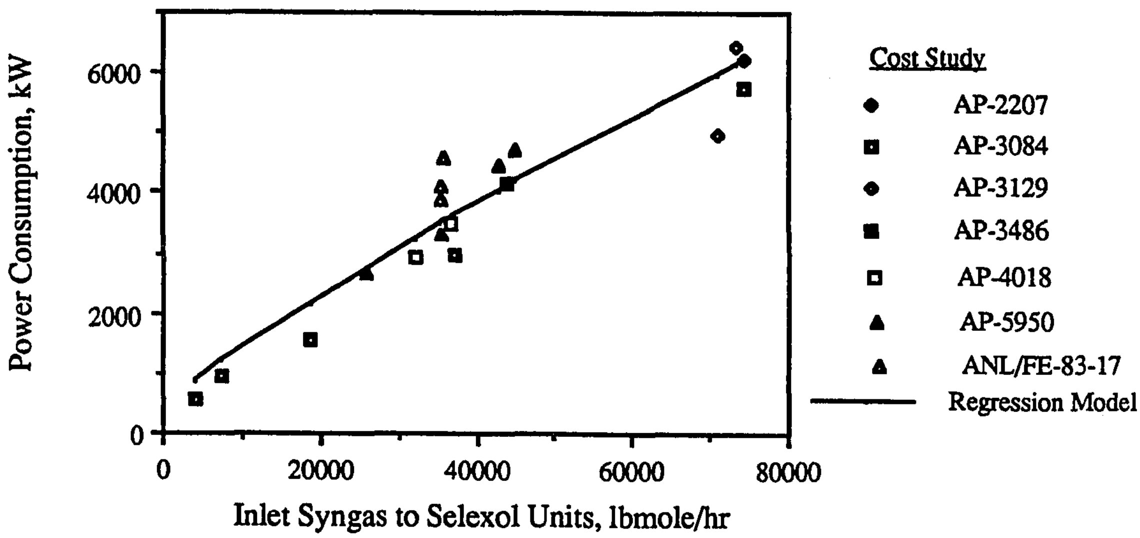

Figure A-56. Power Requirement for the Selexol Process.

Claus Plant

$$W_{e,C} = 0.021\ m_{s,C,o} \qquad (R^2 = 0.870;\ n = 20) \qquad \text{(A-180)}$$

where,

$$1{,}000 \leq m_{s,C,o} \leq 30{,}800 \text{ lb/hr.}$$

The standard error of the estimate is 67 kW. The regression model is shown graphically in Figure A-57.

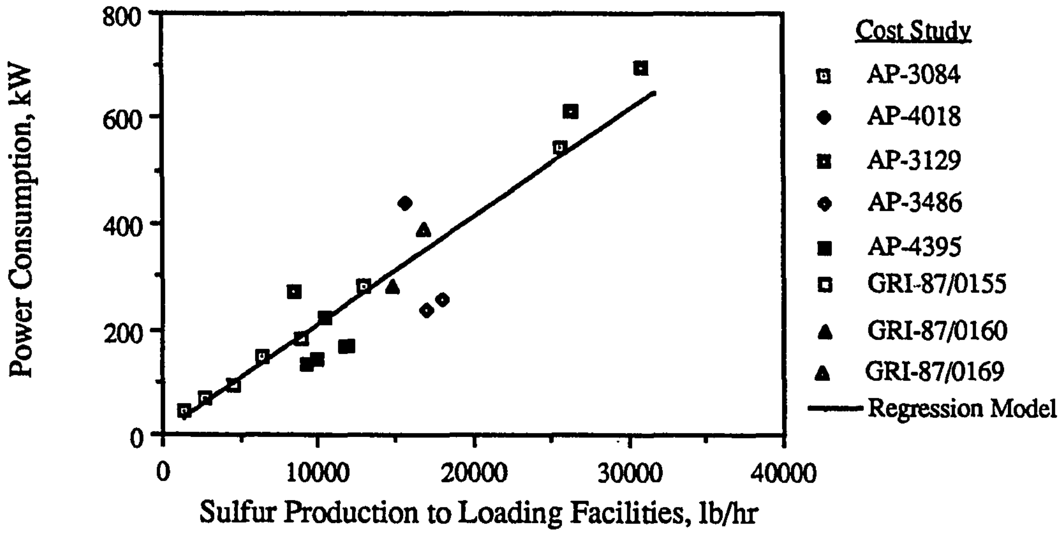

Figure A-57. Power Requirement for Two-Stage Claus Plants.

SCOT Process

$$W_{e,SC} = 2.4 + 0.0408\ m_{s,C,o} \qquad (R^2 = 1.00;\ n = 6) \qquad \text{(A-181)}$$

where,

$$9{,}000 \leq m_{s,C,o} \leq 18{,}000 \text{ lb/hr}$$

The standard error of the estimate for the data points included in the analysis is negligible. Four outlier data points were excluded from the regression analysis. Two of the excluded points were from the least recent study (Fluor, 1983b). The other two outliers are from a separate study (Fluor, 1985). Although one of these points is close to the regression line, the other point is an extreme outlier. The regression model is shown graphically in Figure A-58.

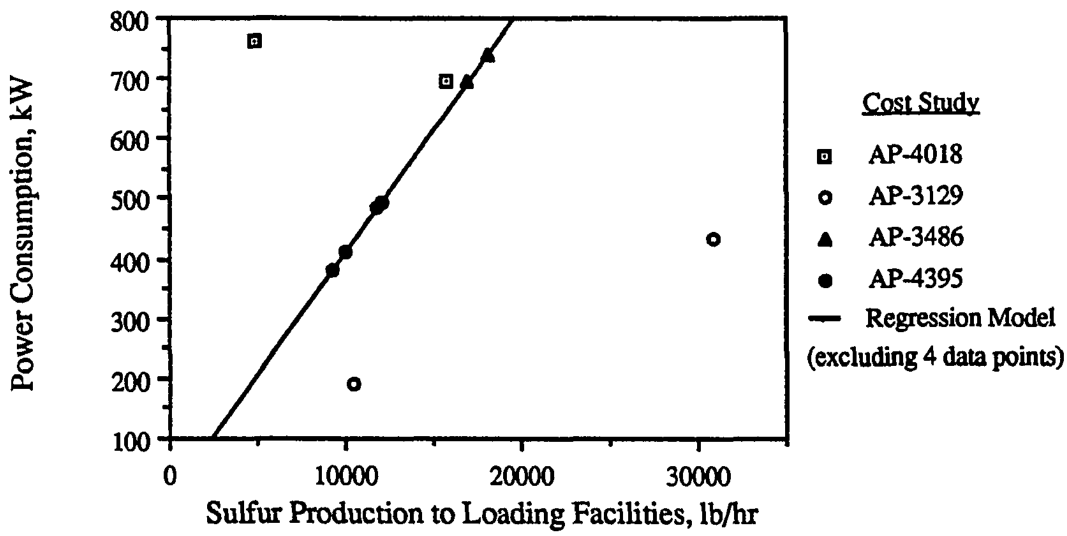

Figure A-58. Power Requirement for the SCOT Process.

<u>Beavon-Stretford Process</u>

$$W_{e,BS} = 44.5 + 1.12\, m_{s,BS,o} \qquad (R^2 = 0.990;\ n = 7) \qquad \text{(A-182)}$$

where,

$$100 \leq m_{s,BS,o} \leq 2{,}000 \text{ lb/hr}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-59.

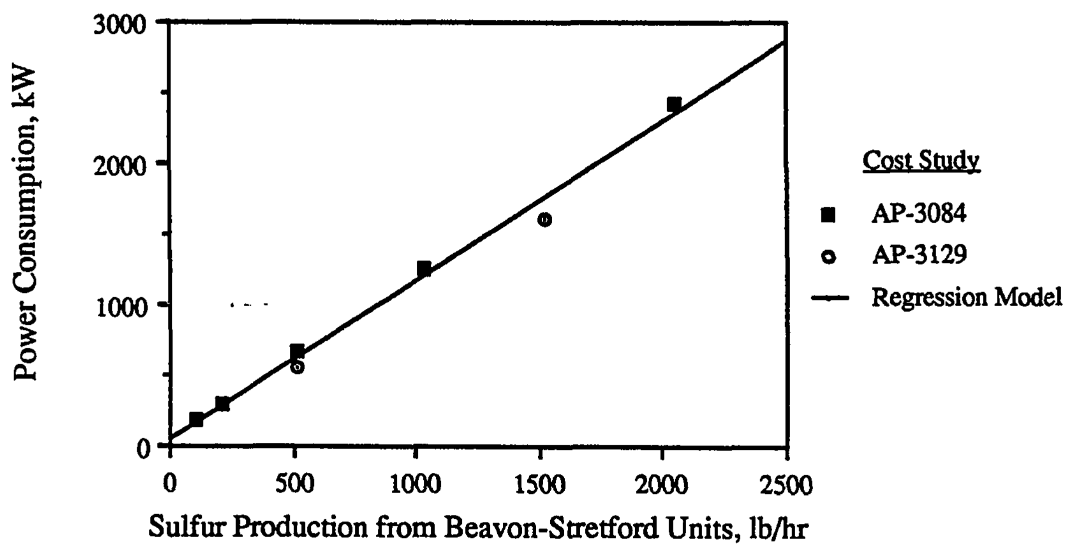

Figure A-59. Power Requirement for the Beavon-Stretford Process.

Boiler Feed Water Treating

$$W_{e,BF} = 20.8 + 2.13\times10^{-4} m_{pw} \qquad (R^2 = 0.975;\ n = 14) \qquad \text{(A-183)}$$

where,

$$234{,}000 \le m_{pw} \le 3{,}880{,}000 \text{ lb/hr.}$$

The standard error of the estimate is 38 kW. The regression model is shown graphically in Figure A-60.

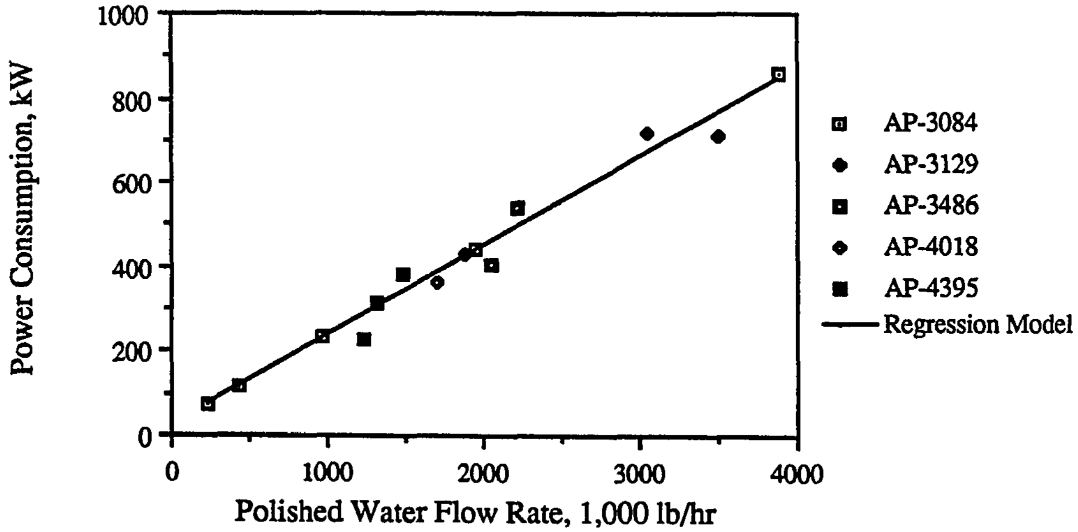

Figure A-60. Power Requirement for Boiler Feed Water Treating.

Process Condensate Treatment

$$W_{e,PC} = 7.34\times10^{-4} m_{sbd} \qquad (R^2 = 1.00;\ n = 3) \qquad \text{(A-184)}$$

where,

$$196{,}000 \le m_{sbd} \le 237{,}000 \text{ lb/r}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-61.

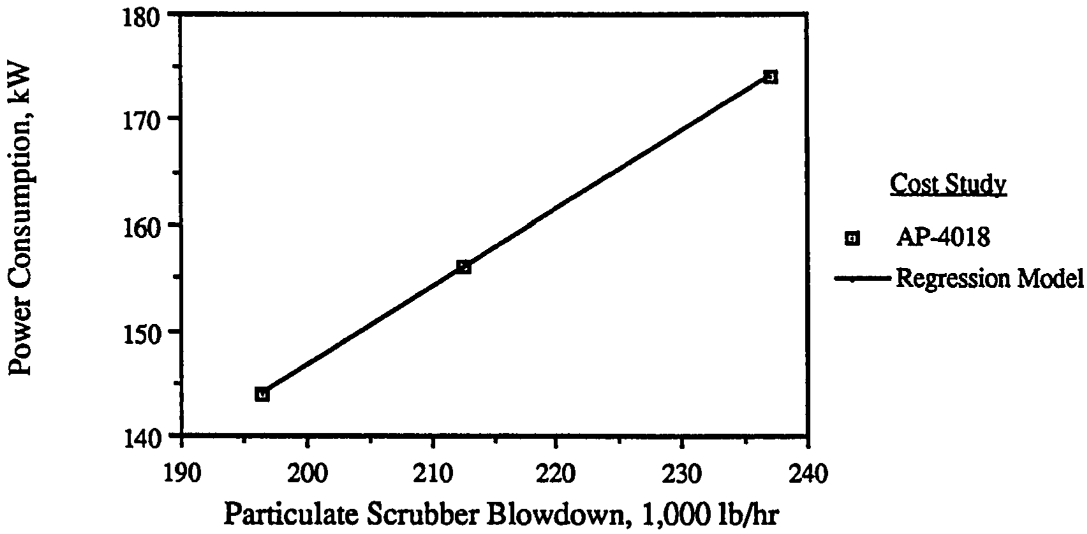

Figure A-61. Power Requirement for Process Condensate Treatment.

General Facilities

$$W_{e,GF} = 1{,}640 + 0.0877 \sum_{i \neq GF} W_{e,i} \qquad (R^2 = 0.999;\ n = 4) \qquad \text{(A-185)}$$

where,

$$58{,}000 \leq \sum_{i \neq GF} W_{e,i} \leq 82{,}000 \text{ kW}.$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-62.

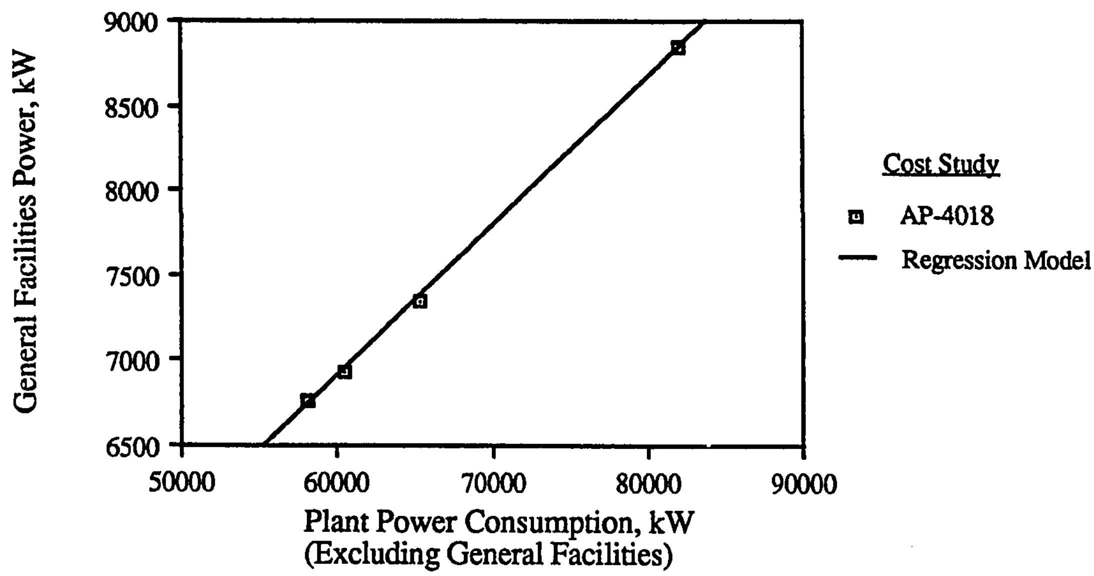

Figure A-62. Power Requirement for General Facilities.

Limestone Handling

$$W_{e,L} = -13.3 + 4.61\times10^{-3}\ m_{L,L,i} \qquad (R^2 = 1.00;\ n = 3) \qquad \text{(A-186)}$$

where,

$$5{,}000 \le m_{L,L,i} \le 104{,}000$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-186.

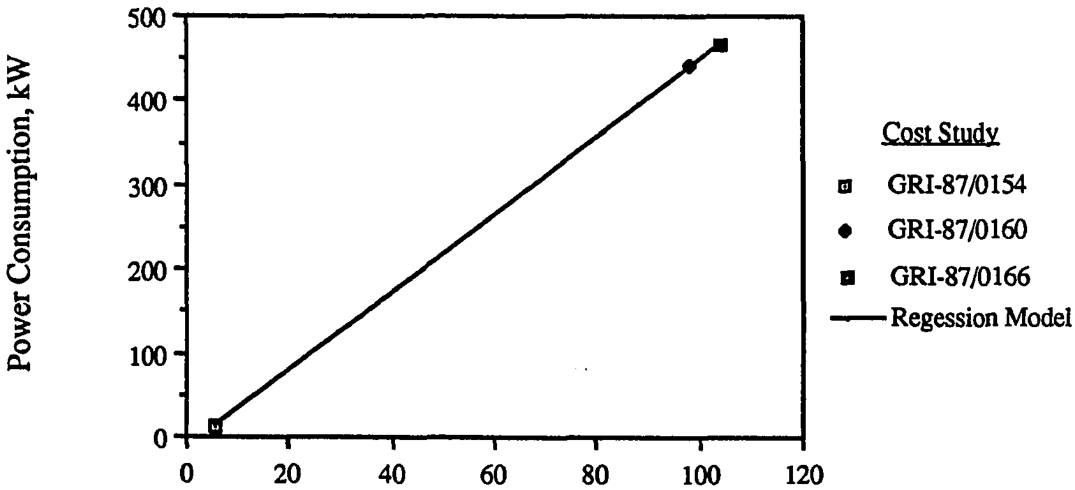

Figure A-63. Power Requirement for Limestone Handling.

KRW Gasification (Hot Gas Cleanup)

$$W_{e,G} = 58 + 8.3\times10^{-3}\ m_{cm,G,i} \qquad (R^2 = 1.00;\ n = 3) \qquad \text{(A-187)}$$

where,

$$30{,}000 \le m_{cm,G,i} \le 600{,}000\ \text{lb/hr.}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-64.

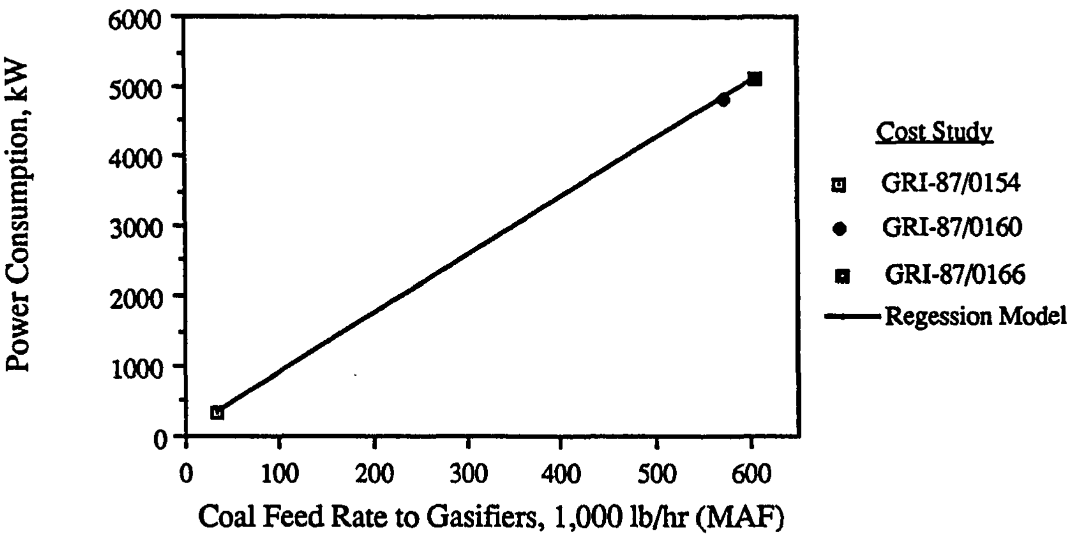

Figure A-64. Power Requirement for the Gasification Section (KRW System with Hot Gas Cleanup).

Sulfation

$$W_{e,SF} = -3.7 + 0.0102\,(m_{fines} + m_{ash}) \qquad (R^2 = 1.00;\ n = 3) \qquad \text{(A-188)}$$

where,

$$10{,}000 \leq (m_{ash} + m_{fines}) \leq 155{,}000 \text{ lb/hr.}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure A-65.

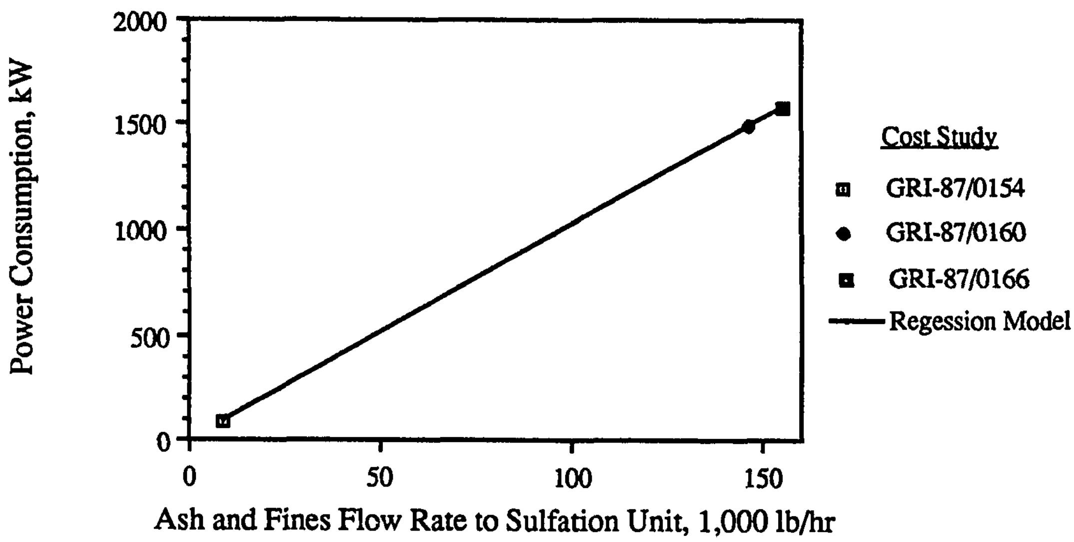

Figure A-65. Power Requirement for the Sulfation Process Area.

Lurgi Coal Handling

The power requirement for the Lurgi coal handling section consists of two components: coal preparation and coal receiving and storage. The coal preparation power requirement is associated with screening of the coal to separate fines prior to feed to the gasifier, as well as transport of the coal from storage to the gasifier coal surge bins. The coal preparation power requirement is given by:

$$W_{e,CH,prep} = 0.00133\ m_{cf,CH,i} \qquad (R^2 = 0.989;\ n = 10) \tag{A-189}$$

where,

$$675{,}000 \leq m_{cf,CH,i} \leq 3{,}380{,}000\ \text{lb/hr.}$$

The standard error of the estimate is 120 kW. The regression model is shown graphically in Figure A-66.

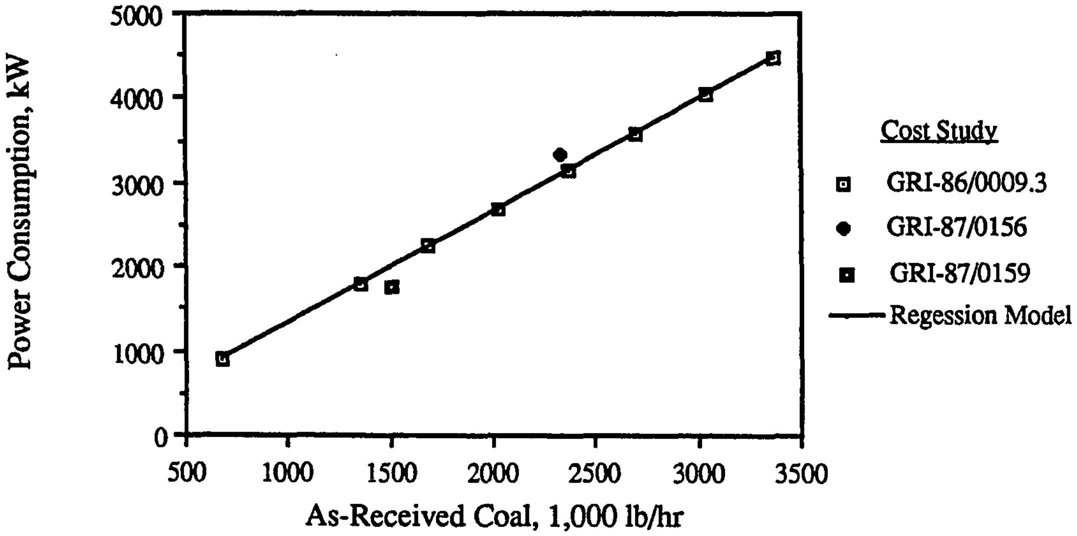

Figure A-66. Power Requirement for Lurgi Coal Handling.

The power requirement for coal receiving and storage was assumed to be directly proportional to the coal feed rate. One data point was available (Smelser, 1986a). The coal receiving and storage power requirement is given by:

$$W_{e,CH,R\&S} = 5.6\times10^{-4}\ m_{cf,CH,i} \tag{A-190}$$

The total power requirement for the coal handling section is the sum of the coal preparation and coal receiving and storage power requirements, and is given by:

$$W_{e,CH} = W_{e,CH,prep} + W_{e,CH,R\&S} \tag{A-191}$$

In Section A.5.1, it is noted that the study upon which the cost model of the Lurgi system is based assumes direct feed of all of the received coal, including fines, to the gasifier (Corman, 1986). Therefore, the power requirement given in Equation (A-191) may be an over-estimate. However, if problems occur as a result of fines in the gasifier, screening of the fines prior to coal feed would be required, and in that event Equation (A-191) is an appropriate model. The use of Equation (A-191) for the coal handling power requirement is therefore conservative.

Lurgi Gasification

The power requirement for Lurgi gasification was estimated by subtracting the coal receiving and storage power requirement estimated from Equation (A-190) and 50 percent of the coal preparation power requirement given in Equation (A-189) from the total power requirement for coal handling and gasification reported in Corman (1986). Only 50 percent of the coal preparation power estimate was used because the Corman study does not assume coal fines separation. The gasification section power requirement was assumed to be directly proportional to the coal feed rate, and is given by:

$$W_{e,G} = 0.005\ m_{cf,G,i} \tag{A-192}$$

Sulfuric Acid Plant

The power requirement for the sulfuric acid plant is assumed to be directly proportional to the inlet gas flow rate. Based on the estimate reported by SMC (1983), the power requirement is given by:

$$W_{e,SA} = 0.9\ M_{SA,i} \tag{A-193}$$

A.7.5 Representing Uncertainty in Operating Cost Estimates

Unlike capital cost estimates, there is no widely accepted notion of contingency costs for operating costs. However, uncertainties in operating costs can have a significant impact on the overall uncertainty in the cost of a new technology process plant. These uncertainties may include: potential difficulties with operation and control of new processes, requiring additional operating labor or more highly trained (and paid) operators; problems with fouling, corrosion, or other material degradation requiring additional maintenance; variability or uncertainty in the unit costs of catalysts and chemicals, especially for new materials that are not yet commercially available (e.g., zinc ferrite sorbent); uncertainty regarding the requirement for catalysts and chemicals due to uncertainty in process performance (e.g., stoichiometric ratio for calcium-to-sulfur for in-bed desulfurization); uncertainty in the availability of ash disposal sites and the cost of ash disposal; uncertainty regarding the availability of byproduct markets and the price of byproducts; and uncertainty or variability in fuel prices. These types of uncertainties can be explicitly characterized and quantified using probability distributions and a Monte Carlo (or similar) simulation technique, as discussed previously in Chapter 2.

The cost of the zinc ferrite sorbent initial charge and annual makeup requirement is a case in point. This sorbent has not been mass produced for commercial use. Therefore, in preparing cost estimates, cost estimators typically rely on the judgment of catalyst suppliers regarding the likely commercial cost of the sorbent. The sorbent unit cost is projected to fall within a range of values, but cannot be predicted with certainty. One study indicates that a major catalyst manufacturer has estimated the sorbent unit cost at between $3.00 and $3.50 per pound, while cost estimators at METC had been using a value of $2.50 per pound (Klett et al, 1986). If the unit cost of the sorbent is nominally estimated at $3.00 per pound, then the range of costs reported (from $2.50 to $3.50 per pound) represents 33 percent of the cost estimate. In another study (Banchik and Cover, 1988), the unit cost of zinc ferrite sorbent was assumed to vary from $4.00 to $10.00 per pound. Therefore, the reported range of unit costs from the open literature vary by a factor of four. If in addition the uncertainty in the *amount* of zinc ferrite sorbent required is considered, which results from uncertainty in the long term sulfur loading capability of the sorbent and the sorbent attrition rate, the overall uncertainty in the initial and annual total cost for zinc ferrite sorbent may be significantly greater than a factor of four. Yet these types of interactions between uncertainties in performance and cost parameters are not characterized in traditional deterministic cost estimates.

The unit cost of the zinc ferrite sorbent can be represented using a probability distribution. Similarly, the long term sorbent loading and the sorbent attrition rate can be represented using probability distributions. The probability distributions must be selected to represent judgments about the most likely values of these parameters, as well as the possible range in values that these parameters may assume. Using a probabilistic modeling approach, the effect of these uncertainties taken *simultaneously* on the total cost for zinc ferrite sorbent can be quantified. Also, a correlation analysis can be done to identify which uncertain model parameters are the most significant in determining the uncertainty in overall cost. Probabilistic analysis is therefore a valuable quantitative tool for identifying research priorities.

As for the capital cost models, an important application of the operating cost models for the selected IGCC systems will be the development of probabilistic estimates of IGCC system cost based on a disaggregated consideration of uncertainties in the parameters of the models. The development of the probabilistic parameter estimates, and application of these in specific case studies, is addressed in a later phase of the work described here.

A.8 TOTAL ANNUALIZED COST MODEL

The total annualized cost is the levelized annual revenue requirement required to cover all of the capital and operating costs for the economic life of the plant. For electric power plants, the total annualized cost is typically expressed as the cost of electricity. The total capital requirement, fixed operating cost, and variable operating cost are used to calculate the cost of producing electricity that is available for sale from the power plant, based on the net electrical output of the power plant. The net power output is the total power generated from the gas turbines and steam turbine less the total auxiliary power demand:

$$MW_{net} = MW_{GT} + MW_{ST} - 1{,}000 \sum_{i} W_{e,i} \qquad \text{(A-194)}$$

The cost of electricity in mills (one-thousands of a dollar) per net kWh is given by:

$$C_{elec} = \frac{\left[1{,}000\, f_{cr}\, TCR + f_{vclf}\,(FOC + VOC)\right]\left(\frac{1{,}000 \text{ mills}}{\text{dollar}}\right)}{MW_{net} \times 8{,}760 \times c_f} \qquad \text{(A-195)}$$

where,

f_{cr} = capital recovery factor

f_{vclf} = variable cost levelization factor

The numerator of Equation (A-195) is the total annual revenue requirement for the plant, and the denominator is the total net kilowatt-hours of electricity generated in a year. The total capital requirement from Equation (A-122) is in units of thousands of dollars. Therefore, a factor of 1,000 is used to convert the total capital requirement to units of dollars. The fixed operating cost from Equation (A-130) and the variable operating cost from Equation (A-168) are both in units of dollars. The annual revenue requirement shown in square brackets in the numerator is converted from dollars to mills, which is a more convenient unit for reporting the cost of electricity.

The capital recovery factor converts the capital cost into the equivalent levelized annual revenue required to provide a return to equity (stock) and debt (loan) financing sources, to pay for a portion of the principal, and to pay associated taxes and insurance (EPRI, 1987). The capital recovery factor, therefore, depends on the economic life of the plant, the type of financing used to supply the capital, and the applicable tax laws. A typical capital recovery factor for a 30 year life based on typical assumptions regarding financing and the federal Tax Reform Act of 1986 is 0.1034 (see EPRI, 1986). The variable cost levelization factor converts the annual costs into an equivalent levelized annual cost. If inflation and escalation are assumed to be zero, the variable cost levelization factor has a value of 1.0. The capital recovery and variable cost levelization factors are typically calculated using the standard method described by EPRI (1986).

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B.0 TECHNICAL BACKGROUND FOR PROCESS-RELATED UNCERTAINTIES

This appendix documents the basis for the estimates of uncertainty in model parameters used in case studies of five clean coal technologies. The estimates of uncertainty were developed for selected performance and cost parameters based on several approaches, depending on the availability of information. These approaches include: (1) review of published information indicating variability or uncertainty in specific parameters; (2) statistical analysis of data, typically as part of regression analysis; (3) elicitation of technical judgments from engineers involved in process development or process evaluation; and (4) judgment by the author based on discussions with technical experts, published information, data analysis, and engineering judgment. For more detail regarding the philosophy and approach to uncertainty analysis, the reader is referred to Chapters 2 and 4 of Volume 1.

The five clean coal technologies for which estimates of uncertainty have been developed are:

1) Pulverized coal (PC) power plant with flue gas desulfurization (FGD) for SO_2 control and selective catalytic reduction (SCR) for NO_x control;
2) PC power plant with the fluidized bed copper oxide process for simultaneous SO_2/NO_x control;
3) Oxygen-blown fluidized bed gasifier-based integrated gasification combined cycle (IGCC) system with cold gas cleanup;
4) Air-blown fluidized bed gasifier-based IGCC system with hot gas cleanup; and
5) Air-blown fixed bed gasifier-based IGCC system with hot gas cleanup.

These technologies are discussed in Chapter 3 of Volume 1.

The estimates of uncertainty in the PC power plant with FGD/SCR are adopted from a previous study (Rubin, Salmento, and Frey, 1988). The basis for the uncertainty estimates for this system, which is assumed as a baseline commercially available technology, are summarized in Section B.1 for the convenience of the reader.

The estimates of uncertainty in performance and cost parameters of the copper oxide process are discussed in Section B.2.

The three IGCC systems have several process areas which are common to more than one system. Therefore, rather than describe each IGCC system separately, the approach here is to separately document uncertainties in key process areas. These process areas are gasification (fixed bed and fluidized bed), fixed bed zinc ferrite desulfurization, and gas turbine and are described in Sections B.3, B.4, B.5, and B.6, respectively. Other IGCC uncertainties not included in these process areas are described in Section B.7.

For several IGCC process areas, a formal approach to eliciting judgments about uncertainties from technical experts was employed. These process areas included the fixed bed gasifier, zinc ferrite desulfurization, and gas turbine. For each of these process areas, a three part briefing package was developed and provided to each technical expert. The briefing package included: (1) a nine page introduction to uncertainty analysis; (2) a process area technical background paper ranging from 12 to 23 pages and citing 16 to 36 references, depending on the process area; and (3) a detailed questionnaire. Part 1 of the briefing package contained information presented in Chapter 2 of Volume 1. The technical discussion of each process area included in this appendix is the same as that in Part 2 of each briefing package. The questionnaires for the fixed bed gasifier, zinc ferrite desulfurization, and gas turbine process areas are given in Section B.8. The approach to the expert elicitations is also discussed in Chapter 4 of Volume 1.

References for all information discussed in this appendix are given in Section B.9.

B.1 Baseline Plant Design

The baseline power plant technology assumed in this study is a pulverized coal-fired power plant with conventional emission control technology. The development of assumptions regarding key plant performance and cost parameters for the conventional emission control system, as well as uncertainties in key parameters, has been previously reported. The assumptions are repeated here and summarized. However, the interested reader may want to refer to previous studies for more details (Rubin et al, 1986; Rubin, Salmento, and Frey, 1988).

Table B-1 summarizes the key design constraints assumed for the analysis, along with uncertainties for selected parameters. Table B-2 summarizes the properties of the coals selected for analysis. The assumed emission constraints for particulate matter and SO_2 are the current federal New Source Performance Standards (NSPS), which allow credit for sulfur removal through coal cleaning prior to combustion (Rubin et al, 1986). These constraints are an emission rate of 0.03 lb/MMBtu and typically 90 percent removal, respectively. The assumed NO_x emission limit, a 90 percent reduction from uncontrolled emissions, however, is nearly an order of magnitude below the current NSPS value, reflecting levels now required for coal-fired power plants in Germany and Japan (OECD, 1987). Higher NO_x emission reductions are also expected for U.S. coal-fired power plants in the next 10 years.

To meet environmental regulations, the pulverized coal-fired plant is assumed to use a conventional wet limestone flue gas desulfurization (FGD) system with forced oxidation for SO_2 control, a cold-side electrostatic precipitator (ESP) for particulate control, and a "high dust" selective catalytic reduction (SCR) system plus low-NO_x burners for NO_x control. Nominal design characteristics for these systems are summarized in Table B-3, along with the uncertainty ascribed to selected model parameters.

The principal uncertainties deal with design aspects of the FGD and SCR systems. For the wet FGD system, probabilities were assigned to three technical innovations currently found in Japanese and European designs, which could reduce the cost of conventional U.S. systems in the coming decade. These innovations are: (1) potential elimination of spare FGD absorber trains; (2) potential elimination of stack gas reheat; and (3) potential use of fewer numbers of large absorber vessels as opposed to current practice of using many small vessels.

For the SCR system, key uncertainties are the lifetime and cost of the SCR catalyst in U.S. plants. Because of the lack of commercial operating experience in the U.S., and because a number of potential problems are anticipated for high sulfur coal applications, the capital cost of SCR is given a positively skewed probability distribution to reflect the possibility of cost growth. However, a small probability has also been assigned that capital costs could be less than estimated due to experience gained in the early stages of commercialization.

Table B-1. Selected Input Parameter Assumptions for Case Studies

Model Parameter	Deterministic (Nominal) Value	Probability Distribution	Values (or σ as % of mean)
Emission Constraints			
Nitrogen Oxides	90% Reduction		
Sulfur Oxides	90% Reduction		
Particulates	0.03 lb/MMBtu		
Power Plant Parameters			
Gross Capacity	522 MW		
Gross Heat Rate	9500 Btu/kWh	-1/2 Normal	(1.8 %)
Capacity Factor	65 %	Normal	(7 %)
Excess Air (boiler/total)	20 %/39 %	Normal	(2.5 %)
Ash to Flue Gas	80 %		
Sulfur to Flue Gas	97.5 %		
Economizer Outlet Temp	700 °F		
Preheater Outlet Temp	300 °F		
Financial Parameters			
Inflation Rate	0 %		
Debt Fraction	50 %		
Common Stock Fraction	35 %		
Preferred Stock Fraction	15 %		
Real Return on Debt	4.6 %	Normal	(10 %)
Real Return on Com. Stock	8.7 %	Normal	(10 %)
Real Return on Pref. Stock	5.2 %	Normal	(10 %)
Federal Tax Rate	36.7 %		
State Tax Rate	2.0 %		
Ad Valorem Rate	2.0 %		
Investment Tax Credit	0 %		
Book Life	30 years		
Real Fuel Escalation	0 %	1/2 Normal	$\sigma = 0.06$ %

Table B-2. Selected Properties of Coals Used for Case Studies (As-Fired Basis)

Coal Property	Illinois No. 6 Coal		Pittsburgh Coal	
	Run-of-Mine	Washed[a]	Run-of-Mine	Washed[a]
Heating Value, Btu/lb	10,190	10,330	13,400	12,900
Sulfur, wt %	4.36	3.09	2.15	1.66
Carbon, wt %	57.0	57.7	74.8	72.1
Hydrogen, wt %	3.7	4.0	4.6	4.5
Oxygen, wt %	7.2	8.4	5.3	5.4
Nitrogen, wt %	1.1	1.1	1.4	1.3
Moisture, wt %	12.3	17.5	2.7	7.9
$/ton (at mine)	26.10	30.68	33.40	34.99
$/ton (transport)	7.90	7.90	7.90	7.90

[a] Model results for a 30 % sulfur reduction on a lb/MMBtu basis using conventional coal cleaning (Level 3 plant design)

Table B-3. Nominal Parameter Values and Uncertainties for the Conventional Environmental Control System

Model Parameter	Deterministic (Nominal) Value	Probability Distribution	Values (or σ as % of mean)[a]
Wet FGD System			
Molar Stoichiometry	(calc)	Normal	(5 %)
No. Operating Trains	4	Chance	10 % @ 1; 20 % @ 2; 40 % @ 3; 30 % @ 4
No. Spare Trains	1	Chance	75 % @ 0; 25 % @ 1
Reheat Energy	(calc)	Chance	75 % @ 0; 25 % @ x
Total Energy Use	(calc)	Normal	(10 %)
Limestone Cost	$15/ton	Uniform	$10-15/ton
Direct Capital Costs	(calc)	Normal	(10 %)
Operating Costs	(calc)	Normal	(10 %)
Selective Catalytic Reduction			
Space Velocity	2,850/hr	Normal	(10 %)
NH_3 Stoichiometry	(calc)	Normal	(10 %)
Catalyst Life	15,000 hrs	Chance	5 % @ 1,275 hrs 30 % @ 5,700 hrs 50 % @ 11,400 hrs 14 % @ 17,100 hrs 1 % @ 28,500 hrs
Energy Requirement	(calc)	Normal	(10 %)
Ammonia Cost	$150/ton	Uniform	$150-225/ton
Catalyst Cost	$460/ft^3	Normal	(7.5 %)
Direct Capital Cost	(calc)	Triangular	0.8x, x, 2x
Operating Cost (excl. Cat.)	(calc)	Normal	(10 %)
Cold-Side Electrostatic Precipitator			
Specific Collection Area	(calc)	Normal	(5 %)
Energy Requirement	(calc)	Normal	(10 %)
Total Capital Cost	(calc)	Normal	(10 %)
Operating Cost	(calc)	Normal	(10 %)
Solid Waste Disposal			
Land Cost	$6,500/acre	Normal	(10 %)
Direct Cost	(calc)	Normal	(10 %)
Operating Cost	(calc)	Normal	(10 %)

[a] For uniform distributions actual values are shown. For triangular distributions, endpoints and median are shown. For chance distributions, the probabilities of obtaining specific values are shown.

B.2 Advanced Pulverized Coal Plant Emission Control

The advanced emission control design for pulverized coal-fired power plants consists of the copper oxide process with integrated coal cleaning, byproduct recovery, and energy recovery via the power plant air preheater. The key inputs and distributions assigned to the copper oxide emission control system are summarized in Table B-4. Initially, uncertainties in key model parameters were developed based on an elicitation of the judgments of a DOE chemical engineer intimately involved in the testing and analysis of the copper oxide process concept, and the assumptions were documented by Frey (1987). A few of the design and economic assumptions were analogous to those for FGD systems, as described by Rubin, Salmento, and Frey (1988). However, more recent model development has lead to the creation of additional model parameters, many of which required probabilistic representation. The uncertainty in parameters in the current study that differ from those reported by Frey (1987) and Rubin, Salmento, and Frey (1988) are described here.

A probabilistic analysis requires input assumptions or data regarding the uncertainties in key process and economic parameters. For this analysis, the selection of parameters for probabilistic representation was based on a review of data, design studies, statistical analysis, and expert judgments by process developers and the author.

In previous modeling, the available copper to sulfur (Cu/S) molar ratio was calculated based on a regression equation, and an uncertainty was assigned to the calculated value based on experimental error. In the newly added sulfation model, uncertainties can be assigned to appropriate parameters used to calculate the Cu/S ratio. Experimental results have shown variability in the density of the expanded sorbent in the fluidized bed, with a nominal value of about 400 kg/m^3 and a coefficient of variation of about 10 percent (Yeh and Drummond, 1986). This variability was modeled using a normal distribution.

An important factor that affects the Cu/S ratio is the regeneration efficiency. Test results for the Cu/S ratio were based on low (e.g., 30 to 50 percent) regeneration efficiencies (Yeh et al, 1984). SMC estimated that a properly designed and sized regenerator, coupled with appropriate heating of the sorbent to reaction temperature, can result in regeneration efficiencies of over 99 percent at a 30 minute residence time (SMC, 1984). However, such a design target has not been achieved in any of the experimental work to date. Therefore, to characterize uncertainty in the regeneration efficiency that may be obtained in a full sclae commercial system, a negatively skewed probability distribution was assumed with a maximum value of 99.2 percent, representative of nominal expectations, with a small probability that the value could go below 50 percent, representative of actual experience to date. The negatively skewed distribution is qualitatively consistent with the notion of performance shortfalls that are characteristic of innovative chemical process plants (Merrow, Phillips, and Myers, 1981).

Of the remaining inputs to the sulfation reaction model, a few are uncertain or variable based on assumptions made regarding the base power plant. For example, uncertainty in the plant heat rate results in uncertainty in the flue gas molar flow rate. However, most of the inputs are design values, which are not appropriate candidates for probabilistic treatment; the effect of variation of these parameters is more appropriately addressed through conventional sensitivity analysis. The standard error of the estimated copper-to-sulfur molar ratio, which was discussed earlier in Chapter 3, is included as an uncertainty in the probabilistic model.

In experiments on a life cycle test unit, the sorbent attrition rate (another key parameter) was reduced to 0.13 weight percent of the sorbent circulation rate after modifications were made to the solids transport system (Williamson, Morici, and LaCosse, 1987). The test results indicated that solids transport was the primary source of sorbent attrition. However, significant improvements in the attrition rate were expected for a commercial process. The judgment of one process developer, elicited for this study, was that the attrition would nominally be 0.06 percent, but could have a 90 percent chance of being between 0.02 and 0.10 percent. This judgment formed the basis for the distribution in Table B-4.

Several new distributions were added to represent uncertainty in the capital cost estimates for specific equipment items. These distributions represent judgments that the costs of these equipment are likely to increase due to potential problems in design or materials. For example, UOP, Inc. had to modify the natural gas distributor plate, receiver hoppers, and vents for the regenerator vessel in the LCTU. The sorbent transport system experienced problems with air leakage (Williamson et al, 1987). In previous PETC work with the LCTU, the fluidized bed flue gas distributor plate had to be redesigned and replaced due to plugging with slag particulates, and a screen and slag knock-out leg were installed upstream of the distributor plate (Plantz et al, 1986). These changes are not included in the original cost estimate used as the basis for the cost model. Therefore, it is likely that the cost of a commercial scale unit will tend to be higher than the values used in preliminary cost estimates.

On a commercial scale, measures such as equipment additions or redesigns could lead to significantly higher costs. Results from a Rand study regarding cost growth in pioneer process plants indicate that solids handling systems pose the greatest difficulties in process design and operation (Milanese, 1987). In deterministic analyses, contingency factors normally are used to represent process risks. However, in probabilistic analyses, process contingency factors are supplanted by directly specifying uncertainties as probability distributions in model parameters affecting cost. To represent the uncertainty in current estimates of capital cost, the capital costs for each major equipment area were assigned uniform probability distributions, with the current estimates at the low end of the range. The high end of the ranges represented probabilities that the actual capital cost might increase by up to 50 percent for most process areas and up to 100 percent for the solids transport system. These uncertainties were intended to be representative of the costs likely to be found in a commercial system (e.g., based on a fifth-of-a-kind plant).

Finally, a number of the distributions for reagent costs were revised, notably methane and sulfuric acid. The cost of ammonia is assumed to be \$150/ton, with a probability of an increase up to \$225/ton. The cost of methane is assumed to be \$4.50 per thousand cubic feet, with an increase to \$6.70 by 2,010 based on an EPRI estimate (EPRI, 1986). The distribution for methane cost is therefore positively skewed with a significant probability of the cost reaching \$6.70. The sale price of sulfuric acid is assumed to have a maximum of about \$40/ton. The distribution for sulfuric acid price is negatively skewed with about a one percent chance that the cost will go below \$7/ton (based on Burns and Roe Services Corp., 1987). Implicit in this assumption is that the extent of sulfuric acid byproduct recovery by flue gas treatment processes will not exceed the available market capacities. The distribution for elemental sulfur sale price is taken to have a maximum value of about \$125/ton, with a negative skewness, to be equivalent to the sulfuric acid cost distribution.

