2009

Techno-economic analysis of biomass-to-liquids production based on gasification

Ryan Michael Swanson
Iowa State University

Techno-economic analysis of biomass-to-liquids production based on gasification

by

Ryan Michael Swanson

A thesis submitted to the graduate faculty

In partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

Co-majors: Mechanical Engineering; Biorenewable Resources and Technology

Program of Study Committee:
Robert C. Brown, Major Professor
Robert P. Anex
Terrence R. Meyer

Iowa State University

Ames, Iowa

2009

TABLE OF CONTENTS

LIST OF ACRONYMS

AGR: Acid Gas Removal
ASU: air separation unit
BTL: biomass to liquids
DCFROR: discounted cash flow rate of return
DME: dimethyl-ether
FCI: fixed capital investment
FT: Fischer-Tropsh
GGE: gallon of gasoline equivalent
HRSG: heat recovery steam generator
HT: high temperature
IC: indirect costs
IRR: internal rate of return
ISU: Iowa State University
LT: low temperature
MJ: megajoule
MM: million
MTG: methanol to gasoline
MW: megawatt
Nm^3: normal cubic meter
NREL: National Renewable Energy Laboratory
PSA: pressure swing adsorption
PV: product value
SMR: steam methane reforming
SWGS: sour water-gas-shift
TDIC: total direct and indirect cost
TIC: total installed cost
TPEC: total purchased equipment cost
TCI: total capital investment
WGS: water-gas-shift

LIST OF FIGURES

LIST OF TABLES

ACKNOWLEDGEMENTS

Many of my friends and colleagues have helped and supported me along during my research. I thank Dr. Brown for giving me the opportunity to join the Center for Sustainable Environment Technologies (CSET) and perform this research. This opportunity allowed me to discover and explore the exciting area of biorenewable resources. I would also like to thank Dr. Anex in Agriculture and Biosystems Engineering and Dr. Meyer in Mechanical Engineering for serving on my committee and providing guidance.

Moreover, I thank ConocoPhillips Company for the financial support of this research and for providing the opportunity to further my learning during a summer visit to the National Renewable Energy Laboratory. I thank Alex Platon of ConocoPhillips company for imparting his expertise in support of this study.

I thank the National Renewable Energy Laboratory for allowing me the opportunity to spend time at its laboratory where I was able to receive knowledge and expertise in support of this study.

Behind the scenes, the administrative staffs of the Mechanical Engineering department and CSET have supported me immensely by keeping things running smoothly. I especially thank Diane Love, Becky Staedtler, and MaryAnn Sherman for frequent assistance.

The talented scientists and fellow graduate students of CSET have provided much guidance and feedback during meetings and presentations. I thank Justinus Satrio for his frequent guidance throughout my research.

Finally, I thank my parents, Robert and Barbara Swanson for continual interest and encouragement in my endeavors. Special thanks go to my good friends Cody Ellens and Andy Herringshaw and my girlfriend Sara Hanenburg. Most of all, I thank God for the talents he has bestowed upon me.

ABSTRACT

This study compares capital and production costs of two biomass-to-liquid production plants based on gasification. The goal is to produce liquid transportation fuels via Fischer-Tropsch synthesis with electricity as co-product. The biorefineries are fed by 2000 metric tons per day of corn stover. The first biorefinery scenario is an oxygen-fed, low temperature (870°C), non-slagging, fluidized bed gasifier and the second scenario an oxygen-fed, high temperature (1300°C), slagging, entrained flow gasifier. Both are followed by catalytic Fischer-Tropsch synthesis and hydroprocessing to naphtha and distillate liquid fractions.

Process modeling software is utilized to organize the mass and energy streams and cost estimation software is used to generate equipment costs. Economic analysis is performed to estimate the capital investment and operating costs. A 20 year discounted cash flow rate of return (DCFROR) analysis is developed to estimate a fuel product value (PV) at a net present value of zero with 10% internal rate of return. All costs are adjusted to the year 2007.

Results show that the total capital investment required for n^{th} plant scenarios are \$610 million and \$500 million, for high temperature and low temperature scenarios, respectively. PV for the high temperature and low temperature scenarios are estimated to be \$4.30 and \$4.80 per gallon of gasoline equivalent (GGE), respectively. The main reason for a difference in PV between the scenarios is because of a higher carbon efficiency and subsequent higher fuel yield for the high temperature scenario. Sensitivity analysis is also performed on process and economic parameters which shows that total capital investment and feedstock cost are among the most influential parameters affecting the PV while least influential parameters include per pass Fischer-Tropsch reaction conversion extent, inlet feedstock moisture, and catalyst cost.

In order to estimate the cost of a pioneer plant (1^{st} of its kind) an analysis is performed which inflates total capital investment and deflates the plant output for the first several years of operation. Base case results of this analysis estimate a pioneer plant investment to be \$1.3 billion and \$1.0 billion for high temperature and low temperature scenarios, respectively. Resulting respective PV are estimated to be \$7.40 and \$7.70 per GGE for pioneer plant.

1. INTRODUCTION

This study investigates economic feasibility of the thermochemical pathway of gasification to renewable transportation fuels. The objective is to compare capital investment costs and production costs for n^{th} plant biorefinery scenarios based on gasification. The selected scenarios are high temperature (slagging) gasification and low temperature (dry-ash) gasification both followed by Fischer-Tropsch synthesis and hydroprocessing. They are designed to produce liquid hydrocarbon fuels from 2000 dry metric ton (2205 dry short ton) per day of agricultural residue, namely, corn stover.

The two scenarios were chosen from many options according to the following criteria. The technology under consideration should be commercially ready in the next 5-8 years. The size of biorefinery should be feasible with current agricultural productivity and within realistic feedstock collection area. In addition, the desired end product should be compatible with the present fuel infrastructure, i.e. gasoline and/or diesel.

The high temperature gasification scenario is based on a steam/oxygen-fed entrained flow, slagging gasifier similar to that described in Frey and Akunuri [1]. The low temperature gasification scenario is based on a pressurized, steam/oxygen-fed fluidized bed gasifier developed by Gas Technology Institute and reported by Bain [2]. The main areas of operation are feedstock preprocessing, gasification, syngas cleaning, syngas conditioning/upgrading, fuel synthesis, power generation, and air separation (for oxygen production) as shown in Figure 1. Process modeling software is utilized to organize the mass and energy streams and cost estimation software is used to generate equipment costs. Economic analysis is performed to estimate the capital investment and operating costs. A 20 year discounted cash flow rate of return (DCFROR) analysis is developed to estimate a fuel product value (PV) at a net present value of zero with 10% internal rate of return. All costs are adjusted to the year 2007.

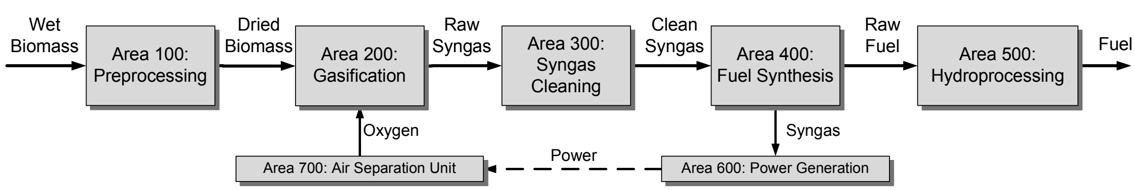


Figure 1. Overall process flow diagram for both scenarios

2. BACKGROUND

The word *economy* is defined in the American Heritage Dictionary as "careful, thrifty management of resources, such as money, material, or labor" and also as "an orderly, functional arrangement of parts; an organized system."[3] The origin of *economy* comes from the Greek word *oikonomiā* meaning "the management of a household."[3] Expanding the word, it can be defined as the careful management of the all of the earth's resources including human beings, monetary systems, and, in regards to this study, energy.

Natural ecosystems are good examples of the earth's economy in action. The earth's economy is evident in the aftermath of forest fires when new growth of forest rises from the ashes. Certain species of conifers flourish the most after a fire because of the heat release of seedlings. Another example is of the annual cycle of plant growth that humans use for sustenance. Year after year the cycle continues as plants utilize the sun's energy and the soil's nutrients to produce new crops. Continued energy from the sun and recycled nutrients from decomposed plants keep the cycle moving. Observation of the earth's cycles lead humans to gain much knowledge of how to practice appropriate *oikonomiā*.

Over the past few decades it has become evident that the appropriate economy of the earth's carbon is important for the direction of human life. A study of history leads to the realization that misuse of resources has serious consequences. During the middle of the last millennium, European misuse of forests led to a near destruction of the forests and demanded better resource management. The United States' misuse of petroleum during the last century led to the high point, or peak usage, of inexpensive, close-to-the-surface domestic petroleum. The balance of energy dependence has now shifted to a high degree of instability. With respect to appropriate *oikonomiā*, usage of carbonaceous energy resources requires careful planning.

2.1 Biorenewable Resources

The world population has long utilized materials that are in close proximity. The nearest resource available to the human population is the organic matter in the environment around them. This organic matter is present for a limited amount of time due to its decomposable nature. Brown [4] defines this material, or biorenewable resources, as organic material of recent biological origin. It is a renewable resource if the rate of consumption is equal to the regeneration or growth and therefore must be used only if preserving biodiversity [5]. As a result, these resources have been important contributors to the world economy serving as foodstuffs, transportation, energy, and construction materials, as well as many other functions.

Biorenewable resources for generating energy can be classified as woody biomass, energy crops, residues, and municipal waste [5]. The first two are primary resources while the remaining are secondary resources meaning their primary use has already occurred. Woody biomass includes logging products and energy crops include short rotation trees (e.g. poplar) and switchgrass. Residues can come from logging processing or agricultural processing (e.g. corn stover). According to Perlack et al. [6], the energy crop and agricultural residue potential in the United States is 1.4 billion annual tons. According to Department of Energy's "Roadmap for Agriculture Biomass Feedstock Supply in the US," there is potential for 2 billion annual tons including municipal waste and biosolids (e.g. manure).

Many end products can be produced from these resources. Aside from the conventional use of biomass for human food consumption, livestock feed, and building materials, there are many new pathways to provide renewable alternatives to our transportation, infrastructure, and energy. Combustion of biomass offers a way to provide heat and power to displace coal and fuel oil. Liquefaction of biomass through fast pyrolysis yields liquid products with the potential to displace petrochemicals. Additionally, gasification of biomass allows for chemical and liquid fuel synthesis, which is the focus of this study.

Developing an economy that involves biorenewable resources, especially biofuels, has many benefits. According to Greene et al. [7], biofuel production has the potential to provide a new source of revenue for farmers by generating $5 billion per year. Additionally, air quality can be improved through the use of biofuels. In the same study Green et al. reports that 22% of our total greenhouse gas emissions could be reduced if biofuels were developed to replace half of our petroleum consumption. Arguably, the most important benefit of biofuel production, when performed intelligently, is the potential for closing the carbon cycle.

2.2 Gasification

Gasification is a high temperature and catalytic pathway to biofuels. It is defined as the partial oxidation of solid, carbonaceous material with air, steam, or oxygen into a flammable gas mixture called producer gas or synthesis gas [4]. The synthesis gas contains mostly carbon monoxide and hydrogen with various amounts of carbon dioxide, water vapor, and methane. Typical volumetric energy content of synthesis gas is between 4-18 MJ/Nm^3 [8]. Comparatively, natural gas (comprised of mostly methane) energy content is 36 MJ/Nm^3 [8]. Much of the energy content of the biomass is retained in the gas mixture by partial oxidation rather than fully oxidizing the biomass which would result in the release of mostly thermal energy. Historically, gasification of coal and wood produced “town gas” where it was subsequently used to burn in street lamps [9]. Additionally, during the World Wars, vehicles were adapted to operate with gasification reactors [9]. During this same time period Germany developed the catalytic synthesis of transportation fuels from synthesis gas [10]. The same concept is still in use today by the South African Coal, Oil, and Gas Corporation (SASOL) to produce motor fuels and liquid byproducts using coal [10].

2.2.1 Reaction

There are four stages that occur during gasification of carbonaceous material: drying, devolatilization, combustion, and reduction [8]. First, the moisture within is heated and removed through a drying process. Second, continued heating devolatilizes the material where volatile matter exits the particle and comes into contact with the oxygen. Third, combustion occurs where carbon dioxide and carbon monoxide are formed from carbon and oxygen. The combustion stage is very exothermic and provides enough heat for the last stage, the reduction reactions, to occur. The last stage includes water gas reaction, Boudouard reaction, water-gas-shift reaction, and methanation reaction (Table 1). As all these stages progress, solid fixed carbon remains present. Fixed carbon amount varies depending on the equivalence ratio.

Table 1. Reactions occurring within the reduction stage of gasification

Name	Reaction
Water gas	$C + H_2O \rightarrow CO + H_2$
Boudouard	$C + H_2O \rightarrow 2CO$
Water-gas-shift	$CO + H_2O \rightarrow CO_2 + H_2$
Methanation	$CO + 3H_2 \rightarrow CH_4 + H_2O$

When equivalence ratio (defined as the actual air/fuel ratio all over the stoichiometric air/fuel ratio) increases, solid fixed carbon (i.e. char) decreases until enough oxidizer is available for complete conversion (Figure 2). This point of complete conversion occurs at approximately 0.25 equivalence ratio. At nearly the same point, the maximum synthesis gas energy content (without accounting for sensible energy) is reached.

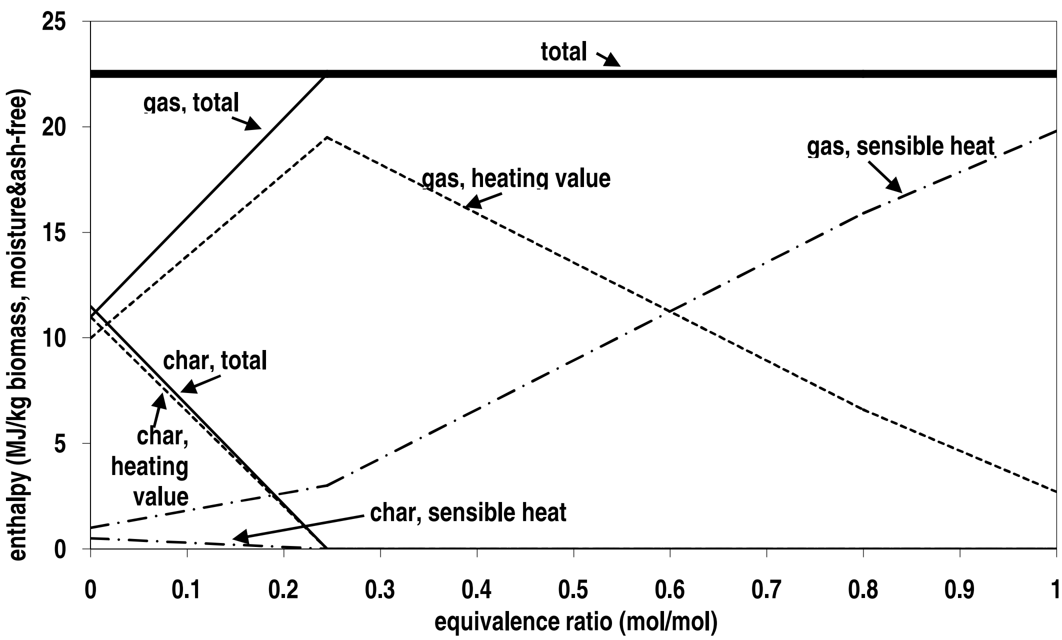


Figure 2. Energy content of the products of gasification of wood using air varied by equivalence ratio [11]

2.2.2 Gasifier Types

Gasification occurs in reactors of three types: fixed bed, fluidized bed, and entrained flow [12]. Fixed beds are fed with biomass from the top of the reactor and form a bed which gasifies as air moves through the bed (Figure 3). As the material releases volatile components, the char and ash exit through a grate at the bottom. Typical operating temperature range is 750-900°C. The two main types of fixed bed gasifiers are updraft and downdraft. The advantage of fixed bed is simplicity, but is limited in scale up and has low heat mixing due to high channeling potential within the reactor [13].

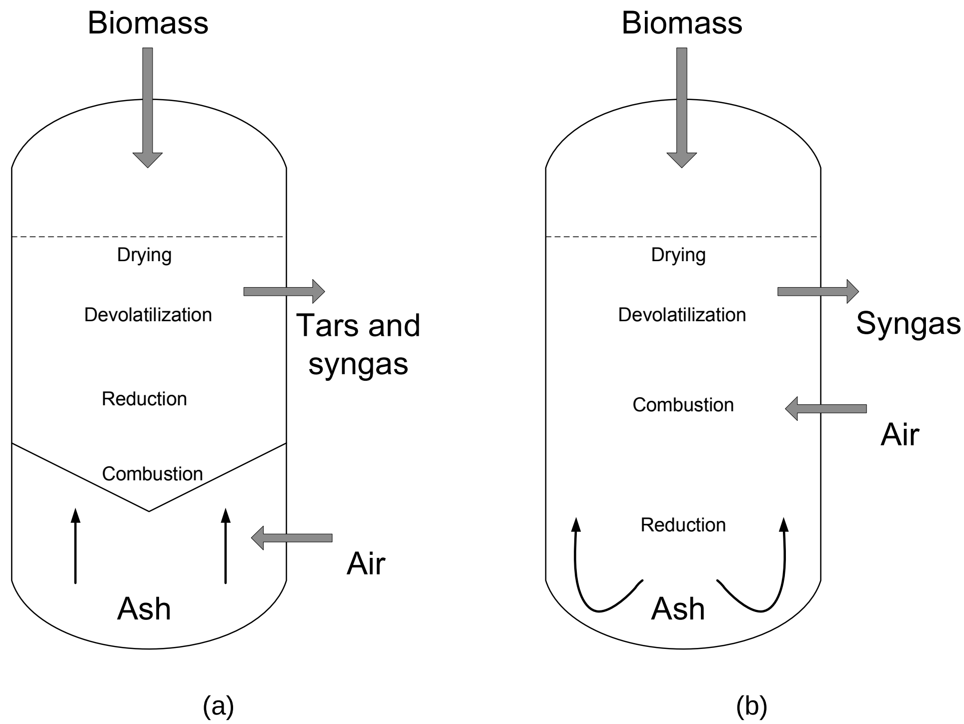
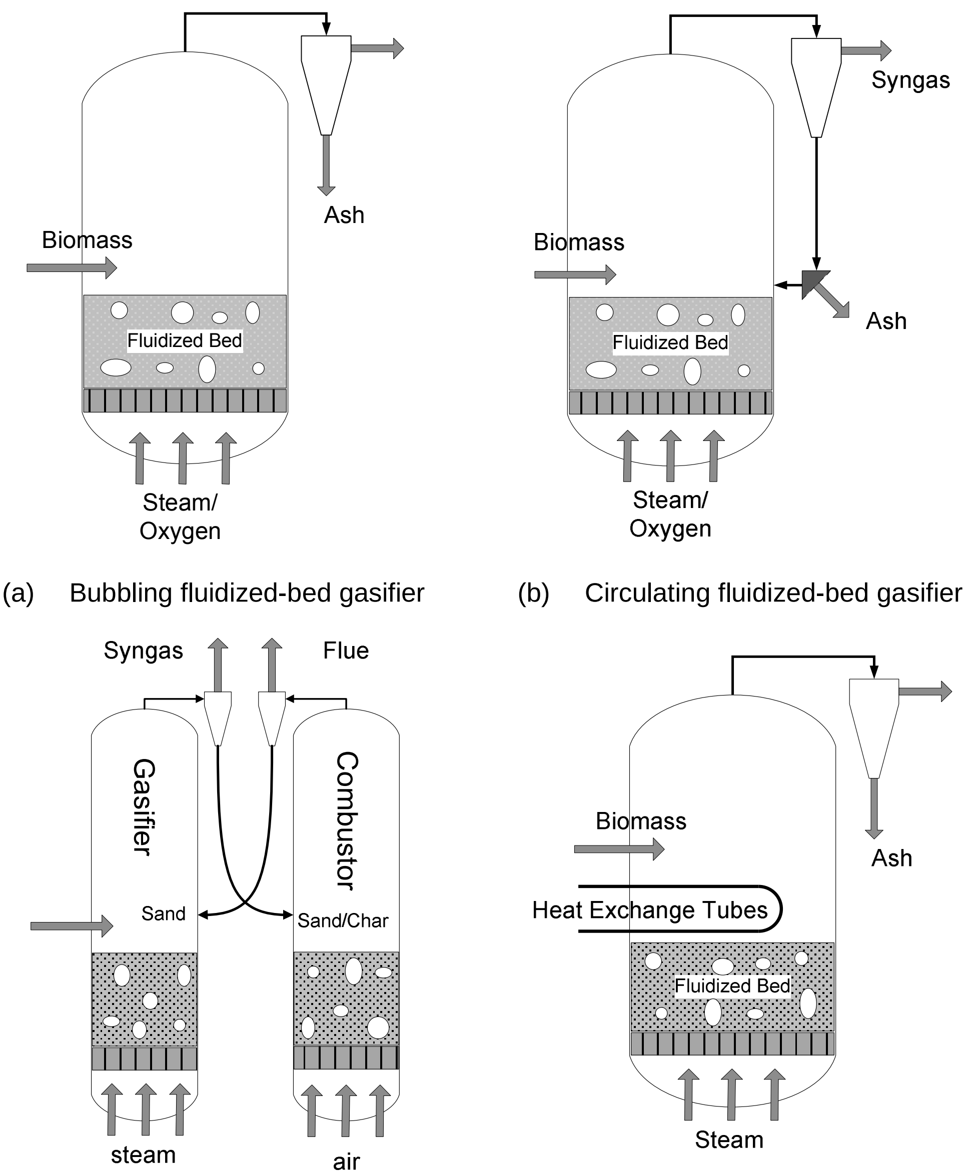


Figure 3. Design of fixed-bed (a) updraft and (b) downdraft gasifiers showing reaction zones [12]

When the volumetric gas flow is increased through the grate the fixed bed becomes a fluidized bed. Fluidized bed gasifiers are named because of the inert bed material that is fluidized by oxidizing gas creating turbulence through the bed material (Figure 4). Biomass enters just above top of the bed and mixes with hot, inert material creating very high heat and mass transfer. Operating temperature range is the same as for the fixed bed. Advantages of the fluidized bed include flexible feeds, uniform temperature distribution across bed, and large volumetric flow capability [14]. The main types of fluidized bed gasifiers are circulating fluidized bed (CFB) and bubbling fluidized bed, which are directly heated from the combustion reactions occurring in the bed. A bubbling bed produces gas and the ash and char falls out the bottom or the side. The CFB recycles the char through a cyclone while the product leaves out the top of the cyclone. Indirectly heated fluidized beds use a hot material such as sand to provide the heat needed for gasification as shown in Figure 4. Fluidized beds have high carbon conversion efficiencies and can scale up easily [13].