Table B-4. Nominal Parameter Values and Uncertainties for the Advanced Environmental Control System

Model Parameter	Deterministic (Nominal) Value	Probability Distribution	Values (or σ as % of mean)[a]
Copper Oxide Process[b]			
Fluidized Bed Height	48 inches		
Sorbent Copper Loading	7 wt-%		
Regeneration Efficiency	99.2 %	-1/2 Normal	(20 %)
Fluidized Sorbent Density	400 kg/m^3	Normal	(10 %)
Standard Error, Cu/S Ratio	0	Normal	$\sigma = 0.39$
Sorbent Attrition	0.06 %	Normal	(41 %)
Ammonia Stoichiometry	(calc)	Normal	(6.25 %)
Regeneration Temp	900 ^oF	Normal	(2 %)
No. Operating Trains	4	Chance	10 % @ 1; 20 % @ 2; 40 % @ 3; 30 % @ 4
No. Spare Trains	1	Chance	50 % @ 0; 50 % @ 1
Sorbent Cost	$5.00/lb	-1/2 Normal	(25 %)
Methane Cost	$4.50/mscf	1/2 Normal	(25 %)
Ammonia Cost	$150/ton	Uniform	$150-225/ton
Sulfuric Acid Cost	$40/ton	-1/2 Normal	(30 %)
Sulfur Cost	$125/ton	-1/2 Normal	(30 %)
Absorber Direct Cap. Cost	(calc)	Uniform	1.0x - 1.5x
Solids Heater DCC	(calc)	Uniform	1.0x - 1.5x
Regenerator DCC	(calc)	Uniform	1.0x - 1.5x
Solids Transport DCC	(calc)	Uniform	1.0x - 2.0x
Sulfur Recovery DCC	(calc)	Uniform	1.0x - 1.2x
Total Capital Cost	(calc)	1/2 Normal	(10 %)
Fabric Filter			
Air-to-Cloth Ratio	2.0 $acfm/ft^2$	-1/2 Normal	(10 %)
Bag Life	(calc)	Normal	(25 %)
Energy Requirement	(calc)	Normal	(10 %)
Bag Cost	$0.80/$ft^2$	Normal	(5 %)
Operating Cost	(calc)	Normal	(15 %)
Total Capital Cost	(calc)	Normal	(15 %)
Solid Waste Disposal			
Land Cost	$6,500/acre	Normal	(10 %)
Direct Cost	(calc)	Normal	(10 %)
Operating Cost	(calc)	Normal	(10 %)

[a] For uniform distributions actual values are shown. For triangular distributions, endpoints and median are shown. For chance distributions, the probabilities of obtaining specific values are shown.

[b] As part of integration of the copper oxide process with the base power plant, the plant air preheater is resized to maintain an exit flue gas temperature of 300 ^oF.

B.3 Fixed-Bed Coal Gasification Process Area

For this process area, technical judgments for uncertainties in performance parameters were elicited from two engineers at the U.S. Department of Energy Morgantown Energy Technology Center (DOE/METC). These engineers were provided with a three-part briefing packet, as discussed in Chapter 4. Part One of the briefing packet was an introduction to uncertainty analysis. Part Two was a review of published information about the fixed bed gasifier process area, all of which is included in Sections 3.1 to 3.3. Part Three was a questionnaire to elicit technical judgments about uncertainties, which is included in Section B.8.1. The responses of the two experts to the questionnaire are summarized in Section B.3.4.

B.3.1 Process Description

The system of interest in this study is a fixed-bed dry-ash Lurgi gasifier as part of a "simplified" IGCC system. This system is shown in Figure B-1 and is based on the design reported by Corman (1986). This technology is discussed in Chapter 3 of Volume 1. The gasifier is assumed to operate in air-blown mode using a caking Illinois No. 6 coal with up to 30 percent coal feed as minus 1/4 inch fines. The coal gas from the gasifier passes through a high efficiency cyclone, which captures most of any coal fines entrained in the coal gas. The captured fines are agglomerated and recycled back to the gasifier. The coal gas exiting the cyclone enters a fixed bed zinc ferrite desulfurization system for removal of H_2S. Regenerated sulfur is recovered as sulfuric acid. The desulfurized coal gas then enters a secondary high efficiency cyclone for removal of entrained sorbent fines and residual particles from the gasifier. The cleaned coal gas is then fed to a gas turbine combined cycle system for power generation. Air is extracted from the gas turbine compressor discharge for use in the gasifier. Gasification steam is obtained from the plant steam cycle.

The Lurgi fixed-bed dry-ash gasification process was developed in the early 1930's in Germany. The first commercial plant was built in 1936. Since then, over 160 Lurgi gasifiers have been built worldwide. Nearly all of these have been oxygen-blown systems operating on lignite or subbituminous coals (Simbeck et al, 1983). Fixed-bed gasifiers have a limited ability to handle coal feed in the form of fines. Lurgi gasifiers have generally been operated with sized coals (i.e. between 2 and 1/4 inches) with no more than about 10 percent of the coal feed as minus 1/4 inch fines. The standard commercial offering for the Lurgi technology is the Mark IV gasifier, which has a working diameter of 12.7 feet (Corman, 1986).

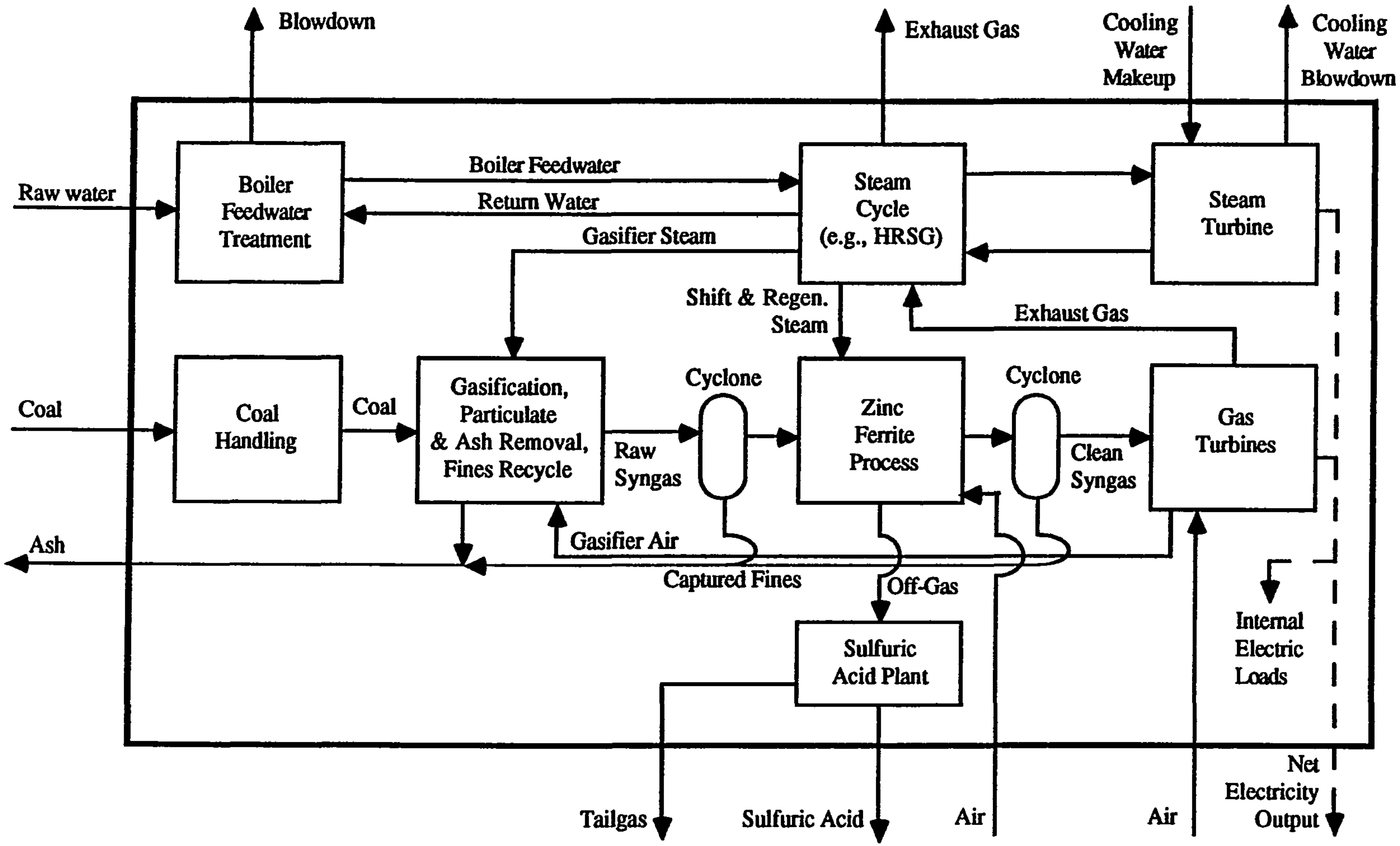

Figure B-1. Schematic of Air-Blown Dry-Ash Fixed-Bed Gasifier IGCC System with Hot Gas Cleanup

A schematic of the Lurgi gasifier is shown in Figure B-2. Coal from a lockhopper enters the top of the gasifier vessel, where it is distributed onto the fixed bed by a rotating grate. The coal proceeds downward through four zones in the gasifier bed: drying, devolatilization, gasification, and combustion. Steam and oxidant (oxygen or air) are introduced at the bottom of the gasifier, and flow upward countercurrently to the coal flow. In the top-most drying zone, surface and some inherent moisture on the coal is evaporated into the exiting gas stream. The temperature of the exiting raw coal gas is typically 1,000 to 1,100 °F for high rank coals and can be as low as 400 to 600 °F for high-moisture low rank coals. In the devolatilization zone, volatile matter is released into the exiting gas stream. The volatile matter typically includes tars, oils, phenols and hydrocarbon gases. The devolatilization products exit with the coal gas. By comparison, in gasifiers with high exit gas temperatures and different bed designs, the devolatilization products would be cracked or reacted with oxygen to form hydrogen, carbon monoxide, or other compounds.

The "cold gas efficiency" of the Lurgi gasifier is approximately 90 percent, if the heating value of hydrocarbon liquids are included. In systems featuring cold gas cleanup, these liquids are condensed out of the fuel gas. In hot gas cleanup systems, these liquids exist in the vapor phase and are contained in the fuel gas delivered to the gas turbine. In the gasification zone, gasification of fixed carbon from the coal with steam and oxidant occurs. The overall gasification reaction is endothermic. Below the gasification zone, combustion of remaining char occurs. The heat released from combustion is required to supply heat required for the endothermic gasification reactions and to raise the reactants to gasification temperature. The resulting ash from combustion falls through a rotating grate. The incoming steam and oxidant cool the ash as it leaves the gasifier vessel. The ash is collected at the bottom of the gasifier in a lockhopper.

The gasifier has a high steam requirement to maintain temperatures in the combustion zone below the coal ash fusion temperature and to avoid slagging or the formation of clinkers, which would plug the gasifier. For an Illinois No. 6 coal, the ash fusion temperature is approximately 2,300 °F. Steam is used a thermal diluent for the purpose of temperature control. As a result, the total steam requirement to the gasifier exceeds the reactant requirements for coal gasification, and the exiting raw coal gas contains a large fraction of water vapor. For air-blown systems, the steam requirements are less than for oxygen-blown systems due to the increased thermal capacity of the oxidant.

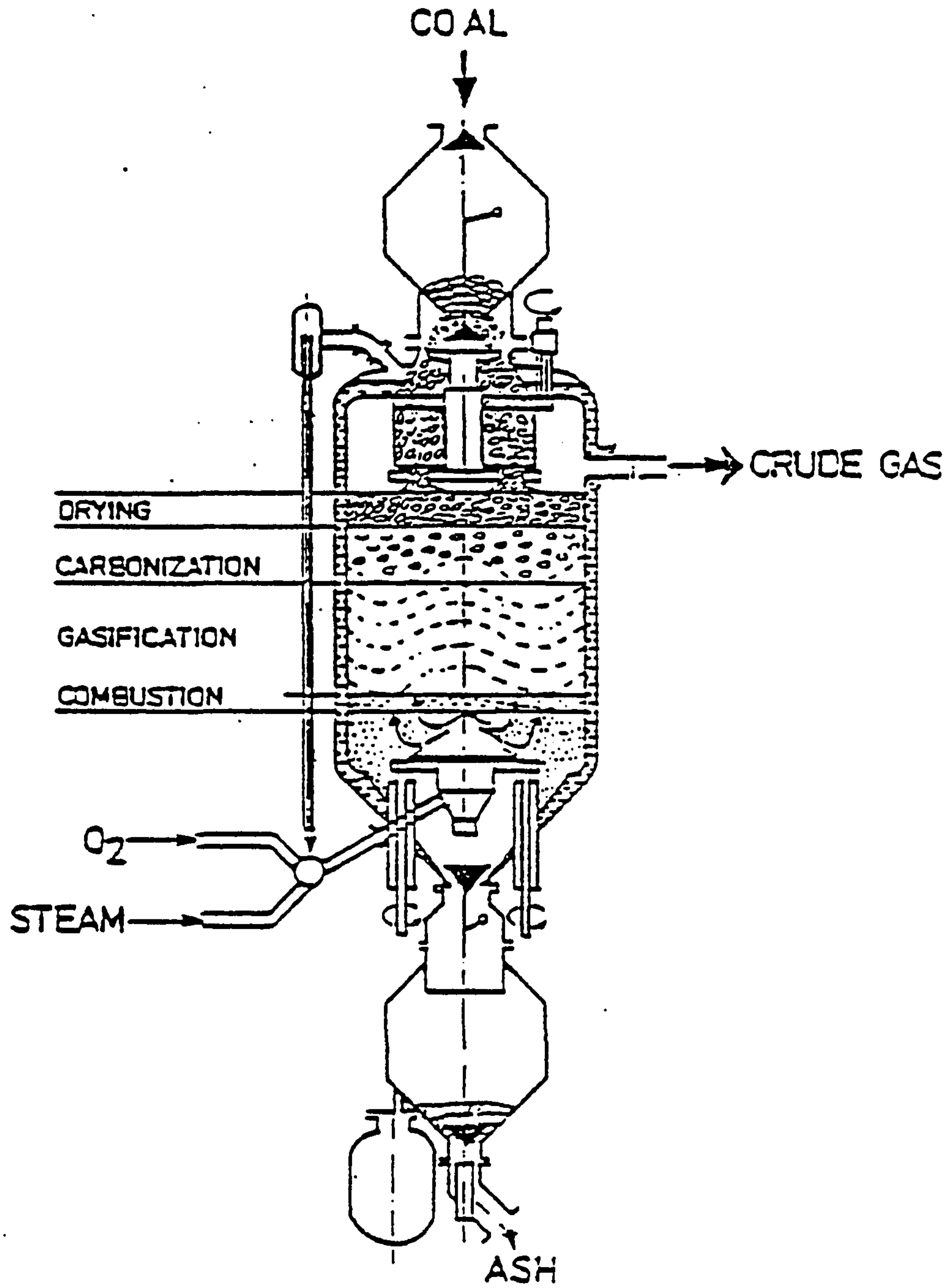

Figure B-2. Diagram of Lurgi Gasifier (Source: Zahnstecher, 1984).

The Lurgi gasifier has a relative low oxidant requirement, due to the excellent heat transfer characteristics of the countercurrent flow of coal and reactants. The excellent heat transfer characteristics presumably reduce the amount of char that must be combusted compared to other gasifier designs to provide heat for the gasification reactions. The reduction in the amount of coal associated with combustion thus reduces the amount of oxygen needed for the combustion reactions.

The advantages of the Lurgi gasifier for simplified IGCC application, compared to other gasifier designs, are (Simbeck et al, 1983; Zahnstecher, 1984; Corman, 1986; Ghate and Longanbach, 1988):

- Stability of operation due to a large bed inventory
- Substantial record of operating experience
- Relatively low oxidant requirements
- High methane yield
- Low gas exit temperature
- High H_2 to CO ratio
- High carbon conversion
- High cold gas efficiency
- Excellent heat transfer within the gasifier
- Good turndown capability and operability

Some of these items are related. For example, high cold gas efficiency implies high carbon conversion. However, the reverse is not true. For example, if the proportion of coal that is combusted is large compared to the amount that is gasified, the carbon conversion would be high, but the resulting gas would have a very low heating value and the cold gas efficiency would be low. Excellent heat transfer within the gasifier facilitates a high cold gas efficiency by reducing the combustion requirements for supplying heat to the gasification zone. The reduced combustion requirement also reduces the amount of oxidant needed, since combustion consumes a larger amount of oxygen than does gasification.

Some potential disadvantages include:

- High steam requirement
- Limited ability to handle coal fines
- Must be modified to handle caking coals
- Long solids residence time results in relatively low capacity
- Product gases contain tars and heavy hydrocarbons

B.3.2 Key Design and Performance Assumptions

The performance and cost case study for this research features air-blown gasification of a caking Illinois No. 6 coal containing up to 30 percent fines (minus 1/4 inch). Table B-5 summarize typical design and performance assumptions

Table B-5. Typical Default Assumptions for the Lurgi-based IGCC System Case Studies[a]

Description	Value
<u>Coal Properties</u>	
Proximate Analysis, wt-%, run-of-mine basis	
Moisture	12.0
Fixed Carbon	47.8
Volatile Matter	31.4
Ash	8.8
Ultimate Analysis, wt-%, dry basis	
Carbon	69.53
Hydrogen	5.33
Nitrogen	1.25
Chlorine	0.00
Sulfur	3.86
Oxygen	10.03
Ash	10.00
Ash Fusion Temperature, ^{o}F	2,300
<u>Gasifier Operating Parameters</u>	
Pressure, psia	308
Coal Throughput, st/day per gasifier MAF basis	480
Air/Coal Ratio, lb air/lb MAF coal	3.1
Oxygen/Carbon Ratio, lb oxygen/lb carbon (oxidant feed only)	0.91
Oxygen/Carbon Ratio, lb oxygen/lb carbon (including oxygen in coal)	1.06
Steam/Air Ratio, lb steam/lb air	0.60
Steam/Coal Ratio, lb steam/lb MAF coal	1.8
Raw Gas/Coal ratio, lb raw gas/lb ROM coal	4.77
Inlet Air Temperature, ^{o}F	800
Inlet Steam Temperature, ^{o}F	618
Fines Carryover, wt-% of coal feed	4
Fines Recycle, wt-% of carryover	98
Fines Carbon Content, wt-%	90
Gasifier Bottom Ash Carbon Retention, wt-% of coal carbon	0.5
Gasifier Bottom Ash Sulfur Retention, wt-% of coal sulfur	3.0

<u>Fuel Gas Compositions</u>

Molar Composition, vol-%	Gasifier Exit	Gas Turbine Inlet[b]
H_2	12.1	17.9
CO	11.0	5.1
CH_4	3.3	3.3
CO_2	8.6	14.6
N_2	31.4	31.4
H_2O	32.8	27.5
H_2S	0.5	10 ppmv
NH_3	0.2	0.2
Tar	200 ppmv	200 ppmv
Phenol	400 ppmv	400 ppmv
Naphtha	500 ppmv	500 ppmv

[a] Based on typical assumptions and results from current ASPEN simulation model.

[b] Water-gas shift reaction is assumed to be in equilibrium just upstream of zinc ferrite reactor.

used in the current model of this system. These assumptions include coal characteristics, key gasifier operating parameters, and raw coal gas composition. Unlike conventional designs in which the raw coal gas is water-quenched, in the simplified IGCC system concept, the raw coal gas is fed directly to the desulfurization and power trains. In the current ASPEN simulation model of the Lurgi-based IGCC system, the water-gas shift reaction is assumed to reach equilibrium in the zinc ferrite reactor vessel, which increases the H_2/CO ratio of the fuel gas to the gas turbine compared to the raw coal gas exiting the gasifier.

B.3.3 Key Technical Issues

The key technical issues which may affect the design and performance of a commercial simplified IGCC system include:

- Modifications for caking coal
- Fines loading/fines carryover
- Carbon conversion
- Sulfur remaining in bottom ash
- Gasifier throughput
- Ammonia production
- Gasifier design pressure
- Gasifier steam requirement
- Gasifier oxidant requirement

Each of these issues will be discussed in turn.

B.3.3.1 Caking Coal Operation

The conventional Lurgi gasifier requires a non-caking sized coal (between 1/4 and 2 inches). Caking coals can cause plugging of the bed due to the formation of agglomerates, which reduces the efficiency of the overall gasification conversion process. Zahnstecher (1984) reported that the conventional Lurgi gasifier is limited to non-caking coals with a free swelling index (FSI) less than 1.0; however, full scale tests using a Mark IV gasifier indicated that bituminous coals with a FSI of 3-4 can be gasified if a water-cooled stirrer is used in the char formation (devolatilization) zone to prevent plugging of the bed. Corman (1986) also reports that commercial tests by Lurgi on both Illinois and Pittsburgh coals with "fixed-position" stirrers (i.e., the vertical location of the stirrer in the gasifier vessel is fixed) indicated feasible operation.

The General Electric Process Evaluation Facility features a fixed-bed gasifier with a vertically-movable stirrer. This gasifier has been operated in air-blown mode with an Illinois No. 6 coal. The 42-inch diameter METC fixed-bed gasifier also features a deep bed stirrer for caking coals (Ghate and Longanbach, 1988). Corman (1986) reports that both

GE and METC tests have demonstrated successful operation on bituminous coal with FSI up to 8-1/2 using deep bed stirring. Corman also indicates that little, if any, capacity derate for the Mark IV gasifier is anticipated when operated with an Illinois No. 6 coal with a FSI of 5-6.

For a cold startup, a noncaking coal is required to properly establish the gasification and devolatilization zones within the fixed bed and to limit the caking region to an area accessible to the stirrers (Corman, 1986). It is unclear from this statement if fixed-position or vertically-movable stirrers are expected for the Mark IV gasifier. The cost of the stirrer and related components is presumably included in the cost estimate reported by Corman.

B.3.3.2 Fines Loading and Carryover

In the conventional Lurgi gasifier, sized coal is required. Excessive fines in the feed coal will lead to entrainment of fines in the coal gas. As a result, commercial Lurgi gasifiers are typically operated with coal feed containing less than 10 weight percent of minus 1/4 inch coal (Simbeck et al, 1983).

Conceptual designs of conventional Lurgi gasifier systems typically assume that sized coal is delivered to the gasifier. The typical size ranges are 1/4-inch to 2-inch or 1/4-inch to 4-inch lumps of coal. For example, Zahnstecher (1984) assumes a nominal coal size of 1/4 to 4 inches, with only three percent of the coal feed less than 1/4-inch. Smelser (1986a) assumes that lignite coal feed from 1/4 to 2 inches in size is used, and this represents perhaps the most typical coal size assumption (Simbeck et al, 1983).

To accommodate the limitations of the Lurgi gasifier to handle minus 1/4 coal fines, four general approaches for utilizing coal in Lurgi-based plants have been assumed in different studies . These include: (1) separation and return of all or a portion of the fines to the coal supplier (e.g., Cover et al, 1985; Smelser, 1986a and 1986b); (2) separation of sized coal for use in Lurgi gasifiers and fines for use in other types of gasifiers (e.g., KRW) or for combustion in conventional boilers (e.g., Cover et al, 1985; Smelser, 1986a and 1986b); (3) separation of a portion of the fines for injection directly into the gasifier bed via "tuyeres" in the case of the slagging BGC/Lurgi gasifier (e.g., Bechtel, 1983); and (4) agglomeration of separated coal fines into briquettes for feed to the gasifier coal surge bin (e.g., Parsons, 1985; Bechtel et al, 1988).

The assumption in the study by Corman (1986) is that the coal is received as minus 2-inch size, and that all of the coal, including fines, is fed directly to the gasifier. Only the fines that are captured from the raw coal gas in a primary cyclone are agglomerated into briquettes using bentonite as a binder. The agglomerated fines are then recycled to the coal surge bin.

For a caking Illinois No. 6 coal, General Electric assumed that up to 30 weight percent of the coal feed could be fed as minus 1/4 inch fines. Lurgi indicated that this amount could be handled if the coal is caking and if the minus 1/8 fraction of the material met certain size criteria which are not reported. Lurgi also indicated that the 30 percent fines loading is a maximum for caking coal. Because caking characteristics of bituminous coal tend to suppress fines carryover due to natural agglomeration, it is expected that operation on a non-caking coal would require a reduction in the fines loading to keep fines carryover at reasonable levels (Corman, 1986).

General Electric estimated that the fines carryover for a Lurgi gasifier operating on Illinois No. 6 coal with 30 weight percent feed as minus 1/4 fines would be 2-4 percent of the coal feed, and that it could be as high as 5 percent. In contrast, if a fully screened coal were used (1/4 to 2 inch) the carryover would be under one percent.

Corman (1986) states that the fines removed by the primary cyclone are expected to contain up to 90 percent carbon as char. However, if there are fines which are entrained prior to drying or devolatilization, such fines would have a carbon content more similar to that of the feed coal.

General Electric proposes the use of a "high efficiency" cyclone for fines capture. Corman (1986) reports that pilot scale tests of such a cyclone indicated over 98 percent collection of the fines leaving the gasifier. Smith et al (1987) report the same result, based on a 200 hour test using a cyclone designed by GE Environmental Systems, Inc. for operation at 300 psig and 1,200 ^oF.

A General Electric cyclone used in a different application achieved lower removal efficiencies, in the range of 92 to 98 percent. The application was for a KRW gasifier, the carryover of which may have a different particle size distribution than that of the Lurgi gasifier. The performance of this cyclone was below the levels predicted by the manufacturer (Haldipur et al, 1988).

Cincotta (1984) performed a similar evaluation of an IGCC system featuring oxygen-blown slagging BGC/Lurgi gasifiers. This study also assumed the use of a primary cyclone for fines capture. When using 1/4 to 2 inch sized coal, Cincotta estimates a fines carryover equivalent to 0.91 percent of the coal feed, while Lurgi estimated the carryover at 0.5 to 0.8 percent. Cincotta assumed that the carryover contained 90 percent coal fines and 10 percent ash, while Lurgi indicated that the carryover would have about the same composition as the feed coal and partly the composition of char formed from the feed coal.

Any fines not captured or not recycled to the gasifier represents a loss of carbon in the gasification process and affects the gasification efficiency.

B.3.3.3 Carbon Conversion

The carbon conversion in the Lurgi gasifier is typically believed to be around 99 percent. However, the assumptions regarding carbon retention in the gasifier bottom ash are usually not reported in most design studies. Simbeck et al (1983) show a typical mass and energy balance for an oxygen-blown Lurgi gasifier using Illinois No. 6 coal which indicates that 0.64 percent of the carbon in the feed coal is contained in the bottom ash. Zahnstecher (1984) uses a carbon conversion rate of 99.9 percent, implying that 0.1 percent of the carbon is contained in the bottom ash. Smelser (1986a) uses a carbon conversion of 99.45 percent, implying that 0.55 percent of the carbon is retained in the bottom ash. The General Electric design study indicates that about one percent of the coal carbon is retained in the bottom ash (Corman, 1986). The ASPEN simulation model developed at METC of the Lurgi-based IGCC system had a default assumption of 97 percent carbon conversion; however, the basis for this assumption was not documented. Also, as mentioned above, unrecycled fines carryover also affects the overall carbon conversion efficiency in the gasifier.

B.3.3.4 Bottom Ash Sulfur Retention

The amount of sulfur retained in the bottom ash is often not explicitly stated in design studies. However, it can be inferred if sufficient detail is provided for the gasifier mass balance. Zahnstecher (1984) indicates that about 2.7 percent of the sulfur in a western subbituminous coal is retained in the bottom ash. Cover et al (1985) and Smelser (1986a) indicate that 18.67 percent of the sulfur in the inlet coal is retained in the bottom ash for lignite coal. Simbeck et al (1983) do not provide sufficient detail in the mass balance to fully account for the fate of the sulfur in the Illinois No. 6 feed coal, but up to about 4.4 percent may be retained in the ash or contained in the tars and oils exiting with the raw coal gas.

B.3.3.5 Gasifier Throughput

Simbeck et al (1983) report that the standard Lurgi Mark IV gasifiers have a nominal capacity of approximately 650 st/day of MAF coal. This is presumably based on oxygen-blown operation. However, the design pressure for this throughput is not stated. The capacity of the gasifiers in the Zahnstecher (1984) study is approximately 620 st/day

MAF coal using subbituminous coal in oxygen-blown mode at 465 psia. The Smelser (1986a) study indicates a capacity of about 630 st/day MAF lignite coal in oxygen-blown mode at a pressure of 430 psia.

Corman (1986) reports that Lurgi estimated the coal throughput in air-blown mode for MAF Illinois No. 6 coal to be 300 lb/hr-ft^2 at 20 atmospheres (294 psia) and 500 lb/hr-ft^2 at 40 atmospheres (588 psia). This corresponds to throughputs of 460 st/day and 760 st/day, respectively, for a Lurgi Mark IV gasifier with a working diameter of 12.7 feet. Corman commented that, compared to pilot-scale data, these throughputs "appear somewhat optimistic for air-blown operation," but that they were used nonetheless in the design study. These throughput estimates for air-blown operation appear to be about the same as the estimates given above for oxygen-blown operation. For example, if the throughput is linearly interpolated to a level of, say, 430 psia, the estimate would be about 640 tons/day, which is slightly more than the throughput used in the Smelser (1986a) study cited above.

B.3.3.6 Ammonia Concentration

One of the key environmental concerns associated with IGCC systems featuring hot gas cleanup up are NO_x emissions from the gas turbine associated with fuel bound nitrogen in the fuel gas. The most prominent form of fuel-bound nitrogen is ammonia, which is one of the products of the gasification process in the Lurgi gasifier. Corman (1986) assumes a fuel gas composition containing 0.2 volume percent ammonia, resulting from the air-blown gasification of Illinois No. 6 coal. Studies featuring oxygen-blown systems typically assume even higher ammonia concentrations. For example, Simbeck et al (1983) show a gas composition from gasification of Illinois No. 6 coal with 0.3 volume percent ammonia. Smelser (1986a) indicates an ammonia concentration of 0.45 volume percent from the oxygen-blown gasification of lignite. Holt et al (1989) indicate that about 50 to 60 percent of coal-bound nitrogen is converted to ammonia in fixed bed gasifiers.

B.3.3.7 Gasifier Pressure

The pressure of the raw gas exiting the gasifier must be large enough to overcome all of the pressure losses between the gasifier exit and the gas turbine combustor. Sources of pressure loss include piping, valving, cyclones, desulfurization vessels, and the gas turbine fuel valve. The pressure in the gas turbine combustor depends on the gas turbine pressure ratio. In this study, the default design strategy is to specify the gasifier pressure as the sum of the gas turbine combustor pressure and the pressure losses between the gasifier outlet and the gas turbine fuel valve.

The typical pressure ratio for the gas turbine most commonly assumed in IGCC design studies is 13.5, which implies a gas turbine combustor pressure of 198.5 psia. A standard gas turbine fuel valve has a pressure drop of about 70 psi. Additional pressure losses in the piping and equipment between the gasifier outlet and gas turbine fuel valve inlet may total approximately 30 to 60 psi for typical design cases. Thus, a typical gasifier pressure may be about 300 to 330 psia. However, possible advances in gas turbine fuel valve design may lead to significant reductions in the system pressure losses, allowing gasifiers to be designed for pressures of, say, 240 to 275 psia.

The gasifier pressure affects the cost of the gasification vessel and the coal throughput of the vessel. At lower pressures, the cost per vessel would be lower, but the number of vessels requirement to handle a given amount of coal may need to be increased due to a reduction in the gasifier coal throughput. This type of trade-off can be evaluated in the performance and cost model of the Lurgi-based IGCC system if sufficient data are available regarding the effect of gasifier pressure on gasifier coal throughput.

B.3.3.8 Gasifier Steam Requirement

Lurgi supplied General Electric with estimates for the gasifier steam/air ratio and the gasifier coal throughput. Lurgi estimated that 0.4 to 0.5 lb steam/lb air would be required to maintain the gasifier combustion zone temperature below the ash fusion temperature of 2,300 °F for the design coal. General Electric used a higher value of 0.6 lb steam/lb air as a conservative estimate (Corman, 1986).

B.3.3.9 Gasifier Oxidant Requirement

For the air-blown Lurgi gasifier, Corman assumed an air-to-coal ratio of 2.41 lb air/lb coal. The Illinois No. 6 coal has a run-of-mine carbon content of 60.1 weight percent and oxygen content of 7.57 weight percent. Thus, the oxygen-to-carbon ratio for the oxidant is 0.93 lb oxygen/lb carbon and the overall oxygen-to-carbon ratio, including oxygen in the feed coal, is 1.06 lb oxygen per lb of carbon. The stoichiometric ratio for combustion is 2.667 lb oxygen/lb carbon.

For comparison, the assumed overall oxygen-to-carbon ratio in a study of an oxygen-blown system featuring lignite coal is 0.76 (Smelser, 1986b), which represents an oxidant oxygen-to-carbon ratio of 0.49. For a system gasifying subbituminous coal, the assumed oxidant oxygen-to-carbon ratio is 0.42, with an overall oxygen-to-carbon ratio of 1.07 including oxygen in the coal (Zahnstecher, 1984). Zahnstecher cites the high reactivity of the Western subbituminous coal and good heat transfer characteristics of the

Lurgi gasifier as key factors leading to the relatively low oxygen-to-coal ratio for this study.

The lower rank lignite and subbituminous coals typically have higher gasification reactivity than high-rank eastern coals, and therefore can be gasified at lower temperatures (Shinnar et al, 1988). Thus, the amount of heat needed from char combustion is reduced, which in turn reduces demand for oxygen in the combustion zone of the reactor. Thus, we would expect a higher oxygen requirement for the Illinois coal than for the other coal types to support combustion of coal char. From the design studies cited above, it appears that the assumed oxygen-to-carbon ratio associated with the oxidant feed is about a factor of two greater for the Illinois coal than for the lower rank subbituminous and lignite coals.

B.3.4 Elicited Technical Judgments About Uncertainties

Technical judgments regarding the performance of a commercial-scale fifth-of-a-kind fixed-bed gasifier based on Lurgi technology were elicited from two engineers at DOE/METC. The two experts will be referred to as LG-1 and LG-2. The engineers were asked to explicitly consider the uncertainty involved in making predictions about a system that has not yet been built or operated. The experts were provided with a three-part briefing paper as described in Chapter 4. The first part was a general introduction to uncertainty analysis. The second part was the review of published information given in the preceding text. The third part was a questionnaire which is given in Section B.8.1.

B.3.4.1 Expert LG-1

Expert LG-1 responded with detailed distributions for all performance parameters included in the questionnaire, but was unable to answer the two questions regarding capital and maintenance costs. In some cases, a brief rationale for the judgments was given; in other cases, presumably due to time constraints, the expert did not indicate qualitatively a basis for the values in the distributions. However, this expert provided detailed and intricate judgments regarding conditional uncertainties in two cases: one case involved coal gasifier throughput which depends on pressure, and the second involved the steam/coal ratio, which is correlated to the air/coal ratio. The expert indicated that the elicitation process was "not too difficult," and that "some relatively easy calculations were required to provide "good" judgments." The expert also explained that "it helped to be knowledgeable of METC's 42-inch gasifier work/data," referring to an in-house research program at DOE/METC on an experimental fixed bed gasifier.

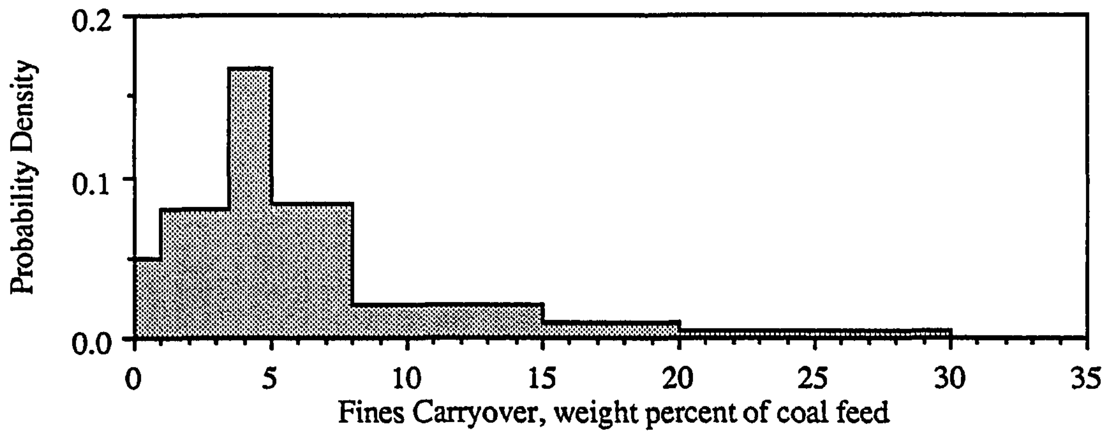

Figure B-3. Judgment of Expert LG-1 Regarding Gasifier Fines Carryover.

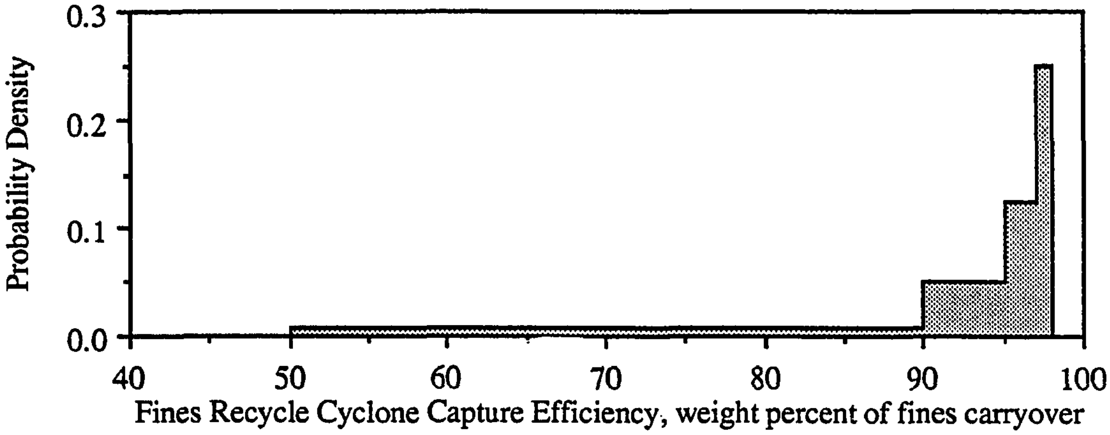

Figure B-4. Judgment of Expert LG-1 Regarding Fines Cyclone Capture Efficiency.

The judgments from expert LG-1 are summarized in Table B-6. With respect to fines carryover, the expert commented that low fines carryover could result from "improvements to [the coal] feed system," while high fines carryover could be associated with "process upsets and/or coal attrition due to coal feed system." A graphical representation of Expert LG-1's judgment regarding fines carryover, in the form of a probability density function (pdf), is shown in Figure B-3. The figure illustrates that the expert considers fines carryover of 3.5 to 5 percent to be most likely, but that there is a small probability that the carryover could be as high as the total fines inlet to the gasifier, which is 30 weight percent of the coal feed. The expert provided a judgment regarding fines capture, but indicated that he felt least knowledgeable about this particular parameter. This judgment is shown graphically as a pdf in Figure B-4. The

Table B-6. Summary of Elicited Lurgi Gasifier Technical Judgments from Expert LG-1

Description	Units	Distribution	Parameters[a]				
Fines Carryover from Gasifier	wt-% of Coal Feed	Fractile	5%:	0	to	1	
			20%:	1	to	3.5	
			25%:	3.5	to	5	
			25%:	5	to	8	
			15%:	8	to	15	
			5%:	15	to	20	
			5%:	20	to	30	
Fines Capture in Recycle Cyclone	% of Carryover	Fractile	25%:	50	to	90	
			25%:	90	to	95	
			25%:	95	to	97	
			25%:	97	to	98	
Fines Carbon Content	wt-% of fines	Fractile	5%:	65	to	70	
			20%:	70	to	75	
			25%:	75	to	79	
			25%:	79	to	84	
			25%:	84	to	87	
Carbon Retention in Bottom Ash	wt-% of coal feed carbon	Triangular	0.75		to	10	(2.5)
Sulfur Retention in Bottom Ash	wt-% of coal feed sulfur	Triangular	1.5		to	6	(3)
<u>Gasifier Coal Throughput</u>							
250 psia	lb DAF/(hr-ft^2)	Triangular	133		to	333	(266)
300 psia	lb DAF/(hr-ft^2)	Triangular	152		to	381	(305)
350 psia	lb DAF/(hr-ft^2)	Triangular	170		to	426	(341)
Gasifier Ammonia Yield	Equiv. fraction of coal N to NH_3	Triangular	0.5		to	1.0	(0.9)
Gasifier Air/Coal Ratio	lb air/lb DAF	Triangular	2.7		to	3.4	(3.1)
<u>Gasifier Steam Requirement</u> (Correlated to Air/Coal Ratio)							
Air/coal = 2.7	lb H_2O/lb DAF	Uniform	0.54		to	1.08	
Air/coal = 3.1	lb H_2O/lb DAF	Uniform	1.24		to	1.86	
Air/coal = 3.4	lb H_2O/lb DAF	Uniform	2.04		to	2.72	

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

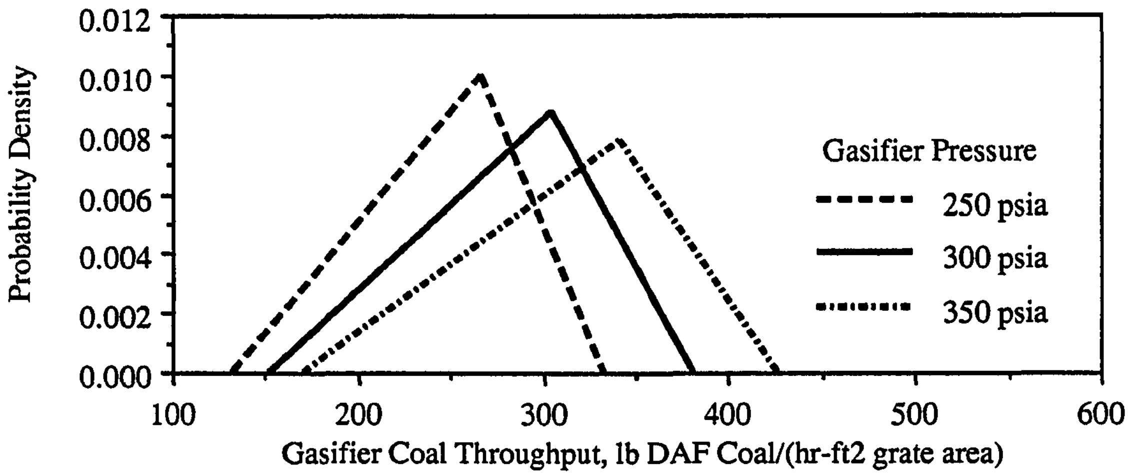

Figure B-5. Judgment of Expert LG-1 Regarding Gasifier Coal Throughput.

judgment indicates a high probability that the fines capture will be 90 percent or greater, with a chance that the capture could go as low as 50 percent.

High carbon retention in the gasifier bottom ash might be associated with "poor distribution of gas flow through [the] bed due to agglomerates and concentrations of fines," whereas low carbon retention "requires good process control, properly designed grate, and smooth operations." High sulfur retention in the bottom ash might result from "poor carbon conversion" and "excessive channeling in [the] bed," while low retention would be related to "good carbon burnout of ash."

In the questionnaire, the experts were asked to provide judgments about gasifier coal throughput at three different pressures. Expert LG-1 provided a complete response, indicating that high throughputs at a given pressure would be associated with "better feed systems" and "poorer gas quality," while low throughputs would be related to "excessive fines carryover" and "excessive carbon content of bottom ash," implying correlation with fines carryover and carbon retention in the bottom ash. Expert LG-1's judgments regarding gasifier coal throughput for three different pressures are shown in Figure B-5. The graph indicates that the variance as well as the mode of the distribution increases as the pressure increases.

Expert LG-1 provided judgments regarding uncertainty in the gasifier steam requirement as a function of the gasifier air/coal ratio, which is also treated probabilistically. The judgments regarding uncertainty in the steam/coal ratio for three different air/coal ratios are shown in Figure B-6. For modeling studies, the uncertainties in the steam/coal ratio for intermediate values of the air/coal ratio can be interpolated. The

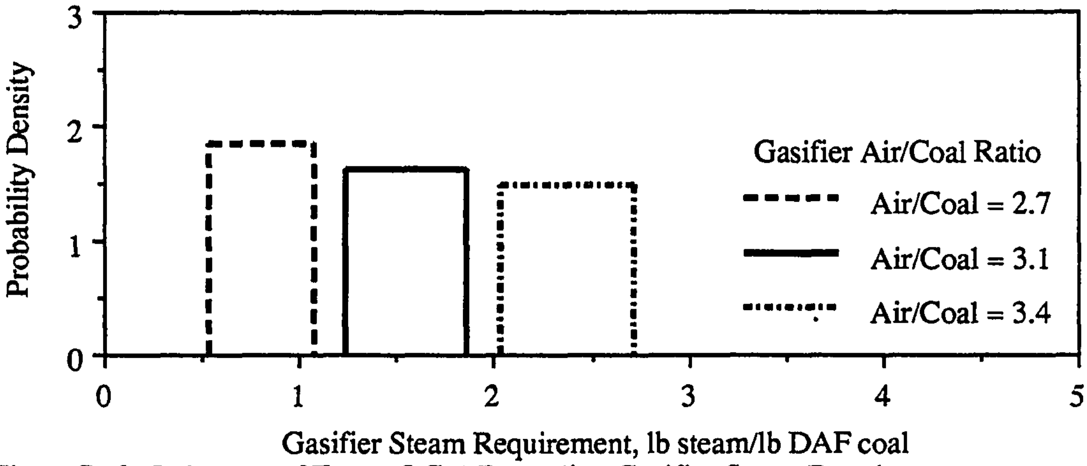

Figure B-6. Judgment of Expert LG-1 Regarding Gasifier Steam Requirement.

interpolation is simplified because the steam/coal ratio is assumed to be uniformly distributed for each air/coal ratio. Therefore, the upper and lower limits of the uniform distribution can be interpolated, and a random number generator (uniform distribution) can be used to sample between the interpolated upper and lower limits.

In a follow-up phone conversation, Expert LG-1 was asked to elaborate on the relationships among the parameters. One pair of correlated parameters are the carbon and sulfur retention in the bottom ash. The expert indicated that the inorganic portion of the sulfur in the coal will "come off" readily in the upper portions of the gasifier, while the organically-bound sulfur can be released only as associated carbon is converted. Therefore, there is a tendency to have high sulfur retention associated with high carbon retention in the bottom ash. The expert indicated that the ratio of sulfur-to-coal in the bottom ash would be similar to the ratio of organic sulfur to carbon in the coal.

High carbon retention in the bottom ash, and high fines carryover, is an indicator of poor carbon conversion. To counter such a situation, a plant operator may choose to change the reagent feed ratios or the coal throughput. Thus, uncertainty carbon retention in the bottom ash may be negatively correlated with uncertainty in the air/coal ratio. Similarly, coal throughput may be positively correlated with carbon retention in the bottom ash, because higher coal throughputs imply less residence time in the gasifier, possibly leading to more incomplete carbon conversion. However, it is possible that coal throughput may be constrained due to gasifier bed stirrer design and possible "hang up" of coal or char along the gasifier walls, leading to channeling. Channeling is a possibility particularly with swelling coals. The result of channeling is that some of the coal will not react to the same degree as the rest. In instances of channeling, even high reagent feed ratios may not be successful at improving carbon conversion in the gasifier.

Thus, several of the parameters for which judgments were obtained are expected to be correlated. The elicitation of correlation structures among these variables was not attempted, because of the time consuming nature of such an exercise. Instead, the approach taken here is to run the model assuming no correlations and then run a "sensitivity" case with assumed correlations. For example, the bottom ash sulfur and carbon retentions are assumed to be closely correlated, and are given a nominal correlation of 0.5 for the purpose of the correlation screening study. High fines carryover is assumed to be associated with situations involving low carbon conversion in the gasifier, which would be correlated with high bottom ash carbon retention. Therefore, to characterize a possible correlation, the fines carryover is given a correlation of 0.5 with the bottom ash carbon retention. Cases of poor gasification efficiency are likely to be associated with high coal throughput. Therefore, a positive correlation is given between bottom ash carbon retention and gasifier throughput. Also, low carbon conversion may be associated with low regent (air, steam) feed ratios. The sign and magnitude of the assumed correlations is intended to be plausibly indicative of correlations among the variables, but these correlations should be interpreted merely as an illustrative example for comparison to an uncorrelated case. The proposed correlations among uncertain parameters are summarized in Table B-7

Expert LG-1 did not provide any comments regarding judgments for fines carbon content, ammonia yield, steam requirement, and oxidant requirement. The expert usually answered questions in the form asked, rather than proposing alternative types of probability distributions or other sets of parameters to consider for probabilistic analysis. For example, some questions were worded to obtain fractile distributions as the default, while others were worded to obtain triangular distributions, although it was clearly indicated that the expert was free to use any distribution that was appropriate. In cases where there were compound questions involving conditional probabilities, the triangular distribution was recommended for simplicity.

Table B-7. Proposed Correlation Matrix for Uncertain Parameters LG-1

Description	Number	Assumed Correlation Coefficients[a] 1	2	3	4	5	6	7	8	9
Fines Carryover from Gasifier	1	1.00								
Fines Capture in Recycle Cyclone	2	--	1.00							
Fines Carbon Content	3	--	--	1.00						
Carbon Retention in Bottom Ash	4	0.5	--	--	1.00					
Sulfur Retention in Bottom Ash	5	--	--	--	0.5	1.00				
Gasifier Coal Throughput	6	--	--	--	0.50	--	1.00			
Gasifier Ammonia Yield	7	--	--	--	--	--	--	1.00		
Gasifier Air/Coal Ratio	8	--	--	--	-0.5	--	--	--	1.00	
Gasifier Steam Requirement (Correlated to Air/Coal Ratio)	9	--	--	--	-0.5	--	--	--	(b)	1.00

[a] For uncorrelated parameters, a dashed line "- -" is shown.

[b] The expert explicitly developed a correlation structure between air/coal ratio and the steam requirement. See Table B-6.

B.3.4.2 Expert LG-2

Expert LG-2 indicated that he did not have previous operating experience with fixed-bed gasifiers, which made the development of judgments "highly difficult." Expert LG-2 also indicated that "it was difficult to develop the range of various values for variables due to lack of abundant actual operating data." The expert also noted that "the failure of a stirrer may present operational risk which should be assessed." The expert provided quantitative judgments for all but one of the parameters for which judgments were requested.

The quantitative judgments provided by Expert LG-2 are given in Table B-8. The expert's judgment regarding fines carryover is shown graphically as a pdf in Figure B-7. This judgment implies that there will be at least some fines carryover, with a most likely single value of 7.5 percent of the coal feed. The fines recycle cyclone capture efficiency uncertainty judgment is shown in Figure B-8. This graph indicates that the mode of the

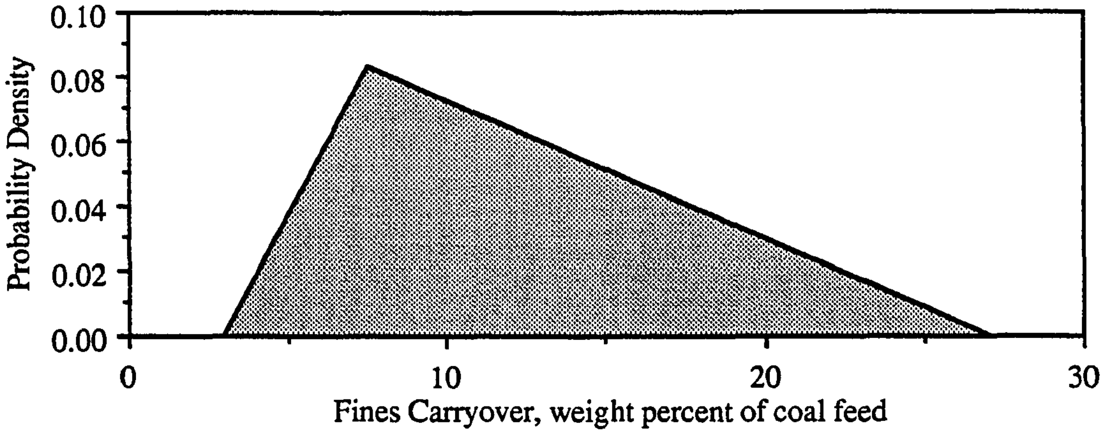

Figure B-7. Judgment of Expert LG-2 Regarding Fines Carryover.

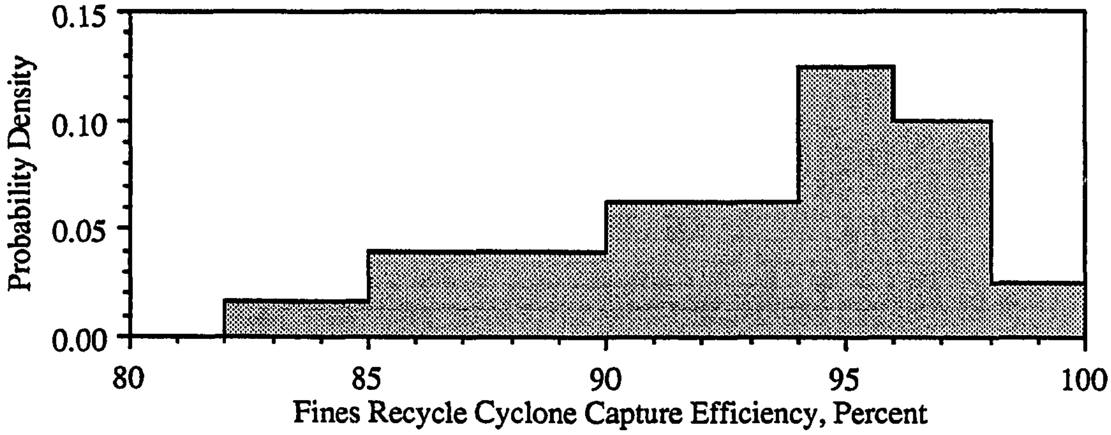

Figure B-8. Judgment of Expert LG-2 Regarding Fines Recycle Cyclone Capture Efficiency.

distribution is a range between 94 and 96 percent fines capture, and that the distribution is negatively skewed.

A low fines carbon content might be associated with "partially gasified chars" that "are broken by stirrer and entrained." A high carbon retention in the bottom ash might be "due to partial agglomeration of caking coal." For gasifier coal throughput, the expert provided "reasonable" numbers for coal throughput and indicated that coal throughput could be 30 percent higher "at the expense of carbon conversion," implying a correlation between coal throughput and carbon conversion. The expert indicated that, in the worst case, the coal throughput would be about 30 percent less than the "reasonable" numbers. Originally, the expert's response was that the minimum value would be zero in the event of a failure of the coal feed system. However, this response was problematic for the current modeling effort. Primarily, this response was more related to an assessment of coal feed process area reliability than it is an assessment of uncertainty in the performance of the

Table B-8. Summary of Elicited Lurgi Gasifier Technical Judgments from Expert LG-2

Description	Units	Distribution	Parameters[a]
Fines Carryover from Gasifier	wt-% of Coal Feed	Triangular	3 to 27 (7.5)
Fines Capture in Recycle Cyclone	% of Carryover	Fractile	5%: 82 to 85 20%: 85 to 90 25%: 90 to 94 25%: 94 to 96 20%: 96 to 98 5%: 98 to 100
Fines Carbon Content	wt-% of fines	Triangular	70.0 to 87.4 (78.7)
Carbon Retention in Bottom Ash	wt-% of coal feed carbon	Triangular	1 to 20 (2)
Sulfur Retention in Bottom Ash	wt-% of coal feed sulfur	Triangular	4 to 30 (10)
<u>Gasifier Coal Throughput</u>[b]			
250 psia	tons DAF/day	Triangular	279 to 517 (398)
300 psia	tons DAF/day	Triangular	382 to 593 (456)
350 psia	tons DAF/day	Triangular	358 to 666 (512)
Gasifier Ammonia Yield[c]	Equiv. fraction of coal N to NH_3	Triangular	0.25 to 0.75 (0.50)
Gasifier Steam/Air Ratio	lb steam/lb air	Fractile	50%: 0.30 to 0.45 50%: 0.45 to 0.70
Gasifier Air/Coal Ratio[d]	lb air/lb coal	Triangular	0.4 to 2.9 (2.41)
Gasifier Maintenance Cost Factor	% of capital investment	Uniform	4.5 to 12.0

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

[b] For this parameter, the expert appears to be combining the concept of process reliability with that of the uncertainty in system operation given that the system is operating. For the purpose of a screening study of uncertainties, the lower bound on the gasifier throughput will be assumed to be 30 percent less than the mode.

[c] The expert provided judgments as a mixture of fractional conversion of coal nitrogen to ammonia and ammonia concentration in the fuel gas. Here, the result is expressed as an equivalent fractional conversion, for consistency with the other expert.

d This judgment includes a mixture of assumptions regarding plant operation. The low value is reported to be associated with "hot standby condition." However, the intent of the modeling exercise is to evaluate full-load operation of the system. Therefore, for a screening study, a lower limit as shown parenthetically will be used.

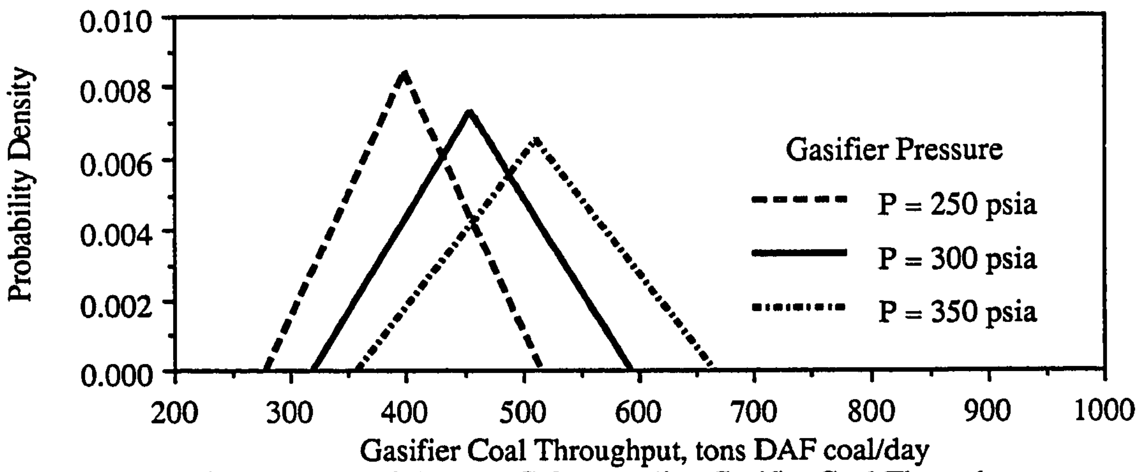

Figure B-9. Judgment of Expert LG-2 Regarding Gasifier Coal Throughput.

gasification process area given that the system is operating at full load. Thus, for the purpose of performing a probabilistic case study of full-load operation, after a follow-up phone call to the expert, it was agreed that the lower limit of the distribution will be assumed to be 70 percent of the mode, symmetric to the upper limit. This set of judgments is shown graphically in Figure B-9.

Similarly, for the oxidant requirement, the expert appeared to mix the requirements for full-load operation with that for hot standby. Upon a follow-up phone call, the expert indicated that the oxidant requirement can be varied at will, but that to get a "good gas," a lower limit would be about 0.4 lb air/lb coal. Expert LG-2 did not provide comments for any of the other parameters for which judgments were made.

Based on the comment of Expert LG-2 that gasifier coal throughput can be increased at the expense of carbon conversion, a correlation structure among the uncertainty judgments was assumed for the purpose of comparison with an uncorrelated case. This proposed correlation structure is given in Table B-9. The fines carryover and the bottom ash carbon retention are both assumed to be positively correlated with the coal throughput, implying reduced carbon conversion as throughput is increased. Furthermore, it is assumed that the carbon retention in the bottom ash will tend to increase as the air/coal ratio is lowered. Hence, a negative correlation is assumed between these two parameters.

Table B-9. Proposed Correlation Matrix for Uncertain Parameters LG-2

Description	No.	Assumed Correlation Coefficients[a] 1	2	3	4	5	6	7	8	9	10
Fines Carryover from Gasifier	1	1.00									
Fines Capture in Recycle Cyclone	2	--	1.00								
Fines Carbon Content	3	--	--	1.00							
Carbon Retention in Bottom Ash	4	--	--	--	1.00						
Sulfur Retention in Bottom Ash	5	--	--	--	--	1.00					
Gasifier Coal Throughput	6	0.9	--	--	0.9	--	1.00				
Gasifier Ammonia Yield	7	--	--	--	--	--	--	1.00			
Gasifier Steam/Air Ratio	8	--	--	--	--	--	--	--	1.00		
Gasifier Air/Coal Ratio	9	--	--	--	-0.9	--	--	--	--	1.00	
Gasifier Maintenance Cost Factor	10	--	--	--	--	--	--	--	--	--	1.00

[a] For uncorrelated parameters, a dashed line "- -" is shown.

B.3.5 Other Uncertainties

While judgments about uncertainties in the performance of a fifth-of-a-kind fixed-bed dry-ash gasifier operating on Illinois No. 6 coal with up to 30 percent fines loading and fines recycle were obtained from technical experts, these experts were generally unable to make judgments about cost-related parameters for this process area. One exception to this is that Expert LG-2 did provide a judgment regarding the maintenance cost factor. This judgment will be included in all Lurgi case studies.

In addition to the maintenance cost factor, a judgment regarding uncertainty is required regarding the gasifier process area capital cost. For example, Corman (1986) commented that operating experience on highly caking coals, such as Illinois No. 6, in the Lurgi gasifier is relatively limited, and that performance estimates are subject to change as more experience is acquired. However, Corman indicates that changes are likely to have more effect on capital costs (e.g., additional vessels for fines handling) rather than on overall system performance. This statement implies that there may be uncertainty in the scope and cost of the gasification process area. This type of uncertainty is traditionally handled using "process contingency factors" as discussed elsewhere in this dissertation (see Chapter 6 of Volume 1). However, Corman assumed a process contingency factor of only 5 percent, which is quite low according to EPRI guidelines. For example, EPRI recommends a factor of 0 to 10 percent for a commercially demonstrated process, 5 to 20 percent for a system in which a full size module has been operated, and 20 to 35 percent for a system for which only small scale test have been performed. However, from the EPRI TAG, it is not clear what probability of cost over-run is associated with these "rule-of-thumb" recommendations. Considering that a system with fines recycle has not yet been operated, and that there may be significant uncertainties not only in the fines recycle aspect of the process area, but also related to the gasifier bed stirrer, it appears plausible that a higher level of contingency should be assumed in a deterministic study, and that a relatively wide range of uncertainty should be assumed in initial probabilistic studies. Furthermore, the available cost estimates from which the cost models were developed do not include gasifier bed stirrers. Thus, a suggested uncertainty factor for a screening study of uncertainties is a uniform distribution from 10 to 30 percent on the same basis as the process contingency factor.

B.4 Fluidized Bed Coal Gasification Process Area

This section reviews uncertainties in, and potential problems of, using fluidized-bed gasifiers in an integrated gasification combined cycle (IGCC) process environment. This review considers uncertainties which arise from: (1) scaling-up of the KRW gasifier process development unit (PDU) to a commercial gasifier design; and (2) applications of the KRW gasifier in two different IGCC process environments. The two IGCC systems under consideration here are:

- Case AKH: Air-blown KRW-based IGCC system with Hot gas cleanup featuring in-bed and external zinc ferrite hot gas desulfurization.
- Case OKC: Oxygen-blown KRW-based IGCC system with Cold gas cleanup.

B.4.1 Process Description

The assumed system configuration for Case AKH is shown in Figure B-10, and the configuration for Case OKC is shown in Figure B-11. The schematic for Case AKH represents process elements based on design and cost studies prepared for the Gas Research Institute (Smelser, 1986; Earley and Smelser, 1988), DOE (Corman, 1986) and the configuration assumed in the ASPEN simulation model developed at METC (Craig, 1988). The schematic for Case OKC is based on elements of design and cost studies prepared for the Electric Power Research Institute (Dawkins et al, 1985) and DOE (Bechtel, 1983) and the configuration assumed in an ASPEN simulation model developed at METC (Stone, 1985).

The primary features of the IGCC system of Case AKH compared to Case OKC are: (1) elimination of an oxygen plant and use of air extracted from the gas turbine for oxidant feed; (2) in-situ desulfurization with limestone or dolomite; (3) external desulfurization using a high temperature removal process; (4) reduced requirement for syngas cooling prior to desulfurization; (5) elimination of sulfur recovery and tail gas treating; and (6) addition of a circulating fluidized bed boiler for sulfation of spent limestone (to produce an environmentally acceptable waste) and conversion of carbon remaining in the ash.

For Case AKH, the key inputs to the gasifier include coal, air, steam, and a calcium-based sorbent such as limestone or dolomite. The assumed coal for this study for both cases is Illinois No. 6. Some characteristics for this design coal are given in Table B-10. Characteristics of typical limestone and dolomite sorbents are given in Table B-11.

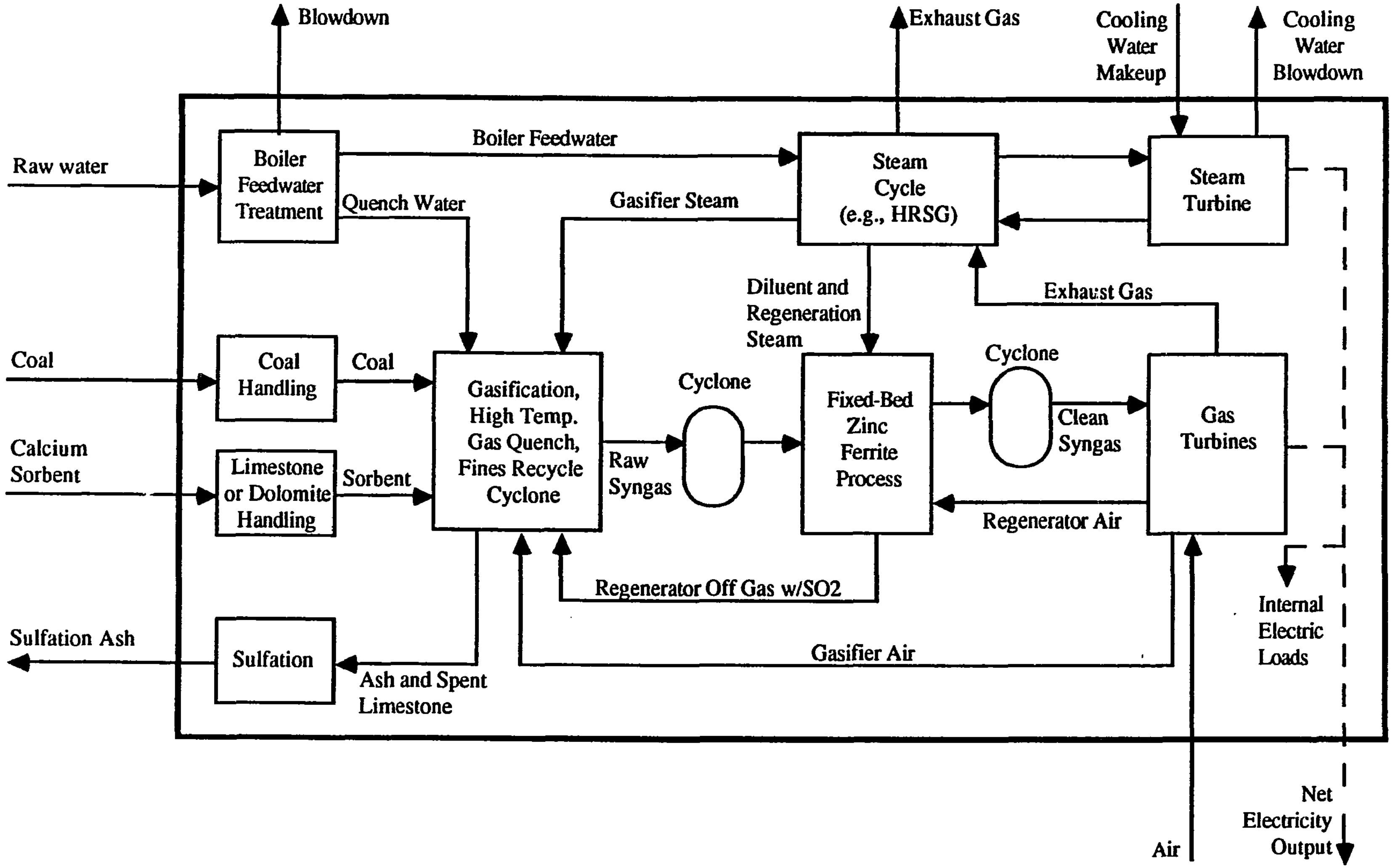

Figure 10. Simplified Schematic of Air-Blown KRW IGCC System with Hot Gas Cleanup

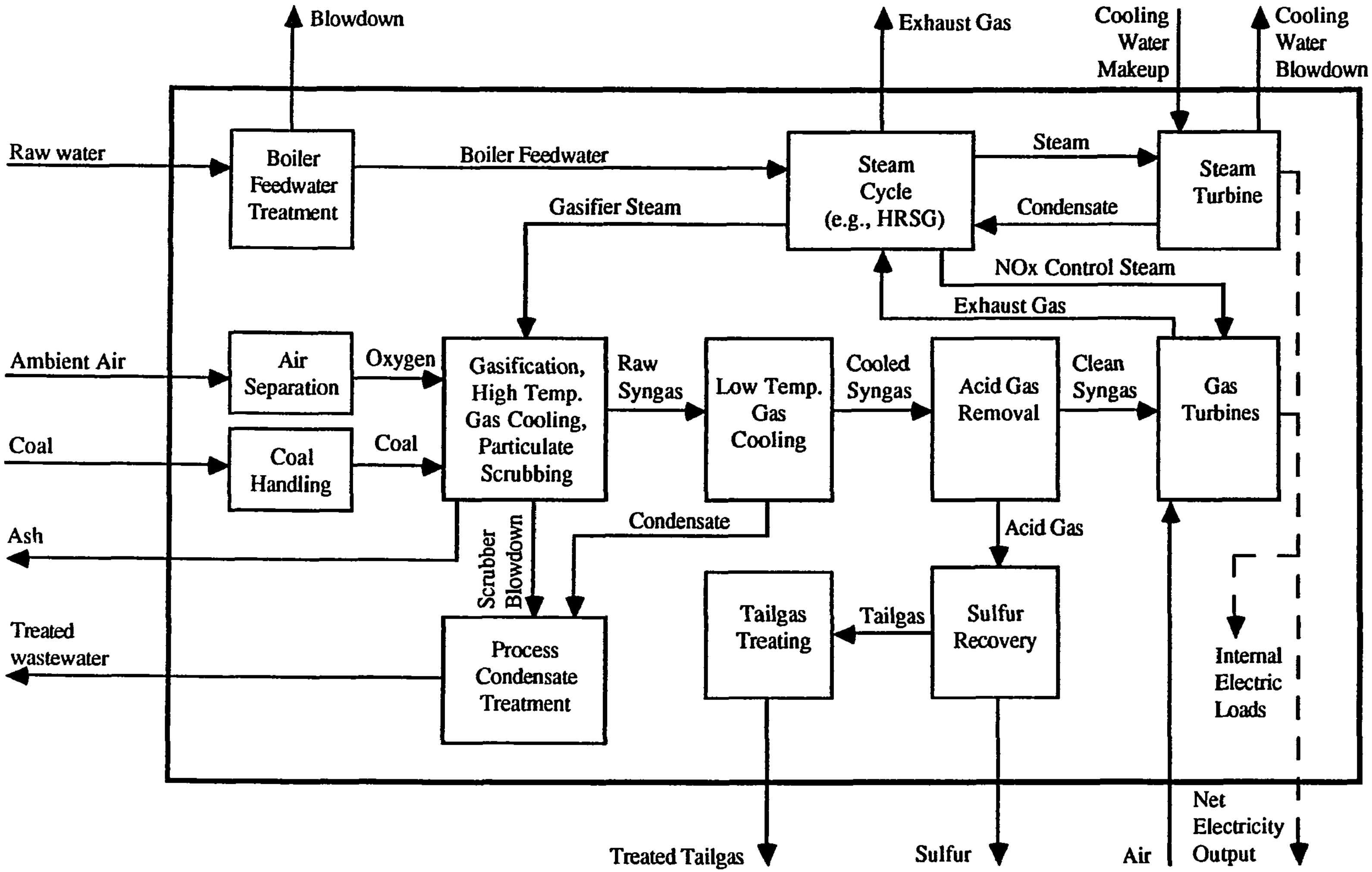

Figure 11. Schematic of Oxygen-Blown IGCC System with Cold Gas Cleanup

Table B-10. Default Characteristics for the Design Illinois No. 6 Coal

General Description	
Rank	Bituminous
Coal Seam	Illinois No. 6
Preparation	Run of Mine
Properties	
Proximate Analysis, as received, wt-%	
Moisture	12.0
Volatile Matter	47.8
Fixed Carbon	31.4
Ash	8.8
Ultimate Analysis, dry, wt-%	
Carbon	69.5
Hydrogen	5.3
Nitrogen	1.3
Oxygen	10.0
Sulfur	3.9
Ash	10.0
Chlorine	0.0
Heating Value of Coal, As Received	
Btu/lb (HHV)	12,774
Form of Sulfur as % of Total Sulfur	
Pyritic	2.0
Sulfate	1.9
Organic	0.0
Ash Fusion Temperature, °F	2,300

Table B-11. Typical Characteristics for High-Calcium Limestone and Dolomite

High-Calcium Limestone, wt-%	
$CaCO_3$	81.7
$MgCO_3$	10.4
Inerts	7.9
Dolomite, wt-%	
$CaCO_3$	69.2
$MgCO_3$	23.9
Inerts	6.9

A simple schematic of a KRW gasifier is given in Figure B-12. The basic features include a large refractory lined carbon steel pressure vessel with several "zones," coal and oxidant feed tubes, ash removal annulus, and fines recycle system. The gasifier zones include low velocity ash cooling and removal, moderate velocity ash separation and gasification, and high velocity combustion and devolatilization. In addition, a freeboard disengaging zone is used to reduce the amount of char entrained in the outgoing syngas. The diameter of the gasifier is largest in the freeboard zone, in order to reduce syngas velocity and entrainment of char (Smith et al, 1986).

Pulverized coal is fed with oxidant and transport gas to a combustion jet. The combustion jet supplies the heat required for the endothermic gasification reactions in the fluidized bed, in which steam and oxidant react with the char remaining after partial combustion. A portion of coal char fines are elutriated with the coal gas leaving the gasifier. Most of these fines are captured in a cyclone and recycled to the gasifier. Ash is removed through an annular area around the coal feed tube. The temperature in the ash-agglomerating zone is above the softening temperature of some of the eutectics in the ash. This leads to ash agglomeration, as ash particles stick together to form larger particles which are defluidized. Cool raw coal gas is recycled to the ash separation zone to cool the ash and for velocity control in the ash annulus (Simbeck et al, 1983).

Two of the key operating parameters of an IGCC power plant are the gasifier oxidant and steam requirements. The gasifier oxidant requirement can be expressed in terms of the oxygen-to-carbon molar ratio based on oxygen in the oxidant and carbon in the coal feed. The gasifier steam requirement can be expressed in terms of the molar ratio of steam-to-oxygen. Some values of the oxygen-to-coal and steam-to-oxygen ratios are summarized in Table B-12. These include values from PDU tests, theoretical modeling, and conceptual design studies.

B.4.2 Commercial Status of the KRW Gasifier

M.W. Kellogg has operated a 15 to 35 ton per day KRW gasifier process development unit (PDU) at Waltz Mill, PA since 1975. This unit has accumulated over 13,000 hours of operation on a variety of feedstocks, including bituminous, subbituminous, and lignite coals, cokes, and non-U.S. coals (Floyd and Agrawal, 1989). The PDU operates at pressures up to 245 psia and has been run in oxygen- and air-blown modes with and without in-bed desulfurization (Haldipur et al, 1989).

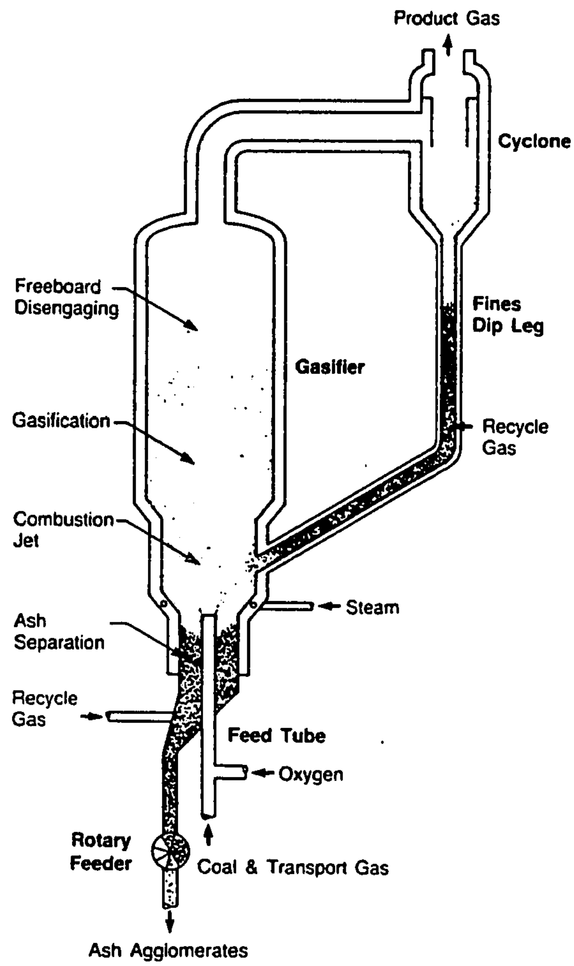

Figure B-12. Schematic of the Fluidized-Bed, Agglomerating Ash KRW Gasifier
(Source: Simbeck et al, 1983)

Table B-12. Summary of Selected KRW Gasifier Design Assumptions

Reference:	Shinnar et al, 1988				Gallaspy et al, 1990	Dawkins et al, 1985	Smith and Smelser, 1990		Earley and Smelser, 1988a	Earley and Smelser, 1988	Shinnar and Weng 1990
Description	KRW test run	KRW test run	Reference commercial design	Model result	SCS EPRI Study	Fluor EPRI Study	Fluor GRI Study	Fluor GRI Study	Fluor GRI Study	Fluor GRI Study	Model Result
Coal	Pgh #8	Pgh #8	Pgh #8	Pgh #8	Ill #6	Ill #6	Pgh #8	Pgh #8	Pgh #8	Pgh #8	Pgh #8
Oxidant	Oxygen	Oxygen	Oxygen	Oxygen	Oxygen	Oxygen	Oxygen	Oxygen	Oxygen	Oxygen	Air
Temp, ^oF	1,800	1,810	1,850	1,850	1,850	1,850	1,875	1,850	1,850	1,850	1,950
Pressure, psia	245	245	465	220	465	465	465	465	465	465	315
O_2/C, mol	0.52	0.39	0.35	0.50	0.34	0.33	0.33	0.34	0.25	0.25	0.45
H_2O/O_2, mol	0.83	0.85	na	5.5	0.36	1.59	1.43	1.12	1.38	1.29	0.5
Carbon Conversion, %	75.9	52.4	94.8	95	95.8	94.9	95.6	94.6	72.3	94.6	95
Cold Gas Efficiency, %	45.2	33.6	87.1	67	na	na	85.1	82.5	67.8	82.5	na
Ca/S, mol	--	--	--	--	--	--	--	1.8	--	1.8	na
Sulfur Removal Efficiency, %	--	--	--	--	--	--	--	90	--	90	85

Simbeck et al (1983) report that in 1981 SASOL of South Africa and Westinghouse (then the owner of the KRW technology) had agreed to fund development and construction of a 1,200 metric ton/day of coal demonstration plant. Dawkins et al (1985) reported that the project was cancelled in mid-1983, but that a detailed performance and economic evaluation of the project had been at least partially completed.

A single train low-BTU gas gasification plant was designed for Fularji Heavy Machinery works in the People's Republic of China in 1985. The plant was to be operated on lignite with a capacity of 350 tons/day. However, the project was cancelled due to "commercial reasons" (Gallaspy et al, 1990).

In April of 1986 M.W. Kellogg submitted an initial proposal to the U.S. Department of Energy (DOE) for an air-blown KRW-based IGCC demonstration plant featuring in-bed desulfurization, external desulfurization, and use of a high-sulfur Eastern bituminous coal. The proposed demonstration is known as the "Appalachian Project." This project was to feature a 500 ton/day KRW gasifier operating at 285 psia (Banchik et al, 1988). However, because Kellogg had difficulties finding a site and negotiating a power supply contract, the project was terminated (Gallaspy et al, 1990). Nonetheless, a substantial amount of design work appears to have gone into the Appalachian Project, supported by testing with the KRW gasifier PDU.

In addition to several attempts at detailed designs for specific construction projects, all of which have been cancelled, there have been a number of conceptual design studies of KRW-based coal-to-substitute natural gas (SNG) and IGCC systems. Some of the most recent studies are mentioned here. These include five studies of IGCC systems (Bechtel, 1983; Dawkins et al, 1985; Smelser, 1986; Earley and Smelser, 1988; and Gallaspy et al, 1990) and four studies of coal-to-SNG systems (Cover et al, 1985; Smith et al, 1986; Smith and Smelser, 1987; Earley and Smelser, 1988b). The Gas Research Institute (GRI) sponsored a study of the KRW PDU (Blinn et al, 1989) to provide supporting data for performance and cost assumptions used in a previous study (Earley and Smelser, 1988). A paper prepared by M.W. Kellogg discussed, in qualitative terms, the design basis for an air-blown IGCC system with in-bed desulfurization using Illinois No. 6 coal (Banchik and Cover, 1988). Other studies or KRW-based systems have been prepared in the past (e.g., Bostwick et al, 1981) but are not considered as reliable as more recent studies. In addition, a comparative evaluation of air-blown and oxygen-blown KRW-based systems by Southern Company Services for DOE/METC is currently in the review phase.

The gasification section definitions are similar in all of the studies which have been reviewed to date. The coal-to-SNG systems are the same as the IGCC systems in the areas of coal pressurization, gasification, and ash removal. In fact, it appears that the physical dimensions of the gasifier are similar across all studies. Typical dimensions are an overall height of about 100 to 115 ft, a maximum outer diameter of about 14 feet, and a minimum outer diameter of about 5.5 ft. All systems use a coal surge bin, coal pressurization lockhopper, coal feed lockhopper, and rotary feed valve to deliver pressurized coal to the gasifier. All systems also use an ash receiving lockhopper and an ash depressurization lock hopper for ash removal. All systems use pneumatic transport of coal from the rotary feed valve to the gasifier, and pneumatic cooling and separation of ash. All systems have a recycle gas compressor and motor.

To accommodate in-bed desulfurization, the gasifier vessel may be slightly increased in size compared to no in-bed desulfurization. For example, one study assumed a gasifier size of 101 feet overall length and 14 feet maximum outside diameter without in-bed desulfurization, and 115 feet overall length and 14 feet maximum outside diameter with in-bed desulfurization (Smith and Smelser, 1987). The fluidized bed height for the in-bed desulfurization case is assumed to be approximately 4 feet higher, due to increased bed volume. The limestone addition results in high levels of ash in the bed and higher bed densities than the conventional gasifier. The higher bed density permits a slightly higher superficial velocity (1.72 ft/s vs. 1.6 ft/s) in the freeboard (uppermost) zone of the gasifier.

The conceptual design studies cited above cover a wide range of coal feedstocks. These include Illinois No. 6, Pittsburgh No. 8, Texas lignite, North Dakota lignite, and Wyodak subbituminous coal as feedstocks, with gasifier coal feed moisture contents ranging from about 1 to 23 percent. For this study, attention is restricted to cases with Eastern bituminous coals.

B.4.3 PDU Data Availability and Applicability

The design and operation of the KRW gasifier PDU differ from the conceptual designs for commercial-scale plants. For example, most design studies assume oxygen-blown gasifiers with nominal capacities of 1,000 tons per day of coal, implying a nominal scale-up factor of 40 from the size of the PDU. Alternatively, some studies assume air-blown gasifiers with a nominal capacity of 500 tons/day. The test facility has a maximum pressure capability of 245 psia, whereas design studies commonly assume 465 psia. However, the design pressure for the KRW gasifier in the Appalachian Project was reported to be 285 psia (Banchik, Buckman, and Rath, 1988). The basis for this design pressure was not reported. However, it may represent a more efficient matching of the gasifier pressure to meet system pressure losses in piping, cyclones, external desulfurization vessels, and gas turbine fuel valves.

The PDU has higher heat losses, higher fines elutriation, higher recycle gas flow rates, higher ash annulus gas velocities, and a lower bed height than most design studies (Shinnar, et al, 1988). Furthermore, Shinnar, et al (1988) assert that the carbon conversions and oxygen-to-coal and steam-to-coal ratios assumed in most design studies cannot be justified based on PDU experience, even accounting for differences in design and operation compared to a full-scale commercial gasifier. This is particularly true, according to the study, for KRW gasifiers operating without in-bed desulfurization:

> "Unfortunately, the envisioned performance was never demonstrated in the pilot-plant and our model shows that it is not achievable in an efficient commercial KRW under any circumstances" (p. 155, Shinnar, et al, 1988)

Shinnar et al (1988) indicate that there are several penalties inherent in the KRW design which pose limitations for its gasification efficiency. For example, coal is fed directly into the combustion zone, where volatiles and reactive char combust preferentially. However, it would be preferable to combust unreacted char, such as that from fines recycle. A more efficient design, according to Shinnar et al, would be to feed the unreacted char from fines recycle into the combustion jet, and to feed coal to the fluidized bed above the jet. However, the caking characteristics of a particular coal may make direct injection into the bed infeasible, unless an additive is used to reduce the caking tendency.

Another penalty involves the use of recycled coal gas in the ash removal annulus. The penalty is manifold. The recycled gas has been cooled prior to entering the combustion zone via the ash annulus, thus imposing a sensible energy penalty on the gasifier. Combustion of recycled coal gas competes for oxygen in the oxidant. Use of recycle gas also increases the amount of char that has to be gasified for a given net coal gas production. The recycle gas also has a diluting effect on the steam concentration in the gasification zone, reducing the gasification reaction rate.

Uncertainty regarding fines carry-over has been compensated for by designing an over-sized freeboard section for solids transport disengagement and over-designing downstream solids collecting equipment (Smith, Hanny, and Smelser, 1986). Over-designing to account for uncertainty may lead to unnecessarily expensive systems.

The basic processes involved in gasification are chemical, thermal, and hydrodynamic. Many of the chemical processes do not depend on scale because they take place at a particle level. However, hydrodynamic processes are generally scale-dependent, and influence the thermal history of particles and gases in the gasifier. The main areas of design that have the least commercial experience are the combustion jet zone and the ash separation zone. Analytic models of the jetting and ash separation zone have been developed by KRW based on a variety of tests at different scales, ranging from four inch to ten foot diameter for cold flow facilities and four inch to 24 inch diameter for hot flow facilities. Kellogg Rust Synfuels, Inc. (KRSI) claims an excellent correlation between the analytic models and the observed test results (Smith et al, 1986).

Perhaps the most significant scale-up uncertainty is in the jet combustion zone. It is important that the jet surface area and the solid recirculation rate near the jet be sufficient to allow for dissipation of the heat from combustion, otherwise agglomeration, clinkering and sintering of bed material will occur. Commercial designs may require the use of multiple jets, rather than a single jet, for better distribution of heat. However, this alternative may introduce problems if the jet velocities are not uniform. Also, multiple jets may interact to form stagnant regions between jets. KRSI recommends more extensive testing using semi-circular and circular models of multiple jets before designing a commercial reactor with multiple jets (Smith et al, 1986).

Gasifier performance for a specific coal is predicted by M.W. Kellogg based on analytic models and empirically-derived data. Experimental data are required to determine reaction rates and the influence of contained mineral matter on coal reactivity. The caking properties of the coal are also important, as is the ash fusion temperature. M.W. Kellogg has devised a number of bench-scale tests that are used to determine the empirical data needed for the analytic models. The combination of bench scale testing and mathematical modeling is reported to yield a good predictive capability for gasifier performance. However, the predictive capability is limited by uncertainties in free-board temperature, bed density, bed carbon content, and bed height (Floyd and Agrawal, 1989).

Other sources of uncertainty in predicting commercial scale gasifier performance stem from the limited understanding of the chemistry of gasification in the KRW gasifier. For example, Shinnar et al (1988) conclude that:

- Kinetics of gasification are insufficiently known, and the reactivity of different chars varies over a wide range;
- Products of coal disintegration due to rapid heating are not well known and influence fines production;
- Insufficient data are available to estimate the maximum bed temperature that can be achieved without defluidization due to sintering of ash;

- It is not currently possible to predict the minimum oxygen requirement needed to prevent clinkering near the inlet nozzle;
- Further work is needed to understand how additives allow increase in bed temperatures;
- Further work is needed to understand how additives reduce the caking tendency of bituminous coals; and
- Further study is needed of the dependence of the gasification rate on carbon conversion, particularly in the 70 to 100 percent conversion range.

B.4.4 Key Technical Issues

B.4.4.1 Oxidant Requirement

Oxygen is required to fuel partial combustion of coal in the KRW gasifier. The heat released during combustion is used to supply the heat of reaction for the endothermic gasification reactions occurring in the fluidized bed. Heat released during combustion is also required to bring the gasification reactants (e.g., steam) to reaction temperature. Thus, any thermal losses, such as conduction through the gasifier vessel walls, radiation to the gasifier freeboard, recycle of cooled recycle gas or fines to the gasifier, and excess steam above reaction requirements, will impose increased demands on oxygen consumption. If oxygen consumption becomes too high, the gasifier will become a partial combustor in which all carbon conversion is achieved by combustion. Steam would then only participate in water-gas shift from carbon monoxide to hydrogen. In gasification reactions, the steam reacts directly with carbon in the coal to form carbon monoxide and hydrogen in an endothermic reaction (Shinnar et al, 1988).

B.4.4.2 Steam Requirement

Steam is required as a reactant for gasification reactions. Because the kinetics of the gasification reactions are finite, excess steam is required for this purpose. In addition, steam is required as a thermal diluent to reduce the temperature in the combustion zone near the oxygen inlet. However, any steam requirement in excess of that needed for combustion zone cooling poses a sensible heat load on the gasifier, which may in turn increase the demand for oxygen to fuel the combustion reactions to heat the excess steam. Also, in the KRW PDU, steam is used as a fluidization gas to prevent large char particles from falling through the ash annulus with the heavier ash agglomerates (Shinnar et al, 1988). The ash removal design of the KRW gasifier thus may impose constraints on the minimum steam requirement necessary for gasifier operability.

The steam requirement may be reduced for air-blown systems, compared to oxygen-blown systems. This is due to the thermal dilution in the combustion jet provided by the nitrogen in the oxidant, which offsets the need for some of the steam.

B.4.4.3 Carbon Conversion

For the oxygen blown KRW-based system with cold gas cleanup, the design assumptions assumed in most studies for eastern coal do not agree with operating maps developed by Shinnar, Avidan, and Weng (1988) for the performance of the KRW gasifier. The key assumptions in the conceptual designs are for carbon conversion efficiency, oxygen-to-coal ratio, and steam-to-coal ratio. Assumptions for several conceptual design studies, theoretical modeling studies, and PDU tests are given in Table B-12. The carbon conversion efficiencies assumed in design studies are typically around 95 percent (e.g., Dawkins et al, 1985; Smith and Smelser, 1987; Earley and Smelser, 1988). The oxygen-to-coal ratios are typically assumed to be 0.3 lbmole oxygen/lbmole carbon in the coal. The steam-to-oxygen ratios vary more widely in the design studies, from about 0.4 to 1.6 lbmole steam/lbmole oxygen in oxygen-blown systems. However, an operating map developed by Shinnar, Avidan, and Weng (1988) for gasifier performance with a Pittsburgh No. 8 coal at 230 psi indicates that at a temperature of about 1,850 ^{O}F and a carbon conversion of about 95 percent, the required feed ratios are 0.5 lbmole oxygen/lbmole carbon in the coal and 5.5 lbmole steam/lbmole oxygen. Alternatively, for the conditions assumed in the conceptual studies (e.g., Dawkins et al, 1985), the carbon conversion is estimated from the operating map (neglecting differences in gasifier pressure) to be about 75 percent.

The carbon conversion depends on many factors. The gasification reaction rate is a complex and non-linear function of many variables, such as temperature, pressure, carbon conversion rate, and reactant and product partial pressures. The reaction rate is a strong function of temperature. The reaction rate is a nonlinear function of pressure and limited knowledge of the pressure dependency makes it difficult to predict gasifier performance at pressures higher than those experienced in the PDU. While increased pressure will allow a greater gas throughput in the gasifier, the reaction rate may not increase proportionately to allow for constant or increased carbon conversion. Thus, overall performance may suffer as pressure is increased; although throughput is increased, overall carbon conversion may be decreased (Shinnar et al, 1988).

As the carbon conversion rate increases, the gasification reaction rate decreases due to finite kinetics. Recycled fines which have already undergone significant carbon conversion in particular are believed to have very low gasification reaction rates. Also, fines introduced above the combustion zone, as in the KRW design, have a low residence time and a high recycle rate (Shinnar et al, 1988). Thus, fines may recycle through the system several times before conversion or, in some cases, escaping past the cyclone.

The bed height in the gasifier influences the carbon conversion rate, presumably due to increased residence time and increased opportunities for reactants to interact in the bed. The carbon conversion is believed to increase linearly with the logarithm of bed height. Thus, there are very small benefits to incrementally increasing the bed height for large beds. Bed heights of 40 to 60 feet are believed to be economical. Baffles may improve mixing and avoid bypassing (channeling) in the bed (Shinnar et al, 1988).

The gasification rate increases as the partial pressure of steam, the key gaseous reactant, increases. Thus, design features such as raw gas recycle, which reduce the steam partial pressure, also reduce the gasification rate.

The carbon retention in gasifier bottom ash appears to be low for the KRW gasifier. An apparently common assumption in conceptual design studies is that the carbon retention in the bottom ash is less than one percent of the carbon in the feed coal (e.g., Earley and Smelser, 1988). These assumptions appear to be supported by KRW PDU testing for cases with in-bed desulfurization using calcium-based sorbents (Haldipur et al, 1989).

The major source of carbon loss in the KRW gasifier is commonly reported to be elutriated fines that escape the recycle cyclone. However, detail on the carbon content of such fines, the fines elutriation rate, and the capture efficiency of the recycle cyclone is somewhat lacking in the open literature. It appears that the typical capture efficiency assumed is 95 percent of the inlet fines on a mass basis (Blinn et al, 1989).

The KRW PDU gasifier does not operate near equilibrium. It appears unlikely that a scaled-up KRW gasifier will operate near equilibrium on eastern bituminous coal without changes in design or the use of a catalyst. For example, using dolomite or limestone as a sorbent for in-bed desulfurization, a catalytic effect of six (a six-fold improvement in relative reaction rate) was realized in the KRW PDU, as reported by Shinnar et al (1988). The increased conversion resulting from the catalytic effect implies that carbon conversion can be increased at a given operating temperature. Alternatively, the operating temperature can be increased to further improve the gasification rate.

By comparison with eastern bituminous coals, western subbituminous coals are non-caking, non-swelling, and up to two orders-of-magnitude more reactive in gasification. The non-caking nature of the western coals would facilitate feeding some or all of the coal directly into the fluidized bed, rather than through the combustion jet. Recycled fines could then be combusted by feeding them to the jet, improving the overall carbon conversion by burning high conversion char which would be slow to gasify.

Eastern coals must be fed through the combustion jet because the intense mixing and devolatilization that occurs during combustion breaks up or prevents formation of char agglomerates. For this reason, the KRW gasifier is capable of directly handling a caking coal with no pretreatment, at the expense of reduced gasifier efficiency.