(c) indirectly heated gasifier via combustor

(d) indirectly heated gasifier via heat exchange tubes

Figure 4. Fluidized bed gasifier designs of (a) and (b) directly heated type and (c) and (d) indirectly heated type [15]

Another type of gasifier is the entrained flow gasifier (Figure 5). Normally operated at elevated pressures (up to 50 bar) it requires very fine fuel particles gasified at high temperatures to ensure complete gasification during the short residence times in the reactor. The Energy Research Centre group of the Netherlands has investigated this gasification type and have reported promise with biomass as long as the biomass is pretreated to certain requirements [16]. To keep the residence time at approximately the time for a particle to fall the length of the reaction zone, small fuel particles below 1 mm and high temperatures (1100-1500°C) are necessary for successful operation.

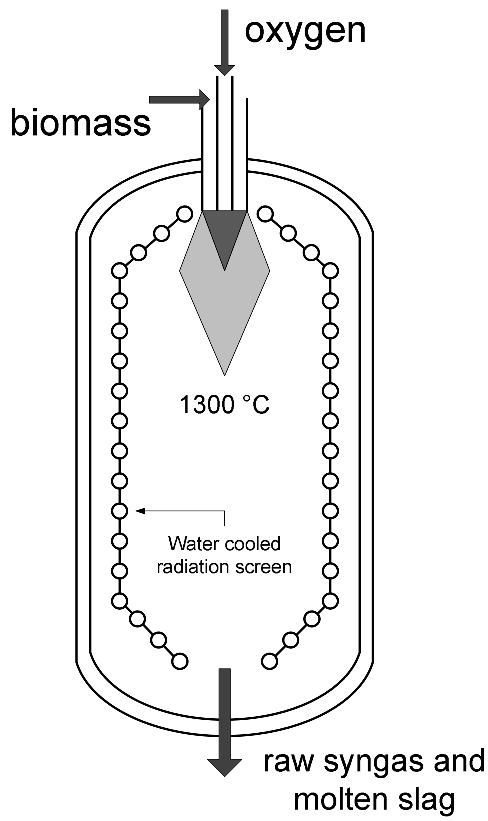


Figure 5. Entrained flow gasifier [17]

Entrained flow gasification mixes the fuel with a steam/oxygen stream and forms into a turbulent flow within the gasifier. Ash forming components melt in the gasifier and form a liquid slag on the inside wall of the gasifier effectively protecting the wall itself. The liquid flows down and is collected at the bottom. To form the slag, limestone can be added as a fluxing material. For herbaceous biomass, such as switchgrass or corn stover, which is high in alkali content, there may be sufficient inherent fluxing material present [17]. Advantages of entrained flow gasification are that tar and methane content are negligible and high carbon conversion occurs due to more complete gasification of the char. Syngas clean up is simplified because slag is removed at the bottom of the gasifier negating the need for cyclones and tar removal [18]. The disadvantages are that very high temperatures need to be maintained and the design and operation is more complex. An entrained flow gasifier co-firing up to 25% biomass with coal has been developed by Shell in Buggenum, Netherlands. Another gasifier developed by Future Energy in Freiburg, Germany uses waste oil and sludges. Both are operating at commercial scale [16].

2.3 Biomass Preprocessing

A degree of processing is required before gasification can occur. Most gasifiers require smaller size feedstock than is typically collected during harvest. Therefore, a significant degree of size reduction needs to be performed. A typical setup for size reduction is using a two-step process where a chipper accomplishes the primary reduction followed by a hammer mill for the secondary reduction [19]. In addition, a maximum moisture content for gasification is between 20-30% (wet basis) and normal operation is less than 15% (wet basis) [8]. Therefore, a drying process is required to prepare the feedstock for gasification.

The main benefit of drying biomass is to avoid using energy within the gasifier to heat and dry the feedstock [20]. Drier biomass makes for more stable temperature control within the gasifier. Rotary dryers typically operate utilizing hot flue gas from a downstream process as the drying medium. They have high capacity, but require high residence times. In addition, rotary dryers have a high fire hazard when using flue gas [20]. To avoid using flue gas, rotary dryers can use superheated steam, essentially an inert gas, when a combined cycle heat and power system is used downstream. That system has significant steam available for use because of the steam produced in the steam cycle. An advantage of using steam for drying is better heat transfer and therefore shorter residence time.

Pretreatment options for entrained flow gasification include torrefaction followed by grinding to 0.1 mm particles, grinding to 1 mm particles, pyrolysis to produce bio-oil/char slurry (bioslurry), or initial fluidized bed gasification of larger particles coupled to an entrained flow gasifier. Torrefaction, essentially an oxygen-free roasting process, causes the biomass particles to be brittle for easy grinding, but releases up to 15% of the energy in the biomass via volatile compounds [16]. The coupled option is attractive because of an overall energy efficiency of 80-85%, but is expensive due to the two gasifiers used in series. The bio-slurry option is illustrated in Figure 6. Basically, a flash pyrolysis process yields bio-oil and char followed by a slagging, entrained flow gasifier. Since this process utilizes an entrained flow gasifier, the feed must be pressurized. Fortunately, the pyrolysis slurry, already in an emulsified liquid state, can be pressurized easily. Technology for slurry feeding is state of the art due to experience with coal slurries [16]. The bioslurry still contains 90% of the energy contained in the original biomass [21]. Another advantage is that no inert gas is needed for solids pressurization, which would dilute the feed and therefore dilute the syngas. In the search for cost effective methods for production of syngas, this option has potential, but isn't as developed as technologies such as fluidized bed gasification. The biggest challenge is constructing and operating a large-scale pyrolysis process since large-scale systems have not been demonstrated [16].

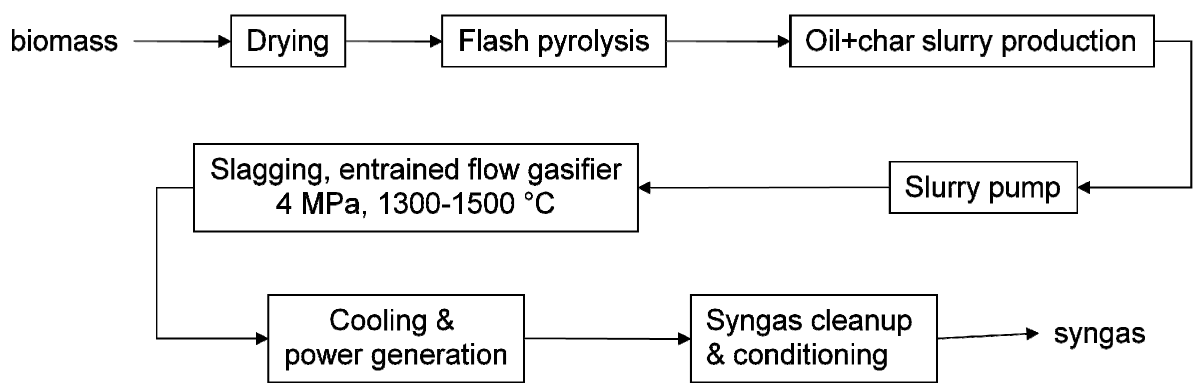


Figure 6. Schematic of a biomass pretreatment via fast pyrolysis followed by an entrained flow gasifier [16].

2.4 Syngas Cleaning

Since the raw syngas leaving the gasifier contains particulate, tars, alkali compounds, sulfur compounds, nitrogen compounds, and other contaminants, those components must be removed or reduced significantly. Particulate and tars have the potential for clogging downstream processes. Sulfur and nitrogen have the potential to poison downstream processes especially catalysts used in fuel synthesis applications. Moreover, another motivation to clean syngas is meeting environmental emissions limits.

Cooling of the syngas must occur before conventional gas clean up is to be utilized. This can happen two ways: direct quench by injection of water and indirect quench via a heat exchanger. Direct quench is less expensive, but dilutes the syngas. The direct quenching also can be used to clean up the gas by removing alkali species, particulate, and tars [22].

Particulate is defined as inorganic mineral material, ash, and unconverted biomass, or char [23]. In addition, bed material from the gasifier is included in the particulate. For feedstock such as switchgrass typically has 10% inorganic material in the form of minerals. Many gasifiers operate with a 98-99% carbon conversion efficiency where 1-2% of the solid carbon is in the form of char [23].

Removal of particulate primarily occurs through physical methods like cyclones where the heavy particles fall down the center while the gases rise up and out of the cyclone. The initial step for particulate removal is usually a cyclone. Important in particulate removal is that they should be removed before the gas is cooled down for cold gas cleaning. If removed after gas cooling, then tars can condense onto particulate and potentially plug equipment. Barrier filters, which operate above tar condensation temperatures use metal or ceramic screens or filters to remove particulate allow the gas to remain hot, but have presented problems in sintering and breaking [23].

Even more critical to downstream syngas applications is tar removal. Tars are defined as higher weight organics, oxygenated aromatics, heavier than benzene 78 and are produced from volatized material after polymerization [23]. A review by Milne et al. [24] of tars produced during gasification covers different removal methods. Physical removal via wet gas scrubbing of tars is accomplished by a scrubbing tower for the “heavy tars” followed by a venturi scrubber for lighter tars. This setup is similar to the direct quench cooling as mentioned previously since cooling occurs as well. Tar concentration is reported to be lower than 10 ppm by volume at the exit of this setup. The disadvantage of this setup is that waste water treatment is required and can be expensive. The other method for tar removal is catalytic or thermal conversion to non-condensable gas. This is also known as hot gas cleaning since it occurs at temperatures at or above gasification temperatures. Catalytic conversion can occur as low as 800 °C and thermal conversion occur up to 1200 C. The energy required for thermal tar cracking may not be cost competitive because of the temperature rise from the gasification temperature to crack the high refractory tars [23].

Alkali compounds such as calcium oxide and potassium oxide are present in biomass and when gasified either become vaporized or concentrated in the ash. Condensation of these compounds begins at 650°C and can deposit on cool surfaces causing equipment clogging, equipment corrosion, and catalyst deactivation [25]. According to Stevens [25], research on alkali adsorption filters using bauxite has been promising, but not demonstrated on a large scale. Stevens concludes that the best current method for alkali removal is using proven syngas cooling followed by wet scrubbing, where the addition of water cools the syngas and physically removes small particles and liquid droplets.

Wet scrubbing also removes ammonia which forms during gasification from the nitrogen in the biomass. Without proper removal, ammonia can deactivate catalysts as well. Complete ammonia removal can be accomplished through wet scrubbing [26]. For gasifiers coupled to a catalytic or thermal tar reformer, most of the ammonia can be reformed to hydrogen and nitrogen [26]. Sulfur in the biomass mostly forms into hydrogen sulfide (H_2S) with small amounts of carbonyl sulfide (COS). Hydrogen sulfide removal occurs by three main ways: chemical solvents, physical solvents, and catalytic sorbents. For chemical removal, amine-based solvents are typically utilized. Chemical removal occurs by the solvent chemically bonding with H_2S. Physical removal takes advantage of the high solubility of H_2S using an organic solvent. Typical setups of both chemical and physical removal involve an absorber unit followed by a solvent regenerator unit, known as a stripper. Operation usually occurs at temperatures lower than 100°C and medium to high pressures (150-500 psi) [26]. Sulfur leaving these two systems is around 1-4 ppm and can require further removal, especially for fuel synthesis. In that case, a syngas polishing step using a fixed bed zinc oxide activated carbon catalyst removes H_2S and COS to parts per billion levels necessary for fuel synthesis. Halides, present in trace amounts in the biomass, can also be removed with the zinc oxide catalyst [26].

2.5 End Use Product

After syngas has been cleaned from particulates, impurities, and contaminants there is sufficient energy content for producing a higher valued product. There are three main large-scale biomass gasification pathways that have been researched and suggested for higher valued product: power generation, liquid fuel synthesis, and chemical synthesis. According to Wender [27], the three largest commercial uses for syngas are ammonia production from hydrogen, methanol synthesis, and hydrocarbon synthesis via Fischer-Tropsch process.

2.5.1 Power Generation

Power generation using gasification occurs by combusting the syngas in a gas turbine to provide mechanical work for a generator. Steam is generated by recovering heat from the hot syngas and the steam in turn provides the means for mechanical work via a steam turbine. This gasifier plus gas and steam turbine setup is known as integrated gasification combined cycle (IGCC) power generation. The level of particulates, alkali metals, and tar can decrease the performance of the gas turbine. Consonni and Larson [28] found that particulate can cause turbine blade erosion and 99% of 10 micron size particles or less should be removed. In addition, they also report that alkali metals corrode the turbine blades and tars condense on the turbine blades both hindering operation and escalating turbine failure. Fortunately, nearly all alkali and tars can be removed using proven wet scrubbing techniques.

Using the IGCC approach to generate power, Bridgwater et al. [29] and Craig and Mann [22] expect biomass to power efficiencies in the range of 35-40% with large scale systems (greater than 100 MW net output) at the high end of the range. Moreover, Craig and Mann suggest that future advanced turbine systems could reach 50% biomass to power efficiency.

2.5.2 Synthetic Fuels and Chemicals

Instead of converting the energy content of the syngas to power, the energy content can be condensed into a liquid energy carrier, or fuel. The conversion of syngas to fuels can only occur in the presence of proper catalysts [30]. The catalytic reactions basically build up the small molecules in the syngas (i.e. carbon monoxide and hydrogen) into larger compounds that are more easily stored and transported. A summary of many catalytic pathways to fuels and chemicals is shown in Figure 7. In most catalytic synthesis reactions, syngas cleanliness requirements are very high. Most impurities and contaminants are removed to low parts per million and even parts per billion. This means that significant cost must be directed towards syngas cleaning.

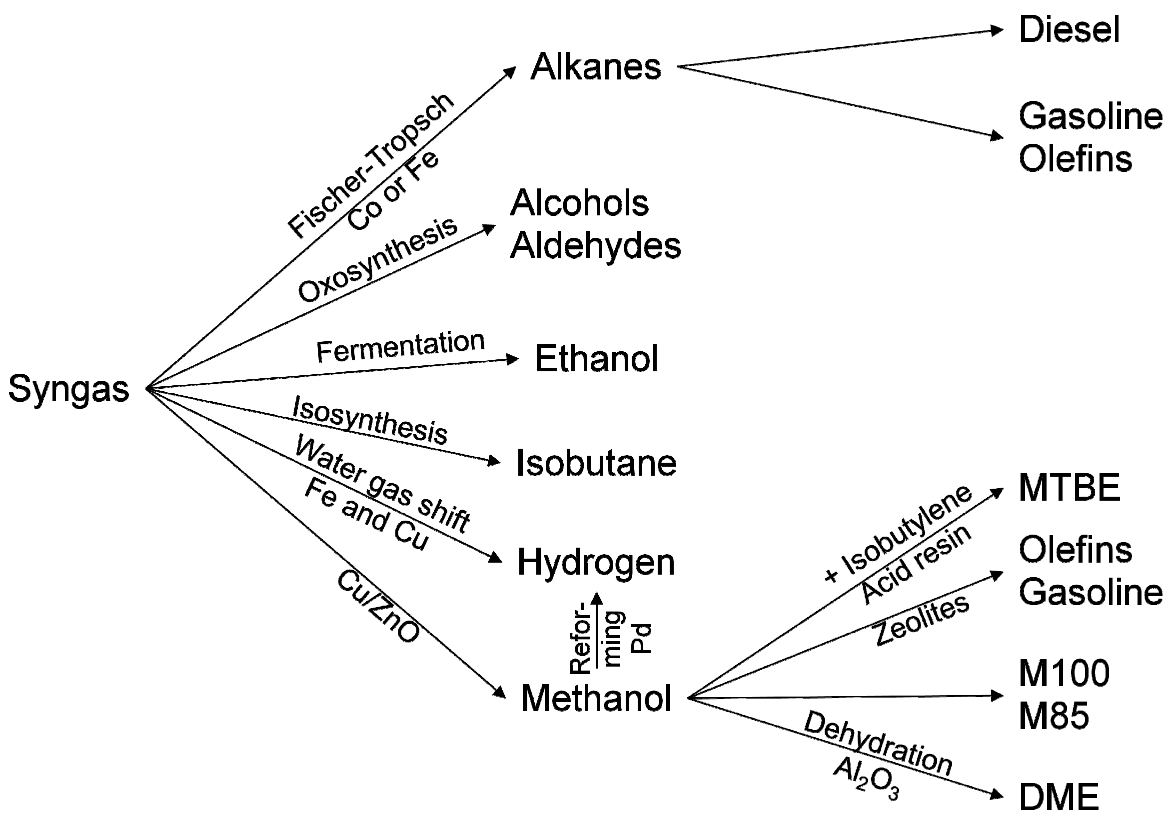


Figure 7. Main syngas conversion pathways [31]

2.5.2.1 Methanol to Gasoline

Methanol is one of the top chemicals produced in the world [31]. Most commercially produced methanol is synthesized via steam methane reforming and autothermal reforming. The synthesis of methanol from syngas is highly exothermic (equation 1). The reaction occurs over a $Cu/ZnO/Al_2O_3$ catalyst between temperatures of 220-275°C and pressures of 50-100 bar and have lifetime of 2-5 years [30]. Wender [27] reports syngas to methanol conversion efficiency can reach 99% with recycle, but per pass efficiency is about 25%.

(eqn. 1)

Although methanol can be used directly as a liquid fuel, it can also be converted into the conventional transportation fuel range. This process is known as the methanol to gasoline (MTG) process and was developed by the Mobil Oil Corporation [30]. In that process, methanol is heated to 300°C and dehydrated over alumina catalyst at 27 atm yielding methanol, dimethyl ether (DME), and water. The exiting mixture reacts with a zeolite catalyst at 350°C and 20 atm to produce 56% water and 44% hydrocarbons by weight. Of the hydrocarbon product, 85% is in the gasoline range and 40% of the gasoline range is aromatic. However, limitations on the aromatic content of gasoline have been proposed in legislation [30]. Thermal efficiency of methanol to gasoline range hydrocarbons is 70% [10]. The overall MTG process usually contains multiple MTG reactors in parallel in order to perform periodic catalyst regeneration by burning off coke deposits [10]. A commercial plant producing 14,500 barrels per day operated in New Zealand during the 1980s by Mobil [31]. The reaction process could stop directly after the methanol synthesis and focus on producing DME because it can be used as a diesel fuel as it has a high cetane number. It is formed from the dehydration reaction of methanol over an acid catalyst γ-alumina. Per pass can be as high at 50%. Overall syngas to DME is higher than syngas to methanol [30]. However, DME is in gaseous form at atmospheric conditions and needs to be pressurized for use in diesel engine [32]. Therefore, engine modification is required and is the main disadvantage for DME use as transportation fuel.

2.5.2.2 Fischer-Tropsch

Fischer-Tropsch catalytic synthesis is a highly exothermic reaction producing wide variety of alkanes (equation 2).

$$CO + 2.1H_2 \rightarrow \; -(CH_2)- \; + H_2O \qquad \text{(eqn. 2)}$$

For gasoline range products, higher temperatures (300-350°C) and iron catalysts are typically used. For diesel range and wax products, lower temperatures (200-240°C) and cobalt catalysts are typically used [33]. Operating pressures are in the range of 10-40 bar. Product distribution can be estimated using the Anderson-Schulz-Flory chain growth probability model where longer hydrocarbon chains form as the temperature decreases. At high temperatures, selectivity favors methane and light gases. This is a disadvantage if liquid fuel production is the focus. At low temperatures, selectivity favors long carbon chain wax products requiring further hydrocracking to the diesel range in a separate unit adding more construction cost, but necessary for liquid fuel production.

Because of the highly exothermic reaction, the heat must be removed or the catalyst can be deactivated. Two main types of reactors have been designed: a fixed bed tubular reactor and slurry phase reactor (Figure 8). Heat removal is crucial to the process and has been the focus of reactor design development [30]. The fixed bed reactor has many catalyst tubes where heat removal is achieved by steam generation on the outside of the tubes [34]. The fixed bed reactor is simple to operate and is well suited for wax production due to simple liquid/wax removal. However, it is more expensive to build because of the many tubes and has a high pressure drop across the reactor [35]. The slurry phase reactor (SPR) operates by suspending catalyst in a liquid and the syngas is bubbled through from the bottom. A disadvantage of a SPR is a more complex operation and difficult wax removal. However, the SPR requires approximately 40% less construction cost [35].