In the air-blown mode, Shinnar and Weng (1989) report modeling results which indicate that, when using a calcium-based sorbent for 85 percent in-bed desulfurization at 315 psia, an oxygen-to-carbon molar ratio of 0.45 and a steam-to-oxygen ratio of 0.5 is required for 95 percent carbon conversion at 1,950 °F.

In a series of KRW PDU tests, bituminous coals ranging from 1.5 to 4.5 percent sulfur content were gasified in air-blown mode using calcium-based sorbents. The carbon conversions estimated over 2,634 hours of testing ranged from 70 to 95 percent, depending on factors such as bed temperature and fines recycle efficiency. The bed temperature ranged from 1,800 to 2,000 °F, and the gasifier pressure was 245 psia. The presence of sorbent in the bed was reported to retard the passage of fines through the bed, presumably increasing fines residence time and increasing the per-pass consumption of carbon in recycle fines. Fines consumption was also improved due to the catalytic effect of the sorbent on the gasification reactions. In addition, losses associated with recycle raw gas were reported to be reduced when using calcium based sorbent, because changes in the fines particle size distribution and reduction in fines carryover reduced the the amount of recycle gas required (Haldipur et al, 1989).

Haldipur et al (1989) report that the fines loading into the recycle cyclone is on the order of 100,000 to 200,000 ppmw and that the fines exiting the recycle cyclone with the coal gas are on the order of 20,000 ppmw. Blinn et al (1989) assert that the carbon conversions in the PDU are low due to less than optimum cyclone performance that allows high fines loss. They report that a cyclone capture efficiency of 95 percent would result in a carbon conversion rate of 97 percent for a gasifier-desulfurizer operating on Eastern coal. Blinn et al also indicate that the particulates captured downstream of the recycle cyclone contain a significant amount of carbon, but do not indicate quantitatively the concentration.

Fines may be generated from fines entering with the coal feed that pass through the gasifier, fracturing of larger bed particles, or chemical destruction of large coal feed particles. Fines may be consumed by chemical consumption (e.g., gasification) or agglomeration with the ash. Fines escaping the cyclone are thus derived from various mechanisms of fines production in the gasifier (Haldipur et al, 1988). Fines which originate from the feed coal or from physical breakup of the feed coal are likely to have a high carbon concentration, while fines which are obtained from chemical consumption may or may not have a high carbon concentration. For example, some fines may be obtained from the combustion zone, in which devolatilization occurs. This may tend to increase the carbon concentration of the remaining particles.

B.4.4.4 In-Bed Desulfurization

Sulfur removal during gasification using calcium-based sorbents has been demonstrated in the KRW process development unit (PDU). The sulfur absorption rate depends on the molar calcium-to-sulfur ratio (Ca/S) of the sorbent, the type of sorbent material used, and the sulfur content of the coal, among other factors. For a given sorbent, the sulfur removal increases as the Ca/S ratio is increased. A few different types of calcium-based sorbents have been used in testing, including high-calcium, magnesium, and dolomitic limestones. The dolomitic limestones have shown the most consistent sulfur removal performance. The sulfur removal efficiency is ultimately limited by equilibrium constraints, and the equilibrium sulfur removal efficiency increases as the coal sulfur content increases (Schmidt, Sadhukhan, and Lin, 1989).

The degree to which the equilibrium limits can be approached depend on kinetic and mass transfer variables. For example, minimizing gasifier temperature promotes sulfur absorption, as does minimizing the water vapor and carbon dioxide content of the fuel gas. A high carbon dioxide content may inhibit calcination of the sorbent, although PDU tests indicate that sorbent calcines almost completely in most cases. Maximizing the bed calcium content and the ratio of bed depth to gas velocity (a measure of gas residence time in the bed) also improves the desulfurization efficiency. Coals with high sulfur content and sorbents with high reactivity also favor high removal efficiency (Haldipur et al, 1989).

KRW Energy Systems, Inc. has developed an in-bed desulfurization model which relies on empirical data regarding reaction kinetics and measurements from PDU testing regarding several key performance parameters. The model is based on adjusting estimates of sulfur removal efficiency based on equilibrium calculations with a correction term involving calcium concentration in the bed, gas residence time, the sulfidation rate constant, the fraction of sulfided calcium in the bed, and an empirical reaction order constant. Prediction of several key parameters, such as the weight fraction of calcium in the bed and the density of the bed, is reported to be difficult for commercial gasifiers. Furthermore, prediction of the H_2O and CO_2 content of the fuel gas, which are required to estimate equilibrium sulfur removal conditions, is also reported to be difficult, requiring the development of a fuel gas composition model. KRW Energy Systems has developed empirically-based regression correlations specific to the operating conditions of the PDU for the given coal to try to analyze experimental data (Haldipur et al, 1989). The model as it is reported thus does not appear directly appropriate for predicting the performance of a commercial scale gasifier.

In the KRW PDU, 85 to 95 percent overall desulfurization is reported, depending on the coal, using either limestone or dolomite. However, not all of the sulfur in the coal is released during gasification; a portion of it leaves the gasifier in the bottom ash. A reported conservatively high assumption is that 10 percent of the sulfur is not released in the gasifier for an Eastern coal (Haldipur et al, 1989). The amount of sulfur not released from the coal is believed to be higher in the case without in-bed desulfurization. Haldipur et al (1989) report that 10 to 20 percent of the sulfur in the coal was not released in PDU tests conducted without sorbent. Furthermore, it is reported that in previous PDU tests, typically 15 to 20 percent of the sulfur in the coal was not released. These data are based on sulfur balances. Based on an environmental characterization of the KRW PDU, Radian reports a combined 20 percent sulfur retention in ash, tertiary cyclone solids, and hot gas particulates for a Pittsburgh No. 8 coal with no sorbent (Scheffel and Skinner, 1988). The assumptions in conceptual design studies appear to be no sulfur retention or perhaps a nominal one percent sulfur retention in the bottom ash. However, more reasonable assumptions appear to be 10 to 20 percent sulfur retention without a sorbent, and a lower sulfur retention with sorbent. Based on the report by Haldipur et al (1989), it appears that the influence of in-bed calcium-based sorbents on coal sulfur release is not fully understood, but is believed to promote sulfur release.

The nominal expected in-bed desulfurization for a high sulfur (e.g., 4.5 percent) coal is 90 percent (Haldipur et al, 1988). For dolomitic limestone, this removal rate is commonly assumed in conceptual design studies to occur at a Ca/S ratio of 1.8 (e.g., Smelser, 1986; Earley and Smelser, 1988). However, the actual Ca/S ratio required to achieve 90 percent sulfur removal for a given coal and sorbent also depends on the residence time in the gasifier. Based on a graph presented by Haldipur et al (1988) for the PDU using a 4.5 percent sulfur coal, 90 percent sulfur removal may be achievable with dolomitic limestone at a Ca/S ratio as low as 1.4 for a gas residence time of near 18 seconds, or a Ca/S ratio of about 1.7 at a residence time of about 13 seconds. In contrast, for a high-calcium limestone, the reported Ca/S ratios for the same residence times are approximately 2.4 and 4, respectively.

In a more recent study, Haldipur et al (1989) estimate the Ca/S ratios for dolomite and high-calcium limestone required to achieve 90 percent sulfur removal for Pittsburgh No. 8 coal as follows:

> "In a commercial-scale gasifier, 90 percent sulfur removal is predicted when feeding a dolomitic limestone at a calcium-to-sulfur molar feed ratio ranging from 1.2 to 1.5. For a high-calcium limestone, a required feed ratio from 2.3 to 2.8 is projected." (p. 66, vol 1)

Haldipur et al (1989) report that the estimated gas residence time in the PDU gasifier is about 15 seconds, but that residence times as high as 20 seconds are expected to be feasible for commercial-scale units. For dolomitic limestones, Haldipur et al (1989) report that Ca/S ratios as low as 1.1 may be possible, with a more likely value of 1.4 expected for either Pittsburgh No. 8 or Upper Freeport coals. Similarly, high-calcium limestones, which have a lower reaction rate constant than dolomitic limestones, are expected to require a minimum Ca/S ratio of 2 and a likely value of about 2.6. The basis for these ranges is not given, and it is unclear whether the ranges are based on possible variations in design or operating conditions in a commercial gasifier, uncertainty in estimates of key performance parameters used to estimate the Ca/S ratio, or statistical error based on regression analysis of PDU test data extrapolated to try to predict commercial-scale gasifier performance.

Reduction of the Ca/S ratio can be achieved by: lowering the gasifier pressure, lowering the gasifier temperature, lower coal ash content, higher coal sulfur content, and smaller sorbent particle diameter, in addition to other effects discussed previously (Haldipur et al, 1989).

B.4.4.5 Effect of Sorbent on Gasifier Performance

There is general agreement that calcium-based sorbents catalyze the gasification reactions, increasing the reactivity of eastern coals. For example, the reactivity of Pittsburgh No. 8 coal appears to triple with the use of a limestone sorbent (Floyd and Agrawal, 1989). This implies that carbon conversion efficiencies should be higher with in-bed desulfurization. A sorbent may also reduce the caking tendency of bituminous coals, thereby allowing the gasifier to operate at higher temperatures, which also would tend to increase carbon conversion as discussed previously (Shinnar, Avidan, and Weng, 1988). Based on PDU tests with limestone, the design operating temperature for the Appalachian clean coal technology demonstration project featuring a KRW gasifier using a high sulfur eastern coal was raised from 1,850 ^{o}F to 1,900 ^{o}F (Banchik, Buckman, and Rath, 1988).

There appears to be less consensus on the effect of calcium sorbents on the environmental performance of the gasifier. For example, a conceptual design study (Earley and Smelser, 1988) reports that ammonia production is less with in-bed desulfurization than without, particularly for high calcium limestone. But an environmental study by Radian (Scheffel and Skinner, 1988), based on testing of the PDU, does not indicate any reduction in ammonia production with dolomite compared to without a sorbent. Earley and Smelser (1988) also report that pilot plant data indicate reduced production of methane with sorbents. The increased gasification rate resulting from use of a calcium sorbent is reported to increase fines consumption in the gasifier and reduce fines elutriation (Banchik, Buckman, and Rath, 1988).

B.4.4.6 Ammonia Production

For the KRW system with cold gas cleanup, ammonia yield from the gasifier is not an air pollution concern because the ammonia is almost completely removed by wet scrubbing. However, for the hot gas cleanup system, the ammonia yield will affect NO_X emissions in the gas turbine combustor because ammonia is not removed in the hot gas cleanup systems assumed in most studies (e.g., Cincotta, 1984). In a conventional gas turbine combustor, about 70 percent of the ammonia in the fuel gas will be converted to NO_X. If the ammonia content of the fuel gas is sufficiently high, alternative gas turbine combustor designs, such as rich/lean staged combustion or catalytic combustion, may be required to maintain NO_X emissions within applicable standards.

Some conceptual design studies report that the gasifier ammonia yield decreases for in-bed desulfurization compared to gasification without a sorbent and depends on the type of sorbent used (e.g., high calcium limestone or dolomite) (Earley and Smelser, 1988). For example, a conceptual design study (Earley and Smelser, 1988) reports that ammonia production is less with in-bed desulfurization than without, particularly for high calcium limestone. In a summary of testing with the KRW PDU, Haldipur et al (1989) report that when using sorbent in the gasifier the nitrogen yield of the gasifier corresponded to an equivalent of conversion of 0.6 to 8.8 weight percent of the nitrogen in the coal.

An environmental study by Radian (Scheffel and Skinner, 1988) based on testing of the PDU does not indicate any reduction in ammonia production with dolomite compared to gasification without a sorbent. In two set points without dolomite injection, the ammonia production was equivalent to a conversion of 9.1 and 7.8 percent of the coal nitrogen. In a set point with dolomite, the ammonia production was equivalent to a conversion of 10.6 percent of the coal nitrogen. The tests were conducted in air-blown mode using Pittsburgh No. 8 coal with a 1.4 percent nitrogen content on a dry basis.

For an oxygen-blown system operating without in-bed desulfurization, a recent design study indicates that an equivalent of about 25 percent of the coal-bound nitrogen is converted to ammonia. Because a wet scrubbing system is employed for gas cooling and cleanup in this design, the ammonia is removed prior to firing the fuel gas in the gas turbine (Gallaspy et al, 1990).

B.4.4.7 Spent Sorbent Sulfation

For the air-blown KRW-based system with in-bed desulfurization, the spent sorbent is a component of the gasifier bottom ash waste stream. This sorbent contains calcium sulfide (CaS) which is not acceptable for landfilling. Untreated spent sorbent can release sufficient quantities of hydrogen sulfide to violate RCRA standards when exposed to sulfuric acid. To comply with RCRA, a portion of the CaS must be oxidized to the more stable calcium sulfate ($CaSO_4$), which is then suitable for disposal. The oxidation of calcium sulfide can occur by roasting in air, at a temperature high enough for rapid conversion but low enough to avoid excessive SO_2 emissions. A pilot plant has achieved 20 to 30 percent conversion of calcium sulfide in spent limestone in the waste stream, which is claimed to be sufficient to meet the nonhazardous waste criteria of the RCRA standards. In the case of dolomite, the conversion rate is significantly higher, and it is reported to be about 55 percent. The leachate from the treated waste also meets the toxicity requirements for Ag, As, Ba, Cd, Cr, Hg, Pb, and Se under RCRA (Haldipur et al, 1989).

For a commercial plant design, the sulfation unit is envisioned to be a circulating fluidized bed combustor, which is expected to operate at temperatures appropriate for good sulfide conversion and for minimal SO_2 emissions. The gasifier ash waste stream also includes unconverted carbon. In addition, nonrecycled fines captured downstream of the gasifier in high efficiency cyclones or other hot gas particulate control devices also contain unconverted carbon. These collected fines can also be sent to the sulfation unit for combustion. Therefore, a sulfation unit can recover the heating value of unconverted carbon. The heat released from the sulfation reaction and combustion of unconverted carbon can be used to generate steam, and thus to reduce the energy penalty associated with incomplete carbon conversion in the gasifier (Earley and Smelser, 1988).

The sulfation unit is an additional emission source. Fluidized bed combustors generally have more uniform flame temperatures than conventional types of systems; therefore, the NO_x emissions from these systems are comparatively low. The sulfur emissions in the sulfation unit can stem from several sources, including sulfur retained in the coal bottom ash, sulfur retained on unconverted fines, and sulfur chemically bound with calcium. By proper control of the sulfation unit combustion temperatures, sulfur release from the sorbent is expected to be minimum. However, according to PDU data, it appears that 10 to 20 percent of the sulfur in the coal feed is retained in the bottom ash and tertiary solids captured from the fuel gas. It is unclear, however, if the sulfur in the ash and fines solids is released in the sulfation unit and, if so, if it is captured by unreacted calcium sorbent in the fluidized bed (recall that the molar feed ratio of calcium to the gasifier exceeds stoichiometric requirements) or emitted as SO_2. It is unclear if perhaps additional sorbent might be required to control SO_2 emissions from the sulfation unit.

According to Earley and Smelser (1988), the SO_2 and NO_x emission rates from a sulfation circulating fluidized bed combustor are expected to be 0.01 lb SO_2/MMBtu and 0.15 lb NO_x/MMBtu, respectively. These are well below the NSPS limits for steam generators fired with bituminous coal. Whether the NSPS standard is applicable to this emission source is not yet known. Particulate emissions are expected to be controlled by a fabric filter to less than 0.03 lb/MMBtu.

To achieve more complete conversion of sulfide to sulfate, additives containing sodium carbonate have ben tested. According to Haldipur et al (1989), under a fairly narrow range of conditions, the additive promotes near complete conversion of calcium sulfide to oxidized products. The conditions evaluated in a small-scale test study include a temperature of 1,831 ^{o}F and an oxygen concentration of 10 percent or higher. According to Earley and Smelser (1988), the cost of the additive would have a small effect on levelized plant costs.

B.4.5 Technical Judgments About Uncertainties

Originally, a three-part briefing package was to be distributed to engineers at DOE/METC to obtain expert judgments regarding uncertainties in selected KRW gasifier and related parameters. The briefing package was prepared. However, because of personnel constraints at DOE/METC, the briefing package could not be distributed. Furthermore, engineers at KRW Energy Systems declined to participate in the current study, citing concerns about the proprietary nature of data that might be released. Therefore, judgments regarding initial uncertainties for a probabilistic screening study are made here by the author based on information in the published literature reviewed in the preceding Sections B.4.1 through B.4.4. Each process parameter which was considered for probabilistic treatment is discussed below. The parameters that are treated probabilistically are given in Table B-13. In most cases, the uncertainty is conditional on whether a sorbent is used for in-bed desulfurization.

Table B-13. Summary of Assumed Uncertainties for the KRW Gasifier Process Area.

Description and Units	Distribution	Parameters[a]			
Overall Carbon Conversion, wt-% of feed coal carbon					
Without Sorbent	Triangular	75	to	95	(95)
With Sorbent	Triangular	90	to	97	(95)
Oxygen/Carbon Ratio, lbmole O_2/lbmole C					
Without Sorbent	Uniform	0.33	to	0.35	
With Sorbent	n/a	0.45			
Steam/Oxygen Ratio, lbmole H_2O/lbmole O_2					
Without Sorbent	Uniform	1.1	to	1.6	
With Sorbent	Uniform	0.4	to	0.5	
Carbon Retention in Bottom Ash, wt-% of coal feed carbon					
Without Sorbent	Uniform	0.5	to	1.0	
With Sorbent	Uniform	0.5	to	1.0	
Sulfur Retention in Bottom Ash/LASH, mol-% of inlet sulfur					
Without Sorbent	Triangular	10	to	20	(15)
With Sorbent	Triangular	85	to	95	(90)
Calcium-to-Sulfur Ratio, lbmole Ca/lbmole S					
Limestone	Triangular	2	to	2.8	(2.6)
Dolomite	Triangular	1.1	to	1.8	(1.4)
Gasifier Temperature, °F					
Without Sorbent	n/a	1,850			
With Sorbent	Triangular	1,900	to	1,950	(1,900)
Gasifier Ammonia Yield, Equiv. fraction of coal N converted to NH_3					
Without Sorbent	Triangular	0.10	to	0.25	(0.25)
With Sorbent	Triangular	0.005	to	0.10	(0.10)
Relative Gasifier Coal Throughput (465 psia, reference to oxygen-blown, no sorbent)					
Oxygen-blown, no sorbent	N/A	1.0			
Oxygen-blown, sorbent	Uniform	1.2	to	1.3	
Air-blown, no sorbent	Uniform	0.5	to	0.6	
Air-blown, sorbent	Uniform	0.7	to	0.8	
Sulfation Unit Emissions, lb /MMBtu					
SO_2	Triangular	0.01	to	0.05	(0.01)
NO_x	Triangular	0.10	to	0.20	(0.15)
Particulate Matter	Triangular	0.02	to	0.03	(0.03)
Sulfation Unit Sulfide Conversion	Uniform	30	to	90	
Sulfation Unit Carbon Conversion	Triangular	90	to	98	(95)

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

Overall Carbon Conversion and Feed Ratios. Carbon conversion is related to many aspects of gasifier design and operation, as discussed in Section B.4.4.3. For the oxygen-blown case, carbon conversions of around 95 percent are typically assumed in the literature. However, for a given oxygen/carbon and steam/oxygen ratio in a commercial scale gasifier, there appears to be uncertainty regarding the carbon conversion that would actually be obtained. The approach taken here is to use typical values for the reactant feed ratios based on published conceptual design studies and to assume a resulting characterization of uncertainty in carbon conversion. Most of the conceptual design studies that have been reviewed assume an oxygen-to-carbon ratio of 0.33 to 0.34 (see Table B-12). These studies also assume steam-to-oxygen ratios of 0.36 to 1.59, with most estimates clustered above 1.1. While carbon conversions of around 95 percent were assumed in these studies, Shinnar et al (1988) indicate that such a high conversion rate cannot be achieved with such low steam/oxygen ratios. In fact, modeling results from Shinnar et al suggest that a carbon conversion of only 75 percent might result from the proposed oxygen and steam feed ratios, although these values are dependent on design and operating details of a particular gasifier. For the initial screening study, the carbon conversion will be assumed to vary from 75 to 95 percent,with values near the upper limit assumed to be more likely than those near the lower limit. A small range of uncertainty for the oxygen/carbon ratio is assumed, representing the variability observed across design studies. The range of values reported in design studies for the steam/oxygen ratio will form the basis for assigning uncertainty to this parameter. However, it is assumed that the steam/oxygen ratio will have a correlation with the carbon conversion rate.

For the air-blown system with in-bed desulfurization, the typically assumed oxygen/carbon ratio is about 0.45, based on a draft of the final report by Southern Company Services to DOE/METC of a study of air-blown KRW-based IGCC systems. According to Shinnar and Weng (1989), a steam-to-oxygen ratio of about 0.5 is required for 95 percent carbon conversion at 315 psia and 1,950 ^{o}F. An estimate used in a conceptual design study of an air-blown system is somewhat lower than this value, at 0.37 lbmole steam/lbmole oxygen. For initial studies, an oxygen/carbon ratio of 0.45 and a steam/oxygen ratio of 0.5 will be used. The steam/oxygen ratio will be assigned an uncertainty ranging from 0.4 to 0.5. A higher average carbon conversion rate is assumed for the in-bed sorbent case compared to without sorbent, because the sorbent is believed to catalyze the gasification reactions, prevent caking of char particles, and allow higher operating temperatures that also increase reaction rates.

The bulk of unconverted carbon is likely to be contained in unrecycled fines leaving the KRW gasifier. These fines could be combusted in the sulfation unit to recover the heating value of the unconverted carbon in the steam cycle. It is assumed that about 0.5 percent of the carbon in the coal feed would be retained in the gasifier bottom ash, with the rest of the unconverted carbon contained in unrecycled fines.

Sulfur Removal. Whether in-bed desulfurization is employed, a portion of the sulfur in the coal is expected to be retained in bottom ash or unrecycled fines. As indicated in the technical background discussion, perhaps 10 to 20 percent of sulfur in the coal is not released even when a sorbent is not used. When a sorbent is used, perhaps 85 to 95 percent of the sulfur entering the gasifier, both in the feed coal and from the recycle off-gas stream from the external desulfurization unit, is captured in the bottom ash and limestone. The removal rate may be variable depending on coal properties. These ranges are used as the basis for characterizing uncertainty, with the midpoint of the ranges assumed to be the mode of a triangular distribution. Therefore, the assumed "most likely" values for sulfur capture are 15 percent without sorbent and 90 percent (per pass) with sorbent.

In-Bed Desulfurization Sorbent. According to the technical information reviewed in Section B.4.4.4, there appears to be uncertainty regarding prediction of the calcium-to-sulfur molar ratio needed to obtain about 90 percent sulfur removal. The reported range of Ca/S values that are predicted for obtaining about 90 percent (per pass through the gasifier) sulfur removal are adopted directly as a basis for characterizing uncertainty. For two types of sorbent, limestone and dolomite, upper, lower and "most likely" values are reported. These are represented using triangular distributions.

Gasifier Bed Temperature. For the oxygen-blown gasifier without in-bed desulfurization, the universally assumed operating temperature for bituminous coal is 1,850 ^{o}F, and that temperature is adopted here without uncertainty. For the case with in-bed desulfurization, the presence of sorbent is believed to reduce the caking tendency of coal and char particles and thus allow higher operating temperatures. Some studies assume a 1,900 ^{o}F temperature, while a modeling study has assumed 1,950 ^{o}F. The actual operating temperature that may be sustained in a commercial gasifier may be somewhere between 1,900 and 1,950 ^{o}F; therefore, this range is assumed as the basis for an uncertainty distribution. There is some indication, based on the reviewed technical information, that an increase in bed temperature will tend to increase carbon conversion and decrease sulfur capture.

Ammonia Yield. Based on the previous technical discussion, it appears that ammonia yield with sorbent is expected to be lower than without sorbent. Without sorbent, ammonia yield may be equivalent to conversion of 25 percent of coal-bound nitrogen, although some tests show it to be around 10 percent. With sorbent, tests have indicated equivalent nitrogen conversions as low as 0.6 percent up to about 10 percent. For systems with cold gas cleanup, virtually all of the ammonia will be removed during wet scrubbing. For the case with hot gas cleanup, the ammonia may be converted to NO_x in the gas turbine combustor. Without sorbent, it is assumed that 10 to 25 percent of the coal-bound nitrogen may be converted to ammonia. With sorbent, a coal nitrogen conversion of 0.5 to 10 percent is assumed. In both cases, the upper limit is assumed as the mode of a triangular distribution.

Gasifier Coal Throughput. Little data exist to estimate the gasifier coal throughput as a function of gasifier pressure. One of the key factors affecting coal throughput is believed to be whether an in-bed sorbent is employed. Assumptions used in the draft final report of a recent design study suggest that coal throughput may be increased as much as 50 percent with a sorbent compared to without. This might be attributable to the catalytic effect of the sorbent on the gasification reaction rate, as well as the increase in reaction rate that is associated with the higher bed temperatures which can be obtained when using sorbent. An increase in gasifier coal throughput can have a substantial effect on the capital cost of the IGCC system, because fewer gasifier vessels would be required.

From a comparison of several design studies, it appears that gasifier coal throughput is greatest for oxygen-blown gasification with sorbent and lowest for air-blown gasification without sorbent. The relative gasifier coal throughputs for several cases are shown in Table B-13. These are based on the ratio of moisture- and ash-free (MAF) that is feed to a single gasifier as indicated in several studies (Dawkins et al, 1985; Earley and Smelser, 1988; Gallaspy, 1990; SCS, 1991; Smith and Smelser, 1987). These ratios should be considered preliminary. However, they are expected to provide at least an appropriate qualitative indication of the effect of oxidant type and sorbent utilization on gasifier coal throughput.

Sulfation Unit. The sulfation unit will emit some amount of SO_2 and NO_x. One estimate is that the emissions will be 0.01 lb SO_2/MMBtu and 0.15 lb NO_x/MMBtu, based on the heating value of the solids fed to the roasting unit. The latter may lead to a significant increase in plant NO_x emissions. Other studies are less specific about these emissions. Particulate emissions can be controlled by a fabric filter to below NSPS requirements.

For SO_2, it is assumed that there is some possibility that the emission rate could be higher than expected. There is no detail in the available literature regarding the fate of unreleased sulfur in the bottom ash or sulfur captured in the limestone when combusted in a fluidized bed combustor. Certainly, if a portion of the unreleased sulfur in the bottom ash were released, SO_2 emissions could increase. A nominal five-fold increase is assumed as possible, with the most likely value equal to the published estimate. This uncertainty range should be considered illustrative for the purpose of determining whether uncertainty in sulfation unit SO_2 emissions might contribute significantly to uncertainty in plant SO_2 emissions.

The ranges of the uncertainties for sulfation unit carbon conversion are based on various values reported in the literature.

Correlations. Proposed correlations structures for both the oxygen-blown case without sorbent and the air-blown case with sorbent are shown in Tables B-14 and B-15, respectively. For the oxygen-blown system without in-bed desulfurization, it is assumed that carbon conversion will tend to improve as the steam/oxygen ratio is increased. Therefore, a nominal correlation of 0.75 is assumed between carbon conversion and the steam/oxygen ratio. The purpose of inducing this correlation between these two parameters is to identify whether such a correlation significantly changes the answer obtained from the modeling. If the correlation has a significant effect on model results, that would imply that further effort should be developed to modeling the interdependence between the two parameters that is not captured by the current model.

For the air-blown system with in-bed desulfurization, a similar correlation is assumed between the carbon conversion rate and the steam/oxygen ratio. The sulfur removal rate is assumed to be positively correlated with the calcium-to-sulfur ratio and negatively correlated with the gasifier temperature. In addition, the carbon conversion is assumed to be positively correlated with the gasifier temperature. These correlations are intended to be representative of trends observed in the literature. To the extent that these correlations have an important effect on model results, they would indicate that further model development is warranted to better characterize these interdependencies.

Table B-14. Proposed Correlation Matrix for KRW Gasification Uncertain Parameters: Oxygen-Blown Case Without Sorbent

Description	Number	Assumed Correlation Coefficients[a] 1	2	3	4	5	6
Carbon Conversion	1	1.00					
Oxygen/Carbon Ratio	2	--	1.00				
Steam/Oxygen Ratio	3	0.75	--	1.00			
Carbon Retention in Bottom Ash	4	--	--	--	1.00		
Sulfur Retention in Bottom Ash	5	--	--	--	--	1.00	
Gasifier Ammonia Yield	6	--	--	--	--	--	1.00

[a] For uncorrelated parameters, a dashed line "- -" is shown.

Table B-15. Proposed Correlation Matrix for KRW Gasification Uncertain Parameters: Air-Blown Case With Sorbent

Description	No.	Assumed Correlation Coefficients[a] 1	2	3	4	5	6	7	8	9
Carbon Conversion	1	1.00								
Oxygen/Carbon Ratio	2	--	1.00							
Steam/Oxygen Ratio	3	.75	--	1.00						
Carbon Retention in Bottom Ash	4	--	--	--	1.00					
Sulfur Retention in Bottom LASH	5	--	--	--	--	1.00				
Calcium-to-Sulfur Ratio	6	--	--	--	--	0.5	1.00			
Gasifier Temperature	7	0.5	--	--	--	-0.5	--	1.00		
Gasifier Ammonia Yield	8	--	--	--	--	--	--	--	1.00	
Gasifier Throughput	9	--	--	--	--	--	--	--	--	1.00

[a] For uncorrelated parameters, a dashed line "- -" is shown.

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B.5 Fixed-Bed Zinc Ferrite Process Area

This section concerns a concept for removing hydrogen sulfide from the high temperature, high pressure raw product gas exiting a gasifier. Hydrogen sulfide is the principle gaseous sulfur species in coal gas and must be removed for compliance with emission regulations and for protection of the gas turbine from deposition of alkali sulfates. The fixed-bed zinc ferrite desulfurization process has been proposed as part of both a fixed-bed dry-ash Lurgi gasifier-based IGCC system and a fluidized-bed agglomerated-ash KRW gasifier-based system. The Lurgi-based system is shown in Figure B-1 (see Section B.3), and the KRW-based system is shown in Figure B-10 (see Section B.4).

B.5.1 Process Description

The zinc ferrite process area consists of multiple trains of two-vessel systems, in which one vessel is in sulfur absorption mode while the other vessel is in regeneration mode, as shown in Figure B-13. Absorption occurs at high pressure (e.g., 300 psia) and an inlet syngas temperature of around 1,100 ^{o}F. Regeneration is assumed to occur at the same pressure. Both IGCC systems are assumed to consume an Illinois No. 6 coal. The product of regeneration is an offgas containing SO_2. This offgas may be recycled to the gasifier when in situ desulfurization is employed with a sorbent fed to the gasifier, or sent to a sulfuric acid plant for byproduct recovery.

Two modes of desulfurization are envisioned for zinc ferrite systems. "Bulk desulfurization" is the term given for application of the zinc ferrite process to gas streams with high concentrations (e.g., 5,000 ppmv) of gaseous sulfur species, and "polishing desulfurization" is the term for application of the process to streams with low sulfur concentrations (e.g., 1,000 ppmv or less). For the Lurgi-based system, the zinc ferrite process is used for "bulk" desulfurization. For the KRW-based system, the zinc ferrite process can be used either in bulk mode or, in combination with gasifier in-bed desulfurization, in polishing mode. One high efficiency cyclone is assumed to be located upstream of the absorber and one cyclone is located downstream of the absorber. The upstream cyclone is required to prevent the build-up of particles in the sorbent bed, which could lead to increased pressure drop across the absorber. The downstream cyclone may be required to prevent entrained sorbent particles and ash particles from entering the gas turbine combustor. The systems use advanced high firing temperature gas turbines.

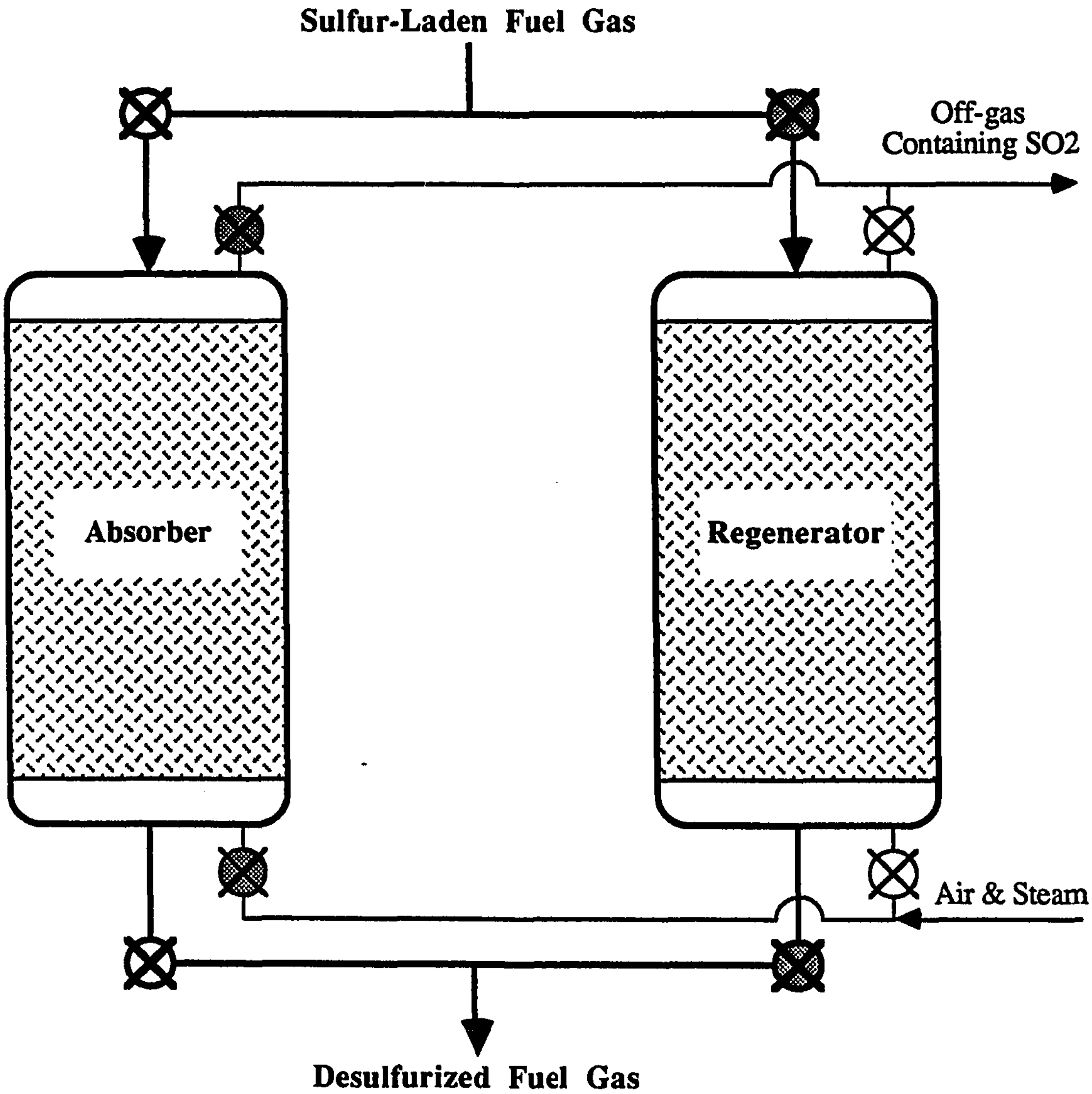

Figure B-13. Simple Schematic of the Fixed-Bed Zinc Ferrite Process.

The zinc ferrite absorber/regenerator reactors are assumed to be vertical cylinders with a maximum diameter of 12.5 feet and a maximum length-to-diameter ratio of 4. These sizing constraints are intended to represent rail-shippable size limits and an economical aspect ratio, based on recommendations by Kasper (1988).

B.5.2 Key Technical Issues

The following is a brief summary of some of the key technical and cost issues which have been identified based on a review of some recent literature and limited discussions with process developers. Various studies of the fixed-bed zinc ferrite process have included:

- Laboratory and bench scale tests at DOE/METC
- Process development unit (PDU) tests at KRW Energy Systems, Inc.
- Sulfidation models by Louisiana State University
- Sorbent characterizations by Amax

Limitations of the test results include a small number of data points, wide scatter in the data, lack of extended life tests, and testing of only relatively small sorbent beds compared to envisioned commercial-scale systems. In the following sections, factors contributing to uncertainty in several aspects of the fixed-bed zinc ferrite process will be discussed.

Sorbent Sulfur Loading. The theoretical sorbent sulfur loading is about 35 weight percent of the fully sulfated zinc ferrite sorbent. Factors such as space velocity, linear velocity, bed depth, number of regeneration cycles, masking by particulate and coke deposition, and flow channeling are expected to affect the achievable capacity for commercial units. In bench-scale and PDU tests the actual loadings have varied widely. Some reported results are:

- About 15 percent sorbent sulfur loading in the METC 6-inch diameter absorber with a side-stream from the METC 42-inch fixed bed gasifier at pressures of about 90 to 150 psig, temperatures of 1,000 ^oF, and space velocities of typically 2000/hr with an Arkwright coal and few regeneration cycles (Underkoffler, 1986).

- About 25 percent sorbent sulfur loading in the METC 6-inch diameter absorber with a side-stream from the METC 42-inch fixed bed gasifier at pressures of about 120 psig, temperatures of 1,000 ^oF, and space velocities of typically 1000/hr with various coals and few regeneration cycles (Underkoffler, 1986). Sorbent capacity appeared to decline after two regenerations to 14 percent.

- 30 percent sorbent sulfur loading in the bulk polishing mode with a superficial gas velocity of only 0.25 ft/sec in the KRW Energy Systems PDU (Smith, Haldipur, and Lucas, 1987)

- 10 percent sorbent sulfur loading in polishing model with a superficial gas velocity of 0.7 ft/sec in the KRW Energy Systems PDU (Haldipur et al, 1988).

In the past few years, the assumptions for sulfur loading used in conceptual performance and cost studies of commercial-size units have decreased. Studies by Cincotta (1984), Corman (1986), Smelser (1986), and Smith and Smelser (1987) apparently used sorbent sulfur loadings around 30 percent (the actual assumptions are not reported). Klett et al (1986) used a loading of 22.3 percent. A more recent study explicitly states an assumption of 10 percent (Earley and Smelser, 1988). In this latter study, it is reported that KRW Energy Systems expects improved sorbent sulfur loading of 17 percent for commercial scale systems operating in polishing mode.

Commercial scale designs must be based on the expected long-term loading associated with many cycles of absorption and regeneration. In a recent METC paper, the recommended long term (i.e. more than 40 absorption and regeneration cycles) assumption for sorbent sulfur loading is 10 percent for bulk desulfurization mode. In the short term (i.e. the first cycle) a sulfur loading of 32 percent may be possible (Kasper, 1988), although few tests have apparently achieved such high loading levels, even in the saturated zone of the reactor closest to the coal gas inlet.

The sorbent loading capability is a function of temperature. To achieve maximum absorption, an absorption temperature window from 1,000 to 1,200 ^{o}F has been suggested by many researchers. As conditions such as reactor size and actual fuel gas composition change, a different optimum temperature may be established. At higher temperatures, sintering of the sorbent and zinc vapor carry-over is expected. Furthermore, the sorbent would no longer be in the magnetite composition region, weakening the sorbent (Underkoffler, 1986).

The sorbent capacity also may be a function of the inlet sulfur concentration. Data obtained by KRW Energy Systems suggest that the capacity in the saturated zone of the reactor increases as the inlet sulfur concentration increases (Haldipur, Schmidt, and Smith, 1989), presumably due to reaction kinetics considerations. However, the average capacity of the sorbent will be less than the capacity of the saturated zone, because in the absorption front zone the sorbent sulfur loading tapers to zero toward the gas exit prior to sulfur breakthrough.

Modeling results suggest that the reaction rate is limited by the effective diffusivity of the sorbent. This is the rate at which reactant gas diffuses through the pores of the sorbent. The effective diffusivity is expected to decrease gradually with sulfidation/regeneration cycling (Wang et al, 1988).

Absorption Superficial Gas Velocity. Bench scale and PDU tests have used relatively low superficial velocities. The superficial velocity in KRW Energy System tests ranged from 0.10 to 1.35 ft/sec, with most tests well below 1 ft/sec. The expectations for commercial scale systems appear to range from 1.2 ft/sec (Haldipur et al, 1988) to a maximum of 2 ft/sec (Kasper, 1988). It is unclear if these numbers depend on the desulfurization mode (e.g., polishing, bulk). However, it appears that no tests have been performed with the higher superficial velocities that are assumed as the basis for commercial scale designs.

Klett et al (1986) assumed superficial gas velocities of less than 1 ft/sec in a conceptual design study.

Absorption Space Velocity. The commonly recommended maximum space velocity is 2,000/hr (Kasper, 1988). The space velocity is related to the sorbent absorption cycle time and to the sorbent sulfur loading. For low sorbent sulfur loadings, a lower space velocity may be required.

Absorption Cycle Time. The design basis for the study by Earley and Smelser (1988) is for 168 hours of continuous sulfidation, followed by a regeneration cycle with a minimum time of 48 hours. Absorption time until sulfur breakthrough during testing by KRW in polishing mode has typically been 55 to 60 hours (Haldipur, Schmidt, and Smith, 1989).

Sorbent Replacement Rate. A recommended design basis value for sorbent attrition is 0.0022 weight fraction of sorbent per cycle (Kasper, 1988). This recommended value is apparently an average value from test results. The statistical or measurement error is not reported. The use of laboratory and PDU data to estimate commercial scale data implies that there are no scale up issues which would affect sorbent attrition. The validity of this assumption has not been discussed in the paper. Furthermore, the assumption appears to be that sorbent attrition is constant over each cycle.

Based on the above assumption, Kasper indicates that about 20 percent of the sorbent must be replaced every 80 cycles of absorption and regeneration. An earlier study assumed complete replacement of the sorbent every 300 cycles, which is equivalent to 0.0033 weight fraction per cycle (Klett et al, 1986). Sorbent may have to be replaced due to physical attrition and due to loss of chemical reactivity.

Sorbent Bulk Density. For the purposes of sizing reactor vessels, a sorbent bulk density of 82 lb/ft^3 has been suggested by Kasper (1988).

Absorber Pressure Drop. Tests by KRW Energy Systems with particle-laden (800 to 1,200 ppmw of char solids) fuel gas entering the absorber resulted in substantial pressure drop increases of 15 to 30 inches of water per hour depending on the superficial gas velocity and solids content. It was concluded that, for application with a KRW-based system with in-bed desulfurization, a nominally particle-free inlet gas is required (Haldipur et al, 1988). A later report concludes that barrier filtration (i.e. with a ceramic filter) is superior to the use of high efficiency cyclones and that the lower solids loading of <15 ppmv obtained from barrier filtration allows desulfurization to the full capacity of the bed without an increase in pressure drop (Haldipur, Schmidt, and Smith, 1989). However, others have assumed that high efficiency cyclones are sufficient. In particular, a clean coal technology program demonstration project featuring a hot gas cleanup system is based on the use of cyclones for particle removal (Hester and Pless, 1990).

In tests with the METC 6-inch zinc ferrite absorber using a sidestream from the 42-inch fixed bed gasifier, pressure drops appeared to routinely increase substantially due to the deposition of particulates and coke in the absorber bed (Underkoffler, 1986).

Absorber pressure drop is also expected to increase as sorbent pellets disintegrate over time (attrition) (Kasper, 1988).

To overcome increases in pressure drop, it is expected that the pressure ratio of the boost air compressor must be increased, according to one report (Haldipur, Schmidt, and Smith, 1989).

Sulfidation Temperature. A temperature window of 1,000 to 1,200 ^oF has been commonly assumed. See section on sorbent sulfur loading.

Fuel Gas Conditioning. The formation of iron carbide (Fe_3C) or wustite (FeO) reduces the sulfur absorption capacity of the sorbent. The formation of soot deposits also interferes with sulfur absorption. The fuel gas must be conditioned such that wustite or carbides do not form and carbon is not deposited from the gas phase on the sorbent pellets (Haldipur, Schmidt, and Smith, 1989). A commonly-used assumption is that the $CO/(CO+CO_2)$ ratio must be less than 0.4, which can be achieved by steam conditioning of the fuel gas if necessary, which promotes the water-gas shift reaction. Because of the high moisture content of the raw fuel gas in the Lurgi-based system, it is not expected to be necessary to add steam if the water-gas shift reaction is in equilibrium.

Outlet Sulfur Concentration. The common assumption is that the zinc ferrite system in commercial scale application will achieve an outlet sulfur concentration of 10 ppmv or less. The KRW Energy Systems tests often achieved outlet concentrations higher than this target. Some tests had outlet sulfur concentrations near 30 ppmv (Haldipur, Schmidt, and Smith, 1989).

The outlet sulfur concentration must remain low enough so that the plant complies with emission permits and to remain above the acid dew point limit of the heat recovery steam generator downstream of the gas turbine combustor (Haldipur, Schmidt, and Smith, 1989).

Lab scale testing using a side-stream of METC's fixed bed gasifier and a 6 inch diameter fixed-bed zinc ferrite reactor indicated that tars and oils or particulates in the fuel gas do not inhibit the level to which H_2S in the gas is removed (Underkoffler, 1986). However, only low molecular weight sulfur compounds were absorbed. For a fixed bed gasifier, Hester and Pless (1990) report that one to four percent of the sulfur in the coal feed is bound in long-chain organic tar vapor, which is not removed by the zinc ferrite process.

Alkali and Halides. The sorbent may remove alkali metals and halides to a large extent and release them during regeneration. Vanadium, calcium, and mercury were similarly removed to some extent. This removal may be associated with particulates in the feed gas that are captured in the zinc ferrite absorber vessel. Ammonia may also be removed to some extent (Underkoffler, 1986). However, there is little documentation of such removal.

Regeneration. While regeneration is often assumed to occur at high pressure in fixed-bed systems, there may be a benefit to regenerating at low pressure in terms of plant efficiency and integration of the regeneration off-gas with a sulfuric acid byproduct recovery system. A two-step regeneration sequence is envisioned by Kasper (1988) and others. In the first step, regeneration occurs by reacting the sulfated sorbent with a stoichiometric amount of oxygen. Because this oxidative phase of regeneration is highly exothermic, steam is to be used as diluent to avoid sintering of the sorbent. It has been assumed that the regeneration bed temperature must not exceed 1,450 ^{o}F to avoid sintering of the sorbent (Kasper, 1988).

During the oxidative regeneration phase, some of the absorbed sulfur may be converted to sulfates, which would be evolved as sulfur dioxide during the beginning of the following absorption cycle. Thus, a second reducing regeneration phase may be required, in which the sulfates react with either hydrogen or carbon monoxide. The reducing cycle may be about one hour of the total regeneration cycle time. Space velocities of 600/hour for both the oxidation and reducing phases have been recommended (Kasper, 1988). Steam and air are assumed to be used for oxidative regeneration, while fuel gas is used for the reductive regeneration step to reduce sulfates in the sorbent.

Off Gas and Byproduct Recovery. Studies by General Electric of IGCC systems with the fixed bed zinc ferrite process have assumed off-gas pretreatment prior to entering a sulfuric acid plant. Pretreatment includes adiabatic cooling by humidification, further cooling in a tower with circulating weak sulfuric acid, acid mist removal, particulate removal, and drying in a drying tower (Cincotta, 1984).

Sorbent Cost. The assumed commercial cost of zinc ferrite sorbent has varied significantly in different studies. For example, in Banchik and Cover (1988), the cost is assumed to be vary from \$4/lb to \$10/lb. Pitrolo and Bechtel (1987) indicate that current sorbent costs are \$5/lb and that the target cost for commercial sorbent is \$2/lb. Kasper (1988) indicates that the projected commercial price is \$3/lb.

Capital and Maintenance Costs. For commercial scale application, the zinc ferrite system must have a safe, reliable, and automatic system for proper valving of the fuel gas for sulfidation, regeneration gases for oxidative and reductive regeneration, inert gases for purging between cycles, and sorbent replacement. Thus, the valving, piping, and controls for the zinc ferrite system are complex. Furthermore, the cycling of pressure, temperature, and oxygen content in the reactor vessels may lead to long-run vessel damage (Koch, 1989).

As an aside, several studies by the Rand Corporation have examined the historical tendency for the capital cost of pioneer (first-of-a-kind) chemical process plants to be underestimated, especially in early stages of process development (e.g., Merrow et al, 1981, Milanese, 1987, Hess and Myers, 1989). The implication of these studies also appears to be that early cost estimates tend to underestimate the capital cost of a fifth-of-a-kind plant, based on limited data about cost improvement as more plants are built (Hess and Myers, 1989). Costs estimates prepared early in process development may not capture all of the costs that would be revealed by a final estimate based on more detailed engineering analysis. Also, potential problems that could be encountered in a commercial-scale plant may not be anticipated. Therefore, one often expects that capital cost estimates are biased low, especially for preliminary cost estimates. This low bias can be compensated for by positively skewed uncertainties applied to the calculated process area direct capital costs in a cost model.

B.5.3 Elicited Technical Judgments About Uncertainties

Technical judgments regarding the performance of a commercial-scale fifth-of-a-kind dual vessel fixed-bed zinc ferrite system were elicited from three engineers at DOE/METC. These engineers will be referred to as ZF-1, ZF-2, and ZF-3. Expert ZF-1 is the same engineer as LG-1, who made judgments regarding the Lurgi gasifier. Thus, the set of assumptions labeled LG-1 and ZF-1 are used together during case studies. The engineers were asked to explicitly consider the uncertainty involved in making predictions about a system that has not yet been built or operated at a commercial scale. The experts were provided with a three part briefing paper as discussed in Chapter 4. The first part was a general introduction to uncertainty analysis. The second part was the review of published information given in the preceding sections here. The third part was a questionnaire, which is reproduced in Section B.8.2. The results of the elicitations of uncertainty in model parameters from the three experts are described in the following sections.

B.5.3.1 Expert ZF-1

Expert ZF-1 provided detailed responses to the written questionnaire, and substantial additional detail during a follow-up phone conversation. In the questionnaire, the expert indicated that uncertainties "were difficult to evaluate." The expert added, "considerable thinking and a number of calculations were required in order to provide "good" judgments." Furthermore, he indicated that the summary of technical issues provided as Part 2 of the briefing packet "was useful, and perhaps essential to the exercise." The questionnaire provided several default assumptions (see Section B.8.2) and asked the expert to comment on them. Expert ZF-1 indicated that the absorption cycle time in the assumptions was excessive at 168 hours, and that a value of 30 hours is more appropriate and recommended. The expert indicated that a maximum superficial gas velocity of 2 ft/sec is reasonable. Also, the expert suggested an absorber vessel length-to-diameter (L/D) ratio of 3 instead of 4, but agreed that the maximum vessel diameter should be around 12.5 feet.

Four parameters (sorbent sulfur loading, sorbent attrition rate, absorber pressure drop, and sorbent unit cost) were selected for probabilistic analysis in preparing the questionnaire. Expert ZF-1 suggested that two additional parameters be treated probabilistically. These are: (1) residual sulfate content of sorbent following oxidative regeneration; and (2) residual sulfide content of sorbent following reductive regeneration.

Table B-16. Summary of Elicited Zinc Ferrite Technical Judgments from Expert ZF-1

Description	Units	Distribution	Parameters[a]			
Residual Sulfate After Oxidative Regeneration	mol-% of captured S	Triangular	3	to	11	(7.5)
Residual Sulfide After Reductive Regeneration	mol-% of S in sulfate	Triangular	50	to	90	(85)
Sorbent Sulfur Loading	wt-% S in sorbent	Normal	2.16	to	31.84	(17)
Sorbent Attrition Rate	wt-% sorbent loss/cycle	Fractile	5%: 0.17	to	0.34	
			20%: 0.34	to	0.5	
			25%: 0.5	to	1	
			25%: 1	to	1.5	
			20%: 1.5	to	5	
			5%: 5	to	25	
Absorber Pressure Drop	psi/ft bed height	Triangular	0.29	to	0.53	(0.4)

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

The sulfate content affects the requirement for fuel gas used as a reductant to convert the sulfate to sulfide or released sulfur. The sulfide content affects the availability of sorbent for absorption in the following cycle. The quantitative judgments of uncertainty obtained from Expert ZF-1 are summarized in Table B-16.

The residual sulfate concentration is highly dependent on the pressure of the system and how regeneration is performed. At 600 psi, 15 percent of the absorbed sulfur has been found to remain in the sorbent as sulfate after oxidative regeneration. At 300 psi, the residual sulfate after oxidative regeneration would be less. Upon reductive regeneration, only about 15 percent of the sulfate is released, leaving residual sulfide. The fraction of sulfate converted to sulfide may be independent of pressure. There is not yet data on how the residual sulfide content would vary with factors such as regeneration space velocity, temperature, or superficial velocity. The reductive regeneration tests at METC have been at 1 atm in packed bed reactors. Dirty gasifier product gas diluted with steam was used as the reductant gas. In contrast, in a "real" plant, clean desulfurized fuel gas would be used as the reductant.

The lower limit on residual sulfate after reductive regeneration is estimated by Expert ZF-1 to be approximately 3 percent of the absorbed sulfur in the sorbent. This outcome might result from optimized proper programming of the timing and temperature profile of the regeneration process. In particular, such an outcome would depend on obtaining high regeneration temperatures without sintering the sorbent, which is a significant control problem. In a separate technology, General Electric has reported good results regenerating zinc ferrite in a semi-batch process. A more likely value is 7 to 8 percent sulfur as residual sulfate. However, it is possible that long term life cycle testing of sorbent may yield residual sulfate contents as high as 10 to 11 percent. Therefore, the expert agreed that a triangular distribution, ranging from 3 to 11 percent, with a mode of 7.5 percent, reasonably represented his judgment about the residual sulfur as sulfate after oxidative regeneration.

Expert ZF-1 was confidant that 85 percent of the sulfate would be converted to sulfide during reductive regeneration. However, after some discussion, the expert indicated that after more research, it may be possible to conduct reductive regeneration in such a way to obtain 30 to 40 percent sulfur release from the sulfate. Under some regeneration conditions, only 10 percent of the sulfur in the residual sulfate may be released, with the rest converted to sulfide. DOE/METC did not pursue research on the dual vessel fixed bed zinc ferrite process, but if it had, it would have looked more carefully at these regeneration issues. The expert agreed that a triangular distribution for the conversion of sulfate to released sulfur during reductive regeneration would have a lower limit of 10 percent, an upper limit of 50 percent, and a mode of 15 percent. The unreleased sulfur would be retained as sulfide, which would reduce the absorption capacity of the sorbent. Thus, the uncertainty distribution for the fraction of sulfate converted to sulfide during reductive regeneration would be triangular with a lower limit of 50 percent, an upper limit of 90 percent, and a mode of 85 percent.

Expert ZF-1 indicated that there is only about a 10 percent chance that the weight percent sorbent sulfur loading could be more than 26 percent for a commercial bed, assuming an ideal reactor, that sorbent does not deactivate, and that breakthrough begins when the effective space velocity reaches 4000/hr. Conversely, there is only about a 5 percent chance that the sorbent sulfur loading would be less than 5 or 6 weight percent, which could occur if there were severe channeling in the bed and deactivation of the sorbent. The most likely value is 17 weight percent sorbent sulfur loading, and the expert suggested that the distribution for this parameter should be normal with a mean of 17. This implies a 99.8 percent range from 2.2 to 31.8 weight percent sorbent sulfur loading, with a standard deviation of 4.8.

Expert ZF-1 indicated that he was not "comfortable" making a judgment about uncertainty in attrition rate but was willing to provide some "off-the-cuff" judgments. The expert indicated that he was more comfortable making judgments about chemistry-related parameters. The expert indicated that the sorbent attrition rate would follow the traditional "bathtub" curve typical of failure rates in many applications. Fresh sorbent may have "a lot of rough edges" and "weak point" that show up as attrition within the first few cycles. Manufacturing steps may be possible to "pre-attrit" the sorbent. In addition, the sorbent may undergo "chemical stabilization" in the first 10 cycles that would be likely to affect its physical stability. During the middle of the sorbent life, it may behave quite well. As the sorbent ages, it may later enter a high attrition period.

Attrition has both physical and chemical dimensions. In physical attrition, the sorbent pellets break up and may leave the bed through entrainment in exiting gas streams. Sorbet deactivation occurs from loss of chemical activity, such as slow zinc loss from the sorbent. Attrition is related to the number of absorption/regeneration cycles, but may also be related to the duration of each cycle. For example, the zinc vapor pressure is a factor that influences chemical deactivation of the sorbent, and this deactivation depends more on duration than number of cycles.

In the questionnaire, the expert indicated that there was only a 5 percent chance that the sorbent attrition rate would be greater than 5 percent per absorption/regeneration cycle. Such a result might be obtained due to carbon deposition from excessive cracking of hydrocarbons and formation of iron carbides. Attrition rate is dependent on coal properties and gasifier type. There is a 10 percent chance, according to Expert ZF-1, that the sorbent attrition rate could be lower than 0.34 percent per cycle, which assumes one complete sorbent replacement every 292 cycles. The median attrition rate was estimated at 1 percent per cycle, while the 25th and 75th fractiles were estimated at 0.5 and 1.5 percent per cycle, respectively. In the follow-up phone conversation, the expert indicated that he "can't imagine it ever being worse than 25 percent--would take one heck of an upset." But such a case might be possible, for example, if there were ever water condensation in the reactor bed. Condensation might occur if the bed was off-line prior to regeneration and not kept warm, and then diluent steam prematurely introduced. The sorbent would "fall apart" and "go to dust" if it had any residual sulfate. As a maximum lower limit, the expert indicated that 0.17 percent attrition/cycle might be assumed.

Expert ZF-1 provided a judgment regarding uncertainty in the absorber pressure drop. He assumed a sorbent bed depth of 37.5 feet, based on an absorber vessel length-to-

Table B-17. Proposed Correlation Matrix for Judgments ZF-1

Parameter	Number	Correlation Coefficients[a] 1	2	3	4	5
Residual Sulfate After Oxidative Regeneration	1	1.00				
Residual Sulfide After Reductive Regeneration	2	--	1.00			
Sorbent Sulfur Loading	3	--	--	1.00		
Sorbent Attrition Rate	4	--	--	0.50	1.00	
Absorber Pressure Drop	5	--	--	0.25	0.50	1.00

[a] For uncorrelated parameters, a dashed line "- -" is shown.

diameter ratio of 3. However, he indicated over the phone that the estimates scale linearly with bed height. Therefore, the reported values were converted to a pressure drop per foot of bed height. The expert indicated that adequate removal of particulates upstream of the absorber is essential for maintaining low pressure drop. METC sidestream tests suggested that the absorber bed can tolerate up to 0.06 lb particles per standard cubic foot of gas if the particles are less than 10 microns in size.

Expert ZF-1 did not provide any explicit indication of possible correlation among the uncertain parameters in Table B-16. However, it is plausible that several of these are correlated. Several "nominal" correlations have been assumed among sorbent sulfur loading, attrition, and absorber pressure drop as a "sensitivity" case. These correlations are shown in Table B-17. The probabilistic simulation of this system can be exercised both with and without the assumed correlations to determine if the model results are sensitive to assumptions regarding parameter correlation.

The sorbent sulfur loading and sorbent attrition rate may be related by chemical deactivation of the sorbent, which would reduce sorbent activity and possibly reduce the physical strength of the sorbent. Sorbent attrition and absorber vessel pressure drop may be related. As physical attrition proceeds, the sorbent breaks into smaller pieces and one effect of this is an expected increase in pressure drop. For example, the "Ergun correlation" indicates that pressure drop in a pellet bed is inversely proportional to the pellet diameter (Smith, 1988). Sorbent sulfur loading and pressure drop may be indirectly related through their relationships with sorbent attrition.

B.5.3.2 Expert ZF-2

Expert ZF-2 provided detailed responses to the questionnaire. The expert commented:

> It was fairly easy to make the judgments of uncertainty; however, much thought was required to arrive at what seemed like meaningful inputs. The parameters are very much interrelated and therefore difficult to separately evaluate.

With respect to the briefing materials, the expert noted the following:

> The briefing information was needed and about the right amount of depth to stimulate thought without being too cumbersome. The uncertainty analysis discussion [Part 1] seemed like a justification for the approach and more than necessary to elicit answers to the questionnaire. The examples were most useful to quickly understand what was expected.

With respect to the approach, the expert had the following concern:

> It is unclear how these inputs will improve cost or economic studies without some method of quantifying the relative importance of performance parameters to the ultimate cost of electricity.

In a follow-up phone conversation, the expert was advised that several approaches are possible to sort out the influence of uncertainty in performance parameters on uncertainty in the cost of electricity. These include probabilistic "sensitivity" analysis, in which the model is run with uncertainties in performance and cost parameters, and then with uncertainties in cost parameters only, and the results compared, or statistical analysis of the samples of the input and output distributions using "regression" type techniques. The latter approach reveals the parameter uncertainties which are most highly correlated or influential to the uncertainty in a model output, such as the cost of electricity.

The expert detailed the time required to review and respond to the briefing packet:

Read briefings, refresh technology issues	1.5 hr
Base questions #1 and #2	3.0 hr
Uncertainty ratings	2.5 hr
Summarize, reflect, edit	2.0 hr
Total	8.0 hr

Expert ZF-2 has significant background and experience with the zinc ferrite system, and provided the following comment:

Table B-18. Summary of Zinc Ferrite System Default Assumptions Used by Expert ZF-2

Description	Units	Proposed[a]	Recommended[b]
Space Velocity	1/hr	2,000	2,000
Superficial Gas Velocity	ft/sec	2	2
Inlet Temperature	^oF	1,100	1,100
Inlet Pressure	psia	300	350
Sulfur Concentration	ppmv	5,000	5,000
Vessel Outer Diameter	ft	12.5	14
Vessel Length	ft	50	35
Cycle Time	hours	168	38
Additional Assumptions			
Sulfur Loading	wt-%	NG	14
Life Cycles		NG	100
Sorbent Replacement (at 85% capacity factor)		NG	yearly
Regeneration Pressure		atm	350 psi
Regeneration Type		NG	2 step
Chloride Guard		No	Yes
Upstream Barrier Filter		No	Yes
Downstream Turbine Guard		NG	No

[a] Set of assumptions proposed by surveyor in the uncertainty questionnaire. NG = Not Given
[b] Set of alternative assumptions recommended by expert.

> The "informed" judgments are based upon first hand KRW experience from pilot plant equipment design through test completion and reporting of results plus follow up application of the test results to several commercial scale designs.

The expert carefully reviewed the design assumptions given in the questionnaire and chose to suggest an alternative set of assumptions. However, he provided judgments both for the set of assumptions proposed in the questionnaire as well as for the set he recommended. These two sets will be referred to as ZF-2P and ZF-2R, respectively. The proposed and recommended assumptions are given in Table B-18.

Expert ZF-2 suggested that vessel height should not exceed 30 to 35 feet in order not to exceed crush strength limits of the sorbent. Above this height, the vessel design must include internal sorbent supports. These supports, though undesirable, are less costly than additional vessels. A design target for sorbent life is one year between sorbent replacement with about 100 absorption/regeneration cycles per year. This would lead to an absorption time of about 38 hours, if the plant is operated 85 percent of the year. This is

Table B-19. Summary of Elicited Zinc Ferrite Technical Judgments ZF-2P from Expert ZF-2

Description	Units	Distribution	Parameters[a]			
Sorbent Sulfur Loading (Bulk Mode)	wt-% S in sorbent	Fractile	50%:	8	to	15
			50%:	15	to	20
Sorbent Sulfur Loading (Polishing Mode)	wt-% S in sorbent	Fractile	50%:	6	to	10
			50%:	10	to	15
Sorbent Attrition Rate	wt-% sorbent loss/cycle	Fractile	50%:	0.05	to	0.20
			50%:	0.20	to	1.00
Absorber Pressure Drop	psi/ft bed height	Fractile	50%:	0.14	to	0.29
			50%:	0.29	to	1.43
Sorbent Unit Cost	$/lb	Fractile	50%:	2.50	to	3.00
			50%:	3.00	to	5.00
Sorbent Life Cycles		Fractile	50%:	10	to	50
			50%:	50	to	100
Chloride Attack	ZF consumption as wt-% of coal	Fractile	50%:	0.005	to	0.05
			50%:	0.05	to	0.5

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

about the same as the minimal time required to perform valve switching and two step (oxidative and reductive) regeneration.

The expert suggested that regeneration at atmospheric pressure is "considered high risk and not practical due to rapid pressure cycling of vessels and/or longer required cycle times." Reductive regeneration is a necessary step in sorbent regeneration.

Illinois No. 6 coal contains high chlorine which "is likely to react with the zinc and render a portion of the zinc ferrite inactive." This leads to reduced sorbent activity over time.

Expert ZF-2 indicated that an upstream barrier filter is required in order for the fixed bed zinc ferrite system to be operable. "Fines escaping a conventional cyclone would lead to an unacceptable pressure drop increase that would control the cycle time and life cycle." The cleanup system might be rendered economically unviable in such a case. However, a downstream cyclone to protect the gas turbine from catastrophic loss of sorbent "is considered unnecessary for a fixed bed system."

Table B-20. Summary of Elicited Zinc Ferrite Technical Judgments ZF-2R from Expert ZF-2

Description	Units	Distribution	Parameters[a]			
Sorbent Sulfur Loading (Bulk Mode)	wt-% S in sorbent	Fractile	50%:	10	to	18
			50%:	18	to	22
Sorbent Sulfur Loading (Polishing Mode)	wt-% S in sorbent	Fractile	50%:	8	to	12
			50%:	12	to	18
Sorbent Attrition Rate	wt-% sorbent loss/cycle	Fractile	50%:	0.01	to	0.10
			50%:	0.10	to	0.20
Absorber Pressure Drop	psi/ft bed height	Fractile	50%:	4	to	5
			50%:	5	to	10
Sorbent Unit Cost	$/lb	Fractile	50%:	2.50	to	3.00
			50%:	3.00	to	5.00
Sorbent Life Cycles		Fractile	50%:	50	to	100
			50%:	100	to	200
Chloride Attack	ZF consumption as wt-% of coal	Fractile	50%:	1.E-4	to	5.E-4
			50%:	5.E-4	to	0.001

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

Expert ZF-2 proposed two parameters for probabilistic analysis, in addition to the ones included in the questionnaire. The additional parameters include sorbent life cycles and sorbent consumption due to chloride attack. The uncertainty distributions for the "proposed" set of judgments ZF-2P are summarized in Table B-19. The distributions for the "recommended" set of judgments ZF-2R are summarized in Table B-20.

For sorbent sulfur loading, the expert indicated that high values would be obtained in large vessels with even temperature distributions, large gas volumes, routine and automated procedures for absorption/regeneration cycling, proven and reliable equipment, improved sorbent strength and porosity, and improved reductive regeneration methods. Low values would be obtained due to channeling, dust deposition, chloride reaction, temperature excursions, crushed pellets due to pressure drop surge, incomplete regeneration, steep absorption wave front (with small effective sorbent volume) and attrition loss.

Low sorbent attrition is associated with the same conditions as for high sorbent sulfur loading, including even temperature distribution, routine procedures and automated controls, proven and reliable equipment and components, improved sorbent strength and shape, dust free gas, and minimal vessel height. High attrition would be associated with channeling, leading to a varying temperature profile, and other factors such as dust deposition, chloride reaction, temperature excursion, crushed pellets due to pressure drop surge or slugging of the bed, gas composition excursions, improper handling methods, and sorbent manufacturing problems.

Low absorber vessel pressure drop would result from the use of barrier filters or advanced cyclones. Also, pressure drop is associated with sorbent attrition. High pressure drop could be obtained due to dust accumulation, temperature excursion, crushed pellets due to slugging of the bed, and due to other causes of attrition with particles accumulating in the bed.

Sorbent cost could be low under conditions of large batch production, competition among suppliers, mature production methods, and many power plants to ensure continued large demand to provide incentive for supplier competition. Sorbent cost could be high due to lack of competition, possible new and more costly formulations needed to achieve higher sorbent strength and other market conditions affecting raw material cost.

A low number of life cycles would be associated with excessive pressure drop and sulfidation periods reduced to less time than required for regeneration, leading to partially unregenerated sorbent. A high number of life cycles would be obtained from stable sulfur capacity over time, minimal attrition, no pressure drop buildup, and no plant parameter excursions to damage sorbent.

Finally, low chloride attack would be associated with the use of a chloride guard (see case ZF-2P vs. ZF-2R for comparison of a case with no guard to one with a guard), chlorine in the coal, and rate of HCl reaction with zinc ferrite. High chloride attach would be associated with no chloride guard, high coal chlorine content, and high rates of reaction of zinc ferrite with HCl.

From the above discussion, it is apparent that there are strong interrelationships among the uncertain parameters, implying that these uncertainties are not uncorrelated. For example, the conditions that favor high sorbent sulfur loading also favor a high number of absorption/regeneration cycles and high sorbent life. The conditions that lead to high sorbent attrition and reduced sorbent life also tend to lead to high sorbent bed pressure drops. Therefore, the sorbent life cycles and sorbent sulfur loading were assumed to be strongly positively correlated, and the sorbent life cycles and pressure drop were assumed to be moderately negatively correlated. These correlations are intended to be used in a

Table B-21. Proposed Correlation Matrix for Judgments ZF-2

Parameter[b]	Number	Correlation Coefficients[a] 1	2	3	4	5
Sorbent Sulfur Loading	1	1.00				
Sorbent Life Cycles	2	0.9	1.00			
Absorber Pressure Drop	3	--	0.5	1.00		
Chloride Attack	4	--	--	--	1.00	
Sorbent Unit Cost	5	--	--	--	--	1.00

[a] For uncorrelated parameters, a dashed line "- -" is shown.
[b] For the case studies, the sorbent attrition rate is not needed because the overall sorbent life cycle has been specified in addition and supersedes the sorbent loss due only to attrition.

"sensitivity" case study compared to an uncorrelated case to determine how significantly these assumed correlations affect the modeling results.

B.5.3.3 Expert ZF-3

Expert ZF-3 did not comment on the uncertainty elicitation procedure. In reviewing the default design assumptions for the zinc ferrite system, the expert indicated that the assumptions were reasonable with the exception of the absorption cycle time. The expert suggested that a cycle time of no more than three days was appropriate, as opposed to the one week cycle time proposed in the questionnaire. The expert pointed out that cycle time could change over the life of the sorbent as the sorbent deactivates. The quantitative judgments regarding uncertainty obtained from Expert ZF-3 are summarized in Table B-22.

The expert provided judgments regarding sorbent sulfur loading indicating that an upper limit for a commercial system would be 20 weight percent of sulfur in the sulfated sorbent, with values around 10 percent more typical. Higher sulfur loadings would be associated with a sorbent formulation with a minimum amount of porous binder material, careful loading of sorbent in the absorber vessel, no tar dropout (presumably due to low temperature excursions during absorption), no carbon deposition, and avoidance of sharp changes in temperature and pressure. There is a small chance very low loadings of one to two weight percent might be obtained. Such a result would be due to swings in operating conditions due to sticky control valves or changes in coal composition. The expert indicated that there may be a difference between sulfur loading capacity in polishing vs. bulk desulfurization mode, but that there is no data to demonstrate such an effect.

Table B-22. Summary of Elicited Zinc Ferrite Technical Judgments from Expert ZF-3

Description	Units	Distribution	Parameters[a]			
Sorbent Sulfur Loading	wt-% S in sorbent	Fractile	1%:	1	to	2
			4%:	2	to	5
			90%:	5	to	15
			5%:	15	to	20
Sorbent Attrition Rate	wt-% sorbent loss/cycle	Fractile	35%:	0.07	to	0.18
			15%:	0.18	to	0.20
			40%:	0.20	to	0.50
			10%:	0.50	to	2.00
Absorber Pressure Drop	psi/ft bed height	Fractile	50%:	0.4	to	0.5
			50%:	0.5	to	1.0
Sorbent Unit Cost	$/lb	Triangular	0.75		to	5.00 (3.00)

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

With respect to sorbent attrition, the expert noted that there was one reported instance of catastrophic loss of nearly all of the sorbent. The cause of this loss was believed to be a sudden change in operating conditions. However, the median value for attrition is expected to be 0.22 weight percent sorbent per cycle, implying a life cycle of about 450 absorption/regeneration cycles. A maximum value for a commercial scale unit would be about two percent attrition per cycle, while a lower limit would be about 0.07 percent per cycle. Favorable attrition rates are expected under stable operating conditions.

Low pressure drop in the absorber vessel is expected in cases where there is no sticking or agglomeration of particles in the fuel gas within the sorbent bed, and in which very fine particles are entrained in the exiting fuel gas. Also, it is assumed that very small particles will be periodically removed from the sorbent bed by removing the sorbent, screening it, separating out fine sorbent, and returning the larger sorbent pellets to the absorber vessel. Pressure drops significantly higher may be encountered if fines build up in the sorbent bed.

The minimum sorbent cost is assumed to be slightly higher than the cost for the iron oxide and zinc oxide which are used to manufacture the sorbent. Such an outcome assumes negligible costs for binder material and manufacture of the sorbent pellets. A high cost would be associated with difficulty in scaling up production machinery.

B.5.4 Other Uncertainties

Two additional uncertainties in the zinc ferrite process area for which judgments are required include the direct capital cost and the annual maintenance cost. In the conceptual design studies from which the cost data used to develop the zinc ferrite direct cost model was taken, the process contingency factor assumed was 40 percent (Smith and Smelser, 1987; Earley and Smelser, 1988). This is a comparatively high contingency factor, representative of the early stage of development of the zinc ferrite process, which has not been built on a commercial scale.

For the purpose of an initial screening study of uncertainties, it is assumed that the uncertainty in the capital cost of the zinc ferrite process is symmetric to the commonly assumed contingency factor. In this case, a uniformly distributed range of 100 to 180 percent of the process area capital cost is assumed, with a median value of 140 percent representative of the deterministic contingency factor. While this range of uncertainty is large compared to other process areas, it is also representative of the types of cost growth that might be expected for a process area currently in an early stage of development. A particular source of cost-related uncertainty in the zinc ferrite process area is the complexity of the valving required to properly control gas flows during absorption and the two stage regeneration. In particular, it is important to properly purge the reactor vessel between oxidative and reductive regeneration steps. The potential effects of contaminants on valve operation may necessitate expensive valves. In addition, there may be significant costs associated with the control system.

The typical maintenance cost factor assumed for the zinc ferrite process is 4.5 percent of the process area capital cost per year. As an initial characterization of uncertainty, it is assumed that this maintenance cost factor could be between 3 and 6 percent per year, with a triangular distribution and a mode at 4.5 percent per year. Maintenance costs might include such things as valve repair and replacement.

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B.6 Gas Turbine Process Area

This document reviews uncertainties in and potential problems of using state-of-the-art gas turbines designed for natural gas firing in an integrated gasification combined cycle (IGCC) process environment. This review considers the uncertainties which may arise from the application of high-firing temperature heavy-duty gas turbine technology (2,300 ^oF turbine inlet temperature) designed for natural gas fuels to low- and medium-BTU coal gas integrated gasification combined cycle (IGCC) systems. In particular, we are interested in three IGCC systems. The key design and performance assumptions for these systems are summarized in Table B-23. These systems are: Case ALH, the Air-blown Lurgi-based system with Hot gas cleanup using external fixed-bed zinc ferrite desulfurization; Case AKH, the Air-blown KRW-based system with Hot gas cleanup using in-bed and external hot gas desulfurization; and Case OKC, the Oxygen-blown KRW-based system with Cold gas cleanup, which is a baseline case. This case is believed to represent the lowest technical risks and, hence, is the lowest priority for developing technical judgements about uncertainty.

B.6.1 Key Design and Performance Assumptions

The assumed system configuration for Case ALH is shown in Figure B-1 in Section B.3. Hot coal gas is fed directly to the gas turbines at a temperature of 1,200 ^oF. Two high efficiency cyclones, one upstream and one downstream of the zinc ferrite absorber, are assumed for particulate removal and for alkali control. The gas turbine will have pressurized air extraction at the compressor outlet for use as gasifier blast air. The gas turbine exhaust is cooled in a heat recovery steam generator.

The system configurations for Case AKH and Case OKC are shown in Section B.4 in Figures B-10 and B-11, respectively. Case AKH is similar to Case ALH in that a low-BTU coal gas is fed to the gas turbine fuel valve at a temperature of 1,200 ^oF, air is extracted from the compressor discharge for use as gasifier blast air, and the gas turbine is downstream of zinc ferrite absorbers and two stages of cyclones. In Case OKC, the gas turbine operates on medium-BTU coal gas, which enters the fuel valve at a much lower temperature (350 ^oF) than Case ALH or Case AKH. In Case OKC, low temperature gas cleaning is used, featuring wet scrubbing for particulates and selective removal of hydrogen sulfide by the Selexol process.

Table B-23. Gas Turbine Default Assumptions and Typical Performance Estimates for Case Studies

Description	CASE ALH	CASE AKH	CASE OKC
Oxidant	Air	Air	Oxygen
Gasifier	Lurgi	KRW	KRW
Particulate Removal	Two Cyclone Stages	Two Cyclone Stages	Wet Scrubbing
Alkali Removal	Two Cyclone Stages	Two Cyclone Stages	Wet Scrubbing
Sulfur Removal	External fixed-bed zinc ferrite	Gasifier in-bed and external fixed-bed zinc ferrite	Selexol
Gas Turbine Fuel Temperature, ^oF	1,200	1,200	350
Typical Fuel Heating Value, BTU/scf	103	93	295
Typical Fuel Composition, mole-percent			
H_2	18.2	19.2	36.0
CO	5.2	8.9	46.2
CH_4	3.2	1.3	5.3
NH_3	2,000 ppm	700 ppm	0
N_2	31.3	40.0	0.7
CO_2	14.5	13.4	10.6
H_2O	27.3	16.5	0.0
H_2S	10 ppm	10 ppm	740 ppm
COS	inc. w/H_2S	inc. w/H_2S	360 ppm
T/O/P	1,000 ppm	N/A	N/A
Particles	10 - 100 ppmw	10 - 100 ppmw	5 - 50 ppmw
Alkali	<1 ppmw	<1 ppmw	0 ppmw
Steam Injection, lb steam/lb fuel	N/A	N/A	0.30
Gas Turbine Air Extraction	Yes	Yes	No
Gas Turbine Mass Flow Ratios			
lb air/lb exhaust	0.89	0.91	0.90
lb fuel/lb comb. air	0.30	0.33	0.09
lb extr. air/lb fuel	0.51	0.63	N/A
Inlet Ducting Pressure Loss, in H_2O	4	4	4
HRSG Pressure Loss, in. H_2O	20	20	20
Combustor Pressure Loss, %	4	4	4
Combustor Heat Loss, %	1	1	1
Heat Loss Recovered in Exhaust, %	50	50	50
Fuel Valve Pressure Drop, psi	70	70	70
Mechanical Shaft Losses	0	0	0
Generator Efficiency, %	98.5	98.5	98.5
Cooling air, % of inlet air flow			
Discharge	6	6	6
Medium pressure stage	3	3	3
Low pressure stage	3	3	3

The key default assumptions for system design are summarized in Table B-23 for all three cases. In addition, typical modeling results, obtained from ASPEN simulations of these systems, are reported for fuel gas heating value, fuel gas composition at the turbine inlet, and mass flow ratios in the gas turbine. These ratios include the mass flow ratio of the compressor inlet air to the turbine exhaust air, the mass flow ratio of the fuel to the combustor inlet air, and the mass ratio of the extraction air flow to the fuel flow. The gas turbine mass balance is illustrated in Figure B-14.

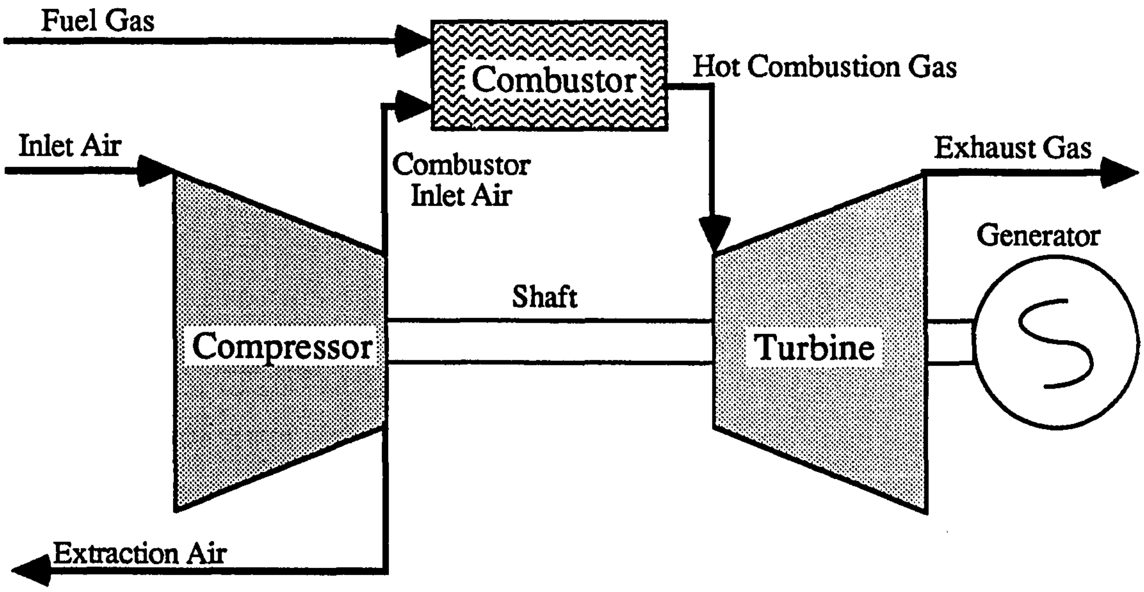

Figure B-14. Simple Schematic of Gas Turbine Mass Balance with Compressor Air Extraction.

B.6.2 Commercial Offerings for 2,300 °F Gas Turbines

In this research, the modeling of IGCC systems is intended to include performance representative of typical high-firing temperature gas turbine technology. However, the intent is not to attempt to model exactly the performance of any one proprietary gas turbine model. Instead, the goal is to achieve reasonable accuracy in reproducing the key performance characteristics of this class of gas turbines.

Currently, there are two 2,300 °F turbine inlet temperature heavy-duty gas turbine models which are expected to be offered commercially in the next year or two. These are the General Electric MS7001F and the Westinghouse/Mitsubishi 501F. Some characteristics and design assumptions for these gas turbines are given in Table B-24.

Table B-24. Representative 2,300 °F Firing Temperature Heavy-Duty Gas Turbine Commercial Offerings

Design Specification (Fuel: Natural Gas)	General Electric MS7001F	Westinghouse/Mitsubishi 501F
Net Power, kW	150,000	145,000
Heat Rate, BTU/kWh	9880	10,000
Compressor Inlet Air, pps	918.7	912
Pressure Ratio	13.5	14.2
Exhaust Temp., °F	1,081	1,061
Compressor Stages	18	16
Inlet Guide Vanes	Yes	Yes
Variable Stator Vanes	No	No
Compressor Cooling Air Extraction (stage no.)	13, 17, discharge	13, 10, 6, discharge
Compressor Bleed (stage no.)	13	6,10,13
No. of Combustor Cans	14	16
Standard Combustor Design (Natural Gas firing)	multiple fuel nozzles, wet injection--NO_x ("quiet" combustor)	pre-mix, two-stage lean-burn low-NO_x
Turbine Stages	3	4
Turbine Cooling:		
Row 1 rotor vanes	internal convection	film, impingement, pin fin
Row 2 rotor vanes	internal convection	similar to Row 1
Row 3 rotor vanes	uncooled	inlet cavity convection
Row 4 rotor vanes	N/A	uncooled

NOTES:
- The GE MS7001F apparently uses film cooling on the turbine stator vanes ("nozzles"), but not on the rotor vanes ("buckets").
- Both offerings use corrosion coatings on the hot gas path components.

Sources: Brandt, 1988; Brandt, 1989; Scalzo et al, 1989.

The gas turbine model commonly assumed for IGCC system studies is the GE MS7001F. The prototype of this model reportedly began commercial operation in June 1990 as part of a natural gas-fired combined cycle unit at Virginia Power's Chesterfield Station in Richmond, VA. In addition, GE has reported sales of at least 10 more of these gas turbines. A proposed IGCC plant, to be located in Freetown, Massachusetts, using the Texaco gasifier, is planned to include a MS7001F (Smock, 1990).

The MS7001F is designed to fire either natural gas or distillate oil at design point conditions of 59°F ambient temperature, 14.7 psia ambient pressure, and 60 percent relative humidity. The use of coal gas represents a departure from the design fuel. Because coal gas has a substantially lower heating value than natural gas, the fuel mass flow rate is significantly larger than the design basis for the gas turbine. Typically, the mass flow at the turbine inlet nozzle is limited by choking. Therefore, an increase in the fuel mass flow rate must be compensated by a reduction in the compressor air flow rate, for a given pressure ratio and firing temperature. This results in off-design operating conditions for the gas turbine, which has implications for gas turbine performance, such as efficiency, exhaust temperature, and other parameters.

Many IGCC studies were developed prior to the testing and delivery of the prototype MS7001F. In these studies, a variety of assumptions regarding the projected performance of this unit were made regarding firing temperature, pressure ratio, efficiency, and other measures of performance. In most cases, these assumptions have proven to be different from the actual unit. This is an example of the difficulty involved in trying to predict the commercial scale performance of an advanced system for which no commercial experience is yet available. In many cases, the assumptions may have been unnecessarily conservative, while in other cases they may have been optimistic.

The studies appear to give only superficial consideration to the off-design nature of gas turbine operation on coal gas. Furthermore, the studies appear to give only superficial consideration to other factors associated with firing coal gas in a gas turbine.

Although a MS7001F is now in commercial service, the performance of this model with coal gas has yet to be demonstrated.

B.6.3 Operating Strategies for Coal Gas Firing

The primary issues discussed in this section are the interactions between fuel flow, compressor performance, and compressor air extraction.

A gas turbine is designed to meet a set of goals for a specific set of operating conditions. When any of these conditions are changed, the turbine is said to be in an "off-design" mode. The response of the gas turbine to changes in operating conditions requires detailed knowledge which is specific to each machine. This type of information is closely held proprietary information. The design of a gas turbine, and prediction of its performance, involves a significant amount of empirical information. In many cases, off-design information must be obtained from testing under various conditions, which is expensive. At a minimum, some testing is required to verify the accuracy of theoretical models. Because of the expense of testing needed to support gas turbine design and to verify the operation the gas turbine once built, detailed information about gas turbine design, such as compressor operating maps, are not published (Eustis and Johnson, 1990). Furthermore, gas turbine manufacturers usually try to adopt existing successful designs where feasible into new models, or to modularize the system (in the case of combustor cans, for example) so that a change in one component requires only a simple substitution and no changes in other components (Cohen et al, 1987; Brandt, 1988; Scalzo et al, 1989).

Because of the expense of developing and testing gas turbines, it is unlikely that, in the near term, the gas turbine industry will develop a machine designed specifically for operation with coal gas. Instead, they will try to develop an understanding of how a machine designed for larger markets (e.g., natural gas firing) will behave when firing coal gas. The manufacturers may be required to offer some modifications, such as for fuel valves or combustors. However, the manufacturers are also likely to impose limitations on fuel composition or gas turbine operation to which a customer must adhere. The development of such limitations is presumably based on some type of technical risk analysis of the gas turbine, supported either by theoretical models, empirical testing, both or neither.

Uncertainties are likely to remain, however, regarding the long term maintainability and performance of the gas turbines when firing coal gas. In particular, problems such as loss of output or shorter maintenance cycles (e.g., more frequent reblading) may be encountered in machines fired with coal gas for long periods of time (a complete life cycle). In some cases, these uncertainties can be represented solely as uncertainties in cost. However, there may be trade-offs between changing operating conditions and maintenance costs. A major concern for reliable operation of an integrated plant is the stability of the compressor and the control system, particularly when air is extracted for use in the gasifier.

A key difference between natural gas firing and coal gas firing is the heating value of the fuel. Natural gas has a heating value of about 1,000 BTU/scf. Medium-BTU coal gas (MBG) has a heating value of 300 to 500 BTU/scf, and low-BTU coal gas (LBG) has heating values around 100 BTU/scf. As a result, the mass flow rate of fuel required to supply a given amount of chemical energy is significantly larger for LBG than for natural gas.

The factor that usually limits the mass flow in a gas turbine is the area of the turbine inlet nozzles (Eustis and Johnson, 1990). When the flow is choked (sonic) the mass flow is at its maximum, and the maximum mass flow for an ideal gas is given by:

$$m_{max} = P\,A^{*}\sqrt{\frac{MW}{T}}\sqrt{\frac{\gamma}{R}\left(\frac{2}{\gamma+1}\right)^{\frac{\gamma+1}{\gamma-1}}} \qquad \text{(B-1)}$$

where,

m_{max} = maximum mass flow rate
P = total pressure
A^* = critical area where flow is choked
MW = molecular weight of gas
T = total temperature
R = universal gas constant
γ = ratio of specific heats for the gas

The molecular weight of the exhaust gas varies within about two percent for all three cases compared to the natural gas design point. The term under the radical varies about 5 percent as the ratio of specific heats varies from 1.2 to 1.4. At 2,000 °F, the ratio of specific heats of nitrogen, the largest component in the exhaust gas, is about 1.3. The mass flow into the gas turbine is proportional to the critical area (which is fixed for a given gas turbine model) for a given pressure ratio and firing temperature.

For natural gas-fired operation, the air flow into the GE MS7001F compressor is about 919 lb/sec. The natural gas flow rate is about 20 lb/sec, yielding an exhaust flow rate of about 939 lb/sec. However, in the case of low-BTU coal gas, the fuel flow rate is likely to be on the order of 200 lb/sec. This would imply a turbine flow rate of over 1,100 lb/sec, or a compressor flow rate of about 720 lb/sec, depending on the operating strategy employed and whether a substantially redesigned gas turbine is assumed.

Eustis and Johnson (1990) discuss several strategies for firing coal gas in a gas turbine. These options include:

1) Increase the pressure ratio. This increases the maximum mass flow rate in the turbine nozzle. However, the compressor may not have enough surge margin to do this. Also, the increased mass flow would increase the thermal loads on the turbine blades and vanes, which may require a reduction in firing temperature.

2) Reduce compressor mass flow using inlet guide vanes (IGV). This reduces the compressor mass flow to compensate for the increased fuel flow. The flow reduction is limited by the compressor design. Compressors with variable stators and intermediate air bleed points in addition to IGVs are better able to achieve flow reductions without inducing stalling in any of the compressor stages.

3) Increase the inlet turbine nozzle critical area. This is a major redesign and would require a new gas turbine model. As a practical matter, it is unlikely that gas turbine manufacturers would develop such a machine.

4) Reduce the turbine inlet temperature. This would reduce the gas turbine efficiency and power output, but allow increased turbine mass flow.

5) Bleed air from the compressor. This is possible only where there is a use for high pressure air elsewhere in the plant. Otherwise, it is wasteful, and reduces plant efficiency.

In this study, a combination of Strategies 2 and 5 is assumed. Both the GE MS7001F and the Westinghouse/Mitsubishi 501F have IGVs. They do not have variable stator vanes. For the low-BTU coal gas systems, a portion of the compressor discharge air is assumed to be extracted for use as gasifier blast air. However, as noted in Table B-23, the ratio of extraction air to the fuel flow is about 0.5 to 0.6. The extraction air does not fully compensate for the increased fuel mass flow. Thus, at full load, the IGVs would have to be partially closed.

IGVs are often used to respond to part load conditions without having to reduce firing temperature. At the point where the IGVs are "fully" closed, firing temperature must then be reduced to further reduce the load. In a coal gasification application, because the IGVs are already partially closed at full load, the gas turbine will be less efficient at part load operation, as the point at which firing temperature must be reduced will be at a higher load condition than for natural gas.

The partial closure of IGVs will slightly affect the gas turbine pressure ratio. However, because the gas turbine model used in these case studies is based on mass and energy balances only, and not the aerodynamic characteristics of a gas turbine, pressure ratio is not predicted. Any change in pressure ratio must be specified by the model user.

Closure of IGVs also affects the compressor surge margin. At surge conditions, the compressor is no longer able to generate a steady high pressure exit stream. Thus, any downstream pressurized gas, such as that in the combustor, will backflow into the compressor, possibly causing severe vibration and damage. Compressors are usually designed to operate at a point sufficiently removed from the "surge line" to reduce the possibility of encountering surge. However, the operation of the machine with IGVs closed may reduce the margin between the operating conditions and surge conditions (Eustis and Johnson, 1990).

The determination of the surge line and the compressor characteristics requires extensive testing under a variety of loads, corrected speeds, IGV settings, and mass flow rates. These data are summarized in compressor "maps." These maps are proprietary information, due to the expense of developing them and the importance of the information to the competitive position of the manufacturer. General Electric reports that the MS7001F has a better surge margin than the MS7001E, which has been commercially available for years. GE reports that no in-service surges of the MS7001E have been reported. Thus, GE expects a superior surge margin for the MS7001F (Brandt, 1989). This may alleviate any concerns about using the IGVs to reduce the compressor mass flow. However, without a compressor map, it is difficult to make any quantitative assertions.

The use of air extraction for the low-BTU coal gas cases helps to improve the surge margin of the compressor, by reducing the amount of IGV closure needed at full load conditions. However, air extraction poses significant control problems for the IGCC plant, because it imposes a coupling between the gas turbine and the gasifier. Changes in coal composition can affect the fuel/air ratio, but can also affect the gasifier blast air requirement. This requires a sophisticated control system to regulate the IGVs, extraction air flow rate, and fuel flow rate. Advanced control systems may be required (Corman, 1986).

B.6.4 Fuel Valve

The pressure drop across the fuel valve system has an important effect on system efficiency. The gasifier pressure must be high enough to compensate for all pressure losses between the gasifier outlet and the gas turbine combustor. The pressure in the combustor is determined based on the gas turbine pressure ratio. Pressure losses in the system include the fuel gas piping, fuel valve, particulate removal devices (e.g., cyclones), and sulfur removal devices (e.g., zinc ferrite absorbers). Increasing the gasification pressure above that required for fuel gas delivery can reduce the system efficiency (Simbeck et al., 1983).

Reduction in the fuel valve pressure drop was reported to be one goal of a proposed demonstration plant. The typical pressure drop in the fuel valve was reported at about 70 psi. The goal was to achieve about 10 psi. The demonstration project proposes to use a GE MS7001E with a fuel gas temperature of about 1,000 ^oF. The material requirements for this system were claimed not to be a major problem (Hester and Pless, 1990).

A design study of an IGCC system with hot gas cleanup assumed a gas turbine fuel inlet temperature of 1,200 ^oF. The basis for this assumption was reported to be GE's expectation that by 1994 a fuel system for 1,200 ^oF gas could be developed, although the highest fuel gas temperature tested to date has been 1,000 ^oF (Earley and Smelser, 1988).

The presence of particles in the fuel gas could lead to erosion or deposition in the fuel nozzles. Based on two-stage high-efficiency cyclones, a GE study concludes that the particle concentration and size distribution in the fuel gas would allow for "adequate" nozzle and control valve lives. However, any solids that deposit in the fuel nozzle can alter flow characteristics. This can result in reduced combustion efficiency. Solids deposits can also interfere with fuel valve operation. Naphthas, tars, and phenols can build up on valve internals (Cincotta, 1984).

Any liquids entering the combustor as large droplets may not burn completely within the combustor. They may carry over to, and burnout in, the first stage turbine nozzle. This can cause damage to the turbine (Cincotta, 1984).

The fuel control system poses a design challenge for an IGCC plant. The control system must account for changes in the heating value of the fuel gas during plant operation, as well as differences in the load-following capability of the gasifier and gas turbine. The fuel control system could potentially depressurize the gasifier by demanding more fuel than the gasifier can supply during ramp-up (Cincotta, 1984). The addition of gas turbine air extraction for gasifier blast air further complicates the control system (Corman, 1986).

In the modeling studies, the effect of pressure drop in the fuel gas valve can be explicitly included in the ASPEN performance simulation. The effect of exotic fuel valve materials or designs on gas turbine cost can be incorporated in the cost model through, for example, a direct capital cost multiplier factor.