FT diesel is very low in sulfur, low in aromatic content, and has high cetane number, making it very attractive as conventional fuel alternative. Emissions across the board decrease when using FT diesel. A South African based company, Sasol, has been producing transportation fuel since 1955 using the FT process and supplies 41% of South Africa's transportation fuel requirements [30].

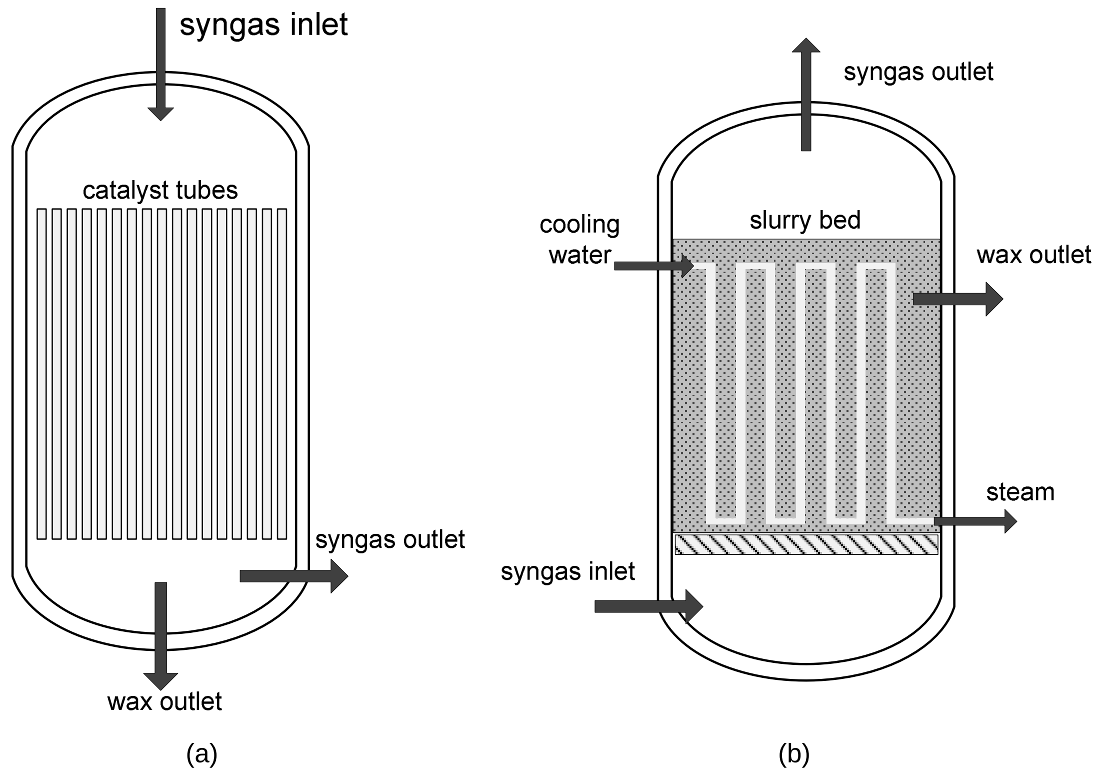


Figure 8. Fischer-Tropsch reactor types (a) Multi-tubular fixed bed and (b) Slurry bed[30]

2.6 Techno-economic Analysis

In order for biofuels technologies to be utilized in commercial applications, the economic feasibility must be determined. A feasibility analysis is also called a techno-economic analysis where the technical aspects of a project are coupled to the economic aspects. First, the basic theoretical configuration is developed and a mass and energy balance is performed. Second, cost estimation allows the investment and production cost of a biorefinery to be determined. With rising interest in biorenewable resources, many techno-economic studies have been performed on power generation and biofuel scenarios. These studies assist in understanding how the physical process relates to cost of producing renewable alternatives. Accuracy of results from these studies is usually ±30% of the actual cost [4].

2.6.1 Economics of Biomass Power

A study by Bridgwater in 1994 [36] demonstrated that an IGCC power generation plant using biomass at 100 MW electric output could produce power for 6 ¢ per kWh and would require $2000 per kW (i.e. $200 million total) in capital investment. That study also compared between various power generation pathways showing that an IGCC could produce power for less compared to combustion and gas engine scenarios. Another study by Craig and Mann [22] using 1990$ compares varying IGCC scenarios with power output between 56-132 MW. Capital investment for these scenarios range between $1100 to 1700 per kW and production cost of power range between 6.5 and 8.2 ¢ per kWh. A study by Larson et al. [37] increases the power generation to 440 MW and shows that the increased size benefits from economies of scale. Capital investment is $1000 per kW and production cost of power is just above 5¢ per kWh.

2.6.2 Economics of Biofuels

Previous studies of gasification based, biomass-to-liquid production plants have estimated the cost of transportation fuels to range from $12-16/GJ ($1.60-2.00 per gallon of gasoline equivalent) [15,38-41]. The same studies have estimated total capital investment in the range of $191 million for 2000 dry metric ton per day input [40] to $541 million for 4500 dry metric ton per day input [39].

A 1650 dry metric ton per day biomass to methanol plant based on gasification, production cost of $15/GJ ($0.90 per gallon of methanol) is reported by Williams et al. [15] in 1991$ for $45 per dry metric ton of biomass. Williams et al. also shows production cost of methanol derived natural gas to be $10/GJ ($0.60 per gallon of methanol). However, that study concludes that if a carbon tax system was developed for lifecycle carbon emissions, then renewable methanol could become competitive to natural gas derived methanol at a tax of approximately $90 per metric ton of carbon. A more recent study by Larson et al. of switchgrass to hydrocarbons production in 2009 reports a production cost of $15.3/GJ ($1.90 per gallon of gasoline) in 2003$ for a 4540 dry metric ton per day (5000 dry short ton per day) plant based on gasification [39].

Table 2 shows a comparison between four biofuel production studies based on gasification. A range of cost year, plant size, and feedstock cost show the diversity of characteristics and assumptions that techno-economic studies use. In addition, resulting capital investment costs of the studies have a large range. For example, the capital investment of the Phillips et al. and Tijmensen et al. studies are $191 million and $387 million, respectively, at similar plant sizes. Reasons for such a significant difference are choice of technologies and level of technology development. The Phillips et al. study is a target study meaning that it estimates future technology improvement and results in lower costs. Direct comparison is difficult because of the varying assumptions used by each study.

Table 2. Previous techno-economic studies of biofuel production plants

	Williams et al. [15]	**Phillips et al. [40]**	**Tijmensen et al. [41]**	**Larson et al. [39]**
Cost Year	1991	2005	2000	2003
Plant Size (dry metric tonne per day)	1650	2000	1741	4540
Feedstock	generic biomass	poplar	poplar	switchgrass
Fuel Output	methanol	ethanol	FT liquids	diesel, gasoline
Feedstock Cost ($/dry short ton)	41	35	33	46
Capital Investment ($MM)	N/A	191	387	541
Product Value ($/GJ)	15	12	16	15
Product Value ($/GGE)	1.90	1.60	2.00	1.85

3. METHODOLOGY

The following steps are undertaken to perform the analysis in this study:

- Collect performance information on relevant technologies for systems under evaluation.
- Perform down selection process with developed criteria to identify most appropriate scenarios
- Design process models using Aspen PLUS™ process engineering software
- Size and cost equipment using Aspen Icarus Process Evaluator®, literature references, and experimental data
- Determine capital investments and perform discounted cash flow analysis
- Perform sensitivity analysis on process and economic parameters
- Perform pioneer plant cost growth and performance analysis

3.1 Down Selection Process

A number of process configurations for the gasification-based, biomass to liquids (BTL) route are initially considered as listed in Table 3 and discussed in the following sections.

Table 3. Process configurations considered in down selection process

Gasifier block	Entrained flow, slagging gasifier
	Fluid bed, dry ash gasifier
	Transport gasifier, dry ash (e.g. Kellog, Brown, and Root)
	Indirect gasifier, dry ash (e.g. Battelle-Columbus Labs)
Syngas cleaning	Water scrubbing
	Catalytic tar conversion/reduction
	Thermal tar conversion/reduction
	Amine-based acid gas removal
	Physical sorbent-based acid gas removal (e.g. Sorbitol, Rectisol)
Fuel synthesis	Fischer-Tropsch
	Mixed alcohols
	Methanol to gasoline (MTG)
	Dimethyl ether
	Syngas fermentation

3.1.1 Preliminary Criteria

The initial technology configuration options are reviewed and screened in accordance with the following criteria. The technology under consideration should be commercially ready in the next 5-8 years and preferably with high technology development. High technology development increases the likelihood of a configuration to perform at the scale in this study. For example, coal gasification has been demonstrated commercially at large-scales [10]. While similar scale biomass gasifiers have not been proven commercially, the technology development on coal is assumed to apply for biomass in 5-8 years. Secondly, the size of biorefinery should be feasible with typical agricultural productivity and within a realistic collection area. For example, if one third of total land use surrounding the biorefinery is for stover collection and each acre provides conservatively one short dry ton per year, then the required collection radius is 35 miles and amount of biomass transported to the biofinery is approximately 2300 short tons (2090 metric tons) per day. The collection area with a 35 mile radius is assumed to be realistic. In addition, previous studies by Tijmensen et al., Phillips et al., and Lau et al. have used a similar plant sizes [40-42]. Thirdly, the desired product should be compatible with the present transportation fuel infrastructure, i.e. gasoline and diesel range hydrocarbons.

3.1.2 Scenarios selection

For the gasification area, two gasifiers were selected for modeling. First, an entrained flow, slagging gasifier is chosen due to its commercial application with coal (GE, Siemens, Shell, and ConocoPhillips) and its potential for use with biomass. Moreover, process modeling of this gasifier is simple since it can be closely approximated at thermodynamic equilibrium [1]. Second, a fluidized bed, dry ash gasifier is chosen due to experience at Gas Technology Institute and because of data availability. A report by Bain [2] at the National Renewable Energy Laboratory contains collected and analyzed data for fluidized bed gasification. In addition, Iowa State University is currently operating an atmospheric pressure, fluidized bed gasifier as either air or oxygen/steam fed.

The syngas cleaning area is chosen to include configurations that have less technological complexity than previous studies. Phillips et al. [40] and Larson et al. [39] both employ an external catalytic tar reforming process for dry-ash gasification. Because of low technological development in tar conversion and its inherent complexity, a direct-contact syngas quenching and scrubbing are chosen for this study. In the case of the slagging gasifier, high temperatures inhibit tar formation, yet still require quenching and particulate and ammonia removal. An amine-based, chemical absorber/stripper configuration is chosen for removal of hydrogen sulfide and carbon dioxide. This configuration is chosen due to data availability as compared to proprietary physical gas cleaning process such as Rectisol® and Selexol®.

Two fuel synthesis configurations under consideration produce liquid hydrocarbons: Fischer-Tropsch (FT) synthesis and MTG. FT synthesis has been proven in operation at commercial scale for many years by Sasol [10]. Due to more accessible data and long industrial experience, FT synthesis is the only fuel synthesis option chosen. A consequence of this selection is a post-synthesis fuel upgrading area since FT products need to be separated and hydroprocessed.

3.1.3 Scenarios not selected

The indirect, dry-ash gasifier and the mixed alcohol synthesis configurations is not considered due to previous work by Phillips et al. [40] The transport gasifier design, though a promising technology, is not considered due to reactor complexity, unproven commercial-scale operation and lack of public domain data. Tar conversion via external thermal or catalytic cracking is not considered due to lack of public domain data and commercial scale experience. Acid gas removal using proprietary technology (e.g. Rectisol™ or Selexol™) is not considered because of a lack of public operational data. MTG, including methanol synthesis, is not considered because of time constraints and limited operational data. DME and syngas fermentation is not considered due to the limited commercial scale experience and because of incompatibility with present fuel infrastructure.

3.1.4 Project Assumptions

Main project assumptions for process and economic analysis are listed in Table 4. A more extensive list can be found in Appendix A.

Table 4. Main assumptions used in nth plant scenarios

Main assumptions
The plant is modeled as n^{th} plant
Plant capacity is 2000 dry metric ton/day
Feedstock is corn stover at 25% moisture
Feedstock ash content at 6%
Feedstock is purchased at plant gate for $75/dry short ton
All financial values are adjusted to 2007 cost year
Plant is 100% equity financed
Fuel PV is evaluated at 10% internal rate of return
Plant initiates operation in 5-8 year time frame
Plant life is 20 years
Plant availability is 310 days per year (85%)

3.2 Process Description

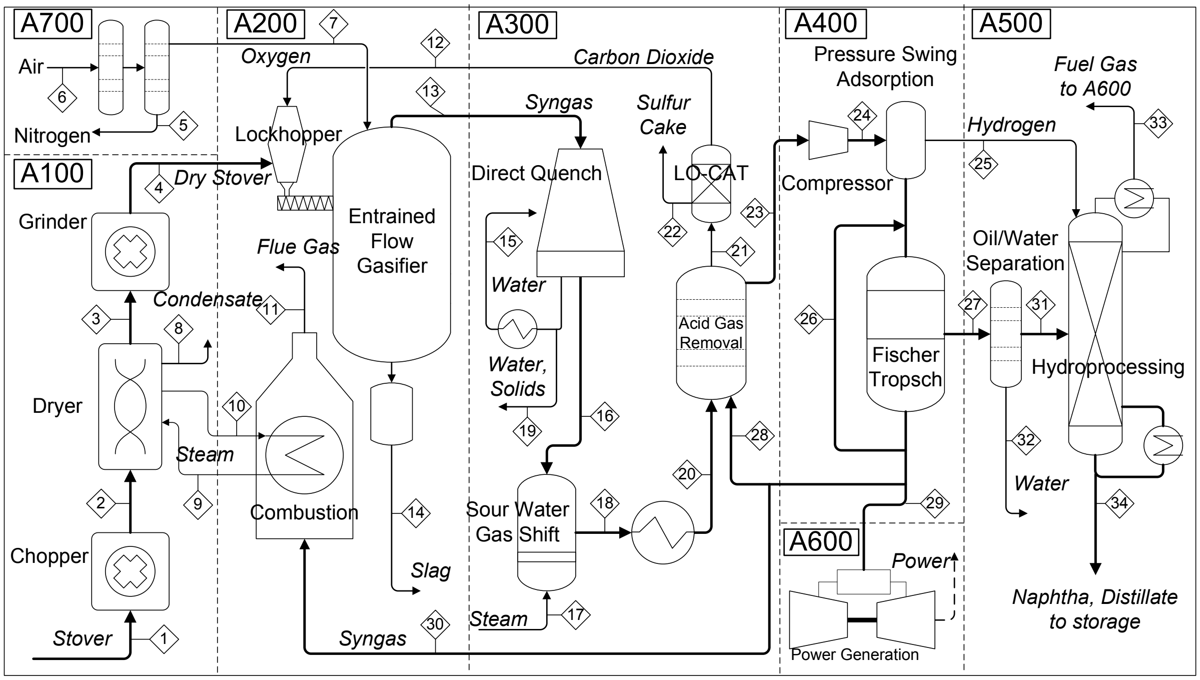
3.2.1 High Temperature Scenario Overview

The high temperature scenario is a 2000 dry metric ton (2205 dry short ton) per day corn stover-fed gasification biorefinery that produces naphtha and distillate to be used as blendstock as well as electricity for export. It is based on pressurized, oxygen blown, entrained flow gasification. The HT scenario is an n^{th} plant design meaning significant design, engineering, and operating experience has been achieved.

The main areas of operation as shown in Figure 9 include feedstock preprocessing (Area 100) where the stover is chopped, dried, and ground to 1-mm, 10% moisture. Gasification (Area 200) contains the stover pressurization for solids feeding, gasification, and slag removal. Synthesis gas cleaning (Area 300) contains cold gas cleaning technologies where the syngas is quenched and scrubbed from particulate, ammonia, hydrogen sulfide, and carbon dioxide. Area 300 also contains the water-gas-shift reaction which occurs before the hydrogen sulfide and carbon dioxide removal in order to adjust the ratio of hydrogen to carbon monoxide for optimal fuel synthesis. Fuel synthesis section (Area 400) contains syngas boost pressurization, contaminant polishing via zinc oxide guard beds, Fischer-Tropsch reactor, and hydrocarbon gas/liquid separation. Hydroprocessing (Area 500) produces the final fuel blend and is treated as a black box utilizing published data. Power generation (Area 600) contains gas and steam turbines along with a heat recovery steam generator. Area 700 contains the Air Separation Unit (ASU) where oxygen is separated from air and pressurized for use in the gasifier. For cost analysis uses only, a balance of plant (BOP) area accounts for cooling tower area, cooling water system, waste solids and liquids handling area, and feed water system. Detailed process flow diagrams can be found in Appendix E and detailed stream data can be found in Appendix F.

Recycle streams are utilized to provide better syngas to FT products conversion. Unconverted syngas in the fuel synthesis area is recycled to the syngas cleaning area to remove carbon dioxide and allows for further conversion in the Fischer-Tropsch reactor. A small portion of unconverted syngas is sent to a steam boiler to raise steam required for drying the biomass. The balance of unconverted syngas is combusted in a gas turbine and waste heat is recovered in a steam generator for steam turbine power. Power generated is used throughout the plant and excess is sold.

Some of the largest consumers of power are the ASU and hydroprocessing area at 11.6 MW and 2.2 MW, respectively. Another consumer of power is the hammermill for grinding the dried biomass in Area 100 requiring 3.0 MW. The amine/water solution recirculation pump in Area 300 requires approximately 0.9 MW. Syngas compressors throughout the plant require significant amount of power as well. Gross plant power production is 48.6 MW and net electricity for export is 13.8 MW.



Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
T (℃)	25	25	90	90	16	32	149	120	200	120	220	249	1300	50	203	203	190
P (bar)	1.01	1.01	1.01	1.01	1.1	1.01	28	1.98	1.98	1.98	1.03	28	26.62	26.62	26.62	25.93	10
m (Mg/day)	2667	2667	2222	2222	2189	2903	744	444	4000	4000	271	180	3823	114	4000	3867	550
Stream	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34
T (℃)	240	40	40	50	50	62	76	30	200	42	45	45	45	35	35	35	37
P (bar)	24.8	24.82	24.82	3.45	1.93	22.75	26	1.01	24.96	23.6	23.58	23.58	23.58	22.2	22.2	22.2	1.03
m (Mg/day)	4417	1502	2965	1755	3.1	3570	3570	4.38	1404	1068	2308	156	33	427	641	52	378

Figure 9. Overall process flow diagram for HT scenario (parallelograms enclosing numbers in the diagram designate individual process streams, which are detailed in the accompanying table).

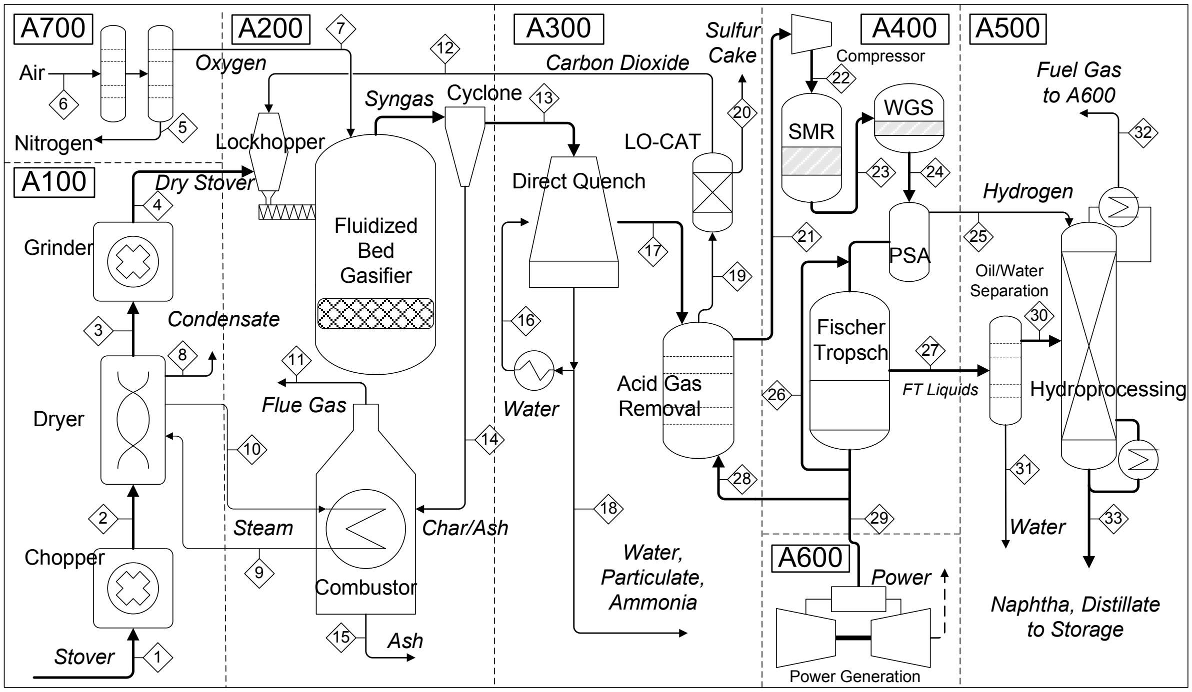
3.2.2 Low Temperature Scenario Overview

The low temperature scenario is a 2000 dry metric ton (2205 dry short ton) per day corn stover-fed gasification biorefinery that produces naphtha and distillate to be used as blendstock as well as electricity for export. It is based on a pressurized, oxygen/steam blown fluidized bed gasifier developed by Gas Technology Institute. The HT scenario is an nth plant design meaning significant design, engineering, and operating experience has been achieved.