B.6.5 Combustion and Emissions

Gas turbine combustors have been developed in an empirical-based manner. Mathematical analysis and scale model testing apparently have been inadequate predictors of full-scale combustor performance (Dawkins et al, 1986). As a result, heavy-duty gas turbines have been developed using multiple modular "can" combustors. Typically, many of these combustors are arranged around the circumference of the machine between the compressor and the turbine. As part of a development program only one combustor can needs to be used in testing (Cincotta, 1984). In a commercial-scale gas turbine, such as the ones summarized in Table B-24, perhaps 16 to 18 combustor cans are utilized. Each one can be changed out for maintenance and repair. The standard combustor can also can be replaced by improved versions as they become available. The same combustor design can be used in different size machines by using an appropriate number of the combustor cans.

There are a number of pollutant species that may be contained in the hot gas exiting the combustor which have received attention in the literature. These are:

- Thermal NO_x resulting from thermal fixation of oxygen and nitrogen in air.
- Fuel NO_x resulting from conversion of chemically bound nitrogen in the fuel (e.g, ammonia).
- SO_2 resulting from hydrogen sulfide, carbonyl sulfate, and sulfur contained in naphtha, tars, oils, and phenol.
- CO resulting from incomplete carbon conversion in the combustor.

- Uncombusted particles passing through the combustor.
- Alkali (sodium and potassium compounds) which may cause turbine blade corrosion.

The design of gas turbine combustors is undergoing changes in response to environmental constraints on NO_x and CO emissions and an increasing array of potential gas turbine fuels. Currently, most efforts are focused on developing low-NOx combustors for natural gas applications (Angello and Lowe, 1989). However, some theoretical studies, bench scale research, and a few commercial-scale demonstrations have involved medium- and low-BTU gases, such as those derived from coal gasification. The design of combustors for coal gas applications may be fundamentally different from those for natural gas applications, particularly with respect to NO_x emissions.

B.6.5.1 NO_x Emissions

NO_x emissions result primarily from the thermal fixation of nitrogen and oxygen in the inlet combustion air and from conversion of chemically-bound nitrogen in the fuel. The former is referred to as "thermal" NO_x, while the latter is referred to as "fuel" NO_x. Thermal NO_x formation is sensitive mainly to the flame temperature of the burning fuel. Poor mixing of fuel and air can lead to localized "hot spots" which generate high flame temperatures and, hence, high thermal NO_x emissions. Uniform mixing of fuel and air leads to more uniform flame temperatures, which reduces thermal NO_x formation. In addition, other measures which reduce flame temperatures, such as staged lean combustion or the addition of diluents such as water or steam, will reduce thermal NO_x emissions (Davis et al, 1987; Touchton, 1984)

Fuel NO_x arises from the conversion of ammonia, HCN, or other nitrogen-containing chemical species in the fuel. The formation of fuel NO_x is relatively insensitive to temperature compared to thermal NO_x formation. Fuel NO_x formation depends primarily on the concentration of fuel-bound nitrogen in the fuel gas and the method of fuel/air contacting (Folsom et al, 1980). To reduce fuel NO_x formation, two-stage rich/lean combustion has been proposed and tested by several (e.g., Folsom et al, 1980; Sato et al, 1989; Unnasch et al, 1988). In the rich combustion stage, fuel bound nitrogen is converted mostly to diatomic nitrogen. In the lean stage, fuel burnout is completed under conditions which minimize the formation of thermal NO_x.

The most widely used gas turbine fuel is natural gas, which contains negligible fuel-bound nitrogen. Most major gas turbine manufacturers are attempting to develop dry low-NO_x combustors, to reduce the formation of "thermal" NO_x by premixing the fuel and air and use of lean-burn or lean-lean two-staged combustion. The

Westinghouse/Mitsubishi 501F will be offered with a low NO_x combustor featuring fuel and air premixing and a lean-burn combustor (Scalzo et al., 1989). The GE MS7001F is offered with a multiple fuel nozzle combustor can (Brandt, 1988). This is not a low-NO_x design per se, but it does allow increased levels of water or steam injection to achieve low NO_x emissions with fuels that do not contain fuel-bound nitrogen. The multiple nozzle design has been referred to as the "quiet" combustor because it has a lower vibration and noise level than GE's single fuel-nozzle combustor. The reduced vibrations permit higher levels of water injection.

Medium-BTU Coal Gas: Case OKC

In the current modeling work, the only medium-BTU coal gas of concern is that produced from an oxygen-blown KRW gasifier in Case OKC. The Case OKC IGCC system features "cold" gas cleanup, which effectively removes any ammonia, the primary fuel-bound nitrogen species, from the raw coal gas. Thus, fuel NO_x emissions are not expected to be a problem for this application. Thermal NO_x emissions are of concern, however. MBG may have flame temperatures similar to that of distillate oil , and thus uncontrolled NO_x emissions from firing MBG may be comparable or greater than uncontrolled emissions from firing distillate oil (Davis et al, 1987).

Most conceptual design studies assume that steam injection and/or fuel gas saturation can be used to reduce the combustor flame temperature and, hence, NO_x emissions to meet current New Source Performance Standards (NSPS) for gas turbines (e.g., Gallaspy et al, 1990 and many of the other EPRI design studies). Wet injection is a standard technique for natural gas and oil fired gas turbines. The thermal diluent, steam or water, results in a reduction in peak combustion temperatures, thus reducing thermal NO_x formation (e.g., Davis et al, 1987; Touchton, 1984; Touchton, 1985). Both steam injection and fuel gas saturation have been tested at the Cool Water demonstration plant, which uses MBG from a Texaco gasifier (Cool Water, 1988; Holt et al, 1989).

The NSPS is often quoted as 75 ppm at 15 percent oxygen on a dry basis, but the standard actually includes a correction for plant efficiency. Thus, the actual allowable emissions under NSPS for a particular gas turbine model may be higher.

However, it is controversial whether the gas turbine NSPS is the applicable standard for IGCC power plants, or whether it is even a relevant standard. More likely, IGCC plants will be subject to local or EPA-mandated procedures such as Best Available Control Technology (BACT), which is determined on a plant-by-plant basis. The procedure for BACT analysis that is becoming increasingly common is known as the "top-down" approach. In this approach, a facility is asked to use the most stringent control system that has been demonstrated unless there are energy, environmental, or economic reasons to do otherwise. For natural gas-fired gas turbines, BACT may include combinations of low-NO_x combustors, wet injection, and post-combustion NO_x control using selective catalytic reduction (Smock, 1989; Moore-Staub et al, 1990). It is likely that an actual IGCC plant will be required to achieve very low NO_x emissions on the order of 10 ppm, rather than the 75 ppm (corrected) often assumed. Thus, SCR may be required. SCR has been applied to or required for a number of natural gas- and oil-fired gas turbines in California and a few other states (Radin and Boyles, 1987; Moore-Staub et al, 1990). SCR is expected to be capable of reducing IGCC system NO_x emissions to 5 ppm (Holt et al, 1989). At least one IGCC plant, a proposed demonstration plant in Florida, is to be permitted with SCR (Hester, 1990). This may set a BACT precedent for other IGCC plants.

For the purposes of the current study, fuel gas saturation and/or steam injection for combustion NO_x control is assumed for medium-BTU coal gases with no fuel-bound nitrogen. The effect of SCR would primarily be to increase the capital and operating costs of the system, with a slight penalty on plant efficiency due to increased HRSG backpressure and the auxiliary power requirements of the SCR ammonia injection and control systems. SCR may be more advantageous for application with fuel gases containing significant concentrations fuel-bound nitrogen.

The applicability or efficacy of dry low-NO_x combustors designed for natural gas when converted to coal gas firing may merit some testing and evaluation. Whether the combustors can be used "as is", other than modifications for the fuel nozzles, might be the subject of further research.

Low-BTU Coal Gas: Cases ALH and AKH

Thermal NO_x is not expected to be a major concern with LBG gases because of their low adiabatic flame temperatures resulting from the presence of thermal diluents in the fuel such as N_2. The thermal NO_x emissions from LBG are often dismissed in the literature as being insignificant, particularly if peak flame temperatures are limited to less than 2,800 °F (Davis et al, 1987; Folsom et al, 1980; Notestein, 1989; Sato et al, 1989; Unnasch et al, 1988). Uncontrolled thermal NO_x emissions from LBG combustion may in fact be on the order of 10 to 50 ppm, as suggested by some small scale combustor tests (e.g., Unnasch et al, 1988).

A confounding factor for thermal NO_x emissions from LBG is the expected high gas turbine fuel valve inlet temperatures associated with hot gas cleanup (HGCU) systems. Also, increasing pressure ratios for gas turbines may promote thermal NO_x emissions (Folsom et al, 1980). Increasing the fuel gas temperature will tend to increase thermal NO_x production because the flame temperatures will be marginally higher. However, this is not expected to significantly increase thermal NO_x emissions for the fuel temperatures of current interest (1,000 to 1,200 °F).

The primary concern regarding NO_x emissions from LBG is fuel NOx resulting from ammonia, HCN, or other fuel bound nitrogen species. LBG is derived from air-blown gasification systems. Air-blown gasification is commonly envisioned in conjunction with HGCU. HGCU systems typically are based on dry pollutant removal processes, such as cyclones or barrier filters for particulate control and chemical sorption for sulfur control. Unlike "cold" gas cleanup wet scrubbing processes, these dry processes do not remove ammonia, the primary fuel-bound nitrogen specie, in the fuel gas. In conventional gas turbine combustors, most of the ammonia would be converted to NO_x. For example, Cincotta (1984) states that the conventional GE MS7001E combustor would convert about 70 percent of ammonia in a Lurgi fuel gas to NO_x. Another study reports a similar finding (Sato et al, 1989). In a conventional combustor, the conversion rate of ammonia to NOx may vary from 50 to 90 percent depending on the concentration of ammonia in the fuel gas (Pillsbury, 1989).

The ammonia concentration in the fuel gas depends on the gasifier type and operating conditions. Notestein (1989) indicates typical ranges of ammonia concentration in coal gas as 200 to 600 ppmv for fluidized bed gasifiers operating at 1,300 to 1,800 °F, 2,000 ppm for entrained flow gasifiers, and up to 5,000 ppm for fixed bed gasifiers operating below 1,200 °F. Holt et al (1989) suggest that about 50 to 60 percent of coal-bound nitrogen is converted to ammonia in fixed bed gasifiers, while only 10 to 15 percent is converted in entrained-flow gasifiers. Some typical concentrations from ASPEN simulation models are given in Table B-23.

The most likely near-term solution for reducing fuel NO_x emissions from LBG combustion appears to be staged rich/lean combustion (Cincotta, 1984; Folsom et al, 1980; Sato et al, 1989; Unnasch, 1988). In rich/lean combustion, the rich stage is used to convert ammonia to nitrogen, and the second stage is used for fuel burnout. The combination of a rich and lean stage also reduces the peak flame temperatures in the combustor, thereby reducing thermal NO_x emissions.

Some of the findings of several combustor research efforts have been:

- Temperature. Fuel NO_x formation is relatively insensitive to temperature (Holt et al, 1989). Variation in fuel heating value appears to have little effect on conversion of ammonia to NO_x (Folsom et al, 1980).

- Fuel-nitrogen concentration. The fraction of fuel-bound nitrogen converted to NO_x decreases with increasing fuel-bound nitrogen concentration (Folsom et al, 1980; Sato et al, 1989; Unnasch et al, 1988). In the Unnasch et al (1988) tests, it was found that above 5,000 ppm ammonia concentration, there was very little marginal increase in NO_x emissions.

- Stoichiometry. Fuel NO_x formation is sensitive to the reaction stoichiometry. In an oxygen-deficient environment, a substantial portion of fuel-bound nitrogen can be converted to diatomic nitrogen. The optimal reactant stoichiometry (fuel/air ratio) in the rich stage to maximize conversion of fuel-bound nitrogen to N_2 (minimize fuel NO_x) is influenced by reaction temperature (Folsom et al, 1980).

- Pre-Mixing. Uniform pre-mixing of fuel and air may be required to assure a uniform fuel/air ratio throughout the reaction mixture (Folsom et al, 1980).

- Hydrocarbons. The presence of hydrocarbons, such as methane, appears to promote the formation of fuel NO_x, due to reactions with intermediate reaction products which interfere with N_2 formation. However, a hydrocarbon gas does appear to promote the conversion of NO to N_2. This may have implications for the second stage (Folsom et al, 1980).

- Burnout. A rich stage for fuel-bound nitrogen "cracking" to N_2 requires a second lean stage for fuel burnout (Folsom et al, 1980).

- Thermal NO_x. The lean mixture in the second stage can be adjusted to reduce or minimize thermal NO_x formation (Folsom et al, 1980). However, the rich/lean combustor may not reduce thermal NO_x as effectively as a lean/lean combustor would for fuels without nitrogen compounds (Holt et al, 1989). Unnasch et al (1988) found that MBG combustion yielded higher thermal NO_x emissions than LBG, and speculated that this was attributable to higher flame temperatures.

- Turbulence. Fuel NO_x formation is expected to increase in turbulent flames. A laminar diffusion flame appears to allow for good conversion of ammonia to N_2 (Folsom et al, 1980).

- Fuel heating value. If fuel heating value is too low, combustion may not start in the fuel-rich zone. If combustion begins in the fuel-lean zone, conversion of ammonia to NO_x may be very high (Sato et al, 1989).

- Pressure. As combustor pressure increases, the conversion of ammonia to NO_x appears to decrease slightly, based on testing from 1 to 14 atm using a half-scale conventional combustor model (Sato et al, 1989).

- Efficacy. Rich/lean combustor tests using small scale combustors at relatively low pressures have achieved up to 95 percent conversion of ammonia to N_2 (Folsom et al, 1980; Unnasch, 1988; Notestein, 1989). Folsom et al attempted to develop ideal combustors of various designs on the bench-scale, but indicated that full-scale commercial designs may not be as successful in achieving NO_x reductions. The tests by Sato et al (1989) did not appear to achieve such high conversion rates. These tests involved perhaps more realistic full- and half-scale gas turbine combustors. In the Sato tests, ammonia conversion to N_2 was increased from a nominal value of 30 percent to a nominal value of 50 percent. This may be contrasted with the value of 30 percent typical of conventional combustors, discussed previously. These results imply that the efficacy of a commercial scale rich/lean combustor in reducing fuel NO_x emissions may be in doubt.

- CO emissions. In the Sato et al (1989) tests, CO emissions were below 100 ppm.

Another concept that has received some attention is catalytic combustion. However, in the near term, rich/lean combustion appears to be receiving more attention and testing. Therefore, for this study, rich/lean combustion is assumed as the most likely alternative for fuel NO_x control.

B.6.5.2 Combustion Efficiency and CO Emissions

CO emissions, which result from incomplete combustion of hydrocarbons or no combustion of CO in the fuel gas, are an indicator of poor combustion efficiency. Many of the measures which reduce NOx emissions, such as reducing flame temperature through wet injection or staged combustion, also tend to increase CO emissions by reducing the combustion efficiency. Most heavy-duty natural gas-fired and distillate oil-fired gas turbines have very low CO emissions (less than 5-10 ppm).

CO emissions increase at part load as the gas turbine combustor firing temperature is reduced during load-following (Entrekin and Edwards, 1987). Becker and Shulten (1985) report on part-load gas turbine combustion of low-BTU blast furnace gas in which it was difficult to achieve conversion of CO in the gas. However, coal gas has a higher hydrogen content than blast furnace gas, and may tend to combust more completely.

At the Cool Water demonstration plant, CO emissions were low with wet injection or fuel gas saturation. However, there are limits to fuel gas moisturization. As moisturization increases, the combustor flame becomes increasingly unstable, leading to pressure oscillations which can reduce the life of the combustor. At very high injection or moisturization rates, the combustion flame will ultimately blow out. Prior to the loss of flame, combustion efficiency will be low and CO emission will be high (Holt et al, 1989). The maximum fuel moisturization level is thus usually determined based on the point at which CO emissions begin to increase significantly.

A post-combustion flue gas CO catalyst can be used to convert CO to CO_2. The CO catalyst is relatively low cost, compared to SCR catalyst for NO_x control. However, the combination of reduced combustion efficiency and the exhaust gas pressure drop across the CO catalyst leads to reduced plant efficiency (Holt et al, 1989). The effects of flue gas from coal gas combustion on CO catalyst, such as catalyst masking or poisoning, may need to be assessed to determine the economics of CO catalysts in an IGCC process environment.

Incomplete combustion may occur due to local chilling of the flame, such as at points of secondary air entry (Cohen et al, 1987) or due to wet injection.

One advantage that coal gases have compared to natural gas or distillate oil with respect to combustion efficiency is the presence of hydrogen, which has a very high flame speed. This results in early ignition and promotes complete combustion (Holt et al, 1989).

<u>CO Emissions With MBG: Case OKC</u>

For a medium-BTU gas, CO emissions are not expected to be a major concern at baseload operation, particularly if there is hydrogen in the fuel gas. CO emissions could become a problem at part load if firing temperature is significantly reduced, or could become significant if high levels of water injection or fuel moisturization are used.

<u>CO Emissions with LBG: Cases ALH and AKH</u>

CO emissions are more of a concern for LBG than MBG. Corman (1986) reports an estimate for baseload CO emissions from a 100-MW class gas turbine firing LBG with a heating value of less than 150 BTU/scf to be approximately 10,000 tons/year. Corman implies the emissions would be higher for part-load gas turbine operation. However, in a phone conversation (1990) Corman appeared to have no concern about CO emissions with LBG. Pillsbury (1989) indicated that heating value is not the proper determinant of combustion efficiency, particularly because hydrogen is highly flammable and will tend to promote complete combustion even in LBG. Pillsbury and Corman both stated that the expected CO emissions are on the order of 10 ppm or less when firing LBG at baseload conditions.

B.6.5.3 Combustor Pressure Drop

The combustor pressure drop is one of the significant losses in the gas turbine system. Pressure losses are due to skin friction and turbulence. The rise in temperature during combustion increases velocity and momentum of the gases in the combustor, which leads to temperature-related pressure losses. However, the pressure drop due to turbulence is usually much higher than the pressure loss associated with the temperature ratio in the combustor. The build-up of carbon or other deposits on the combustor liner may also affect skin friction and/or turbulence-related pressure losses. Furthermcre, aerodynamically excited vibrations in the combustor could lead to deposits breaking away, which could result in turbine damage (Cohen et al, 1987).

B.6.5.4 Particles

The particle loading in the fuel gas may be considered to consist of refractory materials or carbonaceous materials. Refractory particles may pass through the combustor without alternation. They can split into smaller particles, or possibly agglomerate into larger particles. Carbonaceous material may be fully or partially combusted, leaving perhaps ash residues (Cincotta, 1984). The particle discharge from the combustor may affect turbine maintenance.

B.6.5.5 Combustor Life

The combustor life has an effect on maintenance and repair work and, hence, the cost of maintaining the gas turbine. For industrial gas turbines, combustor chamber lives of 100,000 hours are desirable (Cohen et al, 1987). However, deposition, erosion, corrosion, and vibrations can shorten the life of combustor components such as the liners, requiring more frequent liner replacement or more expensive materials. The modular nature of the combustor cans makes this type of maintenance routine. However, the cost will increase with the frequency of maintenance and repair. The possible presence of particulates and alkalis in the coal gas may lead to more costly maintenance compared to clean fuel (e.g., natural gas) fired gas turbines.

B.6.6 Turbine

The heavy-duty high firing temperature gas turbines assumed for this study typically employ three or four turbine rotor stages. The first two or three stages are subject to high thermal loadings due to the high temperature exhaust gas. Improvements in turbine rotor blade cooling technology have made possible increases in gas turbine firing (turbine inlet) temperatures while maintaining essentially constant bulk metal temperatures in the rotors and stators of the first turbine stage. Possible future improvements in materials and manufacturing processes (such as making turbine blades from a single crystal with no grain boundaries) may allow higher blade bulk metal temperatures, due to the improved strength of the material, and further increases in firing temperature (Smock, 1989).

A number of potential problems with the effect of hot combustion gas on the turbine have been identified in various reports. These include:

- Corrosion of hot gas path components from alkali metals
- Erosion of material from airfoils (rotor and stator blades) due to ash particles of sufficient size and quantity. This would likely exacerbate corrosion as well, as the airfoils are often coated with a corrosion resistant layer.

- Deposition of ash on hot gas path components, changing the aerodynamic characteristics of the turbine and resulting in loss of efficiency. This would also affect film cooling and the heat transfer from the hot gas to the airfoils.
- Blockage of film cooling holes, reducing the efficiency of blade cooling. This could lead to localized thermal stresses arising from thermal gradients in the blade material, affecting the operating life and/or sustainable firing temperature of the turbine

All of these possible problems would affect the gas turbine maintenance cycle, thereby affecting maintenance costs. Some or all of these affects could also require changes in gas turbine operation, such as a reduction in firing temperature or strict specifications on fuel gas composition.

B.6.6.1 Advanced Cooling Technology

Aircraft derivative gas turbines, and particularly military engines, have employed a variety of advanced turbine cooling techniques. These machines fire clean jet fuel, and as such are not subject to the exhaust gas contaminants expected in coal gas-fired units. Turbine blades and stator vanes subject to high temperature environments may have hollow internal cooling passages, through which compressed air is passed for convective cooling. These passages may have pin fins, to promote heat transfer from the metal to the cooling air. The cooling air is typically exhausted from the blade through holes in the blade tip or the trailing edge of the blade. The cooling air exhausted at the blade tip does provide some aerodynamic advantages by blocking against external bypass flow of exhaust gases between the blade tip and the rotor shroud. To further promote heat transfer in the internal cooling circuits, high velocity impingement of cooling air against the inside surface of a highly heated area may be used (referred to as impingement cooling). In addition, film cooling, in which some cooling air from inside the blade is vented near the leading edge of the blade, may also be employed. Film cooling results in a boundary layer of cooling air over the blade surface (Cohen et al, 1987; Dawkins et al, 1986).

The amount of cooling air required depends on the firing temperature, cooling air temperature, heat transfer features of the rotor and stator vanes, the material properties, and the design life of the system (Dawkins et al, 1986).

Based on testing of a prototype MS7001F engine with high (2,300 ^{o}F) firing temperature, GE reports that they expect their minimum hot gas component life design requirement to be met. The basis for this assertion is measurement of hot gas path metal temperatures to be 30 to 50 ^{o}F below the design values. The test was conducted with natural gas (Brandt, 1989). The gas path metal temperatures in a coal gas application may be affected by deposition or hole plugging, which is discussed in a later section.

The design of blades is complicated due to the changes in hot gas temperature across the blade surfaces, and the changes in temperature of cooling air inside the blade. Thus, the design must account for thermal gradients. Stresses in the blades may arise from thermal gradients (Cohen et al, 1987).

Any particles or liquid droplets which pass through the combustor and burn-out in the turbine nozzle or turbine first-stage may have deleterious effects on the thermal stresses in the hot gas path components.

B.6.6.2 Turbine Blade Materials

The selection of firing temperature for a gas turbine depends on both the turbine blade cooling technology employed and the blade materials. Three key criteria for selecting hot gas path materials, particularly for rotor blades, are: (1) creep-rupture properties; (2) hot corrosion resistance; and (3) hot oxidation resistance. The creep strength of a metal is a function of the bulk metal temperature. The time to obtain a standard 0.2 percent creep strain decreases as temperature increases. Also, the fatigue strength of a metal subject to cyclic stresses decreases as temperature increases. To provide blade strength, nickel-based superalloys may be used for rotor blades. To provide corrosion and oxidation resistance, coatings may be applied to the blade surfaces. Typical coatings include platinum-chromium-aluminide (Dawkins et al, 1986).

The GE MS7001F is reported to use a first-stage coating alloy containing cobalt, chromium, aluminum, and yttrium (Brandt, 1988). The blades for the GE turbine are reported to be manufactured using a technology called directional solidification that has been used for 20 years to make jet engine blades. In this casting method, the grain boundaries in the crystal structure of the metal are oriented to improve tensile strength, ductility, and fatigue strength. The use of this molding technology has permitted an increase in firing temperature of about 150 ^{o}F. Possible future improvements would be the casting of a single-crystal blade with no grain boundary, which would permit another 50 to 150 ^{o}F improvement in firing temperature (Smock, 1989). Inceases in firing temperature permit increased simple cycle efficiency. Such a design improvement is likely to be a long term development objective.

B.6.6.3 Deposition

Deposition of ash on surfaces in the hot gas path can restrict air flow, thus reducing turbine efficiency. Deposition of ash particles is expected to some extent in coal-fueled gas turbines (Cincotta, 1984). Deposits can also lead to plugging of cooling air outlet holes, particularly those used for film cooling, on the turbine rotor blades (Becker and Schulten, 1985; Dawkins et al, 1986). This can lead to increased localized temperature gradients that can result in thermal stress cracking, and can be exacerbated by the stress riser effect of the cooling air holes themselves. Also, film cooling can be affected by deposits on the turbine blades and hot gas channels. Such deposits, of certain size and consistency, can significantly alter the flow and heat transfer characteristics of the blades (Becker and Schulten, 1985).

Hot gas path blockage is generally expected with any gas turbine application involving a fuel containing ash particles. GE predicted a blockage rate of about 0.4 percent of the first-stage turbine nozzle area per 100 hours of operation at a 2,300 ^oF firing temperature, based on a system with two-stages of high efficiency cyclones (Cincotta, 1984). This implies nozzle cleaning every 2,500 hours, if up to 10 percent blockage is allowed. The assumption appears to be that this cleaning can be accomplished using off-line water washing, for example.

GE conducted some tests with a turbine simulator to determine possible effects of ash deposition. No measurable deposits were found on the airfoils. However, the tests were only 57 hours in duration (Corman, 1986).

Evaluation of deposition appears to require a long term testing program, which in reality may not be realized until a demonstration plant is built and operating. The effect of deposition on the heat transfer characteristics of the turbine blades might be to require a reduction in firing temperature or to increase the frequency of blade replacements. Thus, either performance and/or cost may be affected by these types of problems.

B.6.6.4 Erosion

Erosion occurs due to contact of particles with sufficient mass or velocity to remove material from hot gas path surfaces, particularly rotor and stator vanes. Some possible sources of particles contributing to erosion include: particles not removed from the fuel gas in cyclones or barrier filters; break-away deposits from the fuel nozzle, fuel valves, combustor lining, transition piece, or turbine nozzles; and carry-over of sorbent material from the zinc ferrite sorbent bed and, if included in the system, alkali removal sorbent bed. GE reported that they expect to achieve a particle size distribution and loading using two-stages of high efficiency cyclones to be within the erosion tolerance of the gas turbine materials (Cincotta, 1984).

However, some speculate that cyclones are insufficient to avoid the build up of particles and, hence, pressure drop in the zinc ferrite absorber bed. Therefore, barrier filtration upstream of the zinc ferrite unit may be required, in lieu of a single-stage cyclone.

There is also speculation that a cyclone downstream of the zinc ferrite absorber may not be needed. Most design studies assume a cyclone between the absorber and the gas turbine combustor to capture any catastrophic loss of sorbent or unusual entrainment of sorbent, as well as to provide for additional removal of particles still present from the gasifier.

B.6.6.5 Corrosion

The most widely expressed concern regarding hot gas path corrosion is due to the presence of alkali in the exhaust gas. For systems with cold gas cleanup, alkali are not expected to pose a corrosion threat because it is believed that below 1,200 to 1,400 °F, alkali condense onto particles in the gas stream (METC, 1987; Notestein, 1989), which are in turn removed very effectively by wet scrubbing. For hot gas cleanup systems using the zinc ferrite process, the fuel gas temperature in the particulate removal device is typically expected to be about 1,100 °F. The removal efficiency of alkali which condense on particles depends on the alkali concentration on the particles as a function of particle size, and the particle removal efficiency as a function of particle size. The expectation is that, because the smaller particles have a larger surface area per unit mass, there will be a larger concentration of condensed alkali on the smaller particles (Cincotta, 1984).

Several have reported that there is evidence that the alkali in coal gas may not pose as much of a threat as an equivalent concentration of alkali in petroleum fuels. The suggestion is that alkali in the coal gas are "gettered" by aluminosilicate ash materials (METC, 1987; Notestein, 1989). This, combined with the absence of "catalytic" elements, such as vanadium and molybdenum, are believed to reduce the ability of the coal gas alkali to cause corrosion.

In the event that particulate removal proves to be insufficient for alkali control, several alkali control technologies for hot gas cleanup systems have been explored (Notestein, 1989). Perhaps the most promising of these is an absorber utilizing emathlite, a naturally occurring clay (Bachovchin, 1987).

B.6.7 Judgments About Uncertainties

Compared to the fixed bed gasifier and the zinc ferrite desulfurization process, the gas turbine process area proved to be the most difficult of the three with respect to obtaining expert judgments regarding uncertainties. Technical experts at both DOE/METC and a leading gas turbine manufacturer were approached regarding performance- and emissions-related uncertainties. The technical experts at DOE/METC were selected by DOE/METC management for participation in the survey. The author followed-up the written survey responses with telephone calls. The experts at the gas turbine manufacturer were contacted directly by the author by phone and by letter.

In most cases, it was not possible to obtain quantitative judgments regarding uncertainties from specific experts. The reasons for this are several. A pervading theme is the proprietary nature of gas turbine designs and performance information, particularly that of the GE gas turbine commonly assumed in studies of IGCC systems. Engineers at GE were reluctant to provide information over the phone or directly to Carnegie-Mellon other than what is already published. A second pervading theme is the relative lack of information upon which an expert would base the types of judgments requested. One expert commented that the literature review given in the preceding sections was fairly comprehensive, and that if the review did not uncover the needed information, it is either not available or not published because it is proprietary. Some of the questions posed are the subject of current and ongoing research in preliminary stages, particularly with respect to combustor performance and emissions. The results of such work are not yet available to help an expert make an informed judgment. The experts at DOE/METC, some of whom previously worked for gas turbine manufacturers, may still be bound by confidentiality agreements with their former employers and, hence, may be unable to provide detailed information even if they had access to it.

The gas turbine process area questionnaire was distributed to three experts at DOE/METC, of whom two responded. The responses of these two experts, referred to as Expert GT-1 and Expert GT-2, are summarized in Sections B.6.7.1 and B.6.7.2, respectively. Furthermore, conversations with several other experts are summarized in Section B.6.7.3. The information obtained from the literature review in the previous sections and from the gas turbine experts was used to inform the development of judgments of uncertainties on the part of the author. The basis for assigning uncertainties to gas turbine parameters is discussed in Section B.6.7.4.

B.6.7.1 Expert GT-1

A DOE engineer, Expert GT-1, who was given an uncertainty briefing packet responded, "I do not have any knowledge, data or information germane to answering these questions." However, the engineer did provide extensive comments on the default assumptions provided in the technical background paper, Part 2, of the briefing packet. With respect to estimating uncertainties, the engineer explained that "gas turbine manufacturers would be the prime source of this data, *if it exists.*" However, such data "would be proprietary" because "manufacturers spend millions of dollars of internal R&D funds to get reliable emissions data." Furthermore, the DOE engineer noted that 36 references were cited in Part 2 of the briefing packet. "If the contractor cannot find what they want in these references, manufacturers and users are not publishing the data because it is proprietary or because they don't have the information."

Expert GT-1 did provide gas turbine design information needed for the modeling studies. He indicated that a gas turbine fuel valve pressure drop of 70 psi is reasonable, and that the GE MS7001F gas turbine cooling circuits used about 12 percent of the compressor inlet air. In a follow-up phone conversation, Expert GT-1 indicated that DOE/METC is seeking the development of a high temperature gas turbine fuel valve that would have a substantially lower pressure drop than conventional designs. The expert also indicated that, with proper upstream particulate control, he doesn't anticipate a problem with gas turbine combustor pressure drop build-up.

B.6.7.2 Expert GT-2

Another DOE engineer, Expert GT-2, provided mostly qualitative answers in response to some of the questions posed in the questionnaire. He indicated that "the preparation of Part 2 was excellent, Part 1 was more information than I needed or wanted, and Part 3 needs to be reevaluated and reworked." With respect to Part 2, the engineer stated, "the person or group who prepared this summary is to be commended for their objective and unbiased presentation of the material." However, the engineer indicated that Part 3, the questionnaire, was too time consuming as presented to him: "in Part 3, you are asking the respondents to devote an unreasonable amount of time to answering 12 compound questions in considerable detail."

Expert GT-2 did provide some quantitative information in his response to the questionnaire. He indicated that a "70 psi fuel valve pressure drop is typical of GE only." In a follow-up phone conversation, the expert indicated that Westinghouse and United Technologies claimed they could supply fuel valves with 20 to 30 psi pressure drops. A high pressure drop is preferred for control reasons, but may not be necessary. "A 20 psi pressure drop fuel gas valve is possible with a butterfly valve and if necessary a separate block valve arrangement."

High efficiency cyclones may not be adequate for overall particulate collection due to their pressure drop, which "will not be economically viable for overall particulate removal needs." Expert GT-2 indicated that barrier filtration would be needed upstream of the zinc ferrite process, but that a cyclone between the zinc ferrite bed and the gas turbine may be acceptable for removing entrained sorbent material. If barrier filtration is used, "particulate deposition will not be a factor in [fuel] valve design."

Expert GT-2 indicated that NO_x and CO emissions will be held below the requirements of current regulations. "Testing of rich-lean combustors has indicated this is possible regardless of incoming ammonia concentrations below 8,000 ppm value." With respect to NO_x emissions from rich/lean combustors firing fuel gas containing significant fuel-bound nitrogen, Expert GT-2 said that the "worst case fear" is 50 percent conversion of ammonia to NO_x. However, conversions of 10 percent or less are expected, with CO emissions below 100 to 200 ppm, based on early test results.

Fuel gas quality could be a key determinant of NO_x and CO emissions. For the air-blown systems, the fuel heating value is low. Lack of control over gasifier operations could lead to reductions in fuel gas quality which, in turn, could affect flammability and flame stability. Low flammability or flame instability could lead to problems with emissions. Fuel heating values of less than 80 BTU/scf could create problems with combustion efficiency and emissions. The fuel gas heating value is related to the coal supply: variability in coal properties could cause variation in fuel gas composition and heating value.[1] Fluidized bed gasifiers tend to "bounce and slug," leading to variable fuel gas composition. Problems with fuel gas flammability would tend to be alleviated if partial air separation technology, such as membranes, become economically attractive. This would lead to a greater concentration of oxygen in the blast "air" to the gasifier, and thus increase the fuel gas heating value.

Expert GT-2 stated that the system with cold gas cleanup incorporates less risk than the two air-blown systems considered in this study.

Expert GT-2 indicated that reduction in gas turbine firing temperature could be used to suppress alkali deposition on the turbine blades. However, alkali cleanup using an emathlite absorber vessel would be recommended as a low technical risk alternative to low firing temperature or unconventional and probably uneconomical maintenance. Sulfur in the combustion product gas can react with alkali to form alkali sulfates, which are a source of corrosion on the gas turbine blades. For Case AKH, there may be some calcium in the fuel gas entrained from the calcium-based sorbent used for gasification in-bed desulfurization. Calcium can also form an alkali sulfate that can be "as bad" as the sodium or potassium compounds typically cited as a corrosion concern. The alkali sulfates may condense onto the turbine blades below a certain temperature. The oxygen content of the

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[1] Regarding the variability of coal properties and supplier promises regarding the same, the expert made a colorful reference to similarities between "damn liars" and "coal suppliers."

exhaust gas in the turbine affects the rate of corrosion. For the nickel alloys typically employed on turbine blading, sulfidation of nickel alloys increases as the oxygen content decreases. The oxygen content of exhaust gas from air-blown systems is lower than that for oxygen-blown systems. To avoid turbine blade deposition and corrosion problems, an operator could try to switch to a lower-alkali coal.

Expert GT-2 also described gas turbine compressor air-extraction for gasifier blast air as "a pipe dream or government-funded." Until a large market appears, it is more likely that a separate booster compressor will be used than extracting air from the gas turbine compressor outlet.

B.6.7.3 Other Experts

Several other gas turbine experts were contacted, at two gas turbine manufacturers and an architect/engineer firm. One expert indicated that very little development effort would be made for gas turbines in IGCC service; instead, the manufacturer would provide a fuel specification to which the plant operator must adhere for gas turbine performance guarantees to be valid. However, the expert indicated that even with fuel specifications, it is still not certain what will happen over a 20 to 30 year life cycle with respect to performance and cost. Possibly loss of output or shorter maintenance cycles will be encountered for gas turbines in IGCC service in systems with hot gas cleanup. The same expert indicated that high efficiency cyclones are expected to be sufficient for upstream particulate removal, including alkali control, and that such cyclones are to be tested as part of a clean coal program demonstration project. With respect to NO_x emissions, the expert indicated that it is not yet known what levels will be achieved with the air-blown systems, which feature significant concentrations of ammonia in the fuel gas. In the laboratory, it is possible to get less than 10 ppm NO_x emissions with fuel gas ammonia concentrations of 2,000 ppm. However, combustor technology may not easily scale-up from the lab to commercial applications. For example, a premix lean-burn system for reducing thermal NO_x achieved less than 5 ppm emissions in the lab, but in the field achieves only less than 25 ppm. A standard combustor on the GE Frame 6 model converts 50 to 60 percent of fuel-bound nitrogen to NO_x.

A second expert at a gas turbine manufacturer indicated that tests of new combustor cans typically cost in excess of $100,000 per run. In testing of a staged lean-lean combustor for thermal NO_x control, the manufacturer expected 50 percent NO_x reduction compared to more conventional combustor designs, but obtained no reduction. This is an example of the difficulty of scaling-up combustor technology. Low NO_x technology employs concepts such as premixing and staged combustion. However, these technologies cannot be designed by modeling studies; they must be empirically arrived at through testing, due to their complexity. For current combustor technology, about 90 percent of fuel-bound nitrogen is converted to NO_x. Normally, a customer is told to assume that 100 percent of fuel-bound nitrogen will be converted.

A third gas turbine manufacturer expert indicated that in the current combustor design for the default gas turbine used in this study, almost 100 percent conversion of fuel-bound nitrogen to NO_x may be assumed. A target for development of rich/lean combustor technology would be achievement of fuel-bound nitrogen conversions of less than 10 percent, with 20 percent conversion being a possible worst case achievable value.

With respect to gas turbine cost, an industry engineer indicated that when demand for IGCC systems increases in the future, manufacturers of high efficiency heavy-duty gas turbines will be a position to significantly increase price, even beyond the cost of modifications needed for IGCC application.

B.6.7.4 Discussion of Uncertainties

While a comprehensive set of expert elicitations comparable to those for the Lurgi gasifier and fixed bed zinc ferrite process areas was not obtained for the gas turbine process area, a significant amount of information was gleaned from published literature, as reviewed in Sections B.6.1 through B.6.6. Also, a number of practical insights were obtained from discussions with several engineers, as discussed in the preceding sections. This information can be used to construct plausible estimates of uncertainty that can be used in initial uncertainty screening studies.

In the questionnaire given to the gas turbine experts at DOE, there were questions about eight technical or cost related subjects. These are:

- Fuel valve and nozzle pressure drop
- Thermal NO_x emissions from MBG
- Thermal NO_x emissions from LBG
- Fuel NO_x emissions from LBG
- CO emissions
- Combustor pressure drop
- Turbine inlet temperature
- Maintenance costs

Each of these will be discussed in turn. Not all of these will be treated probabilistically.

Fuel Valve and Nozzle Pressure Drop. While particle deposition could potentially have an effect on pressure drop on the fuel gas path, the experts who commented on this possibility indicated that it is not likely to be a concern. Proper particulate removal upstream of the gas turbine is believed to avoid such a problem. If such a problem were to occur, it could be corrected by increased frequency of maintenance, without significant degradation of performance. Thus, uncertainty due to fuel valve and nozzle deposition may more appropriately be reflected as an uncertainty in maintenance cost.

The primary effect on pressure drop is the design of the fuel valve. Most conceptual design studies have assumed a minimum pressure drop between the gasifier and gas turbine of about 75 psia (e.g., Corman, 1986), with other studies assuming much larger pressure drops. A design improvement sought by DOE is to reduce the fuel valve pressure drop. Therefore, rather than treat this design feature probabilistically, it would be more appropriate to evaluate "conventional" and "advanced" fuel valve designs as separate cases. The assumption used here is that current fuel valves have a pressure drop of about 70 psi, while advanced fuel valves would have a pressure drop of 20 psi. (Of course, uncertainty in the pressure drop that would be obtained with an advanced fuel valve system could be represented probabilistically. Such an uncertainty range might be 10 to 20 psi, based on conversations with several engineers.)

Thermal NO_x Emissions from MBG. The thermal NO_x emissions from medium-BTU gas (MBG) fired gas turbines are expected to be within the current New Source Performance Standards (NSPS) for gas turbines, which require less than 75 ppmv NO_x corrected to a dry, 15 percent oxygen basis and corrected to a gas turbine efficiency of 25 percent. The efficiency correction results in an increase in the allowable emission rate to over 100 ppmv on a dry, 15 percent oxygen basis for typical gas turbines used in IGCC application. The NSPS level can be achieved with either wet injection or low NO_x combustor designs. In the Cool Water demonstration plant, actual NO_x levels of 21 ppmv were measured (Cool Water, 1986). Low emission levels may be achievable without wet injection using dry low NO_x combustor designs.

For MBG, the thermal NOx emissions are assumed to be low, ranging from about 25 to 75 ppmv as an initial assumption.

Thermal NO_x Emissions from LBG. The thermal NO_x emissions from low-BTU gas (LBG) fired gas turbines are expected to be lower than NO_x emissions from MBG. Uncontrolled emissions may be as low as 10 to 50 ppmv (Unnasch, 1988). Thermal NO_x emissions in air-blown systems are expected to be significantly less than uncontrolled fuel NOx emissions. As an initial assumption, thermal NO_x emissions from LBG combustion are assumed to be between 10 and 75 ppmv on an actual basis at the gas turbine exit.

Fuel NO_x Emissions from LBG. For combustors not employing rich/lean combustion, it is expected that between 50 and 100 percent of ammonia in the fuel will be converted to NO_x. The actual conversion rate may depend on the combustor design, fuel ammonia concentration, and other factors. Based on consideration of information in the literature and conversations with various engineers, an initial characterization of uncertainty is proposed. As an initial assumption, the ammonia conversion to NO_x is assumed to range from 50 to 100 percent, with a mode at 90 percent. This characterization of uncertainty is represented with a triangular distribution. As more details regarding the performance of specific combustor cans become available, this estimate can be revised accordingly.

For rich/lean combustor technology, it is possible that very low levels of ammonia conversion can be achieved, according to some experts. Based on discussions with several experts, an initial uncertainty estimate ranging from a low of 0.1 percent ammonia conversion up to 20 percent ammonia conversion is proposed, with a mode at 10 percent conversion. The distribution for this uncertainty is assumed to be triangular.

CO Emissions. CO emissions from systems firing MBG are not expected to be a concern. Data from Cool Water (1986) indicate that CO emissions for a MBG with steam injection for NOx control were 40 to 90 ppmv. CO emissions were higher in another reported emission test at Cool Water (1988), ranging from less than 100 to 190 ppmv on three different coals. Several experts indicated that CO emissions are expected to be below 10 ppmv for mature commercial systems. Therefore, for MBG, CO emissions are assumed to range from 1 to 200 ppmv, with a mode at 10 ppm.

For LBG, less data are available regarding CO emissions. However, there is more concern that CO emissions may be a problem with low heating value fuels, particularly for high firing temperature gas turbines. The expected CO emission level for mature gas turbine combustors in LBG service is 10 ppmv according to several experts. However, it is possible that CO emissions could be several hundred ppmv. One estimate was that CO could be as high as 10,000 tons/year for a Lurgi-based system (Corman, 1986), which is equivalent to about 350 ppmv. Therefore, CO emissions for LBG are assumed to range from about 1 to 350 ppmv.

Combustor Pressure Drop. While particulate deposition may potentially lead to an increase in combustor pressure drop, several experts indicated that this is not likely to be a problem for mature commercial systems. If a problem is encountered, it can likely be handled through increased maintenance. Therefore, it was decided not to treat this parameter probabilistically in the current analysis.

Turbine Inlet Temperature. One expert indicated that there might be conditions under which firing temperature would have to be reduced to avoid alkali deposition problems. However, a change in firing temperature may only change the location in the turbine where deposition occurs due to changing the temperature profile, as opposed to eliminating deposition altogether. Other measures, such as switching coal or changing maintenance procedures, could also be employed. Therefore, it was decided not to treat this parameter probabilistically in the current analysis.

Gas Turbine Capital Cost. Historically, gas turbines have been used mostly in applications involving relatively little risk. These applications involve use of relatively clean fuels for which the gas turbines were designed. IGCC systems, however, represent a new process environment for gas turbines. Oxygen-blown IGCC systems with cold gas cleanup pose relatively little risk to gas turbines compared to air-blown IGCC systems with hot gas cleanup. While the former systems may tend to be expensive, they are believed to result in high removal rates of all fuel gas contaminants that might harm the gas turbine, such as particles and alkali. Also, the relatively high heating value of MBG compared to LBG means that there is a smaller fuel gas volumetric flow rate that must be handled by the fuel valve system, resulting in lower fuel valve-related costs. The air-blown systems propose compressor air extraction for the gasifier blast air, which is an additional complication for the LBG-fired gas turbine. For these reasons, the capital cost of gas turbines for LBG application is expected to be higher than those for MBG.

While many conceptual design studies have been carefully prepared, there are often inherent biases. For example, equipment vendors for key process equipment are usually asked to provide a cost estimate in response to a specification provide by an architect/engineer firm. While these vendors may provide their best estimate at the time of the conceptual design study, such estimates may not fully anticipate the types of problems that would be revealed by a more detailed analysis for an actual, as opposed to conceptual, design project. Furthermore, an conceptual estimate prepared at a time when no machines are actually ordered may be different from the actual cost incurred in a competitive market. If the IGCC gas turbine market becomes supply-limited, the price of gas turbines, and particularly ones requiring special modifications for LBG, could increase substantially.

Because of both technical and market considerations, the capital cost of the gas turbine process area for all three IGCC cases is assigned an uncertainty. Typically, process contingency factors for gas turbines, which are the traditional approach to characterizing uncertainty in capital cost estimates, have been very low. This has been true even for IGCC conceptual design studies, in spite of needed modifications and a different process environment compared to typical gas turbine applications. For example, Corman (1986) uses a contingency of about two percent for gas turbines used in an air-blown system with hot gas cleanup, while a typical Fluor study (Fluor, 1985) uses a contingency of 5 percent for a gas turbine in an oxygen-blown system with cold gas cleanup. However, an industry expert indicated that gas turbine costs in a competitive market could increase as much as 50 percent from values used in conceptual design studies. Furthermore, low levels of contingency are inconsistent with even the rule-of-thumb values suggested by EPRI for systems that do not have commercial experience. As an initial estimate of uncertainty for gas turbine capital cost, for air-blown systems with compressor air extraction and LBG the uncertainty is assumed to be a zero to 50 percent increase in capital cost. For oxygen-blown systems with MBG, the uncertainty is assumed to be 0 to 25 percent.

Of course, the model can be exercised with alternative judgments regarding uncertainties in any given parameter. These initial judgments regarding gas turbine uncertainty can be evaluated to see if they contribute significantly to overall uncertainty in total capital cost and the cost of electricity. If this uncertainty is found to be important, it may be worthwhile to seek more detailed expert judgments for this particular parameter. If it is not significant, then the model results would not be critically dependent on the specific assumptions used here. Thus, even with limited information, it is possible to perform a screening analysis of uncertainties to determine if additional data collection or expert elicitation is warranted.