The main areas of operation as shown in Figure 10 include feedstock preprocessing (Area 100) where the stover is chopped, dried, and ground to 6-mm, 10% moisture. Gasification (Area 200) contains the stover pressurization for solids feeding, gasification, and char and ash removal. Synthesis gas cleaning (Area 300) contains cold gas cleaning technologies where the syngas is quenched and scrubbed from particulate, ammonia, hydrogen sulfide, and carbon dioxide. Fuel synthesis section (Area 400) contains syngas boost pressurization, contaminant polishing via zinc oxide beds, Fischer-Tropsch reactor, and hydrocarbon gas/liquid separation. Also included within area 400 is the steam methane reformer (SMR) to reduce methane content and water-gas-shift (WGS) to adjust ratio of hydrogen and carbon monoxide. Hydroprocessing (Area 500) produces the final fuel blend and is treated as a black box utilizing published data. Power generation (Area 600) contains gas and steam turbines along with a heat recovery steam generator. Area 700 contains the Air Separation Unit (ASU) whereby oxygen is separated from air and pressurized for use in the gasifier. Detailed process flow diagrams can be found in Appendix E and detailed stream data can be found in Appendix F.

Recycle streams are utilized to provide better FT products conversion. Unconverted syngas in the fuel synthesis area is recycled to the syngas cleaning area to remove carbon dioxide and allows for further conversion in the Fischer-Tropsch reactor. The balance of unconverted syngas is combusted in a gas turbine and waste heat is recovered in a steam generator for steam turbine power. Power generated is used throughout the plant and excess is sold. Unconverted carbon within the gasifier in the form of char is collected and combusted in a furnace to produce heat thereby generating steam for the drying of the biomass.

Some of the largest consumers of power are the ASU and hydroprocessing area at 9.1 MW and 1.7 MW, respectively. Another consumer of power is the hammermill for grinding the dried biomass in Area 100 requiring 1.1 MW. The amine/water solution recirculation pump in Area 300 requires approximately 0.7 MW. Syngas compressors throughout the plant require a significant amount of power as well. Gross plant power production is 40.7 MW and net electricity for export is 16.3 MW.



Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
T (℃)	25	25	90	90	16	32	149	120	200	120	200	243	870	870	1500	40	40
P (bar)	1.01	1.01	1.01	1.01	1.10	1.01	28.00	1.98	1.98	1.98	1.00	28.00	27.55	27.55	1.00	26.17	26.17
m (Mg/day)	2667	2667	2222	2222	1744	2313	569	444	4000	4000	1471	180	2930	215	119	6000	2198
Stream	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	
T (℃)	40	50	50	62	78	870	304	60	200	32	32	32	35	35	50	50	
P (bar)	26.17	3.45	2.10	22.89	26.50	25.81	25.12	1.01	25.00	22.89	22.89	22.89	22.89	22.89	22.89	22.89	
m (Mg/day)	1380	1593	2.3	2709	2709	3707	3708	3.4	560	1463	2073	168	331	1132	41	293	

Figure 10. Overall process flow diagram for LT scenario (parallelograms enclosing numbers in the diagram designate individual process streams, which are detailed in the accompanying table).

3.2.3 Area 100 Preprocessing

The preprocessing area contains all the unit operations required for preparing the biomass for feeding into the gasifier. Biomass enters the plant gate at 25 wt% moisture on wet basis in bales. The corn stover composition is shown below in Table 5. Ash content is assumed to be 6% by weight. Char composition, formed in the gasifier, is also shown in Table 5. Forklifts transport the bales to conveyors where the stover is separated from any metal in a magnetic separator. The first modeled operational area is a primary biomass chopper to complete the initial size reduction step and prepare stover for drying.

Table 5. Stover and char elemental composition (wt%)

Element	Stover	Char
Ash	6.00	0
Carbon	47.28	68.05
Hydrogen	5.06	3.16
Nitrogen	0.80	0.29
Chlorine	0	0
Sulfur	0.22	0.15
Oxygen	40.63	28.34

The next area of operation is the direct contact steam drying which is modeled as a rotary steam dryer with exiting biomass moisture of 10% on wet basis. For steam dryers Amos [20] suggests 9:1 steam to evaporated moisture ratio. Therefore, 4000 metric tons per day steam is utilized in a loop and heated to 200°C from the hot combustion flue gases exiting the syngas or char fired combustor in Area 200. Steam mixes with 25°C biomass and enters the drier. At the exit, steam at 120°C returns to the combustor for reheating and dried biomass exits at 90°C and is conveyed to the grinding area.

The grinding area is the same configuration as the chopping area except the grinder requires significantly more power due to the larger size reduction. The grinder reduces the size of the biomass to 1-mm and 6-mm for the HT and LT scenarios, respectively. The power requirement of the grinder for the HT and LT scenarios are 3000 kW and 1100 kW, respectively. Energy requirements for grinding are determined using the correlations for specific energy (kWh per short ton) which has been adapted from Mani et al.[43]

3.2.4 Area 200 Gasification

The gasification area of the plant produces synthesis gas using pressurized gasifiers. Also in this area slag, char, and ash are removed. This area also includes lock hoppers for biomass pressurization and a fired combustor which provides heat to raise steam for drying the stover.

Dried and ground stover enters the area and is immediately conveyed to a lock hopper system for pressurized feeding. Carbon dioxide is used as pressurization gas and arrives from the syngas cleaning area. According to Lau et al. [42] a lock hopper system is the best setup for pressurized feeding of solids, despite higher operating costs due to high inert gas usage. A proven track record with biomass is the main reason for their recommendation. The power requirement of a lock hopper system using biomass is 0.082 kW/metric ton/day resulting in a 180 kW system. Higman and van der Burgt [44] report inert gas usage as 0.09 kg/kg for 25 bar applications. This results in a 180 MT/day carbon dioxide addition into the hopper. It is assumed that only 5% of the inert gas leaks into the gasifier while the rest is vented by the lock hopper.

Pressurized biomass is then conveyed into the gasifier. Oxygen at 95% purity is produced from the Air Separation Unit. A fixed 0.35 mass ratio of oxygen to biomass is used for the entrained flow gasifier as reported by Henrich [17]. Steam addition to the gasifier is set at 0.48 mass ratio of steam to biomass in accordance with Probstein and Hicks [10] and explained further in appendix C.5. This gasifier operates at a temperature of 1300°C meaning that equilibrium can be modeled according to Frey and Akunuri [1]. The reactions shown in equations 3-9 are modeled using equilibrium constants.

$$C + 2H_2 \leftrightarrow CH_4 \qquad \text{(eqn. 3)}$$

$$2C + 1.5O_2 \leftrightarrow CO + CO_2 \qquad \text{(eqn. 4)}$$

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad \text{(eqn. 5)}$$

$$2CO + O_2 \leftrightarrow 2CO_2 \qquad \text{(eqn. 6)}$$

$$S + H_2 \leftrightarrow H_2S \qquad \text{(eqn. 7)}$$

$$0.5N_2 + 1.5H_2 \leftrightarrow NH_3 \qquad \text{(eqn. 8)}$$

$$CO + H_2S \leftrightarrow COS + H_2 \qquad \text{(eqn. 9)}$$

The LT scenario gasifier uses a 0.26 mass ratio of oxygen to biomass at a gasification temperature of 870°C. This ratio is developed from the data found in an IGT gasifier study by Bain [2]. In that study, Bain develops mass balances for an IGT gasifier operating with woody biomass. Steam addition to the gasifier is calculated using a 40/60 steam to oxygen mass ratio consistent with experiments performed at Iowa State University using corn stover feedstock and a steam/oxygen blown, fluidized bed gasifier. Low temperature gasification cannot be modeled at equilibrium with or without approach temperatures for reactions. Instead an elemental mass balance calculation and adjustment is performed to ensure all inlet and outlet streams are accounted for across the gasifier. For details on the LT gasifier mass balance calculation see appendix C.5.

Yield from each gasifier is different. As Table 6 shows, hydrocarbons and tars are not produced in the high temperature gasifier because of near equilibrium conditions. Also, more hydrogen formation occurs in the high temperature gasifier caused by the water-gas-shift reaction (equation 5) and since thermodynamically nearly no methane, ethane, and ethylene are produced. The low temperature gasifier, on the other hand, produces a significant amount of methane, ethane, and ethylene in the syngas requiring downstream reforming. Slag in the HT scenario is formed from the ash when the ash melts and flows on the inside walls, collected at the bottom and removed for storage and subsequent waste removal. In accordance with Frey and Akunuri [1], it is assumed that 95% of the ash in the stover becomes slag while the rest becomes fly ash.

Table 6. Syngas composition (mole basis) leaving gasifier for gasification scenarios evaluated

Component	High temperature (mole fraction)	Low temperature (mole fraction)
Carbon Monoxide	0.264	0.240
Hydrogen	0.310	0.200
Carbon Dioxide	0.137	0.274
Water	0.280	0.194
Nitrogen	0.002	0
Methane	6 ppm	0.055
Ethane	0	6100 ppm
Ethylene	0	0.013
Ammonia	31 ppm	9400 ppm
Hydrogen Sulfide	672 ppm	1120 ppm
Carbonyl sulfide	26 ppm	0
Tar (Anthracene)	0	500 ppm
Oxygen	0	0
Argon	0.006	0.006

Directly after the low temperature gasifier initial syngas cleaning occurs whereby cyclones capture char and ash. The cyclones are split into two trains because of high volumetric gas flow. Each train contains a medium efficiency followed by high efficiency cyclones particulate capture. Overall particulate removal efficiency for cyclone area is 99%. Nearly particulate-free syngas travels to the more rigorous syngas cleaning area. Captured char in the LT scenario is collected and combusted in a fluidized bed combustor providing energy for heating low pressure steam used for drying the stover. Syngas produced in the HT scenario contains fly ash which is subsequently removed in a direct water quench unit. The combustion area in the HT scenario receives unconverted syngas from the fuel synthesis area, since char is not produced. For both scenarios the combustor is assumed to operate adiabatically resulting in an exit flue gas temperature of approximately 1800 °C. Hot flue gas heats 120°C steam to 200°C and loops to the stover drying area.

3.2.5 Area 300 Syngas Cleaning

After the initial particulate removal accomplished by the cyclones, the syngas still contains some particulate and all of the ammonia, hydrogen sulfide, and other contaminants. Area 300 contains the removal of these species using a cold gas cleaning approach, which is presently proven in many commercial configurations. Hydrogen sulfide and carbon dioxide, collectively known as acid gas, is absorbed via amine scrubbing. Separation of carbon dioxide from hydrogen sulfide with subsequent recovery of solid sulfur occurs via the LO-CAT® hydrogen sulfide oxidation process. In addition, the HT scenario contains a sour water-gas-shift process (sour because of the presence of sulfur), whereas the LT scenario situates the water-gas-shift directly upstream from the Fischer-Tropsch reactor.

Due to less than optimal hydrogen to carbon monoxide ratio from the gasifier, a water-gas-shift (WGS) reaction is necessary at some point in the process to adjust to optimum Fischer-Tropsch ratio of 2.1. Therefore, a significant WGS activity is required meaning a sizable amount of carbon dioxide is produced. To keep that carbon dioxide from building up in downstream processes, the sour water-gas-shift (SWGS) reactor is located before the acid gas removal area. This SWGS unit operation is the most significant difference between the HT and LT scenarios in this area.

In the HT scenario, the syngas arriving from the gasifier is cooled by direct contact water quench to the operating temperature of the SWGS unit. In addition to cooling, the direct water quench removes all of the fly ash, sludge, and black water in order to prevent downstream plugging. At this point a portion of the syngas is diverted to the SWGS unit which is modeled at equilibrium conditions and has an exit gas temperature of 300°C. A ratio of 3:1 water to carbon monoxide is reached by addition of steam to the SWGS reactor.

After the syngas is combined, the gas is further cooled to prepare for the acid gas removal. In the LT scenario, the direct quench unit condenses the syngas removing approximately 90% of ammonia and 99% of solids. Tar is condensed in this unit and can be recycled back into the gasifier using a slurry pump, but this configuration is not modeled. A water treatment facility for the direct quench effluent is not modeled, but is accounted for in a balance of plant (BOP) cost.

The next step for cleanup is the removal of acid gas (carbon dioxide and hydrogen sulfide) through the use of an amine-based solvent in a chemical gas absorption system. At this point in the cleaning process, hydrogen sulfide and carbon dioxide content is approximately 900 ppm and 30% on molar basis, respectively. Sulfur must be removed to at least 0.2 ppm for Fischer-Tropsch synthesis [30]. According to the GPSA Engineering Data book [45], amine-based systems are capable of removing sulfur down to 4 ppm. Therefore, a zinc oxide guard bed is required to remove the difference. In this study, 20% concentrated monoethanolamine (MEA), capable of absorbing 0.4 mol acid gas per mole amine, is used as the absorbent. The process setup is based on report by Nexant Inc.[26] Hydrogen sulfide leaves the top of the absorber at 4 ppm and CO2 at 2%, which is 99% and 90% removal, respectively. The clean syngas is now ready for polishing to final cleanliness requirements. A stripper is utilized to desorb the acid gas and regenerate the amine solution. Before the acid gas and amine solution enter the stripper a heat exchanger raises the temperature to 90°C.

Acid gas is brought to the LO-CAT sulfur recovery system to isolate hydrogen sulfide and convert it to solid sulfur. The LO-CAT system sold and owned by Gas Technology Products uses oxygen and a liquid solution of ferric iron to oxidize hydrogen sulfide to elemental solid sulfur [46]. This system is suitable for a range of 150 lbs to 20 ton per day sulfur recovery and also 100 ppm to 10% H_2S concentration in sour gas as reported by Nexant Inc.[26] The sulfur production in this model is approximately 3 metric ton per day and H_2S concentration approximately 150 ppm which is within the reported ranges. First, the H_2S is absorbed/oxidized forming solid sulfur and water while the ferric iron converts to ferrous iron. The second vessel oxidizes the ferrous iron back to ferric iron and the sulfur cake is removed while the iron solution is recycled back into the absorber [47]. The carbon dioxide gas stream from the absorber is split where a portion is compressed and used in biomass pressurization while the rest is vented to the atmosphere.

3.2.6 Area 400 Fuel Synthesis

Conversion from syngas to liquid fuel occurs in the Area 400 Fuel Synthesis area. The major operations in this area are zinc oxide/activated carbon gas polishing, steam methane reforming (only in the LT scenario), water-gas-shift (only in the LT scenario), Fischer-Tropsch (FT) synthesis, hydrogen separation via pressure swing absorption (PSA), FT products separation and unconverted syngas distribution. Another major difference between the LT and HT scenarios is in this area. Area 400 in the LT scenario contains the water-gas-shift reaction and steam methane reformer since recycle streams contain high enough content of methane and ethylene to significantly accumulate and cause dilution.

A compressor is the first operation in Area 400 boosting the pressure to 25 bar for FT synthesis. Then the syngas is heated to 200°C and passes through zinc oxide/activated carbon fixed bed sorbent. This polishing guard bed acts as a barrier to any upstream non-normal contaminant concentrations as well as sulfur removal down to synthesis requirements. To limit downstream catalyst poisoning, the syngas steam must be cleaned of these components. Removal to 50 ppb sulfur is possible with zinc oxide sorbent [26]. To comply with reported requirements the sorbent removes sulfur to approximately 200 ppb. In addition to sulfur, halides are removed by the sorbent. Syngas contaminant level requirements for Fischer-Tropsch synthesis are shown in Table 7.

Table 7. Fischer-Tropsch gas cleanliness requirements[30]

Contaminant	Tolerance Level
Sulfur	0.2 ppm (200 ppb)
Ammonia	10 ppm
HCN	10 ppb
Halides	10 ppb

Methane, nitrogen and carbon dioxide act as inerts in the FT synthesis. At this point in the LT scenario, a steam methane reforming (SMR) step is utilized. Syngas is heated to 870°C through a fired heater and passed through a reformer nickel-based catalyst to reduce methane, ethylene, and ethane content. It is assumed that the SMR can be modeled to operate at equilibrium. Steam is added to bring the steam to methane ratio to approximately

6.0 which at 870°C and 26 bar results in about 1.5% equilibrium methane content in exit stream [48]. For the HT scenario, the SMR step is not necessary. The WGS reaction is now employed for the LT scenario to increase the H_2:CO ratio. A portion of the gas is diverted through the fixed catalyst bed while the rest bypasses the reactor similarly to the SWGS unit in the HT scenario.

The exiting H_2/CO ratio after WGS is slightly above 2.1 in order for the excess hydrogen to be separated and used in the hydroprocessing area. A pressure swing adsorption (PSA) process is employed to isolate a stream of hydrogen. Since only a small amount of hydrogen needs to be separated from the syngas stream for downstream use, a small percentage of the syngas is directed to the PSA unit. Hydrogen removal efficiency within the PSA unit is assumed to be 85% and produces pure hydrogen [42]. After the PSA, the syngas rejoins the main gas line and enters the FT reactor.

The Fischer-Tropsch synthesis reactor operates at 200°C and 25 bar using a cobalt catalyst according to equation 10. Per pass carbon monoxide conversion in the reactor is set at 40%. The product distribution follows the Anderson-Schulz-Flory alpha distribution where chain growth factor, α, depends on partial pressures of H_2 and CO and the temperature of the reactor reported by Song et al. [49] for cobalt catalyst and shown in equation 11 where y is the molar fraction of carbon monoxide or hydrogen and $Temp$ is the reactor operating temperature in kelvin. The reactor is based on a fixed bed type reactor and that choice is reflected by the low per pass CO conversion.

$$CO + 2.1H_2 \rightarrow \ -(CH_2)- \ + H_2O \qquad \text{(eqn. 10)}$$

$$\alpha = \left[0.2332 \cdot \frac{y_{CO}}{y_{CO} + y_{H2}} + 0.6330\right] \cdot [1 - 0.0039(Temp - 533)] \qquad \text{(eqn. 11)}$$

To ensure the hydrocarbon product distribution to lean towards the production of diesel fuel the value of alpha should be at least 0.85 and preferably greater than 0.9 as shown in Figure 11. Reactor operating temperature to achieve chain growth value of 0.9 is approximately 200°C. This produces 30 wt% wax in the FT products requiring hydrocracking before addition to final fuel blend. All exiting effluent is cooled to 35°C and the liquid water and hydrocarbons are separated in a gas/liquid knock-out separator. Unconverted syngas is split into four streams: direct recycle to FT reactor, recycle to acid gas removal area, purge to combustor in area 200, and a stream to the gas turbine in the power generation area. The LT scenario does not contain a syngas stream to combustor in area 200 because char is used. Overall CO conversion is 66% due to recycling syngas. Recycle ratio is approximately 1.95 for both scenarios.

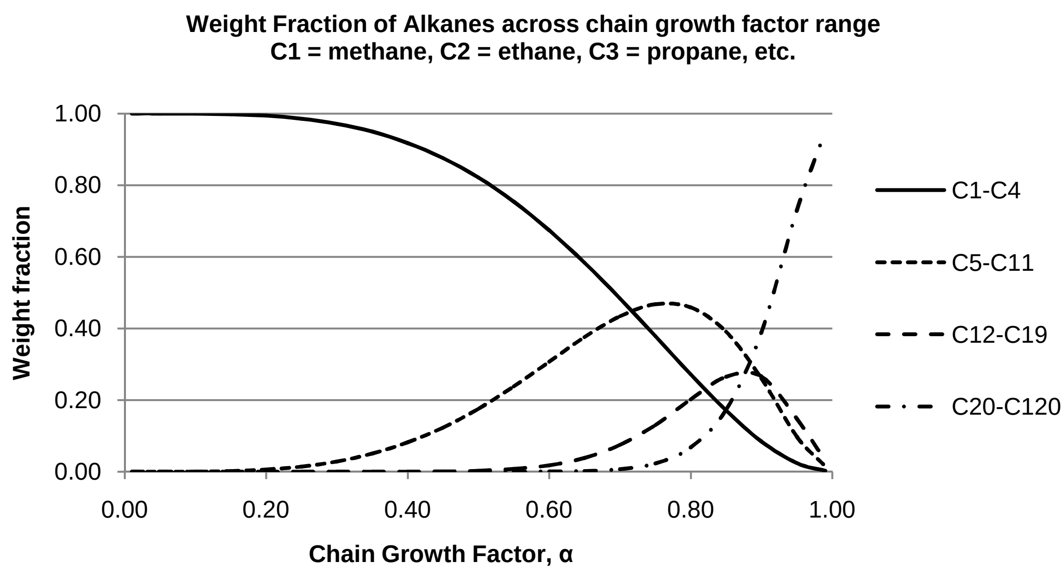


Figure 11. Fischer-Tropsch product distribution as a function of chain growth factor (α) using equation 11 [49]

3.2.7 Area 500 Hydroprocessing

FT products from the fuel synthesis area contain significant amounts of high molecular weight wax which requires hydrogen in order to crack high molecular weight parrafins to low molecular weight hydrocarbons. A product distribution is specified in Table 8 as detailed in Shah et al.[50] It is assumed that the hydroprocessing area contains a hydrocracker for converting the wax fraction and a distillation section for separating naphtha, diesel, and lighter molecular weight hydrocarbon. Also, hydrogen is assumed to be recycled within this area as needed. Methane and LPG are separated and used to fuel the gas turbine in the power generation area. The hydroprocessing area is modeled as a "black box."