<u>Gas Turbine Maintenance Cost</u>. The maintenance cost of the gas turbine may be significantly affected by particle deposition in the fuel valve, combustor, or hot gas path in the turbine. Maintenance cost may also be affected by corrosion on turbine rotor and stator vanes, requiring more frequent reblading than would be necessary for a gas turbine in a more conventional application. Therefore, while in the optimistic case maintenance costs would be similar to that for service with cleaner fuels such as natural gas, there is the possibility that maintenance costs could be higher, and little chance that costs would be lower. A typical maintenance cost factor for gas turbines in conventional service is 1.5 percent per year of the direct capital cost. As an initial characterization of uncertainty, it is assumed that the maintenance cost may be has high as 6 percent per year for the system with hot gas cleanup and as high as 3 percent per year for the system with cold gas cleanup. A most likely value of 2 percent per year is assumed for both systems,

Table B-25. Summary of Assumed Gas Turbine Process Area Uncertainties

Description	Units	Distribution	Parameters[a]			
Thermal NO_x, fraction of air nitrogen fixated						
Oxygen-blown system		Uniform	$2.5x10^{-5}$	to	$7.5x10^{-5}$	
Air-blown systems		Uniform	$1.0x10^{-5}$	to	$7.5x10^{-5}$	
Fuel NO_x, % conversion of NH_3 to NO_x						
Pre-mix lean-burn combustor		Triangular	50	to	100	(90)
Rich/lean staged combustor		Triangular	0.001	to	20	(10)
Unconverted CO, wt-% of CO in fuel gas						
Oxygen-blown system		Uniform	0.9998	to	0.9999	
Air-blown systems		Uniform	0.9772	to	0.9999	
Gas Turbine Capital Cost Uncertainty, % of direct capital cost						
Oxygen-blown system		Uniform	0	to	25	
Air-blown system		Uniform	0	to	50	
Gas Turbine Maintenance Cost, %/yr of direct cost						
Oxygen-blown system		Triangular	1.5	to	3.0	(2.0)
Air-blown system		Triangular	1.5	to	6.0	(2.0)

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

representing the tendency that maintenance costs for a gas turbine in IGCC service would be higher compared to conventional service. As with any other model parameter, this characterization of uncertainty can be revised as more data become available, or to represent the judgment of other experts. As part of an initial uncertainty screening study, the importance of uncertainty in gas turbine maintenance cost to uncertainty in overall operating and maintenance (O&M) costs can be ascertained to determine whether revision to this parameter is warranted.

Table B-25 summarizes the uncertainties which have been assumed for the gas turbine process area.

B.7 Other IGCC Process Uncertainties

In the previous four sections, uncertainties related to the performance and cost of the gasification, hot gas cleanup, and gas turbine process areas of three IGCC systems have been discussed. In this section, uncertainties assigned to additional categories of parameters included in the probabilistic case studies of these IGCC systems are described. These categories include: (1) cost model parameters; (2) direct capital cost uncertainties; (3) maintenance cost factor uncertainty; (4) operating cost uncertainty in unit costs of consumables and unit prices of byproducts; and (5) statistical uncertainty in regression models for process area direct capital costs and auxiliary power requirements. These uncertainties are described in turn in the following sections.

B.7.1 Cost Model Parameter Uncertainties

An important feature of this work is the development of cost estimates for competing technologies on a consistent basis to permit comparative analysis. This section addresses the development of deterministic and probabilistic estimates for cost model parameters that are common to all of the IGCC systems considered here. See Appendix A for more details regarding the structure of the capital, operating and maintenance (O&M), and levelized cost models.

The cost model parameters can be grouped into three categories. These are: (1) capital cost; (2) operating cost; and (3) financial. The latter category of parameters influences the fixed charge factor (also called the "capital recovery factor") and variable cost levelization factor used to calculate the levelized plant cost of electricity production. For each parameter, a "best guess" value is assumed in deterministic modeling studies. In addition, for selected parameters, an uncertainty distribution is ascribed for probabilistic modeling studies. The best guess and uncertainty assumptions are summarized in Table B-26.

Four capital cost parameters are common to all of the IGCC systems. The engineering and home office cost factor is intended to include the costs of: (1) engineering, design, and procurement labor; (2) office expenses during design; (3) licensor costs for basic process engineering services; (4) office burdens, benefits, and overhead costs; and (5) fees or profit to the architect/engineer. In preliminary cost estimates, these costs are represented as a multiplier factor of other process-related capital costs. Standard industry practice, as recommended by the Electric Power Research Institute (EPRI), is to assume that engineering and home office costs is an additional cost ranging from 7 to 15 percent of

Table B-26. Summary of Assumed Values and Uncertainties for IGCC Cost Model Parameters.

Description	Units	"Best Guess"	Distribution	Parameters[a]		
<u>Capital Cost Parameters</u>						
Engineering and Home Office Fee	fraction	0.10	Triangular	0.07	to	0.13 (0.10)
Indirect Construction Cost Factor	fraction	0.20	Triangular	0.15	to	0.25 (0.20)
Project Uncertainty	fraction	0.175	Uniform	0.10	to	0.25
General Facilities	fraction	0.20				
<u>Operating Cost Parameters</u>						
Capacity Factor	fraction	0.65				
Labor Rate	$/hr	19.70	Normal	17.70	to	21.70
Number of Shifts	shifts/day	4.25				
<u>Cost Year and Financial</u>						
Plant Cost Index		351.5				
Chemicals Cost Index		411.3				
Construction Interest	%/yr	10				
Construction Years	years	4				
Booklife	years	30				
Inflation Rate	%/yr	0.0[b]				
Sales Tax	%	5				
Real Return on Debt	%/yr	4.6				
Real Return on Preferred Stock	%/yr	5.2				
Real Return on Equity	%/yr	8.7				
Debt Ratio	fraction	0.50				
Pref. Stock Ratio	fraction	0.15				
Federal and State Tax Rate	fraction	0.38				
Investment Tax Credit	fraction	0.0				
Property Taxes and insurance	%/yr	2.0				

[a] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.
[b] An inflation rate of zero is used in "constant dollar" cost analyses.

the total direct capital cost, indirect capital cost, and sales tax. Most conceptual design studies assume a factor of about 10 to 12.5 percent. These factors may be based either on the judgment of a cost estimator or on the historic cost estimating history of a company. However, in lieu of such experience, a symmetric triangular distribution representing uncertainty in the actual costs that would be incurred for a real project is assumed. The central value of this distribution (which is the median, mean and mode due to symmetry) is taken to be the 10 percent value representative of typical cost estimates. The minimum and maximum values are 7 and 13 percent, respectively. The mode is the single "most likely" value of the distribution, and is taken to be the same as the deterministic "best guess" estimate. The purpose of this distribution is to determine whether uncertainty in engineering and home office costs is an important consideration affecting total capital cost. Clearly, these costs would become better defined for an actual project. However, in a preliminary cost estimate, the details of engineering and home office costs may not be fully known and, hence, there may be uncertainty regarding the actual costs that would be incurred for a particular project.

The indirect construction cost factor used in preliminary cost estimates is intended to account for the costs of workers' benefits, supervision, administration, and other construction related costs which are not part of the permanent capital equipment at the plant (e.g., temporary construction offices). The indirect construction cost factor would be replaced by more detailed cost estimates in later stages of engineering analysis of a particular project. From conceptual design studies of IGCC systems, it appears that indirect construction costs range from about 15 to 25 percent of the total direct capital cost. Here it is assumed that the "best guess" value is 20 percent, with a symmetric triangular distribution. The uncertainty ascribed here is intended to represent the uncertainty in prediction actual indirect costs with the limited information available to develop a preliminary cost estimate.

The project uncertainty is intended to reflect the expected increase in capital cost that would result from a more detailed and comprehensive cost estimate at a later stage of a specific construction project. The project-related uncertainty is expected to be reduced as a particular project progresses to more detailed phases of design and cost estimating. For projects in preliminary stages of development, EPRI (1986) recommends a "project contingency" factor of 15 to 30 percent, while projects for which a "detailed" cost estimate has been developed may require a project contingency of 10 to 20 percent. For an initial estimate of project-related uncertainty, a range of 10 to 25 percent is assumed. This range is slightly lower than that for preliminary estimates, but wider than that for detailed estimates. This uncertainty is taken to be uniformly distributed. For deterministic estimates, the central value of 17.5 percent is used.

The cost of general facilities includes a long list of auxiliary equipment required at a power plant (see Appendix A for more detail). This factor is typically around 20 percent of the direct cost of the major process areas.

In most conceptual design studies of baseload fossil-fuel electric power plants, it is assumed that the power plant produces the equivalent of full-load power 65 percent of the year. This assumption has historical roots in the annual average U.S. capacity factor for fossil fuel power plants.

The operating personnel labor rate is nominally about $19.70 per hour. A modest uncertainty is assigned to this parameter to represent the variability in labor rates from one site to another. The number of shifts per day is based on an eight-hour shift plus some addition shifts to account for vacation/sick time.

The financial assumptions in Table B-26 are taken from the EPRI Technical Assessment Guide (1986) and represent standard assumptions used in many design studies. The costs in this work are reported in 1990 dollars based on the *Chemical Engineering* Plant Cost Index and Chemical Cost Index. Levelized costs are estimated based on constant dollars, in which inflation is taken to be zero. Constant dollars are used to eliminate confusion over often varying assumptions about inflation rates from one study to another, which can lead to large differences in the magnitude of reported costs. The financial assumptions are used to calculate the fixed charge factor for estimating annual capital recovery and the variable cost levelization factor according to EPRI guidelines. For these case studies, the financial assumptions are held at their single point-estimate values.

B.7.2 Direct Capital Cost Uncertainties

For technologies in early stages of development, there is often uncertainty in the cost of particular pieces of equipment or for entire process areas. One source of uncertainty is in the key performance variables that affect equipment design and sizing. For example, if a flow rate is uncertain, then the size of equipment needed to accommodate that flow is also uncertain. This type of interaction between uncertainty in performance and uncertainty in the cost of a process area is explicitly captured in the cost models developed for all of the clean coal technology systems studied in this work.

However, even if key performance variables were known with certainty, uncertainty may still remain in the cost of equipment or a process area. For example, cost estimates developed in early stages of technology development may not capture all of the process area capital costs for two reasons. First, preliminary costs may not capture all of the costs that would be revealed by a finalized cost estimate based on more detailed engineering analysis of the system. This type of uncertainty is usually addressed using "project contingency factors" as discussed in the previous section. Second, even detailed estimates developed early in the development of a technology may not capture all of the costs that will become apparent after a demonstration or the first commercial scale plant has been built.

With respect to the latter source of uncertainty, potential problems that could be encountered in a first full-scale plant might include corrosion, fouling, effects on process chemistry due to trace contaminants or operating conditions not anticipated in design work, and so on. While these types of problems may seriously hamper the operability of a first-of-a-kind plant, design changes can be incorporated into later plants to minimize such problems. However, the capital cost of, say, a fifth-of-a-kind plant may tend to be higher than the estimated cost developed prior to building the first full-scale plant, because of scope changes in the capital cost as the technology matures and because preliminary cost estimates developed for innovative process technologies generally are biased low.

Studies by Rand Corporation have lent a quantitative basis to the notion that capital costs are often severely estimated in preliminary cost estimates for innovative technologies (e.g, Merrow et al, 1981; Milanese, 1987; Hess et al, 1989). Most of the Rand work has addressed difficulties in estimating the cost of the first full scale chemical process plant embodying new technology. However, Rand has also considered the effect of design improvements on the cost of later plants. It appears that, even assuming cost improvement occurs between the first- and fifth-of-a-kind plant, cost estimates developed prior to the first-of-a-kind plant still under-predict the cost of a fifth-of-a-kind plant.

In deterministic cost estimates, process area "contingency factors" are often used to represent the expected increase in cost that usually accompanies unproven technology. However, with the possible exception of propriety information held by architect/engineer firms, there is little historic cost estimating data available to verify the accuracy of any particular value for a contingency factor. For example, an important type of information rarely reported in design studies is the *probability of cost overrun* associated with any given contingency factor. There is no unique value of contingency cost unless the notion of risk is included; the selection of such a value should be based on the risk (in terms of probability of cost overrun or partial mean of a cost overrun) that a decision-maker is willing to take that costs will be higher than the budget estimate. Thus, the selection of a contingency factor is, in itself, an uncertain task. EPRI (1986) suggests some guidelines for ranges of contingency factors depending on the state of technology development. But EPRI does not provide any indication of how a contingency factor should be selected to correspond to a particular probability of cost overrun.

In this work, uncertainties in process area capital costs are expressed, on the same basis as the process contingency factor as defined by EPRI, as a percentage of the process area capital cost using the same formula as would be used to estimate contingency cost. However, rather than select a single point-estimate value for a "contingency factor," instead a range of values described by a probability distribution is used. The purpose of such probability distributions is to more appropriately represent the uncertainty in predicting the process area capital cost for a fifth-of-a-kind plant in a cost estimate developed before even a demonstration plant has been built.

Published values of contingency factors for each process area have been used as a guide in selecting the ranges of values for uncertainty in process area costs, particularly for assigning relative magnitudes of uncertainty between process areas in different stages of development. However, in some cases, such as already discussed for the Lurgi gasifier and the gas turbine process areas, these factors seem to be unreasonably low when considering the new process environment and operating conditions to which these technologies would be subject in an IGCC plant. Therefore, in some instances the author has revised the basis for both process contingency factors used in deterministic estimates and for the process area capital cost uncertainty ranges used in deterministic estimates.

The default approach taken here is to assume that there is some small probability that, for any given process area, costs may not increase significantly. However, for process areas that have not been proven at a full-scale, an upper bound of roughly 50 to 80 percent additional cost has been assumed. While these upper limits are much higher than values typically assumed for deterministic contingency factors, they may actually more accurately reflect the type of cost increase that usually accompanies innovative process technology, and particular technology which involves extensive handling of solids (such as gasifiers or hot gas desulfurization systems). In most cases, because little information was available to develop detailed judgments of uncertainty, uniform probability distributions are assumed. In all cases, the deterministic value of the contingency factor used in deterministic modeling studies is taken to be the mean value of process area capital cost uncertainty distribution. Table B-27 summarizes both the deterministic and probabilistic representations of uncertainty in process area direct cost for all three IGCC systems assumed in this work.

Process area costs with the largest ranges of uncertainty include the KRW gasifier, process condensate treatment for Case OKC, the Lurgi gasifier for Case ALH, the sulfation unit for Case AKH, the zinc ferrite process for Cases AKH and ALH, and the gas turbine for Cases AKH and ALH. The KRW gasifier has not yet been built on the scale currently envisioned for IGCC plants and, as discussed in Section B.4.3 and elsewhere in Section B.4, there is uncertainty regarding the scale-up of the combustion jet and in other aspects of the process area. In EPRI-sponsored design studies, a process contingency of 20 percent has been used for this process area. Assuming that this process contingency is intended to represent a 50 percent probability of cost overrun, and assuming that the range of uncertainty in process area cost could be from zero increase to 40 percent increase, a triangular distribution was assumed. The use of the triangular distribution here places more likelihood on outcomes near the 20 percent cost increase than at the lower or upper extremes, representing a degree of confidence in the judgment of the cost estimating team that developed the EPRI-sponsored estimate.

In Case OKC, process condensate treatment is required to remove contaminants from liquid discharge streams. In an EPRI-sponsored study of a KRW-based system (Gallaspy et al, 1990), this process area was assigned a contingency factor of 30 percent, representing a relatively high level of uncertainty compared to other process areas. In a previous study (Dawkins et al, 1985), a process contingency of 50 percent was used. As an initial characterization of uncertainty, it is assumed that the cost may increase 50 percent compared to the base estimate for this process area, with a chance that there would be little or no cost increase. This uncertainty is represented as a triangular distribution.

The uncertainty in the capital cost of the Lurgi gasifier is discussed in Section B.3.5. Uncertainty in the capital cost of the gas turbine process area is discussed in Section B.6.7.4.

The sulfation unit for Case AKH is one of the least certain aspects of this particular IGCC flowsheet, perhaps because it appears to be one of the least studied areas. In conceptual design studies prepared for the Gas Research Institute (e.g., Earley and Smelser, 1988), this process area has been assigned process contingencies of 35 to 60

Table B-27. Summary of Assumed Values and Uncertainties for IGCC Process Area Direct Capital Cost.

Description	Units[a]	"Best Guess"	Distribution	Parameters[b]			
Oxygen-Blown KRW-based System with Cold Gas Cleanup							
Coal Handling	% of DC	5	Uniform	0	to	10	
Oxidant Feed	% of DC	5	Uniform	0	to	10	
Gasification	% of DC	20	Triangular	0	to	40	(20)
Selexol	% of DC	10	Triangular	0	to	20	(10)
Low Temperature Gas Cooling	% of DC	0	Triangular	-5	to	5	(0)
Claus Plant	% of DC	5	Triangular	0	to	10	(5)
Beavon-Stretford	% of DC	10	Triangular	0	to	20	(10)
Boiler Feed Water	% of DC	0					
Process Condensate Treatment	% of DC	30	Triangular	0	to	50	(30)
Gas Turbine	% of DC	12.5	Uniform	0	to	25	
HRSG	% of DC	2.5	Uniform	0	to	5	
Steam Turbine	% of DC	2.5	Uniform	0	to	5	
General Facilities	% of DC	5	Uniform	0	to	10	(5)
Air-Blown KRW-Based System with Hot Gas Cleanup							
Coal Handling	% of DC	5	Uniform	0	to	10	
Limestone Handling	% of DC	5	Uniform	0	to	10	
Oxidant Feed	% of DC	10	Uniform	0	to	20	(10)
Gasification	% of DC	20	Triangular	0	to	40	(20)
Sulfation	% of DC	40	Triangular	20	to	60	(40)
Zinc Ferrite	% of DC	40	Uniform	0	to	80	
Sulfuric Acid Plant	% of DC	10	Uniform	0	to	20	
Boiler Feed Water	% of DC	0					
Gas Turbine	% of DC	25	Uniform	0	to	50	
HRSG	% of DC	2.5	Uniform	0	to	5	
Steam Turbine	% of DC	2.5	Uniform	0	to	5	
General Facilities	% of DC	5	Uniform	0	to	10	
Air-Blown Lurgi-Based System with Hot Gas Cleanup							
Coal Handling	% of DC	5	Uniform	0	to	10	
Oxidant Feed	% of DC	10	Uniform	0	to	20	
Gasification	% of DC	20	Uniform	10	to	30	
Cyclones	% of DC	5	Uniform	0	to	10	
Zinc Ferrite	% of DC	40	Uniform	0	to	80	
Sulfuric Acid Plant	% of DC	10	Uniform	0	to	20	
Boiler Feed Water	% of DC	0					
Gas Turbine	% of DC	25	Uniform	0	to	50	
HRSG	% of DC	2.5	Uniform	0	to	5	
Steam Turbine	% of DC	2.5	Uniform	0	to	5	
General Facilities	% of DC	5	Uniform	0	to	10	

[a] The "best guess" values represent deterministic "contingency factors" as defined by EPRI (1986) and others. For probabilistic studies, uncertainty in capital cost is represented by an uncertainty factor, which is described by a probability distribution. DC = process area direct cost.

[b] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

percent. Here, a "nominal" best guess value of 40 percent is assumed, with a chance that the increase in the process capital cost could be as low as 20 percent or as high as 60 percent.

The zinc ferrite process has been evaluated also as part of studies prepared for the Gas Research Institute. The process contingency assumed for this process area has been 40 percent. Here, it is assumed that there is a chance there would be no cost growth and a chance that the cost could increase up to 80 percent. This wide range of uncertainty reflects the early stage of development of this process area, the potential troublesome nature of solids handling, potential difficulties with the complex piping and valving needed, and potential difficulty in developing a control system.

The other process areas shown in Table B-27 represent more conventional technologies which have seen applications hor ave been demonstrated at a full scale in similar process environments. For these process areas, the process contingency factors typically reported in the literature were assumed as the median values of uncertainty factors for capital cost. In most cases, a uniform distribution ranging from zero to twice the reported contingency factor was assumed as an initial estimate of uncertainty for each process area. The cost of the boiler feedwater process area was assumed to be certain; this process area is common to any steam power plant and represents standard, commercial, and proven technology. Low contingencies and uncertainties were assumed for process areas that are well-proven or not substantially affected by an IGCC process environment, such as coal handling, limestone handling, air separation, heat recovery steam generator, and steam turbine.

B.7.3 Maintenance Cost Uncertainties

While considerable attention is often devoted to representing uncertainty in capital cost estimates using "contingency factors," usually no attention is given to uncertainties in predicting maintenance costs. Uncertainty in maintenance cost may be particularly important for new technology involving solids handling and facing potential problems from trace contaminants.

In preliminary cost estimates, the typical approach to estimating maintenance cost is to use "maintenance cost factors," which are a multiplier based on process area costs. Typical values, based on previous experience with a process area, are assumed. For a new process area, a maintenance cost factor may be assumed based on experience with analogous systems or judgment about the cost of maintenance that may be required with the new system. In the probabilistic simulations developed here, rather than use a single value of maintenance cost factors, ranges of possible values are assumed for selected process areas. In all cases, triangular distributions have been assumed. These distributions require judgment about the lower and upper bounds and the "most likely" or modal value. The mode has been assumed to be the same as the maintenance cost factors commonly assumed for each process area in published design studies (see Appendix A.7). The deterministic and probabilistic values for maintenance cost factors used here are summarized in Table B-28.

In cases where there appears to be little uncertainty regarding maintenance cost, such as for process areas with which there is a long history of commercial experience, the maintenance costs are assumed to be known with certainty. An examples of this type of process area is coal handling. In cases where a new technology is used, or where an existing technology is adopted for the first time in a process environment like that of an IGCC system, wider ranges of uncertainty are assumed. Furthermore, for some process areas involving solids handling or that might be seriously affected by trace contaminants, positively skewed distributions for maintenance cost are assumed. These process areas include Selexol, process condensate treatment, sulfation, zinc ferrite desulfurization, gas turbines in IGCC systems with hot gas cleanup, and Lurgi gasification. (Maintenance costs for the Lurgi gasifier and the gas turbine process areas have been discussed in Sections B.3.5 and B.6.7.4, respectively.)

Table B-28. Summary of Assumed Values and Uncertainties for IGCC Process Area Maintenance Cost.

Description	Units[a]	"Best Guess"	Distribution	Parameters[b]			
Oxygen-Blown KRW-based System with Cold Gas Cleanup							
Coal Handling	% of TC	3					
Oxidant Feed	% of TC	2					
Gasification	% of TC	4.5	Triangular	3	to	6	(4.5)
Selexol	% of TC	2	Triangular	1.5	to	4	(2)
Low Temperature Gas Cooling	% of TC	3	Triangular	2	to	4	(3)
Claus Plant	% of TC	2	Triangular	1.5	to	2.5	(2)
Beavon-Stretford	% of TC	2	Triangular	1.5	to	2.5	(2)
Boiler Feed Water	% of TC	1.5					
Process Condensate Treatment	% of TC	2	Triangular	1.5	to	4	(2)
Gas Turbine	% of TC	1.5	Triangular	1.5	to	2.5	(1.5)
HRSG	% of TC	1.5					
Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					
Air-Blown KRW-Based System with Hot Gas Cleanup							
Coal Handling	% of TC	3					
Limestone Handling	% of TC	3					
Oxidant Feed	% of TC	2	Triangular	1	to	3	(2)
Gasification	% of TC	4.5	Triangular	3	to	6	(4.5)
Sulfation	% of TC	4	Triangular	3	to	6	(4)
Zinc Ferrite	% of TC	3	Triangular	3	to	6	(3)
Sulfuric Acid Plant	% of TC	2					
Boiler Feed Water	% of TC	0					
Gas Turbine	% of TC	2	Triangular	1.5	to	6	(2)
HRSG	% of TC	1.5					
Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					
Air-Blown Lurgi-Based System with Hot Gas Cleanup							
Coal Handling	% of TC	3					
Oxidant Feed	% of TC	2	Triangular	1	to	3	(2)
Gasification	% of TC	3	Triangular	2	to	12	(3)
Cyclones	% of TC	3	Triangular	1.5	to	4.5	(3)
Zinc Ferrite	% of TC	3	Triangular	3	to	6	(3)
Sulfuric Acid Plant	% of TC	2					
Boiler Feed Water	% of TC	1.5					
Gas Turbine	% of TC	2	Triangular	1.5	to	6	(2)
HRSG	% of TC	1.5					
Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					

[a] TC = process area total cost, including indirects and contingency

[b] For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

B.7.4 Variable Operating Cost Uncertainties

There are two types of uncertainty in variable operating cost that are characterized in this modeling effort. The first relates to uncertainty in the quantity of consumables that are required to satisfy plant requirements. For example, uncertainty in factors which affect plant efficiency lead to uncertainty in the required flow rate of materials needed to produce a given power output. Uncertainty in the long term chemical and physical properties of zinc ferrite sorbent, used in the two IGCC systems with hot gas cleanup, leads to uncertainty in the annual requirement for makeup sorbent. Uncertainty in the coal feed rate required to produce a given amount of power also leads to uncertainty in the amount of sulfur that needs to be recovered from the coal gas and converted to a byproduct. These types of uncertainties are performance related and are explicitly estimated as part of probabilistic modeling of process flowsheets.

A second uncertainty relates to the unit cost of consumables or the unit price of byproducts. These costs are likely to be affected by site-specific market conditions. In generic evaluations of process technologies, which are not intended to be site-specific, it is appropriate to represent the variability of these costs from one location to another using probability distributions. To the extent that variation in, say, the unit price of a byproduct might lead to significant variation in the variable cost of the plant, then the modeling can provide insight into market niches in which the technology is likely to have competitive costs compared to other alternatives.

Judgments regarding uncertainties in variable operating cost parameters are summarized in Table B-29. The judgment about uncertainty regarding the zinc ferrite unit cost was obtained from Expert ZF-3, as discussed in Section B.5.3.3. The cost of limestone, which is required for Case AKH, is assumed to have a nominal value of \$18/ton, which is a typical value used in published studies. However, depending on the availability of limestone, the cost could be higher, as represented by the positively skewed triangular distribution. The cost of ash disposal is often assumed to be \$10/ton, assuming readily a available landfill. However, this cost is also likely to increase. Increased costs of land, permitting, and potential concerns about trace species in the ash could increase the costs of disposal. Again, a positively skewed triangular distribution is used.

The price that may be obtained for either byproduct sulfur (Case OKC) or sulfuric acid (Case ALH) depends on plant location and proximity to customers. Many design

Table B-29. Summary of Assumed Values and Uncertainties for IGCC Variable Operating Cost Parameters.

Description	Units[a]	"Best Guess"	Distribution	Parameters[b]
Zinc Ferrite Sorbent	\$/lb	3.00	Triangular	0.75 to 5.00 (3.00)
Limestone	\$/ton	18	Triangular	18 to 25 (18)
Ash Disposal	\$/ton	10	Triangular	10 to 25 (10)
Sulfuric Acid Byproduct Price	\$/ton	40	Triangular	0 to 60 (40)
Sulfur Byproduct	\$/ton	125	Triangular	60 to 125 (125)
Byproduct Marketing	fraction	0.10	Triangular	0.05 to 0.15 (0.10)

[a] Costs are in 1990 dollars

[b]For Uniform distributions, the lower and upper bounds are given. For the triangular distribution, the mode is given in parentheses. For the fractile distribution, the lower and upper bounds for each range are given, along with the probability of sampling within that range.

studies assume that obtainable prices will be the same as current market prices for these products. However, such assumptions ignore the effect that additional production of these commodities would have on market prices, as well as the effect that large transportation distances would have on price that a particular plant could obtain. In addition, sulfuric acid sale is likely to be disadvantageous compared to sulfur, because currently the sulfur market in the U.S. is "structurally larger" than the sulfuric acid market (Reiber, 1982). Over 80 percent of the sulfur consumed in the U.S. is used to produce sulfuric acid. However, a large portion of the sulfur comes from "discretionary" production at natural mines (Manderson and Cooper, 1982), which can be easily displaced by sulfur byproduct recovery. Moreover, solid sulfur is relatively easy to store and transport compared to sulfuric acid. Sulfur can be sold to sulfuric acid producers as well as to other end-users, and conceivably byproduct sulfur could be exported internationally.

In contrast, the costs associated with shipping sulfuric acid are likely to limit sale to markets relatively near the power plant. There are relatively few concentrations of industry were sulfuric acid is required in large quantities. This is particularly a concern in regions where high sulfur coal is likely to be consumed, where markets for sulfuric acid are weak (Burns and Roe, 1987). Furthermore, byproduct sulfuric acid production from smelters in Canada is likely to depress market prices for sulfuric acid (Burns and Roe, 1987, Manderson and Cooper, 1982).

Because sulfur is believed to be a less risky alternative economically, the uncertainty in byproduct price for sulfur is not as pessimistic as for sulfuric acid. While some studies (e.g., Corman, 1986) assume relatively high sulfuric acid byproduct prices of around \$60/ton, a more reasonable assumption would be \$40/ton, and it is possible that, at a particular site, a plant might obtained almost nil for the byproduct (Burns and Roe, 1987). In contrast, sulfur prices are expected to be relatively stable. A typically assumed value for sulfur price is about $125/ton. However, at any given site, there may be some change that price could be lower, due either to expenses of transport or due to fluctuations in the market price. For both sulfur and sulfuric acid, it is assumed that a portion of the byproduct proceeds are required for activities associated with byproduct marketing, as opposed to power plant operation. Therefore, a byproduct marketing factor is assumed.

B.7.5 Regression Model Error Terms

Regression analysis was used to develop models for performance and cost of a number of process areas in the IGCC systems. See Chapter 2 for a more detailed discussion of the use of regression analysis in model development. For all regression models, the standard error of the estimate can be used to measure the variance in the dependent variable that is not captured by the functional relationship to the selected independent variables. Thus, the standard error is a measure of how well the models predict the dependent variable. In many cases, the standard error from the regression models developing in this work is very small and negligible. The more significant standard errors are summarized in Table B-30. These standard errors may be included as uncertainties in the probabilistic simulation for specific process flowsheets. In Table B-30, the standard errors are grouped depending on whether they are associated with direct capital cost models or with auxiliary power load models. In two cases, non-linear regressions were used, resulting in lognormally distributed error terms.

For probabilistic analysis, several of the standard errors reported in Table B-30 were judged to be negligible and so were excluded. These are: Claus plant, Beavon-Stretford, boiler feedwater, and boost air compressor direct capital cost standard errors, and the Lurgi coal handling and KRW low temperature gas cooling auxiliary power standard errors. The standard errors for models of consumable requirements affecting variable operating cost were generally very small and are not reported in Table B-30 nor included in the probabilistic simulations.

Table B-30. Summary of Regression Model Standard Errors for IGCC Cost and Auxiliary Power Equations.

Description	Units	Mean Error	Distribution[a]	Parameters[b]		
Direct Capital Cost Models						
KRW Coal Handling	$ Million	0	Normal	- 10	to	10
Oxygen Plant	multiplier	1.012	Lognormal	0.78	to	1.29
KRW Gasification (Cold Gas System)	$ Million	0	Normal	- 20.5	to	20.5
Low Temperature Gas Cooling	$ Million	0	Normal	- 1.5	to	1.5
Selexol	$ Million	0	Normal	- 5.1	to	5.1
Claus	$ Million	0	Normal	- 0.25	to	0.25
Beavon-Stretford	$ Million	0	Normal	- 0.26	to	0.26
Boiler Feedwater	multiplier	1.002	Lognormal	0.90	to	1.10
Process Condensate	$ Million	0	Normal	- 0.01	to	0.01
HRSG	$ Million	0	Normal	- 17.3	to	17.3
Steam Turbine	$ Million	0	Normal	- 15.8	to	15.8
Boost Air Compressor	$ Million	0	Normal	- 0.66	to	0.66
Lurgi Coal Handling	$ Million	0	Normal	- 14.4	to	14.4
Sulfuric Acid Plant	$ Million	0	Normal	- 4.0	to	4.0
Auxiliary Power Load Models						
KRW Coal Handling	MW	0	Normal	- 1.6	to	1.6
Oxygen Plant	MW	0	Normal	- 6.6	to	6.6
KRW Gasification (cold gas system)	MW	0	Normal	- 0.52	to	0.52
Low Temperature Gas Cooling	MW	0	Normal	- 0.24	to	0.24
Selexol	MW	0	Normal	- 0.55	to	0.55
Claus Plant	MW	0	N/A			
Beavon-Stretford	MW	0	N/A			
Boiler Feedwater	MW	0	N/A			
Process Condensate	MW	0	N/A			
General Facilities	MW	0	N/A			
Limestone Handling	MW	0	N/A			
KRW Gasification (hot gas system)	MW	0	N/A			
Sulfation	MW	0	N/A			
Lurgi Coal Handling	MW	0	Normal	- 0.35	to	0.35

[a]N/A = not applicable. For these cases the standard error was sufficiently small to be judged negligible.
[b]For Normal and Lognormal distributions, the upper and lower limits of the 99.8 percent probability range are given.

B.8 Questionnaires Used in Elicitation Briefing Packets

For a few of the process areas for which judgments about uncertainties were required, a formal set of briefing materials was prepared and given to selected experts. These materials included a 9-page introduction to uncertainty analysis, a technical background paper (ranging from 9 to 23 pages, depending on the process area), and a questionnaire. The material in the introduction to uncertainty analysis is covered in Chapter 2. The technical background is presented in previous sections of Appendix B. The questionnaires are reproduced here. These include questionnaires for the fixed-bed gasifier, dual vessel fixed bed zinc ferrite process, and the gas turbine process areas. A questionnaire was developed for the fluidized bed gasifier, but was not distributed as discussed in Section B.4.

B.8.1 Questionnaire for the Fixed Bed Gasifier Process Area

Here, you are asked to provide technically-informed judgments about probability distributions for parameters of a fixed-bed dry-ash Lurgi gasifier performance and cost model. You are asked to consider the possibilities of potentially poor performance as well as the probability of obtaining favorable performance, based on current information about the system. The preceding sections provide an overview of uncertainty analysis and some of the technical considerations which might be used as the starting point for your own thinking about technical uncertainties. We are interested in the use of the Lurgi gasifier as part of a "simplified" air-blown integrated gasification combined cycle (IGCC) system with hot gas cleanup, as explained in Part 2. In addition, some of the typical modeling assumptions for the Lurgi-based IGCC system are given in Table 1 of Part 2. Some of the key assumptions are repeated here for convenience:

- The gasifier operating pressure depends on system pressure losses and the gas turbine combustor pressure. The gasifier operating pressure is expected to be between about 240 and 330 psia.
- The raw coal gas exiting the gasifier is at 1,100 ^{o}F.
- Minus 2 inch Illinois No. 6 coal, up to 30 weight percent as minus 1/4 inch fines.
- Fines carryover is about 4 percent of feed coal flow rate
- About 98 percent (or less) of the fines are captured, agglomerated, and recycled
- The fines contain up to 90 percent carbon
- About 0.5 percent of the carbon in the coal is retained in the bottom ash
- About 3 percent of the sulfur in the coal is retained in the bottom ash
- The gasifier throughput varies with pressure. A simple assumption is to assume it varies linearly between two points given by Corman (1984).
- The ammonia concentration of the coal gas is around 2,000 ppm, which represents approximately 50 percent conversion of coal-bound nitrogen to ammonia
- The gasifier steam requirement is about 0.6 lb steam/lb air.
- The gasifier oxidant requirement is about 3.1 lb air/lb MAF coal, or about 0.93 lb oxygen/lb carbon.
- The gasifier steam inlet temperature is 618 ^{o}F
- The gasifier air inlet temperature is 800 ^{o}F

- The gasifier is a modification of the standard Mark IV model, with 12.7 feet working diameter.
- The gasifier has a deep bed stirrer to prevent agglomeration and plugging of the bed
- About half of the sulfur in the coal is organic, and the other half is pyritic.

We are interested in your technically-based judgments about uncertainties in key performance and cost parameters of the gasifier process area, such as the ones given above. We intend to model the uncertainty in performance and cost associated with a fifth-of-a-kind, or mature, system. Thus, we are asking you to make predictions about systems that have not yet been built or operated. We are asking you to express the range of possible outcomes for these systems using probability distributions, as discussed in Part 1.

Several questions follow. You may respond to the questions on these pages, or use additional paper as needed. See the Introduction (Part 1) for examples of how you might estimate uncertainty in each parameter.

Question #1. Comments on Default Assumptions

Do the default assumptions seem reasonable? If not, adjust accordingly and explain the basis for the changes. Are there additional assumptions that should be specified for these systems? If so, please add these assumptions and explain why they are needed. Use your updated set of assumptions as the basis for answering the following questions.

Question #2. Uncertain Parameter Identification

The following is a list of the specific parameters for which uncertainty distributions are desired.

- Fines Carryover
- Fines Capture
- Fines Carbon Content
- Carbon Retention in Bottom Ash
- Sulfur Retention in Bottom Ash
- Gasifier Throughput
- Ammonia Yield
- Gasifier Steam Requirement
- Gasifier Oxidant Requirement
- Gasifier Direct Capital Cost
- Gasifier Maintenance Cost

Are you comfortable making estimates of uncertainty for these parameters? If you are not, who do you think should be approached (preferably within METC) to obtain these estimates?

Are there other parameters which you believe also should be treated probabilistically (whether or not you feel comfortable making the judgment yourself) that are not included in the above list? If so, please specify what these parameters are and supply your judgments about them if you are comfortable doing so (see the following questions for examples of the types of judgments we are looking for). If not, who can we ask to estimate uncertainties for these additional parameters?

Question #3. Fines Carryover

What is the uncertainty in the long term typical fines carryover over the life cycle of a Lurgi gasifier fed with Illinois No. 6 coal containing up to 30 weight percent minus 1/4 inch fines? Consider the possibility of high carryover due to, say, entrainment of the incoming coal fines in the outgoing raw gas. Also consider the possibility of low fines carryover due to, say, the caking characteristics of the Illinois No. 6 coal. Express the fines carryover as a weight percentage of the coal feed rate to the gasifier.

- Consider the best possible (lowest) fines carryover that could occur. Explain how such a result could be achieved (e.g, agglomeration of fines in the gasifier reduces entrainment of small fines). How likely is it that the fines carryover could be less than this amount?

- Consider the worst possible (highest) fines carryover that could occur. Explain, as above. How likely is it that the fines carryover could be equal to or more than this value?

- What do you think is the median (or if you prefer, mean) value of the fines carryover (recall that median implies a 50-50 percent chance that the fines carryover could be higher or lower than this value, while mean implies a probability-weighted average of possible outcomes)? Note that the median value does not have to equally divide the best and worst possible values, nor does it have to be the same as the average (mean) rate that you expect. Alternatively, if you want to express your judgment as a triangle distribution, what is the most likely value (mode) that you expect?

- Can you draw a probability distribution to represent your judgment? You may draw the distribution as either a pdf or a cdf. Can the distribution be represented by one of the functions shown in Figure 1? Please be sure to completely specify the range of possible outcomes in your distribution function (i.e. the distribution must consider the 100 percent range of possible outcomes). If your "worst" and "best" cases above bound only 80 or 90 percent of all outcomes, please consider how the range of outcomes is widened when considering 100 percent percent of all possible outcomes. What are the absolute best and worst possible outcomes (0th and 100th percentiles)?

<u>Question #4. Fines Capture</u>

What is the uncertainty, if any, in the long term typical life cycle cyclone collection efficiency for fines which carryover from a Lurgi gasifier using Illinois No. 6 coal with up to 30 weight percent as minus 1/4 fines? If your judgment depends on the fines loading entering the cyclone, please pick three representative values of fines carryover (e.g., worst, median or mean, and best) based on your answer to Question 3 and use these values as a basis for answering this question (in such a case, your answer would include three, rather than one, probability distributions).

- Consider the highest possible fines capture efficiency that might occur? What is this rate and how might it be achieved (e.g., it may depend on a certain particle size distribution, or on agglomeration of fines leaving the gasifier, or design features of the cyclone)? How likely is the capture efficiency to be better than the number you have just estimated?

- Consider the lowest possible fines capture efficiency that might occur? What is this rate and how might it be achieved (e.g., very small fines, very high fines loading). How likely would it be to obtain an actual capture efficiency less than this rate?

- What is the median (50-50 percent of getting higher or lower) capture efficiency?

- Can you estimate the 25th fractile (i.e. there is a one in four chance that the emission rate is less than this number)? What about the 75th fractile (i.e. a one in four chance that the value is higher than this number)?

- Can you draw a probability distribution to represent your judgement? You may draw the distribution as either a pdf or a cdf. Can it be represented by one of the distributions in Figure 1 of Part 1?

Question #5. Fines Carbon Content

We are interested in your judgment about uncertainty in the composition of the fines carryover from the Lurgi gasifier. What is the uncertainty, if any, in the carbon composition (weight percent) of the fines leaving the gasifier in the raw coal gas?

- Consider the highest possible carbon composition that might occur. What is this composition and how might it occur (e.g., perhaps coal char is entrained into the raw gas from the devolatilization or gasification zones of the gasifier). How likely is the carbon composition to be higher than the number you have just estimated?

- Consider the lowest possible fines carbon content that might occur. What is this rate and how might it be achieved (e.g, incoming coal fines are entrained prior to drying, or ash is entrained from the bottom of the gasifier due to, say, channeling). How likely would it be to obtain actual carbon content below this value?

- What is the median (50-50 percent of getting higher or lower) carbon content? If you prefer to specify a particular type of probability distribution model, such as a triangle distribution, then provide appropriate judgments in lieu of the median (e.g., mode or "most likely" value for a triangle distribution).

- If you haven't already specified and defined the parameters of a probability distribution model (e.g., triangle, normal, lognormal), can you estimate the 25th fractile (i.e. there is a one in four chance that the carbon content is less than this number)? What about the 75th fractile (i.e. a one in four chance that the value is higher than this number)?

- Can you draw a probability distribution to represent your judgement? You may draw the distribution as either a pdf or a cdf. Can it be represented by one of the distributions in Figure 1 of Part 1?

Question #6. Carbon Retention in Bottom Ash

Please provide your judgment about the fraction of carbon in the feed coal that will be retained in the gasifier bottom ash. If there are any key (significant) functional dependencies that you expect, such as with respect to the air/coal ratio, feed coal fines loading, coal throughput, or other factors which you may believe to be uncertain or variable over the life of the plant, please appropriately caveat your judgments or, if you wish, provide a set of judgments for each combination of independent variable values that you select (e.g., one uncertainty distribution for each value of air/coal ratio).

Because the set of judgments may involve several probability distributions, we suggest you might want to use a triangular distribution for simplicity, which requires estimates only for the lowest, highest, and most likely conversion rates. However, feel free to use any type of distribution which best represents your judgment.

- What is the worst (highest) bottom ash carbon retention fraction that you would expect for a Lurgi gasifier operating on Illinois No. 6 coal? How might this high bottom ash carbon retention be realized (e.g., relatively low gasification reactivity of high rank coal, inefficient gasification or char combustion due to agglomeration of caking coal)? What do you think is the probability of obtaining a bottom ash carbon retention higher than the value you have specified?
- What is the best (lowest) bottom ash carbon retention fraction that you would expect? How might this best or optimistic value be explained? How likely would it be to measure a carbon retention rate lower than this value?
- What is the most likely bottom ash carbon retention (as a fraction of the feed coal carbon) that you expect. The most likely value is the mode, or "peak", of the probability distribution. Alternatively, if you prefer to express your judgment as a median, please indicate.

Question #7. Sulfur Retention in the Bottom Ash

Please provide your judgment about the fraction of sulfur in the feed coal that will be retained in the gasifier bottom ash. If your answer depends on the type of sulfur in the coal (e.g., organic or pyritic), please explain the dependency (see page 1 for assumptions). As a default, we would like you to express your answer as the fraction of total sulfur in the coal that is retained in the bottom ash. However, if you feel it is more appropriate to use another approach (e.g., fraction of pyritic sulfur retained in the bottom ash), please explain.

- What is the worst (highest) bottom ash sulfur retention fraction that you would expect for a Lurgi gasifier operating on Illinois No. 6 coal? How might this high bottom ash sulfur retention be realized? What do you think is the probability of obtaining a bottom ash sulfur retention higher than the value you have specified?

  What is the best (lowest) bottom ash sulfur retention fraction that you would expect? How might this best or optimistic value be explained? How likely would it be to measure a sulfur retention rate lower than this value?
- What is the most likely bottom ash sulfur retention that you expect. The most likely value is the mode, or "peak", of the probability distribution. Alternatively, if you prefer to express your judgment as a median, please indicate.

Question #8. Gasifier Coal Throughput

Some studies reviewed in Part 2 suggest that the coal throughput for the Lurgi Mark IV gasifier depends on the oxidant (air or oxygen) and the operating pressure. For example, it is suggested that in air-blown mode, the coal throughput at 300 psia is about 456 tons/day on a dry, ash-free basis (Corman, 1986--see Part 2 for more discussion). Is this value optimistic for air-blown operation? How would the coal throughput differ at 250 psia? 350 psia? Can the throughput at intermediate pressures be interpolated? Could you provide a judgment about the dry, ash-free Illinois No. 6 coal throughput in a Mark IV gasifier (12.7 foot working diameter) for each of the following pressures: 250, 300, and 350 psia? For simplicity, you may want to use a triangular distribution to represent your judgments for each case; however, feel free to use whatever distribution best reflects your judgment. Please note the units you are assuming for coal throughput (e.g., tons DAF coal/day, lb DAF coal/ft^2 of grate area)

- For each gasifier operating pressure, consider the best (highest) possible coal throughput rate? Can you explain how such values might be obtained?
- What is the worst (lowest) possible coal throughput that you expect for each operating pressure? Why might these outcomes occur?

- What is the most likely coal throughput that you expect for each operating pressure?

Question #9. Ammonia Yield

Consider the formation of ammonia in the fixed-bed dry-ash Lurgi gasifier. We are interested in estimating how much ammonia is contained in the coal gas for the purpose of estimating NO_x emissions from the gas turbine combustor. Please provide your judgment about how much ammonia is generated in the Lurgi gasifier in an air-blown system operating on Illinois No. 6 coal. Please explain the units you are using. For example, you may wish to express ammonia production in terms of a fractional conversion of coal-bound nitrogen to ammonia. Or, alternatively, you may wish to express your judgment in terms of the ammonia volume concentration in the raw coal gas exiting the gasifier.

- What is the highest amount of ammonia you would expect to be produced in the Lurgi gasifier when operating on Illinois No. 6 coal? How likely would it be to obtained a measurement of a production rate higher than what you have estimated as the "highest amount"?
- What is the least amount of ammonia you would expect to be produced? How likely would it be to measure an ammonia production rate less than what you just estimated?
- What is the most likely amount of ammonia that you expect to be produced?

Question #10. Steam Requirement

In Part 2, one of the key performance factors that was discussed was the steam requirement for the air-blown dry-ash Lurgi gasifier. The key determinant of the steam requirement is reported to be the ash characteristics of the coal. Over the lifetime of a fifth-of-a-kind plant using Illinois No. 6 coal, what do you think is the uncertainty or variability in the steam requirement? Please explain the units you are using for the steam requirement (e.g., lb steam/lb DAF coal). If your judgment depends on assumptions about the variability in the ash characteristics of the coal over the life of the plant, could you please share your assumptions? Also, if your assumption is tied to a specific air/coal ratio, please indicate. If so, could you provide a set of three judgments based on a low, middle, and high air/coal ratio. (In the next question, you are asked for your judgment about uncertainty or variability in the air/coal ratio as well).

- Over the life of the plant, what is the highest steam requirement that you would expect (i.e. there is a negligible probability of a higher steam requirement)?
- What is the lowest (i.e. little probability that it would be lower) steam requirement that you would expect?
- What is the median (50th percentile) steam requirement that you expect? Alternatively, if you which to express your judgment as a triangular distribution, what is the most likely steam requirement?
- If you are not using a triangular distribution, please indicate the 25th and 75th percentiles of the uncertainty or variability in the steam requirement.

Question #11. Oxidant Requirement

We are interested in your judgments about uncertainty or variability in the oxidant requirement over the lifetime of a fifth-of-a-kind air-blown Lurgi-based IGCC system operating on Illinois No. 6 coal. Please indicate what units you are using for your judgment (e.g., lb air/lb DAF coal, lb oxygen in the air/lb carbon in the coal). Also, please indicate if you are including coal-bound carbon in your estimate of the gasifier oxygen requirement. Recall that we are modeling an air-blown system.

- Over the life of the plant, what is the highest air requirement that you would expect (i.e. there is a negligible probability of a higher air requirement)?
- What is the lowest (i.e. little probability that it would be lower) air requirement that you would expect?
- What is the median (50th percentile) air requirement that you expect? Alternatively, if you which to express your judgment as a triangular distribution, what is the most likely air requirement?
- If you are not using a triangular distribution, please indicate the 25th and 75th percentiles of the uncertainty or variability in the air requirement.

Question #12. Direct Capital and Maintenance Costs

The proposed design for the Lurgi gasifier operating on Illinois No. 6 coal includes a deep bed stirrer to avoid plugging of the gasifier bed. This adds to the gasifier direct capital cost. In addition, the potential of high fines carryover may have implications for potentially increased maintenance costs due to, for example, erosion or deposition in the exiting nozzle. The gasifier maintenance cost is typically estimated as a percentage of the gasifier direct capital cost.