Table 8. Hydroprocessing product distribution [50]

Component	Mass Fraction
Methane	0.0346
LPG (propane)	0.0877
Gasoline (octane)	0.2610
Diesel (hexadecane)	0.6167

3.2.8 Area 600 Power Generation

A gas turbine and steam turbine provide the means to producing power that is required throughout the plant and also generate excess power for export. Unconverted syngas from Fisher-Tropsch synthesis and fuel gas from hydroprocessing are combusted in a gas turbine producing hot flue gas and shaft work. The flue gas exchanges heat with water in a heat recovery steam generator to produce steam for the steam turbines which subsequently produce more shaft work. Electric generators attached to both the gas turbine and steam turbine produce electricity from the shaft work.

3.2.9 Area 700 Air Separation

Since 95% purity oxygen is used for both scenarios, a cryogenic air separation unit (ASU) is employed rather than purchasing oxygen. A two-column cryogenic oxygen/nitrogen separation system is employed with subsequent oxygen compression and nitrogen vent. Air pre-cooling is accomplished by exchanging heat with exiting nitrogen. This area requires a significant amount of power, as explained in the results section, which is provided by the power generation area.

3.3 Methodology for Economic Analysis

Capital investment and PV of each scenario is determined by finding all equipment costs and operating costs for the construction and operation a plant for 20 years. Total capital investment is based on the total equipment cost with the additional installation costs and indirect costs (such as engineering, construction, and contingency costs). Annual operating costs are determined and a discounted cash flow rate of return analysis is developed. PV per unit volume of fuel is determined at a net present value of zero and 10% internal rate of return. The major economic assumptions used in this analysis are listed in Table 9. A detailed list of assumptions can be found in appendix A.

Table 9. Main economic assumptions for n^{th} plant scenarios

Parameter	Assumption
Financing	100% equity
Internal rate of return (after taxes)	10%
General plant depreciation period	7 years (all areas except area 600)
Steam plant depreciation period	20 years (area 600 only)
Construction period	2.5 years with total capital investment spent at 8%, 60%, and 32% per year during years before operation
Start up time	0.5 years where during that time revenues, variable operating costs, and fixed operating costs are 50%, 75%, and 100% of normal, respectively.
Income tax rate	39%
Contingency	20% of fixed capital investment
Electricity cost	5.4 cents/kWh
Working capital	15% of fixed capital investment
Land purchase	6% of total purchased equipment cost
Plant availability	310 days per year (85%)

Unit operations from the scenarios are sized and costs are estimated using Aspen Icarus Process Evaluator based on the Aspen Plus simulation data. Unique equipment costs for such equipment as the gasifier and Fischer-Tropsch synthesis reactor are estimated externally using literature references. Additionally, some equipment such as the biomass dryer and lock hoppers require literature references to determine the sizing whereby their costs are subsequently estimated using Aspen Icarus. The hydroprocessing plant area is modeled as a “black box” and therefore its costs are estimated as an overall scaled area cost from literature.

The costs of each equipment or area are scaled based on a scaling stream and scaling size factor (n) according to equation 12 where the size factor is between 0.6-1.0 depending on the equipment type.

$$Cost_{new} = Cost_0 * \left[\frac{Stream\ Size_{new}}{Stream\ Size_0}\right]^n \qquad \text{(eqn. 12)}$$

All purchased equipment costs determined via Aspen Icarus contain an installation factor that accounts for piping, electrical, and other costs required for installation. However, this installation factor tends to be significantly lower than metrics suggested by Peters et al.[51] Therefore, rather than using the software-derived installation factors, an overall installation factor is applied to most equipment. A 3.02 overall installation factor is used as suggested by Peters et al. for solid-liquid plants. Basically, the purchased equipment cost of a piece of equipment is multiplied by the installation cost to determine its installed cost. For the gasification unit, a 2.35 installation factor is used according to a National Energy Technology Laboratory study by Reed et al.[52] It is assumed that all gas compressors receive a 1.2 installation factor which is consistent with Aspen Icarus. The Chemical Engineering Plant Cost Index is used to bring the cost to $2007 wherever a source for an estimated cost is from a previous year [53]. For multiple unit operations that operate in parallel or in trains, a train cost factor is applied. The reason for the factor, as reported by Larson et al. [39], is because those units share some of piping, electrical, and other installation costs. It is applied as shown in equation 13 where n is the number of units in the train and m is the train factor with value of 0.9.

$$Cost_{train} = Cost_{unit} * n^m \qquad \text{(eqn. 13)}$$

Table 10 explains the methodology undertaken to estimate capital investment. After total purchased equipment cost (TPEC) and total installed cost (TIC) are determined, indirect costs are applied. Indirect costs (IC) include engineering and supervision, construction expenses, and legal and contractor's fees at 32%, 34%, and 23% of TPEC, respectively [51]. Project contingency is added as 20% of total direct and indirect cost (TDIC). TDIC is set as the sum of TIC and total installed costs (TIC). With project contingency added the Fixed

Capital Investment is determined. Total Capital Investment (TCI) is determined by adding working capital to Fixed Capital Investment and thereby represents the overall investment required for each scenario.

Table 10. Methodology for capital cost estimation for n^{th} plant scenarios

Parameter	Method
Total Purchased Equipment Cost (TPEC)	Aspen Icarus Process Evaluator®, references
Total Installed Cost (TIC)	TPEC * Installation Factor
Indirect Cost (IC)	89% of TPEC[a]
Total Direct and Indirect Costs (TDIC)	TIC + IC
Contingency	20% of TDIC
Fixed Capital Investment (FCI)	TDIC + Contingency
Working Capital (WC)	15% of FCI
Total Capital Investment	FCI + WC

(a) indirect costs are broken down into engineering and supervision, construction expenses, and legal and contractor's fees at 32%, 34%, and 23%, respectively, for a total of 89% of TPEC.

Raw material costs are inflated to 2007$ using the Industrial Inorganic Chemical Index also used by Phillips et al. Annual variable operating costs are determine from material stream flows. Variable operating costs and respective cost method is shown in Table 11. Natural gas for use in the gas turbine to produce power during startup and backup periods is assumed to be employed 5% of the annual operating time. Solids disposal costs are for the handling and removal of ash in the LT scenario and slag in the HT scenario. Wastewater disposal cost is applied to the sludge and black water produced during direct syngas quench. Catalyst costs are not calculated on an annual basis since the catalysts for all reactors are assumed to be replaced every 3 years. Instead they are accounted for in the discounted cash flow analysis.

Table 11. Variable operating cost parameters adjusted to 2007$

Variable Operating Costs	Cost information
Feedstock	$75/dry short ton
LO-CAT Chemicals	$176/metric ton of sulfur produced as reported in Peters et al. [40]
Amine make-up	$1.09/lb as reported in Phillips et al. and set as 0.01% of the circulating rate [40]
Process Steam	$8.20/ton (Peters et al.) [51]
Cooling water	$0.31/ton (Peters et al.)
Hydroprocessing	$4.00/barrel produced as reported by Robinson and Dolbear [54]
Natural gas (for backup)	$6.40/thousand standard cubic feet as the average wellhead price for 2007 [55]
Ash/Char disposal	$23.52/ton[40]
Wastewater disposal	$3.30/hundred cubic feet [40]
Electricity	$0.054/kWh [a]
Sulfur	$40.00/ton [40]
Fischer-Tropsch catalyst (cobalt)	$15/lb and 64lb/ft^3 density; applied on first operation year and then every three years [a]
Water-gas-shift catalyst (copper-zinc)	$8/lb and 900kg/m^3; applied on first operation year and then every three years. Sour shift and normal WGS are assumed to operate with same catalyst [a]
Steam methane reforming catalyst (nickel-aluminum)	$15/lb and 70lb/ft^3; applied on first operation year and then every three years [a]
Pressure swing adsorption	$2/lb [a]

(a) assumed

Fixed operating costs include employee salaries, overhead, and maintenance, and insurance and taxes. Salaries are calculated similarly to Phillips et al. [40] where employees include a plant manager, shift supervisors, lab technician, maintenance technician, shift operators, yard workers, and office clerks. The labor index developed by the Bureau of Labor Statistics [56] is used to adjust the labor cost to 2007$. Overhead is calculated as 60% of total salaries; maintenance cost and taxes/insurance cost are both 2% of total installed equipment cost as in accordance with Aden et al.[57]

For the DCFROR analysis, the capital investment is spent over a 2.5 year construction period, with 8% in the first half year, followed by 60% and 32% for the next two years. Working capital is applied in the year before operation and recovered at the end of the plant life. A standard modified accelerated cost recovery system (MACRS) is used, with the steam plant depreciating over 20 years and the rest of the plant over a 7 year period consistent with IRS allowances. The project life is 20 years. Plant availability of 310 days per year (85%) is assumed and affects raw materials purchase as well as fuel production. The PV per gallon of gasoline equivalent is calculated for a set net present value of zero including a 10% internal rate of return.

3.3.1 Methodology for Major Equipment Costs

The software used for determining equipment costs is not capable of estimating every unit in this study. Some units such as the gasifiers and Fischer-Tropsch reactors are unique pieces of equipment that are underestimated if estimated as a simple vertical pressure vessel. Therefore, literature sources have been used to help estimate sizes and costs of many units. The following section details a few of the more important units.

The biomass dryer costs are estimated by determining the drying contact area. According to Couper [58], typical rotary dryers have a diameter of 6 feet and solids holdup of 8%. Assuming a bulk density of 100 kg/m^3 for ground stover and 1000 kg/m^3 for moisture in the stover, the resulting total surface area required for drying is 1880 m^2. The surface area provides enough information for estimating the costs since rotary dryer costs are estimated based on surface area in Aspen Icarus. Details on dryer sizing can be found in section 5 of appendix C.

The lock hopper system sizes are estimated by referring to a Department of Energy report completed by Combustion Engineering, Inc. [59] where residence times and operating pressures are given. The biomass receiving bin, lock hopper, and feed bin costs are then estimated with Aspen Icarus. Details on lock hopper sizing can be found in section 5 of appendix C.

The high temperature gasifier cost is estimated from Reed et al. [52] The total bare erected cost (installed cost) of a train of 8 high temperature E-Gas™ gasifiers (2500 metric ton per day coal) including syngas cooling costs is $638 million (2006$). It is assumed that the syngas cooling accounts for 20% of that cost and therefore the estimated installed cost in millions of 2006$ for a 2000 metric ton per day high temperature gasifier follows the formula in equation 14 resulting in $57 million installed.

$$Cost_{HTgasifier} = \frac{(638 \cdot 80\%)}{8} \cdot \left[\frac{2000MT}{2500MT}\right]^{0.7} \quad \text{(eqn. 14)}$$

A fluidized bed gasifier installed cost is described in Larson et al. [39] and is calculated as shown in equation 15 where $Cost_{0\_gasifier}$ is $6.41 million ($2003), $Stream\ Size_0$ is 41.7 metric ton per hour, and n is 0.7. The gasifier is evaluated at 300 short tons per day because that appears to be the highest proven capacity for GTI gasifier. Therefore, seven fluidized bed gasifiers are used in parallel. It is assumed that the gasifier train follows the train cost formula (equation 13) resulting in $19 million installed.

$$Cost_{LTgasifier} = Cost_{0\_gasifier} * \left[\frac{Stream\ Size}{Stream\ Size_0}\right]^n \quad \text{(eqn. 15)}$$

In a similar manner the FT reactor is estimated as described in Larson et al. [39] where base installed cost is $10.5 million ($2003), base sizing value is 2.52 million standard cubic feet per hour of synthesis gas flow, and sizing exponent of 0.72. A installation factor of 3.6 is assumed for the FT reactor as found in Peters et al. [51] for liquid production plants. This allows the purchased cost of the unit to be back calculated.

The acid gas removal (AGR) area cost is evaluated using information from Phillips et al. [40] following equation 12 where the base stream size is 4000 short tons per day and base cost is $5.45 million. The stream size is the mass flow of the synthesis gas entering the AGR as the sum of fresh syngas from gas scrubbing and unconverted syngas from fuel synthesis area.

Capital investment for the hydroprocessing area is found in Robinson et al.[54] That study reports a volumetric unit cost of $4,000 per barrel per standard day. Assuming the typical hydroprocessing refinery produces 25,000 barrels per day the base cost, C_0, is $100 million. Assuming a scaling exponent of 0.65, the cost of area 500 is found using equation 12. The cost details of both gasifiers, AGR area, FT reactor, and hydroprocessing area can be found in section 5 of appendix C.

3.3.2 Methodology for Sensitivity Analysis

Sensitivity parameters are chosen to reflect the change in PV. The parameters are either economic or process parameters. The sensitivity bounds are chosen as what is expected to be observed in the construction and operation of a biomass-to-liquids production plant. The chosen favorable, baseline, and unfavorable sensitivity variables are shown in Table 12.

Table 12. Sensitivity parameters for n^{th} plant scenarios

Parameter	Favorable	Baseline	Unfavorable
Availability (hours/year)	8000	7446	7000
Balance of Plant (% of TPEC)[a]	8	12	16
Catalyst cost (%)[b]	50	100	200
Catalyst lifetime (year)	5	3	1
CO conversion in FT reactor (%)	30	40	50
Compressor Install factor	1.0	1.2	3.0
Contingency (% of TDIC)[c]	10	20	30
Feedstock Cost ($/dry short ton)	50	75	100
Feedstock Moisture (%wet)	20	25	30
Price of Electricity (¢/kWh)	7.0	5.4	3.0
Total Capital Investment (% of baseline)	70	100	130

(a) TPEC=total purchased equipment cost
(b) All catalyst costs are varied over this range
(c) TDIC=total direct and indirect cost

3.3.3 Methodology for Pioneer Plant Analysis

Economic analysis is based on an n^{th} plant design and before a project is undertaken the pioneer (1^{st}) plant cost is important to estimate. This method begun by the RAND Corporation estimates pioneer plant costs and plant performance. Using this methodology, two main areas of the n^{th} plant economic analysis are adjusted: capital investment and plant performance. Through a series of parameters, a cost inflation factor is generated to inflate the capital investment. In addition, a plant performance factor is calculated which reduces the fuel sales, feedstock purchase, and variable operating costs for first several years that the plant is in operation. Each year the plant performance factor is increased until full performance is attained. For the purpose of determining a range of pioneer plant costs baseline, optimistic, and pessimistic values are chosen. The details of the RAND methodology can be found in Merrow et al.[60] The following section explains the reasoning behind the parameters chosen for the scenarios.

Cost growth and plant performance factors are calculated as shown in equations 16 and 17 in accordance with Merrow et al. [60] The

$$\begin{aligned} Cost\ Growth = 1.1219 - 0.00297 * PCTNEW - 0.02125 \\ * IMPURITIES - 0.01137 * COMPLEXITY + 0.00111 \\ * INCLUSIVENESS - 0.06351 \\ * PROJECT\ DEFINITION \end{aligned} \qquad \text{(eqn. 16)}$$

$$\begin{aligned} Plant\ Perf. = 85.77 - 9.69 * NEWSTEPS + 0.33 * BALEQS - 4.12 \\ * WASTE - 17.91 * SOLIDS \end{aligned} \qquad \text{(eqn. 17)}$$

The factors are applied to the capital investment and plant performance as shown in equations 18 and 19. Expenses and revenues affected by the plant performance factor are fuel sales, feedstock purchase, co-product credits, and variable operating costs.

$$TCI_{pioneer} = \frac{TCI_{nth}}{Cost\ Growth} \qquad \text{(eqn. 18)}$$

The $Cost\ Growth$ factor causes the TCI of the pioneer plant to increase from nth plant.

$$Cost_{pioneer}(t) = Cost_{nth}(t) * \frac{(Plant\ Perf. + 20 * (t - 1))}{100} \qquad \text{(eqn. 19)}$$

$Cost_{nth}(t)$ is the nth plant expense or revenue at year t. The plant performance factor is applied at year 1 and increases by 20% each year until 100% performance is reached. The chosen parameters and calculated factors for baseline, optimistic, and pessimistic are shown in Table 13. Details of variables found in equation 16 and 17 and the chosen values are explained in section 5 of appendix B.

Table 13. Pioneer plant analysis parameters and factors

Parameter	Baseline	Optimistic	Pessimistic	Range
Plant Perf.	**38.18**	**49.93**	**22.31**	**0-100**
Cost Growth(HT)	**0.47**	**0.63**	**0.30**	**0-1**
Cost Growth(LT)	**0.50**	**0.65**	**0.31**	**0-1**

4. RESULTS AND DISCUSSION

4.1 Process Results

Along with lower fuel yield, the LT scenario consumes less power (Table 14). The LT scenario and HT scenario total power usage is 15 and 22 MW, respectively. Major contributions to this result are a lower grinder power due to less strict biomass size requirement, lower pressurized oxygen consumption in gasifier, and generally lower downstream mass flow rates throughout the plant for the LT scenario. A lower syngas yield also means that there is less unconverted syngas and fuel gas from the hydroprocessing area available for the gas turbine. Therefore, the LT scenario generates 31 MW compared to 36 MW as generated by the HT scenario. Due to unoptimized flow rates of the recycle streams, the LT scenario actually generates a net 16 MW of power, which is more than the 14 MW produced in the HT scenario. Reducing the net power generation is achievable by increasing the recycle ratio and thereby increasing conversion, but a consequence is higher flow rates and therefore larger and more expensive equipment. The focus of this study is to produce liquid fuels. However, procedures to optimize recycle ratios, equipment sizes, and fuel production rates are not within the scope of this study and are not undertaken.

Table 14. Power generation and usage

Power (MW)	HT Scenario	LT Scenario
USAGE		
Chopper	0.50	0.50
Grinder	2.96	1.10
Lock hopper system	0.18	0.18
Lean Amine Solution Pump	0.86	0.69
Syngas Booster Compressor	1.25	0.96
PSA Compressor	0.15	0.11
Recycle Compressor	0.39	0.29
Hydroprocessing Area	2.24	1.73
Oxygen compressor (ASU)	3.61	2.80
Air Compressor (ASU)	7.94	6.31
Sour Gas Shift Steam Compressor	1.59	0
CO2 Compressor	0.39	0.39
Total Usage	**22.06**	**15.06**
GENERATION		
Gas Turbine	26.25	21.02
Steam Turbine	9.63	10.40
Total Generated	**35.88**	**31.42**
Net Export	**13.82**	**16.36**

An energy balance of the scenarios shows that the biomass to fuels efficiency for the LT and HT scenarios is 39% and 50% on a LHV basis, respectively (Table 15). When the net electricity is added the efficiencies are 43% and 53% on LHV basis, respectively. The LT scenario is expected to be lower since mass and energy loss occurs in the production and removal of char and tar. Char and tar energy loss sums to 7.5% of the energy in the biomass. In this scenario char is combusted in a fluidized bed combustor to provide heat for biomass drying. Biomass drying in the HT scenario is accomplished by a syngas purge. The most significant energy loss in the LT scenario, about 25%, occurs across the gasifier. One reason for high energy loss is because thermodynamic efficiency increases with increasing operating temperature. The second reason is due to loss of energy during the cooling of the syngas after the gasifier. More effective capture of the energy in the hot syngas would increase the overall energy efficiency.

High exothermicity of the FT reaction causes a significant portion of the chemical energy in the syngas to leave as thermal energy in both scenarios. A higher loss across the FT reactor is observed in the HT scenario due to higher flowrates. Energy closure as shown in Table 15 is approximately 90% for both scenarios. It is assumed that the last 10% is due mostly to heat loss from the cooling of the syngas by direct quench rather than capturing the heat and raising steam.

Table 15. Overall energy balance on LHV basis

	High Temperature	**Low Temperature**
IN		
Biomass	1.000	1.000
OUT		
Fuel	-0.497	-0.385
Net Electricity	-0.035	-0.042
Power Gen Losses	-0.042	-0.031
FT reactor losses	-0.162	-0.125
Gasifier losses	-0.121	-0.249
Char	0.000	-0.063
Tar	0.000	-0.012
Syngas Purge	-0.018	0.000
Total[a]	-0.875	-0.907

[a]The balance of energy is assumed to come from various heating and cooling losses.

A carbon balance analysis shows that 26 and 34 percent of the carbon in the biomass is passed on to the fuels for the LT and HT scenarios, respectively (Table 16). Approximately 99% of the carbon is accounted for. Major carbon losses include carbon dioxide flue gases, LO-CAT venting and lock hopper venting. Char leaving the LT scenario is accounted for in the A200 flue gas since the char is combusted for process heat. Also since the LT scenario produces low molecular weight hydrocarbons in the gasification process, a small fraction become dissolved in the liquid effluent of the wet scrubber. Carbon dioxide also dissolves in wet scrubber effluent stream. Another carbon loss comes from the hydrocarbons that dissolve in the acid gas removal area.

Table 16. Overall carbon balance

	HT scenario		**LT scenario**	
	kmol/hr	%	kmol/hr	%
IN				
Biomass	3280.60	1.000	3280.60	1.000
OUT				
Fuel	1111.28	0.339	861.60	0.263
A300 CO2 Vent	1458.41	0.445	1293.87	0.394
A600 Flue Gas	334.13	0.102	301.92	0.092
A200 Flue Gas	39.35	0.012	226.77	0.069
Lock hopper Vent	159.14	0.049	161.89	0.049
Wet Scrubber Effluent	154.38	0.047	318.30	0.097
Tar	0.00	0.000	34.58	0.011
Dissolved Hydrocarbons	0.00	0.000	45.90	0.014
Total	3256.69	0.993	3244.83	0.989

Throughout the scenarios steam and cooling water are required as utilities. Since a pinch analysis (a method to optimize heat exchange) is not undertaken for this study, integration of the heat streams is not optimized. Therefore, it is assumed that the resulting heating and cooling requirements within the model represent steam and cooling water utilities whereby they are recycled at a ratio of 9:1. In other words, fresh steam and cooling water utility input to the scenarios are assumed to be calculated at 10% of the required circulating rate.