- Can you provide a judgment about the percentage increase in gasifier direct capital cost associated with modifications needed to gasifier caking Illinois No. 6 coal with high (up to 30 weight percent) minus 1/4 inch fines loading? Please explain the basis for your estimate (e.g., additional cost of deep bed stirrers, bearings, motor). What is the range of the cost increase (e.g., highest possible, lowest possible, most likely)? How might these values be obtained?
- Can you provide a judgment about the maintenance cost, as a percentage of the capital investment, for the Lurgi gasifier. For example, typical maintenance cost factors might be 4.5 to 6 percent of the plant facilities investment for the gasifiers. What are the highest, lowest, and most likely maintenance cost factors? (Please give units you are assuming). How might this different outcomes be obtained?

Question #13. Other Experts

Please suggest other experts whom we should contact for judgments about uncertainties in this system. Please supply their names, titles, area of expertise, phone numbers, and addresses.

Question #14. Feedback

We would like your comments on how easy/difficult it was to develop judgments about uncertainties and on these briefing materials. Is there any other information about uncertainty analysis you would like to see in Part 1? Was the summary of technical information in Part 2 a useful starting point for your thinking about uncertainties for this process? Was it difficult for you to develop estimates of the range or likelihood of various values for variables which you believe to be uncertain? Please discuss these or any other comments you may have.

Thank you for your contribution to this project.

B.8.2 Questionnaire for Fixed Bed Zinc Ferrite Desulfurization

Here, you are asked to provide technically-informed judgments about probability distributions for parameters of the zinc ferrite performance and cost model. You are asked to consider the possibilities of potentially poor performance as well as the probability of obtaining favorable performance, based on current information about the system. The preceding sections provide an overview of uncertainty analysis and some of the technical considerations which might be used as the starting point. The default assumptions for the case studies are:

Space Velocity:	2,000/hour
Superficial Gas Velocity:	2 ft/sec
Inlet Gas Temperature:	1,100 °F
Inlet Gas Pressure:	300 psia
Inlet Sulfur Concentration:	~5,000 ppmv H_2S
Maximum Vessel Diameter (D):	12.5 ft.
Maximum Vessel Length:	4 D
Absorption Cycle Time:	168 hours

Fifth-of-a-kind plant
No additional research programs
Sorbent pellets are 3/16 inch diameter extrudates of 1/2 inch length containing equal molar amounts of zinc and iron.

Several questions follow. You may respond to the questions on these pages, or use additional paper as needed. See the Introduction for an example of how you might estimate uncertainty in each parameter.

Question #1.

Do the default assumptions seem reasonable? For example, is a superficial velocity of 2 ft/sec a reasonable target for commercial operation? Is an absorption cycle time of one week a reasonable target? If not, adjust accordingly and explain the basis for the changes. Are there additional assumptions that you should be specified for this system? If so, please add these assumptions and explain why they are needed. Use your updated set of assumptions as the basis for answering the following questions.

Question #2.

The following is a list of the specific parameters for which uncertainty distributions are desired.

- Sorbent sulfur loading
- Long term sorbent attrition rate
- Absorber pressure drop
- Sorbent unit cost

Are you comfortable making estimates of uncertainty for these parameters? If you are not, who do you think should be approached (preferably within METC) to obtain these estimates?

Are there other parameters which you believe also should be treated probabilistically (whether or not you feel comfortable making the judgment yourself) hat are not included in this list? If so, please specify. Who can we ask to estimate uncertainties for these additional parameters?

Question #3.

(a) What is the uncertainty in the long term sorbent sulfur loading (averaged over the entire sorbent bed) at breakthrough in a commercial-scale absorber over a large number of absorption and regeneration cycles (e.g., over 100) for bulk desulfurization at 1,100 ^oF and 300 psia? Consider that a typical commercial scale bed may be 12.5 feet in diameter and perhaps 25 to 50 feet deep. Consider also that the superficial gas velocity of 2 ft/sec assumed here and in other design studies is higher than any of the tests conducted to date. Be sure to explain the basis for your assumptions.

- Consider the best possible (highest) sorbent loading that could occur. Explain how such a result could be achieved (e.g., no channeling in the reactor or problems with flow distribution at the gas inlet, no deposition of particulates or other contaminants over the sorbent, etc.). How likely is this outcome?
- Consider the worst possible (lowest) sorbent loading that could occur. Explain, as above. How likely is it that the sulfur loading could be equal to or less than this value?
- What do you think is the median value of the sorbent sulfur loading (i.e. there is a 50-50 percent chance that the loading could be higher or lower than this value)? Note that the median value does not have to equally divide the best and worst possible values, nor does it have to be the same as the average (mean) rate that you expect. Alternatively, if you want to express your judgment as a triangle distribution, what is the most likely value that you expect?
- Can you draw a probability distribution to represent your judgment? You may draw the distribution as either a pdf or a cdf. Can the distribution be represented by one of the functions shown in Figure 1?

(b) Now consider a fuel gas with only 1,000 ppmv of H_2S. What is your judgment about the uncertainty in sorbent loading (if different from above)? (Use the same approach to estimate the best, worst, and median values, and to draw the probability distribution).

Question #4.

What is the uncertainty in the long term sorbent attrition rate for this system (bulk desulfurization mode)?

- Consider the worst possible (highest) attrition rate that might occur. What is this rate and how could it happen? (e.g., thermal cycling of the sorbent, changes in chemical composition, effects of contaminants, etc.) Is attrition constant over the life of the sorbent? Is attrition uncertain, variable, or both (i.e. with ideal instrumentation would we always obtain the same rate, or is there variability in attrition related to, say, variability in coal properties?). How likely is the attrition to be worse than the number you have just estimated?
- Consider the best possible (lowest) attrition rate. What is this rate and how might it occur? What is the likelihood of obtain a lower rate than this?

- What is the median attrition rate?
- Can you estimate the 25th fractile (i.e. there is a one in four chance that the sorbent attrition is less than this number) What about the 75th fractile (i.e. a one in four chance that the value is higher than this number)?
- Can you draw a probability distribution to represent your judgment? You may draw the distribution as either a pdf or a cdf.

Question #5.

What is the uncertainty in absorber pressure drop prior to sorbent replacement after a long period of absorption and regeneration cycles? Consider various factors contributing to pressure drop buildup, if any. Also consider the size of the commercial-scale bed. Thus, you must estimate the pressure drop for a commercial scale bed. Along with this, estimate the uncertainty associated with deposition of contaminants and the possible breakup of sorbent pellets into smaller pieces, as well as any other mechanisms you suggest.

- Consider the best possible (lowest) pressure drop that might occur. What factors are important to your judgment?
- Consider the worst possible (highest) pressure drop.
- What is the most likely (mode of the distribution) pressure drop?

Question #6.

What is the uncertainty in the unit cost of the sorbent in $/lb. Assume that the cost remains the same in constant dollars over the life of the power plant. For commercial use, the sorbent will have to be mass produced. A single 500 MW plant may require several million pounds of sorbent initially and several hundred thousand pounds annually for replacement.

- What is the lowest possible cost. Explain?
- What is the highest possible cost?
- What is the median or most likely (state which) cost?

Question #7.

Please suggest other experts whom we should contact for judgments about uncertainties in this system. Please supply their names, titles, area of expertise, phone numbers, and addresses.

Question #8

We would like your comments on how easy/difficult it was to develop judgments about uncertainties and on these briefing materials. Is there any other information about uncertainty analysis you would like to see in Part 1? Was the summary of technical information in Part 2 a useful starting point for your thinking about uncertainties for this process? Was it difficult for you to develop estimates of the range or likelihood of various values for variables which you believe to be uncertain? Please discuss these or any other comments you may have.

B.8.3 Questionnaire for Gas Turbine

Here, you are asked to provide technically-informed judgments about probability distributions for parameters of a gas turbine performance and cost model. You are asked to consider the possibilities of potentially poor performance as well as the probability of obtaining favorable performance, based on current information about the system. The preceding sections provide an overview of uncertainty analysis and some of the technical considerations which might be used as the starting point. We are interest in three IGCC cases, as defined in Part 2:

- Case ALH: Air-blown dry-ash fixed bed Lurgi gasification with hot gas cleanup and gas turbine air extraction.
- Case AKH: Air-blown fluidized bed KRW gasification with hot gas cleanup and gas turbine air extraction.
- Case OKC: Oxygen-blown fluidized bed KRW gasification with cold gas cleanup.

The deterministic default assumptions for the three IGCC case studies are summarized in Table 1 of Part 2. The performance and cost modeling of the gas turbine process area is intended to be representative of current or near-term commercial offerings for high firing temperature (2,300 ^{o}F turbine inlet temperature gas turbines) as discussed in Part 2. Some performance information about this class of gas turbines is given in Table 2 of Part 2. The model considers mass and energy balances, but does not include gas turbine aerodynamics.

We are interested in your technically-based judgments about uncertainties in key performance and cost parameters of the gas turbine process area. We intend to model the uncertainty in performance and cost associated with a fifth-of-a-kind, or mature, system. Thus, we are asking you to make predictions about systems that have not yet been built or operated. We are asking you to express the range of possible outcomes for these systems using probability distributions, as discussed in Part 1.

Several questions follow. You may respond to the questions on these pages, or use additional paper as needed. See the Introduction for examples of how you might estimate uncertainty in each parameter.

Question #1. Comments on Default Assumptions

Do the default assumptions seem reasonable? For example:

- Is an ammonia concentration of 2,000 ppm for the air-blown dry-ash Lurgi system reasonable?
- Are the fuel gas inlet temperatures reasonable for a fifth-of-a-kind plant?
- Are the pressure and heat losses in the gas turbine reasonable?
- Are "high efficiency" cyclones sufficient for particle control in the gas turbine hot gas path?

If not, adjust accordingly and explain the basis for the changes. Are there additional assumptions that should be specified for these systems? If so, please add these assumptions and explain why they are needed. Use your updated set of assumptions as the basis for answering the following questions.

Question #2. Uncertain Parameter Identification

The following is a list of the specific parameters for which uncertainty distributions are desired.

- Fuel valve and nozzle pressure drop
- Thermal NO_x emissions from MBG
- Thermal NO_x emissions from LBG
- Fuel NO_x emissions from LBG
- CO emissions
- Combustor pressure drop
- Turbine inlet temperature
- Maintenance costs

In the questions that follow, you are asked to provide estimates of uncertainty for each of these parameters? If you are not comfortable making a particular estimate, who do you think should be approached (preferably within METC) to obtain such estimates?

Are there other parameters which you believe also should be treated probabilistically (whether or not you feel comfortable making the judgment yourself) that are not included in the above list? If so, please specify what these parameters are and supply your judgments about them if you are comfortable doing so (see the following questions for examples of the types of judgments we are looking for). If not, who can we ask to estimate uncertainties for these additional parameters?

In addition, in Question 10 we ask you to comment on other aspects of gas turbine application with coal gas firing that may pose problems for commercial applications:

- Compressor surge margin and IGV closure
- Controllability of the fuel flow, extraction air, and compressor inlet air
- Fuel valve material requirements
- Particulate control system requirements

If you are not comfortable discussing any of these issues, could you suggest someone else who might be approached?

Question #3. Fuel Valve and Nozzle Pressure Drop

(a) What is the uncertainty in the long term fuel valve and nozzle pressure drop over the life cycle of a gas turbine firing LBG in a Lurgi-based system (Case ALH). Consider the possibility of deposition in the valves or nozzle. Also consider the possibility of improved fuel valve and nozzle designs or control strategies that reduce the pressure drop (e.g, down to, say, 10 psi instead of the 70 psi or so typical of current designs).

• Consider the best possible (lowest) pressure drop that could occur. Explain how such a result could be achieved (e.g., improved designs, high efficiency particulate control upstream, no deposition to block nozzle area). How likely is this outcome?

• Consider the worst possible (highest) pressure drop that could occur. Explain, as above. How likely is it that the pressure drop could be equal to or less than this value?

- What do you think is the median (or if you prefer, mean) value of the fuel valve pressure drop (recall that median implies a 50-50 percent chance that the pressure drop could be higher or lower than this value, while mean implies a probability-weighted average of possible outcomes)? Note that the median value does not have to equally divide the best and worst possible values, nor does it have to be the same as the average (mean) rate that you expect. Alternatively, if you want to express your judgment as a triangular distribution, what is the most likely value (mode) that you expect?

- Can you draw a probability distribution to represent your judgment? You may draw the distribution as either a pdf or a cdf. Can the distribution be represented by one of the functions shown in Figure 1 of Part 1? Please be sure to completely specify the range of possible outcomes in your distribution function (i.e. the distribution must consider the 100 percent range of possible outcomes). If your "worst" and "best" cases above bound only 80 or 90 percent of all outcomes, please consider how the range of outcomes is widened when considering 100 percent, or 99 percent of all outcomes.

(b) Does your judgment change if we are considering a LBG gas from the KRW system in Case AKH? Or if we are considering a MBG gas from a KRW system as in Case OKC? If so, could you provide your judgments for these cases in a similar manner to that for Case ALH?

Question #4. Thermal NO_x Emissions from MBG

What is the uncertainty, if any, in the thermal NO_x emission rate (please define the units that you are using, if not lb NO_x as NO_2 per million BTU of coal feed) that can be achieved in a fifth-of-a-kind system firing MBG (as in Case OKC)? You may consider possible improvements in the next few years in either dry or wet NO_x control, such as lean-lean combustors or high levels of fuel gas moisturization or steam injection, that would be expected to be employed in a fifth-of-a-kind plant.

- Consider the highest possible NO_x emission rate that might occur. What is this rate and how might it be achieved (e.g., perhaps current combustor technology, slightly modified for MBG, and steam injection). How likely is the emission rate to be worse than the number you have just estimated?

- Consider the lowest possible NO_x emission rate that might occur. What is this rate and how might it be achieved (e.g., development of a lean-lean combustor for MBG applications, or use of existing combustor technology with modifications to allow for increased steam injection). How likely would it be to obtain actual emissions below this rate?

- What is the median (50-50 percent of getting higher or lower) emission rate?

- Can you estimate the 25th fractile (i.e. there is a one in four chance that the emission rate is less than this number)? What about the 75th fractile (i.e. a one in four chance that the value is higher than this number)?

- Can you draw a probability distribution to represent your juugment? You may draw the distribution as either a pdf or a cdf. Can it be represented by one of the distributions in Figure 1 of Part 1?

Question #5. Thermal NO_x Emissions from LBG

This is similar to Question 4, except now we are interested in your judgment about uncertainty in the thermal NO_x emissions from LBG in systems such as Case ALH and Case AKH. What is the uncertainty, if any, in the thermal NO_x emission rate (please define the units that you are using, if not lb NO_x as NO_2 per million BTU of coal feed) that can be achieved in a fifth-of-a-kind system firing LBG (as in Case ALH)? In this case, because we anticipate that rich/lean combustors may be employed for fuel-NO_x control, we ask you to consider the effect of this technology on thermal NO_x emissions. First, consider Case ALH, a gas turbine using LBG from an air-blown dry-ash Lurgi gasifier.

- Consider the highest possible NO_x emission rate that might occur. What is this rate and how might it be achieved (e.g., perhaps fuel burns poorly in the rich combustion stage, leading to initiation of combustion in the lean zone resulting in high flame temperatures). How likely is the emission rate to be worse than the number you have just estimated?
- Consider the lowest possible NO_x emission rate that might occur. What is this rate and how might it be achieved. How likely would it be to obtain actual emissions below this rate?
- What is the median (50-50 percent of getting higher or lower) emission rate? If you prefer to specify a particular type of probability distribution model, such as a triangle distribution, then provide appropriate judgments in lieu of the median (e.g., mode or "most likely" value for a triangle distribution).
- If you haven't already specified and defined the parameters of a probability distribution model (e.g., triangle, normal, lognormal), can you estimate the 25th fractile (i.e. there is a one in four chance that the emission rate is less than this number)? What about the 75th fractile (i.e. a one in four chance that the value is higher than this number)?
- Can you draw a probability distribution to represent your judgement? You may draw the distribution as either a pdf or a cdf. Can it be represented by one of the distributions in Figure 1 of Part 1?

Does your answer differ in any way for a gas turbine firing LBG from an air-blown KRW gasifier (Case AKH)? If so, please provide similar information as for Case ALH.

Question #6. Fuel-NO_x Emissions from LBG

Please provide your judgment about the fraction of ammonia in the fuel gas that will be converted to NOx in the exhaust gas for Case ALH (air-blown Lurgi-based system). If your judgment is a function of the ammonia concentration of the inlet gas, we would appreciate if you could provide judgments for several ammonia concentrations: e.g., 500 ppm, 2,000 ppm, and 5,000 ppm. If there are any other key functional dependencies that you expect, such as hydrocarbon concentration in the fuel gas, pressure, or others, please appropriately caveat your judgments or, if you wish, provide a set of judgments for each combination of independent variable values that you select.

Because the set of judgments may involve several probability distributions, we suggest you might want to use a triangular distribution for simplicity. This requires estimates only for the lowest, highest, and most likely conversion rates. However, feel free to use any type of distribution which best represents your judgment.

- First, consider "conventional" combustors, such as those that are expected to be offered as standard equipment with the high-firing temperature gas turbines.

What is the expected ammonia conversion rate for Case ALH? Please consider the worst, best, and most likely values that could be obtained for an ammonia concentration of 2,000 ppm. Can you provide some technical basis for your judgment? Does your answer change if the ammonia concentration is 500 ppm? 5,000 ppm? If so, could you provide your judgments for these cases as well?

- Now consider a possible commercial rich/lean combustor in combination with the high-firing temperature gas turbines. What is the expected ammonia conversion rate for Case ALH? Please consider the worst, best, and most likely values that could be obtained for an ammonia concentration of 2,000 ppm. What is the basis for your judgment (e.g., bench scale tests indicate very low conversion rates are possible, but in larger scale systems mixing problems may contribute to a higher emission rate). Does your answer change if the ammonia concentration is 500 ppm? 5,000 ppm? If so, could you provide your judgments for these cases as well?

- Finally, consider Case AKH, which involves a fuel gas from a KRW gasifier. Do your judgments for this system differ from those for the Lurgi based system for either the conventional or rich/lean combustors? If so, could you please provide a similar set of judgments and the basis for them?

Question #7. CO Emissions from LBG

We would like your judgment about the possible emission rate of CO associated with both conventional and rich/lean combustor designs when firing LBG at baseload, as in Case ALH and Case AKH. Please define the type of unit you are assuming (e.g., fraction of CO in the fuel gas that is unconverted in the combustor, CO emissions in ppmv in the exhaust gas, uncorrected, etc).

- First, consider "conventional" combustors, such as those that are expected to be offered as standard equipment with the high-firing temperature gas turbines (see Table 2 of Part 2). What is the expected CO emission rate for Case ALH? Please consider the worst, best, and most likely values that could be obtained. Can you provide some technical basis for your judgment?

- Does your answer differ for Case AKH? If so, please provide similar information.

- Consider now the use of a rich/lean combustor in a fifth-of-a-kind commercial plant. What is the uncertainty in the CO emission rate associated with Case ALH? Again, consider the worst, best, and most likely values (or otherwise specify the appropriate probability distribution to represent your judgment). Please explain the basis for your judgment.

- Does your answer for the CO emissions from a rich/lean combustor change if the gas turbine burns the fuel gas of Case AKH rather than Case ALH? If so, please provide your judgment for this case also.

Question #8. Combustor Pressure Drop

We would like your judgment about the uncertainty involved in predicting the long term life cycle combustor pressure drop. The build up of any deposits on the combustor walls or in the transition piece may lead to an increase in pressure drop. The likelihood of this type of buildup may depend on upstream particulate control as well as combustion efficiency and conditions in the combustor. Please express your judgment about pressure drop either as a percentage of the compressor outlet pressure or as a pressure loss in psi.

- For Case ALH, consider the pressure drop expected in a rich/lean combustor? What might be the worst pressure drop? The lowest? The most likely, mean, or median? Can you explain how these different values might be obtained?
- Would the pressure drop in a standard combustor for Case ALH differ in any way than that for a rich/lean combustor? Could you provide your judgments for this case?
- Do the answers to the two questions above differ for Case AKH? If so, could you provide your judgments for this case also?
- For Case OKC, consider the type of combustor you expect for a fifth-of-a-kind plant with this fuel (e.g., lean/lean combustor, multiple fuel nozzles with steam injection, one fuel nozzle per can, etc). What system are you assuming? What is the pressure drop you expect for this system? What is the worst case? Best case? Most likely pressure drop? Why?

Question #9. Turbine Inlet Temperature/Maintenance Costs

In Part 2, a number of factors were discussed which might interfere with the advanced cooling systems in high firing temperature gas turbines firing coal gas. A possible effect of these problems might be to require a reduction in gas turbine firing temperature or a derating of the gas turbine (e.g., reduction in mass flow). Alternatively, more expensive maintenance may be required.

- Do you think a reduction in gas turbine firing temperature may be required under some conditions for high firing temperature (2,300 °F) gas turbines firing coal gas (consider Case ALH first)? If so, what are these conditions? How much might the firing temperature have to be reduced over the life of the plant? Can you provide a probability distribution to represent your judgment?
- If you don't think a reduction in firing temperature would ever be required, but an increase in maintenance cost may be incurred for more frequent blade cleaning or reblading, could you provide a judgment about what the uncertainty in maintenance cost might be? Or, if you think both a reduction in firing temperature and increased maintenance cost might occur, we would also like your judgment about maintenance costs. For screening studies, it is typical to express the annual maintenance cost as a fraction of the process area direct capital cost. For example, maintenance cost might be assumed to be 1.5 percent per year of direct capital cost with a clean fuel, but might increase to, say 3 or 4 percent per year if significant plugging, deposition, erosion, and corrosion occurs. What is your judgment about uncertainty in maintenance cost? (As with previous parameters, consider worst and best possible values and then consider the type of probability distribution which describes your judgment).
- Do your answers differ for Case AKH or Case OKC. For example, Case OKC might have less of a problem with plugging of film cooling air holes because a wet scrubber system is used, in contrast to the two-stages of cyclones assumed for Case ALH and Case AKH. This might reduce the plugging or deposition rates, and thus affect your answer.

Question #10. Other Considerations

In Part 2, a number of concerns were discussed regarding using coal gas in gas turbines originally designed for natural gas firing. One concern is the possibility of compressor surging or stalling when operating in an off-design mode (e.g., firing coal gas at baseload). Another concern is the controllability of a gas turbine in an IGCC system, particularly with compressor discharge air extraction for gasifier blast air. The availability and reliability of materials for high temperature fuel valves may limit IGCC system design by imposing a ceiling on maximum fuel gas temperatures. The possibility of deposition, erosion, corrosion, and plugging in the hot gas path may require stringent controls for particulates and alkali, beyond the capability of components commonly assumed in conceptual design studies.

Do you foresee that these types of problems will impose serious limitations on the development of IGCC concepts such as Case ALH? If so, in what ways? How realistic is it to expect that commercial technology for this type of system will be developed and "in-hand" within the next ten years? Please feel free to add any comments you think are relevant to our modeling effort.

Question #11. Other Experts

Please suggest other experts (preferably at METC) whom we should contact for judgments about uncertainties in this system. Please supply their names, titles, area of expertise, phone numbers, and addresses.

Question #12. Feedback

We would like your comments on these briefing materials and how easy/difficult it was to develop judgments about uncertainties. Is there any other information about uncertainty analysis you would like to see in Part 1? Was the summary of technical information in Part 2 a useful starting point for your thinking about uncertainties for this process? Was it difficult for you to develop estimates of the range or likelihood of various values for variables which you believe to be uncertain? Please discuss these or any other comments you may have.

Thank you for your help and insights.

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C.0 SAMPLE OUTPUTS FOR IGCC COST MODELS

C.1 Oxygen-blown KRW IGCC with Cold Gas Cleanup

A. COST MODEL PARAMETERS

Plant Capacity Factor:	0.65	Cost Year:	January 1989
General Facilities Factor:	0.20	Plant Cost Index:	351.5
Indirect Construction:	0.20	Chemicals Cost Index:	411.3
Sales Tax:	0.05	Escalation:	0.00
Engr & Home Office Fee:	0.10	Interest:	0.10
Project Contingency:	0.17	Years of construction:	4
Number of Shifts:	4.25	Byproduct marketing:	0.10
Fixed Charge Factor:	0.1034	Average Labor Rate:	19.70
Variable Levelization Cost Factor:	1.0000	Book Life (years):	30

B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Oxidant Feed	0.050	0.020
Gasification	0.200	0.045
Low Temperature Gas Cooling	0.000	0.030
Selexol	0.100	0.020
Claus Plant	0.050	0.020
Beavon-Stretford	0.100	0.020
Boiler Feedwater Treatment	0.000	0.015
Process Condensate Treatment	0.300	0.020
Gas Turbine	0.125	0.015
Heat Recovery Steam Generator	0.025	0.015
Steam Turbine	0.025	0.015
General Facilities	0.050	0.015

C. DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS ($1,000)

Plant Section	Number of Units Operating	Total	Direct Capital Cost	Process Contingency
Coal Handling	1	1	29053.	1986.
Oxidant Feed	2	2	63254.	4323.
Gasification	6	7	121030.	33089.
Low Temperature Gas Cooling	2	2	24852.	0.
Selexol	2	2	16777.	2293.
Claus Plant	2	3	6784.	464.
Beavon-Stretford	2	3	8820.	1206.
Boiler Feedwater Treatment	1	1	4685.	0.
Process Condensate Treatment	1	1	2886.	1184.
Gas Turbine	3	3	96000.	16404.
Heat Recovery Steam Generator	3	3	29992.	1025.
Steam Turbine	1	1	43160.	1475.
General Facilities	N/A	N/A	89459.	6114.

D. TOTAL CAPITAL REQUIREMENT ($1,000)

Description	Annual Cost

Total Direct Cost	536752.
Indirect Construction Cost	107350.
Sales Tax	22007.
Engineering and Home Office Fees	66611.
Environmental Permitting	1000.
Total Indirect Costs	196968.
Total Process Contingencies	69562.
Project Contingency	140574.
Total Plant Cost	943857.
AFDC	151253.
Total Plant Investment	1095110.
Preproduction (Startup) Costs	27460.
Inventory Capital	16264.
Initial Catalysts and Chemicals	5177.
Land	2395.
TOTAL CAPITAL REQUIREMENT ($1,000) -------->	1157357.

E. FIXED OPERATING COSTS ($/year)

Description	Annual Cost
Operating Labor	7836660.
Maintenance Costs	23261312.
Administration and Supervision	5142355.
TOTAL FIXED OPERATING COST ($/year) -------------->	36240328.

F. VARIABLE OPERATING COSTS

1. CONSUMABLES ($/year)

Description	Unit Cost	Material Requirement	Annual Operating Cost
Sulfuric Acid:	110.00 $/ton	1394.3 ton/yr	153373.
NaOH:	220.00 $/ton	317.0 ton/yr	69741.
Na2 HPO4:	0.70 $/lb	1291.2 lb/yr	904.
Hydrazine:	3.20 $/lb	6207.2 lb/yr	19863.
Morpholine:	1.30 $/lb	5765.7 lb/yr	7495.
Lime:	80.00 $/ton	537.8 ton/yr	43024.
Soda Ash:	160.00 $/ton	597.4 ton/yr	95580.
Corrosion Inh.:	1.90 $/lb	106984.4 lb/yr	203270.
Surfactant:	1.25 $/lb	106984.4 lb/yr	133730.
Chlorine:	250.00 $/ton	17.2 ton/yr	4293.
Biocide:	3.60 $/lb	19101.7 lb/yr	68766.
Selexol Solv.:	1.80 $/lb	41307.3 lb/yr	74353.
Claus Catalyst:	440.00 $/ton	8.4 ton/yr	3690.
Sul.. Acid Cat:	1.90 $/liter	0.0 liter/yr	0.
SCOT Catalyst:	230.00 $/ft3	0.0 ft3/yr	0.
SCOT Chemicals:	0.36 $/ft3	0.0 ft3/yr	0.
B/S Catalyst:	170.00 $/ft3	39.5 ft3/yr	6710.
B/S Chemicals:	N/A	N/A	78386.
Fuel Oil:	42.00 $/bbl	37761.9 bbl/yr	1585998.
Plant Air Ads.:	2.80 $/lb	2832.1 lb/yr	7930.
Raw Water:	0.73 $/Kgal	353851.4 Kgal/yr	258312.
Waste Water:	840.00 $/gpm ww	145578.6 lb/hr	158802.
LPG - Flare:	11.70 $/bbl	3304.2 bbl/yr	38659.
TOTAL CONSUMABLES ($/year) ---------------------->			3012880.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)

Coal:	1.61 $/MMBtu	524716.4 lb/hr	61446186.

Ash Disposal: 10.00 $/ton 554.1 ton/day 1314604.
Byprod. Credit: 125.00 $/ton 7.1 ton/hr (4527726.)

TOTAL VARIABLE OPERATING COST ($/year) ------------> 61245943.

G. COST OF ELECTRICITY

Power Summary (MWe)		Auxiliary Loads (MWe)			
Gas Turbine Output	473.11	Coal Handling	3.74	Claus	0.28
Steam Turbine Output	271.96	Oxidant Feed	56.46	B/S	0.84
Total Auxiliary Loads	79.39	Gasification	1.93	Proc. Cond	0.10
		Low T Cool.	1.01	Steam Cycle	3.27
Net Electricity	665.67	Selexol	3.85	General Fac	7.91

Capital Cost: 1738.62 $/kW
Fixed Operating Cost: 54.44 $/(kW-yr)
Incremental Variable Costs: 1.14 mills/kWh
Byproduct Credit: 1.19 mills/kWh
Fuel Cost: 16.21 mills/kWh
Variable Operating Cost: 16.16 mills/kWh
COST OF ELECTRICITY ---------------------------------> 57.29 mills/kWh

Heat Rate is: 8867. BTU/kWh. Efficiency is: 0.3851

H. ENVIRONMENTAL SUMMARY

INPUTS:	Coal	0.788	lb/kWh
	Water	0.779	lb/kWh
OUTPUTS	Water	0.098	lb/kWh
	Ash	0.069	lb/kWh
	WstWater	0.219	lb/kWh
	CO2	1.681	lb/kWh
	CO	0.000	lb/kWh
	SO2	0.342174	lb/MMBtu
	NOx	0.142402	lb/MMBtu
	COS	0.000000	lb/MMBtu
	NH3	0.000123	lb/kWh

C.2 Air-blown KRW IGCC with Hot Gas Cleanup

COST SUMMARY
Air Blown KRW-Based IGCC System with Hot Gas Cleanup

A. COST MODEL PARAMETERS

Plant Capacity Factor:	0.65	Cost Year:	January 1989
General Facilities Factor:	0.20	Plant Cost Index:	351.5
Indirect Construction:	0.20	Chemicals Cost Index:	411.3
Sales Tax:	0.05	Escalation:	0.00
Engr & Home Office Fee:	0.10	Interest:	0.10
Project Contingency:	0.17	Years of construction:	4
Number of Shifts:	4.25	Byproduct marketing:	0.10
Fixed Charge Factor:	0.1034	Average Labor Rate:	19.70
Variable Levelization Cost Factor:	1.0000	Book Life (years):	30

B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Limestone Handling	0.050	0.030
Oxidant Feed	0.100	0.020
Gasification	0.200	0.045
Zinc Ferrite	0.400	0.030
Sulfation	0.400	0.040
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.250	0.020
Heat Recovery Steam Generator	0.025	0.015
Steam Turbine	0.025	0.015
General Facilities	0.050	0.015

C. DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS ($1,000)

Plant Section	Number of Units Operating	Total	Direct Capital Cost	Process Contingency
Coal Handling	1	1	29637.	2026.
Limestone Handling	1	1	5522.	378.
Oxidant Feed	3	3	8869.	1213.
Gasification	5	6	99777.	27288.
Zinc Ferrite	10	24	11124.	6085.
Sulfation	1	1	28154.	15399.
Boiler Feedwater Treatment	1	1	4368.	0.
Gas Turbine	3	3	96000.	32818.
Heat Recovery Steam Generator	3	3	28303.	968.
Steam Turbine	1	1	47099.	1610.
General Facilities	N/A	N/A	71771.	4907.

D. TOTAL CAPITAL REQUIREMENT ($1,000)

Description	Annual Cost
Total Direct Cost	430624.
Indirect Construction Cost	86125.
Sales Tax	17656.
Engineering and Home Office Fees	53440.
Environmental Permitting	1000.
Total Indirect Costs	158221.

Total Process Contingencies	92691.
Project Contingency	119269.
Total Plant Cost	800804.
AFDC	128329.
Total Plant Investment	929133.
Preproduction (Startup) Costs	25590.
Inventory Capital	19098.
Initial Catalysts and Chemicals	18140.
Land	2464.
TOTAL CAPITAL REQUIREMENT ($1,000) -------->	1003716.

E. FIXED OPERATING COSTS ($/year) ------------------------------------

Description	Annual Cost
Operating Labor	6617624.
Maintenance Costs	21596591.
Administration and Supervision	4576878.
TOTAL FIXED OPERATING COST ($/year) -------------->	32791094.

F. VARIABLE OPERATING COSTS --

1. CONSUMABLES ($/year)

Description	Unit Cost	Material Requirement	Annual Operating Cost
Sulfuric Acid:	110.00 $/ton	1449.9 ton/yr	159491.
NaOH:	220.00 $/ton	295.7 ton/yr	65050.
Na2 HPO4:	0.70 $/lb	1313.9 lb/yr	920.
Hydrazine:	3.20 $/lb	6316.7 lb/yr	20213.
Morpholine:	1.30 $/lb	5868.2 lb/yr	7629.
Lime:	80.00 $/ton	590.9 ton/yr	47271.
Soda Ash:	160.00 $/ton	654.5 ton/yr	104726.
Corrosion Inh.:	1.90 $/lb	117674.8 lb/yr	223582.
Surfactant:	1.25 $/lb	117674.8 lb/yr	147094.
Chlorine:	250.00 $/ton	18.6 ton/yr	4640.
Biocide:	3.60 $/lb	20631.0 lb/yr	74272.
Zinc Fer Sorb:	3.00 $/lb	451422.2 lb/yr	1354267.
Limestone:	18.00 $/ton	477627.0 ton/yr	8597285.
Fuel Oil:	42.00 $/bbl	41216.9 bbl/yr	1731108.
Plant Air Ads.:	2.80 $/lb	3091.3 lb/yr	8656.
Raw Water:	0.73 $/Kgal	360456.2 Kgal/yr	263133.
Waste Water:	840.00 $/gpm ww	0.0 lb/hr	0.
LPG - Flare:	11.70 $/bbl	3606.5 bbl/yr	42196.
TOTAL CONSUMABLES ($/year) ---------------------->			12851531.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)

Coal:	1.61 $/MMBtu	536617.5 lb/hr	62839840.
Ash Disposal:	10.00 $/ton	1991.6 ton/day	4725150.

TOTAL VARIABLE OPERATING COST ($/year) ------------> 80416521.

G. COST OF ELECTRICITY --

Power Summary (MWe)		Auxiliary Loads (MWe)			
Gas Turbine Output	486.60	Coal Handling	3.81	Sulfation	1.69
Steam Turbine Output	296.78	Limestne Hdlg	0.76	Steam Cycle	3.25
Total Auxiliary Loads	56.80	Oxidant Feed	37.61	General Fac	6.09
		Gasification	3.59		
Net Electricity	726.58	Zinc Ferrite	0.00		

```
                                    Capital Cost:    1381.43 $/kW
                            Fixed Operating Cost:      45.13 $/(kW-yr)
      Incremental Variable Costs:   4.25 mills/kWh
      Byproduct Credit:             0.00 mills/kWh
      Fuel Cost:                   15.19 mills/kWh
                         Variable Operating Cost:      19.44 mills/kWh
COST OF ELECTRICITY ---------------------------------> 52.45 mills/kWh

     (Heat Rate is:  8308.   Efficiency is: 0.4111)

H.  ENVIRONMENTAL SUMMARY ---------------------------------------------
             INPUTS:   Coal            0.739 lb/kWh
                       Water           0.727 lb/kWh
                       Limestone       0.231 lb/kWh
             OUTPUTS   Ash             0.228 lb/kWh
                       Water           0.051 lb/kWh
                       CO2-GT          1.670 lb/kWh
                       CO2-Sulfa       0.044 lb/kWh
                       CO2-Tot         1.714 lb/kWh
                       CO              0.005 lb/kWh
                       SO2-GT         0.0124 lb/MMBtu
                       SO2-Sulfa      0.0004 lb/MMBtu
                       SO2-Tot        0.0129 lb/MMBtu
                       NOx-GT         0.7076 lb/MMBtu
                       NOx-Sulfa      0.0067 lb/MMBtu
                       NOx            0.7143 lb/MMBtu
             ENERGY IN:Coal        6036.2657 MMBtu
                       LASH         270.7655 MMBtu
```

C.3 Air-blown Lurgi-based IGCC with Hot Gas Cleanup

*** SUMMARY OF SAMPLED ASPEN FLOWSHEET PERFORMANCE PARAMETERS ***

PLANT SECTION	FLOWSHEET PERFORMANCE PARAMETERS	VALUE	UNITS
Coal Feed	Mass flow of coal to gasifier	498966.72	lb/hr
Limestone Feed	Mass flow of limestone (mixed)	0.00	lb/hr
	Mass flow of limestone (cisolid)	3.91	lb/hr
Oxidant Feed	Work to boost air compressor	0.133921E+08	Watts
	First precooler inlet air temp.	713.21	F
	First precooler outlet air temp.	668.88	F
	First precooler inlet BFW temp.	505.56	F
	First precooler outlet BFW temp.	605.29	F
	Heat transfer in first precooler	0.153414E+08	BTU/hr
	Heat trans. to BFW from HRSG	0.515623E+09	BTU/hr
	Heat trans. to BFW from Regen.	0.160659E+09	BTU/hr
	Heat leaving from economizer	0.538128E+09	BTU/hr
Gasification	Gasifier output syngas pres.	285.45	psia
	Gasifier output syngas temp.	1100.00	F
	Gasifier output syngas density	0.169991E-01	lbmole/ft3
	Gasifier recycle fines flow rate	16590.64	lb/hr
Zinc Ferrite	Zinc Ferrite inlet syngas flow	99144.89	lbmole/hr
	Off-gas to sulfuric acid plant	23947.05	lbmole/hr
	SO2 to sulfuric acid plant	512.43	lbmole/hr
	Off-gas temp. to sul. acid plant	1232.08	lbmole/hr
	Zinc ferrite inlet H2S flow	513.51	lbmole/hr
	Zinc ferrite inlet COS flow	0.00	lbmole/hr
Sulfation	Gasifier ash removal	43680.92	lb/hr
	Gasifier ash removal	7822.12	lb/hr
	Gasifier fines removal	218.30	lb/hr
	Gasifier fines removal	654.89	lb/hr
BFW Treating	Raw water to power plant	1014700.25	lb/hr
	Steam turbine condensate	760138.28	lb/hr
Gas Turbine	Gas turbine inlet air flow	9012893.70	lb/hr
Steam Cycle	HRSG outlet HP steam pres.	1465.00	psia
	HRSG outlet HP steam flow	1150860.83	lb/hr
	Gas turbine net shaft work	-.497930E+09	Watts
	Steam turbine net shaft work	-.169644E+09	Watts
	Steam cycle auxiliary power	0.317899E+07	Watts
	Acid gas auxiliary power	0.000000E+00	Watts
Miscellaneous	Moisture in coal feed	12.00	wt %
	Ash in coal	10.00	wt %
	Temp. of ambient air	59.00	F
	Heating value of coal	12774.00	BTU/lb
Environmental	High Pressure Blowdown	23486.96	lbmole/hr
	Low Pressure Blowdown	23486.96	lbmole/hr
	CO2 from gas turbine	24792.71	lbmole/hr
	CO from gas turbine	66.70	lbmole/hr
	SO2 from gas turbine	1.08	lbmole/hr
	COS from gas turbine	0.00	lbmole/hr
	CH4 from gas turbine	0.00	lbmole/hr
	H2S from gas turbine	0.00	lbmole/hr
	NH3 to gas turbine	352.68	lbmole/hr
	NO from gas turbine	317.52	lbmole/hr

NO2 from gas turbine		16.71 lbmole/hr
Actual coal heating value		11248.73 BTU/lb
COST VAR WARNING ---- Variable MCFGI	value of	498966.719
in DCSS	above the upper limit of	433000.000
COST VAR WARNING ---- Variable VSNZFI	value of	8836.921
in DCCY	above the upper limit of	6000.000
COST VAR WARNING ---- Variable MRW	value of	1014700.248
in DCBF	above the upper limit of	614000.000

COST SUMMARY

Air Blown Lurgi-Based IGCC System with Hot Gas Cleanup

A. KEY INPUT ASSUMPTIONS --

Performance Assumptions:

Gasifier Availability:	0.87	Sorb. Sulfur Loading:	0.168
Max. Desulf. Vessel Diameter:	12.50	Superficial Velocity:	2.00
L/D Ratio:	3.00	Absorption Cycle Time:	30.0
Maximum Space Velocity (1/hr):	2000.	Sorbent Bulk Density:	82.0
Sorb. Attrition Rate/80 cycles:	0.800	Generator Efficiency:	0.9850
Gasifier Coal Throughput:	1.01		

Economic Assumptions:

Cost Year:	January 1989	Inflation Rate:	0.000
Plant Cost Index:	351.5	Real Escalation Rate:	0.000
Chemicals Cost Index:	411.3	Plant Booklife:	30
Plant Capacity Factor:	0.65	Sales Tax Rate:	0.05
General Facilities Factor:	0.20	Real Return on Debt:	0.046
Indirect Construction Factor:	0.20	Real Ret. on Pref.:	0.052
Engr & Home Office Fees:	0.10	Real Ret. on Equity:	0.087
Project Contingency Factor:	0.17	Debt Ratio:	0.500
Byproduct Marketing Factor:	0.10	Pref. Stock Ratio:	0.150
Average Operating Labor Rate:	19.70	Fed. & State Taxes:	0.380
Number of Shifts:	4.25	Investment Tax Credit:	0.000
Construction Interest Rate:	0.10	Prop. Taxes & Insur.:	0.020
Years of Construction:	4		

Process Contingency and Maintenance Cost Factors:

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Limestone Handling	0.000	0.030
Oxidant Feed	0.100	0.020
Gasification	0.200	0.030
Coke, Ash, & Bent. Subsystems	0.050	0.020
High Temp. Cyclones	0.050	0.030
Zinc Ferrite	0.400	0.030
Sulfuric Acid Plant	0.100	0.020
Sulfation	0.600	0.045
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.250	0.020
Heat Recovery Steam Generator	0.025	0.015
Steam Turbine	0.025	0.015
General Facilities	0.050	0.015

B. CALCULATED DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS ($1,000) --

Plant Section	Number of Operating	Units Total	Direct Capital Cost	Process Contingency
Coal Handling	1	1	21379.	1462.
Limestone Handling	0	0	0.	0.
Oxidant Feed	3	3	4703.	643.
Gasification	11	13	83663.	22885.
Coke, Ash, & Bent. Subsystems	11	13	13769.	942.
High Temp. Cyclones	22	26	6175.	422.
Zinc Ferrite	11	26	11491.	6287.
Sulfuric Acid Plant	1	1	23210.	3174.
Sulfation	0	0	0.	0.
Boiler Feedwater Treatment	1	1	5293.	0.
Gas Turbine	3	3	96000.	32825.
Heat Recovery Steam Generator	3	3	27691.	947.
Steam Turbine	1	1	26519.	907.
General Facilities	N/A	N/A	63978.	4375.

C. CALCULATED TOTAL CAPITAL REQUIREMENT ($1,000) ---------------------

Description	Capital Cost
Total Direct Cost	383870.
Indirect Construction Cost	76774.
Sales Tax	15739.
Engineering and Home Office Fees	47638.
Environmental Permitting	1000.
Total Indirect Costs	141151.
Total Process Contingencies	74869.
Project Contingency	104981.
Total Plant Cost	704871.
AFDC	112956.
Total Plant Investment	817826.
Preproduction (Startup) Costs	22225.
Inventory Capital	17570.
Initial Catalysts and Chemicals	23606.
Land	2246.
TOTAL CAPITAL REQUIREMENT ($1,000) -------->	891651.

D. CALCULATED FIXED OPERATING COSTS ($/year) ------------------------

Description	Annual Cost
Operating Labor	8533252.
Maintenance Costs	15377427.
Administration and Supervision	4405267.
TOTAL FIXED OPERATING COST ($/year) -------------->	28315946.

E. CALCULATED VARIABLE OPERATING COSTS ------------------------------

1. CONSUMABLES ($/year)

Description	Assumed Unit Cost	Calc. Material Requirement	Calc. Annual Operating Cost
Sulfuric Acid:	110.00 $/ton	2025.6 ton/yr	222816.
NaOH:	220.00 $/ton	426.1 ton/yr	93748.
Na2 HPO4:	0.70 $/lb	2455.7 lb/yr	1719.
Hydrazine:	3.20 $/lb	11820.1 lb/yr	37824.
Morpholine:	1.30 $/lb	11023.7 lb/yr	14331.

Lime:	80.00	$/ton	509.0	ton/yr	40722.
Soda Ash:	160.00	$/ton	566.4	ton/yr	90620.
Corrosion Inh.:	1.90	$/lb	101187.7	lb/yr	192257.
Surfactant:	1.25	$/lb	101187.7	lb/yr	126485.
Chlorine:	250.00	$/ton	16.4	ton/yr	4105.
Biocide:	3.60	$/lb	18272.4	lb/yr	65781.
Zinc Fer Sorb:	3.00	$/lb	2627299.5	lb/yr	7881899.
Limestone:	18.00	$/ton	0.0	ton/yr	0.
Sul Acid. Cat.:	1.90	$/l	26341.7	l/yr	50049.
Coke:	150.00	$/ton	1393.6	ton/yr	209038.
Bentonite:	0.03	$/lb	4062086.6	lb/yr	117801.
Fuel Oil:	42.00	$/bbl	35888.4	bbl/yr	1507315.
Plant Air Ads.:	2.80	$/lb	2691.6	lb/yr	7537.
Raw Water:	0.73	$/Kgal	692583.7	Kgal/yr	505586.
Waste Water:	840.00	$/gpm ww	0.0	lb/hr	0.
LPG - Flare:	11.70	$/bbl	3140.2	bbl/yr	36741.
TOTAL CONSUMABLES ($/year) ---------------------->					11206371.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)

Coal:	1.61	$/MMBtu	498966.7	lb/hr	58430800.
Ash Disposal:	10.00	$/ton	628.5	ton/day	1491151.
Sulfuric Acid:	40.00	$/ton	153732.8	ton/yr	(5534382.)

TOTAL VARIABLE OPERATING COST ($/year) ------------> 65593941.

F. CALCULATED COST OF ELECTRICITY ------------------------------------

Power Summary (MWe)		Auxiliary Loads (MWe)			
Gas Turbine Output	490.46	Coal Handling	0.94	Sulfation	0.00
Steam Turbine Output	167.10	Limestne Hdlg	0.00	Acid Rem.	1.51
Total Auxiliary Loads	24.91	Oxidant Feed	13.39	Steam Cycle	3.18
		Gasification	2.49	General Fac	3.39
Net Electricity	632.65	Zinc Ferrite	0.00		

Capital Cost: 1409.39 $/kW
Fixed Operating Cost: 44.76 $/(kW-yr)
Incremental Variable Costs: 3.52 mills/kWh
Byproduct Credit: 1.54 mills/kWh
Fuel Cost: 16.22 mills/kWh
Variable Operating Cost: 18.21 mills/kWh
COST OF ELECTRICITY ----------------------------------> 51.66 mills/kWh
Fixed Charge Factor: 0.1034 Variable Cost Levelization Factor: 1.0000

The plant heatrate (HHV) is: 8872. BTU/kWh. Efficiency: 0.3849

G. ENVIRONMENTAL SUMMARY --

INPUTS:	Coal	0.7887	lb/kWh
	Water	1.6039	lb/kWh
OUTPUTS:	Blowdown	0.0742	lb/kWh
	Ash	0.0828	lb/kWh
	Sulfuric Acid	0.0854	lb/kWh
	CO2 emissions	1.7243	lb/kWh
	CO emissions	0.0030	lb/kWh
	SO2 emissions	0.0415	lb/MMBtu
	COS emissions	0.0000	lb/MMBtu
	CH4 emissions	0.0000	lb/kWh

H2S emissions	0.0000 lb/MMBtu
NOx emissions	2.7392 lb/MMBtu