4.2 Cost Estimating Results

4.2.1 Capital and Operating Costs for n^{th} Plant

The breakdown of costs by area and resulting total capital investment is shown in Table 17. Total capital investment for the HT and LT scenarios are $606 million and $498 million, respectively. Major areas of investment are the gasification area in the HT scenario and the fuel synthesis area in the LT scenario. Moreover, these two areas contain significant differences in capital investment between the scenarios. The installed cost of the entrained flow gasifier is significantly higher than the fluidized bed gasifiers even when seven are used in parallel. Area 400 costs of the LT scenario are higher than the HT scenario due to steam methane reformer and additional heat exchange equipment required for the high operational temperature. A significant portion of the capital cost is due to gas compression such as the air compressor in the air separation unit and syngas booster compressor. Due to high purchase costs, compressors make up approximately 18% of the TPEC for each scenario. Detailed accounting of equipment found in each process area can be found in section 2 and 3 of appendix B.

Table 17. Capital investment breakdown for nth plant scenarios

Area	**High Temperature**		**Low Temperature**	
	Installed Cost		Installed Cost	
	($MM)	%	($MM)	%
A100: Preprocessing	22.7	7	22.7	9
A200: Gasification	67.8	22	28.2	11
A300: Syngas Cleaning	33.5	11	29.3	12
A400: Fuel Synthesis	49.4	16	58.7	23
A500: Hydroprocessing	33.0	11	29.5	12
A600: Power Generation	45.6	15	38.9	15
A700: Air Separation Unit	24.3	8	19.5	8
Balance of Plant	33.1	11	27.2	11
Total Installed Cost	**309.4**		**253.9**	
Indirect Cost	**129.7**		**107.2**	
Total Direct and Indirect Cost	**439.1**		**361.1**	
Contingency	**87.8**		**72.2**	
Fixed Capital Investment	**526.9**		**433.3**	
Working Capital	**79.0**		**65.0**	
Total Capital Investment	**605.9**		**498.3**	

Annualized costs for operation of the plant are shown in Table 18. The percentage displayed also represents percentage of PV. The largest annual incurred costs for both scenarios are the average return on investment and feedstock purchase. Utilities such as steam and cooling water are higher for the LT scenario due to heating and cooling of the syngas before and after the SMR and steam input to the SMR. Waste disposal costs are equal since equal amount of ash or slag are by-products of the plants. Annual hydroprocessing area costs and income taxes are higher for HT scenario because of higher fuel production rate. Fixed costs and capital depreciation are higher due to higher TCI.

Catalyst costs are not determined on an annual basis since they are assumed to be replaced every three years. Table 19 contains catalyst replacement costs. The catalyst cost the ZnO guard bed and PSA unit are equal across the scenarios because the volumes of the units are assumed to be the same. FT catalyst for the HT scenario is significantly more expensive because of a higher gas flow rate and hence more catalyst. Using a DCFROR analysis, the PV at a net present value of zero for the LT and HT scenarios are $4.83 and

$4.27 per gallon of gasoline equivalent, respectively. Further detail of the yearly cash flow of the life of the plant can be found in section 4 of appendix B.

Table 18. Annual operating cost breakdown for n^{th} plant scenarios

	High Temperature		**Low Temperature**	
	Annual cost (2007$)	%	Annual cost (2007$)	%
Average Return on Investment	$58,200,000	32.7%	$48,300,000	31.0%
Feedstock	$51,300,000	28.9%	$51,300,000	32.9%
Capital Depreciation	$26,300,000	14.8%	$21,700,000	13.9%
Average Income Tax	$21,900,000	12.3%	$18,000,000	11.6%
Fixed Costs	$14,300,000	8.1%	$12,400,000	8.0%
Hydroprocessing	$4,400,000	2.5%	$3,000,000	2.0%
Steam	$2,700,000	1.5%	$3,500,000	2.2%
Cooling Water	$2,300,000	1.3%	$3,500,000	1.6%
Waste Disposal	$1,500,000	0.3%	$1,500,000	0.3%
Other Raw Matl. Costs	$1,400,000	0.8%	$1,300,000	0.8%
Co-product credits	-$5,600,000	-3.1%	-$6,600,000	-4.2%

Table 19. Catalyst replacement costs for both scenarios (3 year replacement period)

Catalyst	**HT scenario**	**LT scenario**
Water-gas-shift (copper-zinc)	$114,621	$104,732
Steam reforming (nickel-aluminum)	N/A	$103,412
ZnO guard bed	$424,410	$424,410
PSA packing	$497,135	$497,135
Fischer-Tropsch (cobalt)	$7,686,720	$6,127,680

4.2.2 Sensitivity Results for n^{th} Plant

The results of sensitivity analysis are summarized in Figures 13 and 14 for the HT and LT scenarios, respectively. Total capital investment and feedstock purchase cost have the highest effect on the PV at approximately ±$0.80 and ±$0.40 per GGE, respectively, for both scenarios. Due to the high percentage of equipment cost for compressors, the compressor installation factor has a very high effect on PV as well. When the compressor installation factor is increased to 3.0, which is the usual installation factor for most of the equipment, the PV increases by $0.71 and $0.78 per GGE for the LT and HT scenarios, respectively. Parameters with a lesser but still significant effect are the contingency factor (as percentage of total direct and indirect costs) and plant availability both with approximately ±$0.20 per GGE. Parameters with the least effect are generally characteristic of the process rather than of the economics. For example, catalyst life, feedstock moisture, and carbon monoxide conversion in the FT reactor affect the PV less than ±$0.15 per GGE.

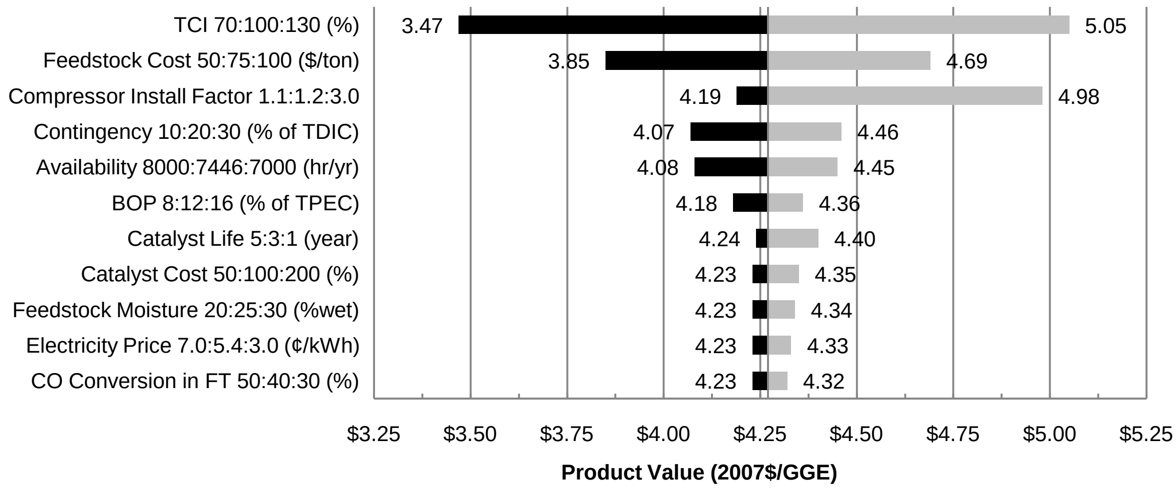


Figure 12. Sensitivity results for HT n[th] plant scenario

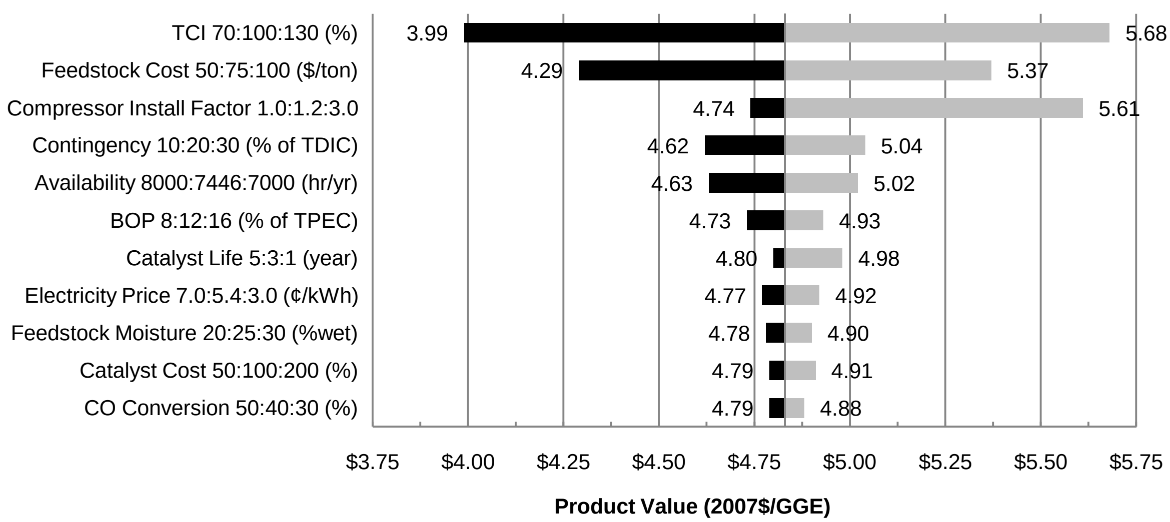


Figure 13. Sensitivity results for LT n[th] plant scenario

Additionally, the plant size of the plants can be varied by feedstock input rate. The effect of plant size on PV and TCI are shown in Figures 15 and 16, respectively. When the plant size is reduced to 500 MT/day the two scenarios approach equal PV. Also, as the plant size is reduced from the baseline, the difference in capital investment decreases. As the plant size increases past the baseline the slope of PV levels out suggesting that the benefits of lower PV may not be worth the significant increase in capital cost (Figure 14).

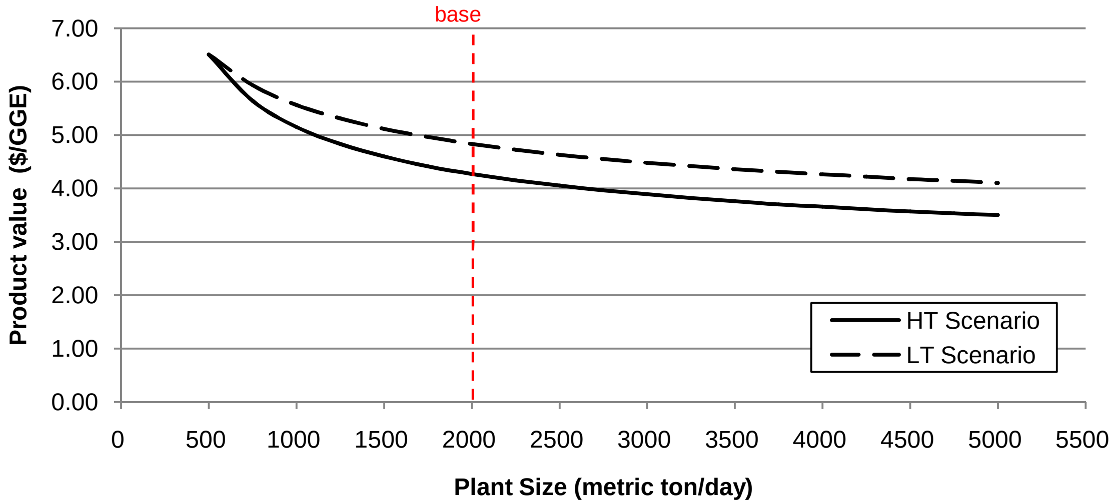


Figure 14. The effect of plant size on product value (per gallon of gasoline equivalent) for nth plant scenarios

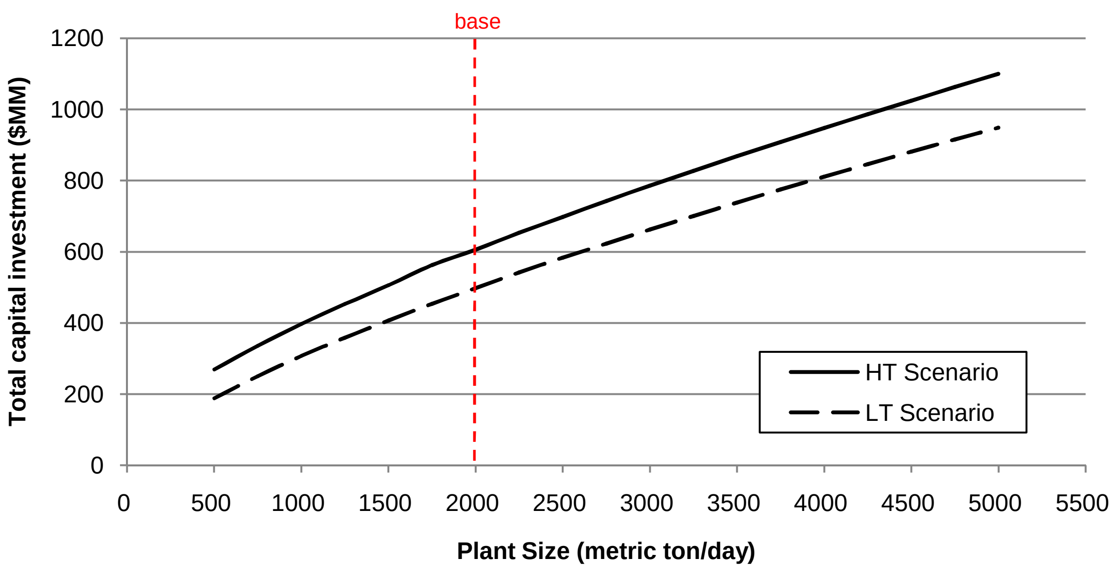


Figure 15. The effect of plant size on total capital investment for nth plant scenarios

4.2.3 Pioneer Plant Analysis Results

The total capital investment for a base case pioneer plant is expected to double from the nth plant scenarios as detailed in Table 20. PV for a base case pioneer plant of the LT and HT scenario are estimated to increase to \$7.20 and 7.70 per GGE, respectively. Table 20 presents further shows estimates of the optimistic and pessimistic cases. An important observation is that the PV for the LT scenario is actually lower than the HT scenario. The reason behind this inverted result is because of the higher capital cost inflation (cost growth factor) in the HT scenario due to higher gasification area capital costs.

Table 20. Pioneer Plant Analysis Results

Analysis	**HT Scenario**		**LT Scenario**	
	TCI (\$MM)	PV (\$/GGE)	TCI (\$MM)	PV (\$/GGE)
nth Plant	606	4.27	498	4.83
1st Plant Base	1290	7.70	997	7.20
1st Plant Optimistic	960	6.00	768	6.00
1st Plant Pessimistic	2050	11.80	1602	10.80

4.3 Comparison with Previous Techno-economic Studies

Two previous BTL studies that specifically use biomass feedstock, low temperature gasification, and Fischer-Tropsch synthesis technology are Tijmensen et al [41]. and Larson et al. [39] In order to compare, major economic and process parameters from the present nth plant LT scenario are adjusted to reflect similar values to the previous studies. First, the plant size of the present study is adjusted to increase equipment costs and raw materials purchases. As a result the annual biomass input and TCI is affected. Second, availability in hours per year, rate of return, cost year, and feedstock cost is adjusted. The combined effect of all adjusted parameters causes the present study's product value to reflect the comparison study.

A comparison to the IGT-R scenario (which employs a low temperature, IGT gasifier and a steam methane reformer) in Tijmensen et al. shows that fuel product value is higher in the present study as summarized in Table 21. Of all the scenarios developed by Tijmensen et al, the IGT-R scenario is most similar to the present study because of the reformer. The IGT-

R scenario has a TCI of $387 million, feedstock cost of $33 per short ton, and a product value of $1.90/GGE. An important characteristic of the Tijmensen et al. study is that it does not include a hydroprocessing area. Therefore, it is expected that the TCI would be higher for the present study since hydroprocessing is included. However, that is not the case since the TCI of the present study using Tijmensen et al. parameters is $339 million which is lower than the reported $387 million. Another important observation is that the annual fuel production for the present study with adjusted parameters is 30.2 million gallons per year compared to 39.8 million gallons per year of FT products reported by Tijmensen et al. One reason for lower annual fuel production in the present study is because of a loss during hydroprocessing. Therefore, due to lower annual fuel production and hence lower fuel revenue, the present study has higher product value compared to Tijmensen et al.

Table 21. Comparison of n^{th} plant LT scenario to Tijmensen et al. study [41]

Parameter	Tijmensen et al. study (IGT-R scenario)	Present study n^{th} plant LT scenario	Present study w/ Tijmensen et al. parameters
Plant Size (dry tons / day)	1920	2205	1920
Annual Biomass Input (tons)	640000	684100	640000
Total Capital Investment ($MM)	387	498	339
Availability (hour/year)	8000	7446	8000
Rate of Return (%)	10	10	10
Cost Year	2000	2007	2000
Feedstock Cost ($/short ton)	33.00	75.00	33.00
Efficiency (%, LHV, incl. elec.)	50.1	42.7	42.7
Fuel Yield (MMGGE/yr)	39.8	32.3	30.2
Product Value ($/GJ)	16.50	39.80	**25.17**
Product Value ($/GGE)	2.00	4.83	**3.05**

A comparison to the FT-OT-VENT scenario (which is low temperature gasification with carbon dioxide vent and once through FT synthesis) reported by Larson et al. is summarized in Table 22. In a similar fashion to the previous comparison the parameters were adjusted to approximate the comparison study. Some important observations are made from this comparison. First, the TCI of the present study with adjusted parameters is significantly higher. Second, the net electricity is significantly lower for the present study.

Third, the PV is significantly higher for the present study. Essentially, the Larson et al. study generates more revenue from selling electricity and recovers the capital investment in less time. In addition, annual operating costs for the Larson et al. study are lower than the present study. Therefore, the present study has a higher fuel product value when compared on a similar basis to Larson et al.

Table 22. Comparison of nth plant LT scenario to Larson et al. study [39]

Parameter	**Larson et al. study (FT-OT-VENT scenario)**	**Present study nth plant LT scenario**	**Present study with Larson et al. parameters**
Plant Size (dry tons / day)	5000	2205	5001
Annual Biomass Input (tons)	1458000	684000	1459000
Total Capital Investment ($MM)	541	498	678
Availability (hour/year)	7000	7446	7000
Debt/Equity (% Equity)	60	100	60
Rate of Return (%)	12	10	12
Cost Year	2003	2007	2003
Electricity Price (cents/kWh)	4.0	5.4	4.0
Net Electricity (MW)	207	16.3	37.1
Feedstock Cost ($/short ton)	46.00	75.00	46.00
Plant Yield (MMGGE/yr)	63.3	32.3	68.9
Product Value ($/GJ)	15.25	39.80	**26.80**
Product Value ($/GGE)	1.85	4.83	**3.25**

4.4 Summary of nth plant scenarios

The HT scenario requires more power and capital investment, yields more fuel per ton of feedstock, and subsequently produces more fuel per year compared to the LT scenario. The total capital investment for the LT and HT scenarios are $498 million and $606 million, respectively. Despite higher capital investment for the HT scenario, the product value (PV) is lower. PV for the LT and HT scenarios are $4.83 and $4.27 per gallon of gasoline equivalent, respectively. The main reason for a lower PV is because of increased fuel revenue. The main nth plant scenario results are shown in Table 23. A detailed summary of costs can be found in section 1 of appendix B.

Table 23. Main scenario nth plant results (TCI=total capital investment; TPEC=total purchased equipment cost; MM=million; GGE=gallon of gasoline equivalent)

Scenario	TCI ($MM)	TPEC ($MM)	Fuel Yield (GGE/metric ton)	Annual Fuel Output (MMGGE/yr)	Net Electricity Export (MW)	PV ($/GGE)
High Temperature	605.9	145.7	61.0	41.7	13.8	4.27
Low Temperature	498.3	120.4	47.2	32.3	16.4	4.83

5. CONCLUSIONS

This analysis compares capital and operating cost for two biomass-to-liquids scenarios: high temperature (HT) gasification and low temperature (LT) gasification. The selection of these scenarios allow for direct comparison between two modes of gasification: slagging and non-slagging. The slagging, entrained flow gasifier employed for the HT scenario results in higher plant costs (about 20%) than the LT scenario, which employs a fluidized bed gasifier. The higher carbon conversions for the HT gasifier, on the other hand, results in a lower PV compared to the LT scenario. Biomass-to-liquids is expected to produce fuels costing in the range of \$4-\$5 per gallon gasoline equivalent using present gasification and Fischer-Tropsch synthesis technology. The factors chiefly responsible for this relatively high PV is feedstock costs and investment return on the capital to build a \$500 million to \$650 million plant to process 2000 metric tons per day. A pioneer plant analysis estimates that the total capital investment for a pioneer plant would double and PV would increase by approximately 60% from the n^{th} plant scale. This uncertainty suggests that economics are yet to be a major challenge for biomass-to-liquids production plants.

The most sensitive effects on PV are total capital cost, feedstock purchase cost, and compressor installation factor affecting the PV between ±\$0.40-0.80 per gallon. Less expensive biomass feedstock that is lower in ash content than used in the present study will have higher fuel yield and have the potential to significantly decrease PV. Gas compression is a major portion of capital investment and sensitivity analysis shows installation costs of compressors have a high effect on PV. Factors with little effect on the PV are mostly related to the process such as carbon monoxide conversion in the FT reactor, feedstock inlet moisture, and catalyst lifetime.

Due to time and resource constraints, the technoeconomic study presented includes a few shortcomings. The process configuration is not fully optimized by means of heat integration. While some recycle streams are included, a complete heat exchange network for heat recovery is not conceptualized. In addition, some areas such as FT product separation and hydroprocessing are not modeled rigorously and can be improved with detailed mass and energy flows.

REFERENCES

1. Frey HC. Akunuri N. *Probabilistic Modeling and Evaluation of the Performance, Emissions, and Cost of Texaco Gasifier-based Integrated Gasification Combined Cycle Systems using ASPEN*. North Carolina State University: Raleigh, January, 2001.

2. Bain RL. *Material and Energy Balances for Methanol from Biomass Using Biomass Gasifiers*. NREL/TP-510-17098; National Renewable Energy Laboratory: January, 1992.

3. "economy." In *The American Heritage® Dictionary of the English Language*, Fourth ed.; Houghton Mifflin Company 2004.

4. Brown RC, *Biorenewable resources : engineering new products from agriculture*. Iowa State Press: Ames, Iowa, 2003.

5. de Swaan Arons J, van der Kooi H. Sankaranarayanan K, Biomass Production and Conversion. In *Efficiency and Sustainability in the Energy and Chemical Industries*, Marcel Dekker, Inc.: New York, 2004.

6. Cushman JH, et al. *Roadmap for Agriculture Biomass Feedstock Supply in the United States*. DOE/NE-ID-11129; Department of Energy: 2003.

7. Greene N, et al. *Growing energy: How biofuels can help end America's oil dependence*. National Resources Defense Council: December, 2004.

8. Knoef H, Practical Aspects of Biomass Gasification. In *Handbook Biomass Gasification*, Knoef, H., Ed. BTG Biomass Technology Group: Enschede, Netherlands, 2005.

9. Knoef H, History of Biomass Gasification. In *Handbook Biomass Gasification*, Knoef, H., Ed. BTG biomass technology group: Enschede, Netherlands, 2005.

10. Probstein RF. Hicks RE, *Synthetic Fuels*. Dover Publications: Mineola, NY, 2006.

11. Prins MJ. Thermodynamic analysis of biomass gasification and torrefaction. Technische Universiteit Eindhoven, Eindhoven, 2005.

12. McKendry P. Energy production from biomass (part 3): gasification technologies. *Bioresource Technology* 2002, 83, (1), 55-63.

13. Warnecke R. Gasification of biomass: comparison of fixed bed and fluidized bed gasifier. *Biomass and Bioenergy* 2000, 18, (6), 489-497.

14. Ciferno JP. Marano JJ. *Benchmarking Biomass Gasification Technologies for Fuels, Chemicals, and Hydrogen Production*. National Energy Technology Laboratory: June, 2002.

15. Williams RH, et al. Methanol and Hydrogen from Biomass for Transportation. *Energy for Sustainable Development* 1995, 1, (5), 18-34.

16. van der Drift H, et al. *Entrained Flow Gasification of Biomass: Ash Behavior, Feeding Issues, and System Analyses*. ECN-C--04-039; Energy Research Center of the Netherlands: April, 2004.

17. Henrich E. Weirich F, Pressurised Entrained Flow Gasifiers for Biomass. In *Annual International Conference on Thermal Treatment Technologies*, New Orleans, Louisiana, 2002.

18. Henrich E, et al., Clean Syngas from Bio-oil/Char Slurries. In *Science in Thermal and Chemical Biomass Conversion*, Bridgwater, A. V.; Boocock, D. G. B., Eds. CPL Press: Newbury, UK, 2006; Vol. 2, pp 1565-1579.

19. Naimi LJ, et al., Cost and Performance of Woody Biomass Size Reduction for Energy Production. In *CSBE/SCGAB 2006 Annual Conference*, American Society of Agricultural and Biological Engineers: Edmonton, Alberta, 2006.

20. Amos WA. *Report on Biomass Drying Technology*. NREL/TP-570-25885; November, 1998.

21. Zwart RWR, Boerrigter H. vanderDrift A. The Impact of Biomass Pretreatment on the Feasibility of Overseas Biomass Conversion to Fischer-Tropsch Products. *Energy & Fuels* 2006, 20, (5), 2192-2197.

22. Craig KR. Mann MK. *Cost and performance analysis of biomass-based integrated gasification combined-cycle (BIGCC) power systems*. NREL/TP--430-21657; 1996.

23. Stevens DJ. *Hot Gas Conditioning: Recent Progress with Larger-scale Biomass Gasification Systems*. NREL/SR-510-29952; National Renewable Energy Laboratory: Golden, CO, August, 2001.

24. Milne TA, Evans RJ. Abatzoglou N. *Biomass Gasifier "Tars": Their Nature, Formation, and Conversion*. NREL/TP-570-25357; National Renewable Energy Laboratory: Golden, CO, November, 1998.

25. Stevens DJ. *Review and Analysis of the 1980-1989 Biomass Thermochemical Conversion Program*. NREL/TP-421-7501; National Renewable Energy Laboratory: September, 1994.

26. Nexant Inc. *Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup, and Oxygen Separation Equipment*. NREL/SR-510-39945; National Renewable Energy Laboratory: May, 2006.

27. Wender I. Reactions of synthesis gas. *Fuel Processing Technology* 1996, 48, (3), 189-297.

28. Consonni S. Larson ED. Biomass-Gasifier/Aeroderivative Gas Turbine Combined Cycles: Part A - Technologies and Performance Modeling. *Journal of Engineering of Gas Turbines and Power* 1996, 118, (3), 507-515.

29. Bridgwater AV, Toft AJ. Brammer JG. A techno-economic comparison of power production by biomass fast pyrolysis with gasification and combustion. *Renewable & Sustainable Energy Reviews* 2002, 6, (3), 181-246.

30. Spath PL. Dayton DC. *Preliminary Screening -- Technical and Economic Assessment of Synthesis Gas to Fuels and Chemicals with Emphasis on the Potential for Biomass-Derived Syngas*. NREL/TP-510-34929; December, 2003.

31. Huber GW, Iborra S. Corma A. Synthesis of transportation fuels from biomass: Chemistry, catalysts, and engineering. *Chemical Reviews* 2006, 106, (9), 4044-4098.

32. Cocco D, Pettinau A. Cau G. Energy and economic assessment of IGCC power plants integrated with DME synthesis process. *Proceedings of the Institution of Mechanical Engineers - Part A - Power & Energy* 2006, 220, (2), 95-102.

33. Dry ME. The Fischer-Tropsch process: 1950-2000. *Catalysis Today* 2002, 71, (3-4).

34. Espinoza R, et al. Low temperature Fischer-Tropsch synthesis from a Sasol perspective. *Applied Catalysis A: General* 1999, 186, (1), 13-26.

35. Dry ME. Practical and theoretical aspects of the catalytic Fischer-Tropsch process. *Applied Catalysis A: General* 1996, 138, (2), 319-344.

36. Bridgwater AV. The technical and economic feasibility of biomass gasification for power generation. *Fuel* 1995, 74, (5), 631-653.

37. Larson ED, Jin H. Celik FE. *Gasification-based Fuels and Electricity Production from Biomass, without and with Carbon Capture and Storage*. Princeton University: Princeton, NJ, October, 2005.

38. Hamelinck CN, et al. Production of FT transportation fuels from biomass; technical options, process analysis and optimisation, and development potential. *Energy* 2004, 29, (11), 1743-1771.

39. Larson ED, Jin H. Celik FE. Large-scale gasification-based coproduction of fuels and electricity from switchgrass. *Biofuels, Bioproducts, and Biorefining* 2009, 3, (2), 174-194.

40. Phillips S, et al. *Thermochemical Ethanol via Indirect Gasification and Mixed Alcohol Synthesis of Lignocellulosic Biomass*. NREL/TP-510-41168; National Renewable Energy Laboratory: April, 2007.

41. Tijmensen MJA, et al. Exploration of the possibilities for production of Fischer Tropsch liquids and power via biomass gasification. *Biomass and Bioenergy* 2002, 23, (2), 129-152.

42. Lau FS, et al. *Techno-Economic Analysis of Hydrogen Production by Gasification of Biomass*. DE-FC36-01GO11089; Gas Technology Institute: December, 2002.

43. Mani S, Tabil LG. Sokhansanj S. Grinding performance and physical properties of wheat and barley straws, corn stover and switchgrass. *Biomass and Bioenergy* 2004, 27, (4), 339-352.

44. Higman C. van der Burgt MJ, *Gasification*. Elsevier/Gulf Professional Publishing: Boston, 2003.

45. *Engineering Data Book*. 12th, FPS. ed.; Gas Processors Suppliers Association: Tulsa, Oklahoma, 2004.

46. Merichem Company, LO-CAT H2S Removal. http://www.gtp-merichem.com/products/lo-cat/process.php (June 20, 2008)

47. Kohl AL. Nelson RB, *Gas Purification*. 5th ed.; Gulf Professional Publishing: Houston, 1997.

48. Imperial Chemical Industries, ltd., Agricultural Division, *Catalyst handbook: with special reference to unit processes in ammonia and hydrogen manufacture*. Wolfe: London, 1970.

49. Song H, et al. Operating strategies for Fischer-Tropsch reactors: A model-directed study. *Korean Journal of Chemical Engineering* 2004, 21, (2), 308-317.

50. Shah PP, et al. *Fischer-Tropsch Wax Characterization and Upgrading*. DOE/PC/80017-T1; UOP, Inc.: Des Plaines, IL, June 6, 1998.

51. Peters MS, Timmerhaus KD. West RE, *Plant design and economics for chemical engineers*. McGraw-Hill: New York, 2003.

52. Reed ME, et al. *Baseline Technical and Economic Assessment of a Commercial Scale Fischer-Tropsch Liquids Facility*. DOE/NETL-2007/1260; National Energy Technology Laboratory: 2007.

53. Chemical Engineering Plant Cost Index. In Chemical Engineering: 2008.

54. Robinson PR. Dolbear GE, Commercial Hydrotreating and Hydrocracking. In *Hydroprocessing of Heavy Oils and Residua*, Ancheyta, J.; Speight, J. G., Eds. CRC Press: Boca Raton, FL, 2007.

55. Energy Information Administration, U.S. Natural Gas Wellhead Price. http://tonto.eia.doe.gov/dnav/ng/hist/n9190us3m.htm (March 1 2009)

56. United States Bureau of Labor Statistics. Labor Index. 2008.

57. Aden A, et al. *Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis for Corn Stover*. National Renewable Energy Laboratory: Golden, CO, June, 2002.

58. Couper JR, *Process engineering economics*. Marcel Dekker: New York, 2003.

59. *CE IGCC Repowering Project Bins and Lockhoppers: Topical Report*. Under contract DE-FC21-90MC26308; Combustion Engineering, Inc.: 1993.

60. Merrow EW, Phillips KE. Myers CW. *Understanding Cost Growth and Performance Shortfalls in Pioneer Process Plants*. Contract no. DE-AC01-79PE70078; Rand Corporation: 1981.

APPENDIX A. ASSUMPTIONS

A.1 Technoeconomic Model Assumptions

A.1.1 Financial Assumptions

- Capital Investment
  - Equity: 100%
  - Working Capital (% of FCI): 15%
- Depreciation Model
  - Zero Salvage Value for both general plant and steam/ power plant
  - Type of Depreciation: Double-Declining-Balance Depreciation Method (DDB) as per IRS Modified Accelerated Cost Recovery System (MARCS) guidelines
    - Depreciation Period (Years):
      - General Plant: 7
      - Steam/Power System: 20
- Construction & Start-up:
  - Construction Period (Years): 2.5
    - % Spent in Year "-3": 8%
    - % Spent in Year "-2": 60%
    - % Spent in Year "-1": 32%
  - Start-up Time (Years): 0.5
    - Revenues (% of Normal): 50%
    - Variable Costs (% of Normal): 75%
    - Fixed Cost (% of Normal): 100%
- Other
  - Internal Rate of Return: 10%
  - Income Tax Rate: 39%
  - Operating Hours per Year: 8,406

A.1.2 Capital Costs

- Cost Year for Analysis: 2007; cost escalation is applied using the Chemical Engineering Plant Cost Index
- The plant is designed based on the State of the Technology, at the n^{th} plant level of experience
- Most equipment installation factors are applied using Peters et al. for solid-fluid plants (i.e. 3.02 installation factor);
- Materials of construction are carbon steel, stainless steel, alloys and refractory where necessary
- Sensitivity parameter involving changes in equipment size or capacity are use scaling exponents available in literature.

A.1.3 Operating Costs

- Working capital is assumed to be 15% of the total capital investment
- Annual maintenance materials are 2% of the total installed equipment cost
- Boiler feedwater and wastewater treatment costs are derived from prior NREL work.

- Fresh cooling water and steam costs are calculated at 10% of the required circulation rate meaning a 9:1 ratio of water recycling.
- Employee salary estimation is same as that chosen by Phillips, et al.
- Employee salaries are indexed to the year of 2007 following the data of the Bureau of Labor Statistics

A.1.4 Feedstock, Products and By-Products

- Feedstock is corn stover (comprising stalks, leaves, cobs and husks)
  - Moisture content in the feedstock is 25%
- Feed rate is 2000 dry metric ton per day
  - The feedstock delivery logistics are not considered
  - The feedstock is delivered to the feed handling area of the plant
- Feed cost is assumed to be $75/dry short ton at the gate
- Gasoline and diesel products are sold for over the fence
- Gasoline energy content is 115000 BTU/gallon
- Fly ash and slag incur a solids waste disposal cost
- Solid sulfur and electricity are sold as by-product

A.1.5 Process Assumptions

For both scenarios, most of the process was modeled with the aid of Aspen Plus™ software. The process was divided by logical process areas which are named below:

Area 100 - Preprocessing

- Biomass is dried down to 10%
  - Steam raised from hot flue gas is used to dry the feedstock
  - Steam to moisture removal ratio is set at 9:1 in accordance with Amos.
  - Heat is provided by combusting char and unreacted syngas
- Grinder reduces biomass to 6-mm or less
  - The energy required for grinding is calculated separately using literature correlations by Mani et al.

Area 200 - Gasification

- Scenario 1: Entrained flow gasifier is modeled using thermodynamic equilibrium
- Scenario 2: Fluidized bed gasifier is modeled using a mass balance calculation
- 95% purity oxygen produced from Air Separation Unit provides oxidizer
- Carbon dioxide is used as solids pressurization gas
- All char produced in LT scenario is combusted for process heat

Area 300 - Syngas Cleaning

- Particulates, tar and partial ammonia removal via wet scrubbing
  - Scrubbing water is recycled at 90% rate
  - Particulate handling (not modeled)
    - High temperature gasifier: particulate decant slurry is sent back into slagging gasifier
    - Low temperature gasifier: particulate decant slurry is piled and landfilled; excess water is sent to aerobic water treatment (not modeled)
  - Makeup water compensates for water lost via particulate slurry

        - Process water condensate is used as makeup water
- Sour water-gas-shift occurs at equilibrium and is modeled as such.
- Carbon dioxide, hydrogen sulfide and excess ammonia removal via amine scrubbing acid gas removal (AGR) at pressure:
    - 99% of sulfur is removed and 90% of carbon dioxide
    - Monoethanolamine (MEA) is the scrubbing solvent
    - Carbon dioxide is vented following LO-CAT™ removal of H_2S.
- Hydrogen Sulfide is converted to solid sulfur via LO-CAT™ oxidation (99% conversion)
- Ammonia can be disposed of by decomposition (not modeled) in
    - Gasifier burner (slagging gasifier)
    - Char and syngas combustor (fluidized bed gasifier)
- Zinc oxide and activated carbon guard bed polishing assumed (not modeled in detail)

Area 400 - Fuel Synthesis

- Water-gas-shift occurs at equilibrium and is modeled as such.
- Pressure swing adsorption (PSA) is employed to remove excess H_2 at an efficiency of 85% and 99% purity.
    - The PSA system employs two trains with 6 reactors each to account for all stages of pressurization, depressurization, purging etc.;
    - PSA adsorbers are filled 2/3 with activated carbon and 1/3 with molecular sieve
- Syngas is catalytically converted to fuels by one step Fischer-Tropsch synthesis followed by wax hydrocracking and fuel separation
    - FT synthesis employs cobalt catalyst
    - 40% syngas conversion to fuels
    - Part of the unconverted syngas is recycled
        - A fraction of the recycle is sent to the AGR to prevent CO_2 buildup.
        - The overall recycle ratio is about 1.9
- A syngas purge is used as fuel in the combustor side of the biomass dryer (only in HT scenario)
- Excess syngas is sent to a gas turbine for power production

Area 500, 600, 700

- Hydroprocessing and product distillation costs are estimated as a "black box" based on literature capital cost and operating cost information from Robinson et al.
    - Literature yield data is used for estimating the relative yields of gasoline and diesel

A.1.6 Miscellaneous

- Combustion occurs with 120% excess oxygen

APPENDIX B. DETAILED COSTS

B.1 Cost Summary

B.1.1 High Temperature Scenario Summary

HT Biomass-to-Liquids Scenario Summary

2,000 Dry Metric Tonnes Biomass per Day
High Temperature Entrained Flow Gasifier, Sulfur Removal, Fischer-Tropsch Synthesis, Hydroprocessing, Combined Cycle Power
All Currency in 2007$ and Volume in Gallons Gasoline Equivalent (GGE)

Product Value ($/gal)	**$4.26**
Total Production at Operating Capacity (MM gal / year)	41.7
Product Yield (gal / Dry US Ton Feedstock)	61.0
Delivered Feedstock Cost $/Dry US Ton	$75
Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	100%

Capital Costs		
Area 100: Pretreatment	$22,700,000	7%
Area 200: Gasification	$67,800,000	22%
Area 300: Syngas Cleaning	$33,500,000	11%
Area 400: Fuel Synthesis	$49,400,000	16%
Area 500: Hydrocracking/Hydrotreating	$33,000,000	11%
Area 600: Power Generation	$45,600,000	15%
Area 700: Air Separation	$24,300,000	8%
Balance of Plant	$33,100,000	11%
Total Installed Equipment Cost	$309,400,000	
Indirect Costs	129,700,000	
(% of TPI)	21.4%	
Project Contingency	79,000,000	
Total Project Investment (TPI)	$605,900,000	
Installed Equipment Cost per Annual Gallon	$7.42	
Total Project Investment per Annual Gallon	$14.52	
Loan Rate	N/A	
Term (years)	N/A	
Capital Charge Factor	0.176	
Gasifier Efficiency - HHV %	82.1	
Gasifier Efficiency - LHV %	87.9	
Overall Plant Efficiency (incl. electricity) - HHV %	52.7	
Overall Plant Efficiency - LHV %	53.0	
Availability (%)	85.0%	
Plant Hours per year	7446	

Operating Costs (cents/gal product)		
Feedstock	123.0	28.9%
Steam	6.4	1.5%
Cooling Water	5.5	1.3%
Other Raw Materials	3.4	0.8%
Waste Disposal	1.3	0.3%
Hydroprocessing	10.6	2.5%
Fixed Costs	34.4	8.1%
Co-product credits	-13.3	-3.1%
Capital Depreciation	63.0	14.8%
Average Income Tax	52.4	12.3%
Average Return on Investment	139.5	32.7%

Operating Costs ($/yr)	
Feedstock	$51,300,000
Steam	$2,700,000
Cooling Water	$2,300,000
Other Raw Matl. Costs	$1,400,000
Waste Disposal	$1,500,000
Hydroprocessing	$4,400,000
Fixed Costs	$14,300,000
Co-product credits	-$5,600,000
Capital Depreciation	$26,300,000
Average Income Tax	$21,900,000
Average Return on Investment	$58,200,000

Total Plant Electricity Usage (KW)	22,065
Electricity Produced Onsite (KW)	35,880
Electricity Purchased from Grid (KW)	0
Electricity Sold to Grid (KW)	13,815
Plant Electricity Use (KWh/gal product)	6.1

Figure 16. Economic Analysis Summary for HT Scenario

B.1.2 Low Temperature Scenario Summary

LT Biomass-to-Liquids Process Engineering Analysis

2,000 Dry Metric Tonnes Biomass per Day
Low Temperature Fluidized Gasifier, Sulfur Removal, Fischer-Tropsch Synthesis, Hydroprocessing, Combined Cycle Power
All Currency in 2007$ and Volume in Gallons Gasoline Equivalent (GGE)

Product Value ($/gal)	**$4.83**
Total Production at Operating Capacity (MM gal / year)	32.3
Product Yield (gal / Dry US Ton Feedstock)	47.2
Delivered Feedstock Cost $/Dry US Ton	$75
Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	100%

Capital Costs		
Area 100: Pretreatment	$22,700,000	9%
Area 200: Gasification	$28,200,000	11%
Area 300: Syngas Cleaning	$29,300,000	12%
Area 400: Fuel Synthesis	$58,700,000	23%
Area 500: Hydrocracking/Hydrotreating	$29,500,000	12%
Area 600: Power Generation	$38,900,000	15%
Area 700: Air Separation	$19,500,000	8%
Balance of Plant	$27,200,000	11%
Total Installed Equipment Cost	$253,900,000	
Indirect Costs	107,200,000	
(%of TPI)	21.5%	
Project Contingency	65,000,000	
Total Project Investment (TPI)	$498,300,000	
Installed Equipment Cost per Annual Gallon	$7.86	
Total Project Investment per Annual Gallon	$15.43	
Loan Rate	N/A	
Term (years)	N/A	
Capital Charge Factor	0.177	
Gasifier Efficiency - HHV %	64.3	
Gasifier Efficiency - LHV %	68.8	
Overall Plant Efficiency - HHV %	43.0	
Overall Plant Efficiency - LHV %	43.3	
Availability (%)	85.0%	
Plant Hours per year	7446	

Operating Costs (cents/gal product)		
Feedstock	158.9	32.9%
Steam	10.9	2.2%
Cooling Water	7.8	1.6%
Other Raw Materials	4.1	0.8%
Waste Disposal	1.5	0.3%
Hydroprocessing	9.4	2.0%
Fixed Costs	38.4	8.0%
Co-product credits	-20.4	-4.2%
Capital Depreciation	67.2	13.9%
Average Income Tax	55.9	11.6%
Average Return on Investment	149.5	31.0%

Operating Costs ($/yr)	
Feedstock	$51,300,000
Steam	$3,500,000
Cooling Water	$3,500,000
Other Raw Matl. Costs	$1,300,000
Waste Disposal	$1,500,000
Hydroprocessing	$3,000,000
Fixed Costs	$12,400,000
Co-product credits	-$6,600,000
Capital Depreciation	$21,700,000
Average Income Tax	$18,000,000
Average Return on Investment	$48,300,000

Total Plant Electricity Usage (KW)	15,044
Electricity Produced Onsite (KW)	31,420
Electricity Purchased from Grid (KW)	0
Electricity Sold to Grid (KW)	16,376
Plant Electricity Use (KWh/gal product)	5.4

Figure 17. Economic analysis summary for LT scenario

B.2 High Temperature Equipment List

Table 24. Detailed equipment list for Areas 100 and 200 of HT scenario

Equipment Number	Number Required	Number Spares	Equipment Name	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaled Uninstalled Cost in 2007$	Installed Cost Base Year	Installed Cost in 2007$	Cost Source
A100.CONV1	2		Bale Transport Conveyor	$400,000	2000	$800,000	$1,066,531	$1,296,000	$1,727,781	Aden et al. 2002
A100.CONV2	2		Bale Unwrapping Conveyor	$150,000	2000	$300,000	$399,949	$357,000	$475,940	Aden et al. 2002
A100.CONV3	1		Belt Press Discharge Conveyor	$50,000	2000	$50,000	$66,658	$94,500	$125,984	Aden et al. 2002
A100.SCALE	2		Truck Scales	$34,000	2000	$68,000	$90,655	$167,960	$223,918	Aden et al. 2002
A100.FORK1	4	1	Truck Unloading Forklift	$18,000	2000	$90,000	$119,985	$90,000	$119,985	Aden et al. 2002
A100.FORK2	4		Bale Moving Forklift	$18,000	2000	$72,000	$95,988	$72,000	$95,988	Aden et al. 2002
A100.SLAB	1		Concrete Feedstock-Storage Slab	$450,655	2000	$450,655	$600,797	$991,441	$1,321,754	Aden et al. 2002
A100.MAGSEP	1		Magnetic Separator	$13,863	1998	$13,863	$18,700	$18,022	$24,310	Aden et al. 2002
A100.A100CHOP.CHGRIN01	4		Chopper	$105,100	2007	$420,400	$420,400	$1,105,258	$1,105,258	Aspen Icarus
A100.A100CHOP.CHMIX01	1		Chopper Conveyor	$61,400	2007	$61,400	$61,400	$185,428	$185,428	Aspen Icarus
A100.A100CHOP.CHSEP01	1		Chopper Screen with Recycle Conveyor	$20,800	2007	$20,800	$20,800	$62,816	$62,816	Aspen Icarus
A100.A100DRY.DRDRY01	10		Dryer	$633,700	2007	$6,337,000	$6,337,000	$15,201,647	$15,201,647	Aspen Icarus
A100.A100GRIN.GRGRIN01	4		Grinder	$167,100	2007	$668,400	$668,400	$1,757,266	$1,757,266	Aspen Icarus
A100.A100GRIN.GRMIX01	1		Grinder Conveyor	$61,400	2007	$61,400	$61,400	$185,428	$185,428	Aspen Icarus
A100.A100GRIN.GRSEP01	1		Grinder Screen with Recycle Conveyor	$20,800	2007	$20,800	$20,800	$62,816	$62,816	Aspen Icarus
A100					**Subtotal**	**$9,434,718**	**$10,049,464**	**$21,647,582**	**$22,676,317**	
A200.A200COMB.CBREAC01	1		Combustor - Steam Boiler	$1,450,500	2007	$1,450,500	$1,450,500	$4,380,510	$4,380,510	Aspen Icarus
A200.A200SLAG.SLREAC01	1		Entrained Flow, Slagging Gasifier	$23,234,043	2006	$23,234,043	$24,433,879	$54,600,000	$57,419,616	Reed et al. 2007
A200.A200SLAG.SLSEP01	1		Slag collector/separator	$35,100	2007	$35,100	$35,100	$106,002	$106,002	Aspen Icarus
A200.A200SLAG.SLSEP03	3		Direct Quench Syngas Cooler	$396,200	2007	$1,188,600	$1,188,600	$3,589,572	$3,589,572	Aspen Icarus
A200.GSHOP01	1		Biomass Receiving Hopper	$151,400	2007	$297,900	$297,900	$899,658	$899,658	Aspen Icarus
A200.GSTANK01	1		Lockhopper	$229,100	2007	$229,100	$229,100	$691,882	$691,882	Aspen Icarus
A200.GSTANK02	1		Biomass Feeding Bin	$228,900	2007	$228,900	$228,900	$691,278	$691,278	Aspen Icarus
A200					**Subtotal**	**$26,664,143**	**$27,863,979**	**$64,958,902**	**$67,778,518**	

Table 25. Detailed equipment list for Areas 300, 400, and 500 of HT scenario

Equipment Number	Number Required	Number Spares	Equipment Name	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaled Uninstalled Cost in 2007$	Installed Cost Base Year	Installed Cost in 2007$	Cost Source
A300.A300AGR.AGRarea	1		High Pressure Amine System	$6,949,800	2005	$6,949,800	$7,798,857	$20,988,396	$23,552,549	Phillips et al. 2007
A300.A300SGS.SGCOMP01	2		Sour Water Gas Shift Steam Compressor	$1,381,900	2007	$2,763,800	$2,763,800	$3,316,560	$3,316,560	Aspen Icarus
A300.A300SGS.SGREAC01	1		Sour Water Gas Shift Reactor	$66,600	2007	$66,600	$66,600	$201,132	$201,132	Aspen Icarus
A300.A300SUL.SUCOL01	1		LO-CAT Absorber	$23,800	2007	$23,800	$23,800	$71,876	$71,876	Aspen Icarus
A300.A300SUL.SUREAC01	1		LO-CAT Oxidizer Vessel	$1,000,000	2007	$1,000,000	$1,000,000	$3,020,000	$3,020,000	Phillips et al. 2007
A300.A300SUL.SUSEP01	1		Sulfur Separator	$15,900	2007	$15,900	$15,900	$48,018	$48,018	Aspen Icarus
A300.CLCOMP01	2		Carbon Dioxide Compressor	$1,181,200	2007	$2,362,400	$2,362,400	$2,834,880	$2,834,880	Aspen Icarus
A300.CLDRUM01	1		Liquid Collection Tank	$29,600	2007	$29,600	$29,600	$89,392	$89,392	Aspen Icarus
A300.CLHEAT03	1		Direct Quench Syngas Cooler	$91,500	2007	$91,500	$91,500	$276,330	$276,330	Aspen Icarus
A300.CLMIX01	1		Venturi Scrubber	$27,100	2007	$27,100	$27,100	$81,842	$81,842	Aspen Icarus
A300					**Subtotal**	**$13,330,500**	**$14,179,557**	**$30,928,426**	**$33,492,579**	
A400.FSCOMP01	2		Booster Syngas Compressor	$1,007,100	2007	$2,014,200	$2,014,200	$2,417,040	$2,417,040	Asen Icarus
A400.FSCOMP02	1		Recycle Syngas Booster Compressor	$748,400	2007	$748,400	$748,400	$898,080	$898,080	Asen Icarus
A400.FSCOMP03	1		PSA Booster Compressor	$1,461,700	2007	$1,461,700	$1,461,700	$1,754,040	$1,754,040	Asen Icarus
A400.FSHEAT01	1		Syngas Heater	$73,400	2007	$73,400	$73,400	$221,668	$221,668	Asen Icarus
A400.FSHEAT03	1		Syngas Cooler	$137,400	2007	$137,400	$137,400	$414,948	$414,948	Asen Icarus
A400.FSHEAT04	1		Recycle Syngas Pre-heater	$21,500	2007	$21,500	$21,500	$64,930	$64,930	Asen Icarus
A400.FSREAC01	1		Fischer-Tropsch Reactor	$8,888,889	2003	$8,888,889	$11,617,468	$32,000,000	$41,822,886	Larson et al. 2005
A400.FSSEP01	2		ZnO Sulfur Removal Beds	$61,000	2007	$122,000	$122,000	$368,440	$368,440	Asen Icarus
A400.FSSEP02	12		Pressure Swing Absorption Unit	$33,300	2007	$399,600	$399,600	$1,206,792	$1,206,792	Asen Icarus
A400.FSSEP03	1		FT knock-out Column	$39,600	2007	$39,600	$39,600	$119,592	$119,592	Asen Icarus
A400.FSSEP04	1		Water Separator	$47,900	2007	$47,900	$47,900	$144,658	$144,658	Asen Icarus
A400					**Subtotal**	**$13,954,589**	**$16,683,168**	**$39,610,188**	**$49,433,074**	
A500.HYREAC01	1		Hydroprocessing Unit	$9,377,483	2007	$9,377,483	$9,377,483	$28,320,000	$28,320,000	Robinson & Dolbear 2007
A500.HYTANK01	1		Diesel 30-day Storage Tank	$1,167,600	2007	$1,167,600	$1,167,600	$3,526,152	$3,526,152	Aspen Icarus
A500.HYTANK02	1		Gasoline 30-day Storage Tank	$371,900	2007	$371,900	$371,900	$1,123,138	$1,123,138	Aspen Icarus
A500					**Subtotal**	**$10,916,983**	**$10,916,983**	**$32,969,290**	**$32,969,290**	

Table 26. Detailed equipment list for Areas 600 and 700 of HT scenario

Equipment Number	Number Required	Number Spares	Equipment Name	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaled Uninstalled Cost in 2007$	Installed Cost Base Year	Installed Cost in 2007$	Cost Source
A600.COMBB	1		Combustion Turbine - Electric Generator	$22,404,000	2007	$22,404,000	$22,404,000	$26,884,800	$26,884,800	Aspen Icarus
A600.CWPUMP	1	1	Cooling Water Pump	$5,900	2007	$11,800	$11,800	$35,636	$35,636	Aspen Icarus
A600.ECON1_HRSG	1		Heat Recovery Steam Generator	$202,200	2007	$202,200	$202,200	$610,644	$610,644	Aspen Icarus
A600.HPPUMP	1	1	High Pressure Steam Pump	$266,700	2007	$533,400	$533,400	$1,610,868	$1,610,868	Aspen Icarus
A600.HPSEP	1		High Pressure Steam/Water Separation	$107,400	2007	$107,400	$107,400	$324,348	$324,348	Aspen Icarus
A600.LPEXP_ELECGEN	1		Combined Steam Turbine - Electric Gen.	$4,709,600	2007	$4,709,600	$4,709,600	$5,651,520	$5,651,520	Aspen Icarus
A600.LPSEP	1		Low Pressure Water/Steam Separation	$108,800	2007	$108,800	$108,800	$328,576	$328,576	Aspen Icarus
A600.O2COMP	1		Air Compressor	$8,431,900	2007	$8,431,900	$8,431,900	$10,118,280	$10,118,280	Aspen Icarus
A600					**Subtotal**	**$36,509,100**	**$36,509,100**	**$45,564,672**	**$45,564,672**	
A700.COMP1	2		Air Compressor	$3,346,500	2007	$6,693,000	$6,693,000	$8,031,600	$8,031,600	Aspen Icarus
A700.COOLER	1		Air Cooler	$27,200	2007	$27,200	$27,200	$82,144	$82,144	Aspen Icarus
A700.GOXCLR-1	1		Oxygen Compressor Cooler	$23,300	2007	$23,300	$23,300	$70,366	$70,366	Aspen Icarus
A700.GOXCLR-2	1		Oxygen Compressor Cooler	$23,000	2007	$23,000	$23,000	$69,460	$69,460	Aspen Icarus
A700.GOXCMP-1	2		Oxygen Compressor	$1,489,600	2007	$2,979,200	$2,979,200	$3,575,040	$3,575,040	Aspen Icarus
A700.HIGH-P.cond	1		High Pressure Column Condenser	$20,300	2007	$20,300	$20,300	$61,306	$61,306	Aspen Icarus
A700.HIGH-P.cond acc	1		High Pressure Column Condenser Accumulator	$40,500	2007	$40,500	$40,500	$122,310	$122,310	Aspen Icarus
A700.HIGH-P.reflux pump	1	1	High Pressure Column Reflux Pump	$14,300	2007	$28,600	$28,600	$86,372	$86,372	Aspen Icarus
A700.HIGH-P.tower	1		High Pressure Column Tower	$314,300	2007	$314,300	$314,300	$949,186	$949,186	Aspen Icarus
A700.INTRC1	1		Air Compressor Intercooler	$338,300	2007	$338,300	$338,300	$1,021,666	$1,021,666	Aspen Icarus
A700.INTRC2	1		Air Compressor Intercooler	$304,500	2007	$304,500	$304,500	$919,590	$919,590	Aspen Icarus
A700.INTRC3	1		Air Compressor Intercooler	$222,500	2007	$222,500	$222,500	$671,950	$671,950	Aspen Icarus
A700.LOW-P.reb	1		Low Pressure Column Reboiler	$19,600	2007	$19,600	$19,600	$59,192	$59,192	Aspen Icarus
A700.LOW-P.tower	1		Low Pressure Column Tower	$2,581,600	2007	$2,581,600	$2,581,600	$7,796,432	$7,796,432	Aspen Icarus
A700.TSA	1		Water Knock-out Drum	$35,900	2007	$35,900	$35,900	$108,418	$108,418	Aspen Icarus
A700.TURB-1	2		Gas Expander	$86,100	2007	$172,200	$172,200	$520,044	$520,044	Aspen Icarus
A700.WK01	1		Water Knock-out Drum	$57,700	2007	$57,700	$57,700	$174,254	$174,254	Aspen Icarus
A700					**Subtotal**	**$13,881,700**	**$13,881,700**	**$24,319,330**	**$24,319,330**	
				Total		$124,691,733	$130,083,951	$259,998,390	$276,233,779	
				Total (with BOP)		**$139,654,741**	**$145,694,026**	**$291,198,196**	**$309,381,833**	

B.3 Low Temperature Equipment List

Table 27. Detailed equipment list for Areas 100 and 200 of LT scenario

Equipment Number	Number Required	Number Spares	Equipment Name	Original Equip Cost (per unit) in Base Year	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaled Uninstalled Cost in 2007$	Installed Cost Base Year	Installed Cost in 2007$	Cost Source
A100.CONV1	2		Bale Transport Conveyor	$400,000	2000	$800,000	$1,066,531	$1,296,000	$1,727,781	Aden et al. 2002
A100.CONV2	2		Bale Unwrapping Conveyor	$150,000	2000	$300,000	$399,949	$357,000	$475,940	Aden et al. 2002
A100.CONV3	1		Belt Press Discharge Conveyor	$50,000	2000	$50,000	$66,658	$94,500	$125,984	Aden et al. 2002
A100.SCALE	2		Truck Scales	$34,000	2000	$68,000	$90,655	$167,960	$223,918	Aden et al. 2002
A100.FORK1	4	1	Truck Unloading Forklift	$18,000	2000	$90,000	$119,985	$90,000	$119,985	Aden et al. 2002
A100.FORK2	4		Bale Moving Forklift	$18,000	2000	$72,000	$95,988	$72,000	$95,988	Aden et al. 2002
A100.SLAB	1		Concrete Feedstock-Storage Slab	$450,655	2000	$450,655	$600,797	$991,441	$1,321,754	Aden et al. 2002
A100.MAGSEP	1		Magnetic Separator	$13,863	1998	$13,863	$18,700	$18,022	$24,310	Aden et al. 2002
A100.A100CHOP.CHGRIN01	4		Chopper	$105,100	2007	$420,400	$420,400	$1,105,258	$1,105,258	Aspen Icarus
A100.A100CHOP.CHMIX01	1		Chopper Conveyor	$61,400	2007	$61,400	$61,400	$185,428	$185,428	Aspen Icarus
A100.A100CHOP.CHSEP01	1		Chopper Screen with Recycle Conveyor	$20,800	2007	$20,800	$20,800	$62,816	$62,816	Aspen Icarus
A100.A100DRY.DRDRY01	10		Dryer	$633,700	2007	$6,337,000	$6,337,000	$15,201,647	$15,201,647	Aspen Icarus
A100.A100GRIN.GRGRIN01	4		Grinder	$167,100	2007	$668,400	$668,400	$1,757,266	$1,757,266	Aspen Icarus
A100.A100GRIN.GRMIX01	1		Grinder Conveyor	$61,400	2007	$61,400	$61,400	$185,428	$185,428	Aspen Icarus
A100.A100GRIN.GRSEP01	1		Grinder Screen with Recycle Conveyor	$20,800	2007	$20,800	$20,800	$62,816	$62,816	Aspen Icarus
A100					**Subtotal**	**$9,434,718**	**$10,049,464**	**$21,647,582**	**$22,676,317**	
A200.A200COMB.CBCYC01	3		Combustor Cyclone (medium efficiency)	$35,400	2007	$106,200	$106,200	$320,724	$320,724	Aspen Icarus
A200.A200COMB.CBCYC02	3		Combustor Cyclone (high efficiency)	$6,700	2007	$20,100	$20,100	$60,702	$60,702	Aspen Icarus
A200.A200COMB.CBMIX01	1		Ash Storage Vessel	$142,800	2007	$142,800	$142,800	$431,256	$431,256	Aspen Icarus
A200.A200COMB.CBREAC01	1		Combustor - Steam Boiler	$1,450,500	2007	$1,450,500	$1,450,500	$4,380,510	$4,380,510	Aspen Icarus
A200.A200CYC.CYCYC01	2		1st train, medium efficiency cyclone	$20,300	2007	$40,600	$40,600	$122,612	$122,612	Aspen Icarus
A200.A200CYC.CYCYC02	4		1st train, high efficiency cyclone	$24,900	2007	$99,600	$99,600	$300,792	$300,792	Aspen Icarus
A200.A200CYC.CYMIX02	1		Char Collector and conveyor	$84,400	2007	$84,400	$84,400	$254,888	$254,888	Aspen Icarus
A200.GSREAC01	7		Fluidized Bed Gasifier (Pressurized)	$1,096,170	2003	$7,673,191	$10,028,594	$14,843,424	$19,399,838	Larson et al. 2005
A200.GSTANK01	7		Biomass Receiving Hopper	$71,700	2007	$501,900	$501,900	$1,247,712	$1,247,712	Aspen Icarus
A200.GSTANK02	7		Lockhopper	$47,700	2007	$333,900	$333,900	$830,068	$830,068	Aspen Icarus
A200.GSTANK03	7		Biomass Feeding Bin	$47,700	2007	$333,900	$333,900	$830,068	$830,068	Aspen Icarus
A200					**Subtotal**	**$10,787,091**	**$13,142,494**	**$23,622,756**	**$28,179,170**	

Table 28. Detailed equipment list for Areas 300, 400, and 500 of LT scenario

Equipment Number	Number Required	Number Spares	Equipment Name	Original Equip Cost (per unit) in Base Year	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaled Uninstalled Cost in 2007$	Installed Cost Base Year	Installed Cost in 2007$	Cost Source
A300.A300AGR.AGRarea	1		High Pressure Amine System	$6,050,000	2005	$6,050,000	$6,789,129	$18,271,000	$20,503,168	Phillips et al. 2007
A300.A300SUL.SUCOL01	1		LO-CAT Absorber	$16,200	2007	$16,200	$16,200	$48,924	$48,924	Aspen Icarus
A300.A300SUL.SUREAC01	1		LO-CAT Oxidizer Vessel	$1,000,000	2007	$1,000,000	$1,000,000	$3,020,000	$3,020,000	Phillips et al. 2007
A300.A300SUL.SUSEP01	1		Sulfur Separator	$16,200	2007	$16,200	$16,200	$48,924	$48,924	Aspen Icarus
A300.CLCMP01	2		Carbon Dioxide Compressor	$1,176,900	2007	$2,353,800	$2,353,800	$2,824,560	$2,824,560	Aspen Icarus
A300.CLHEAT01	2		Direct Quench Recycle Cooling	$188,800	2007	$377,600	$377,600	$1,140,352	$1,140,352	Aspen Icarus
A300.CLHEAT02	1		Venturi Recycle Cooling	$91,500	2007	$91,500	$91,500	$276,330	$276,330	Aspen Icarus
A300.CLMIX01	1		Venturi Scrubber	$26,800	2007	$26,800	$26,800	$80,936	$80,936	Aspen Icarus
A300.CLSEP03	2		Direct Quench Syngas Cooler	$188,800	2007	$377,600	$377,600	$1,140,352	$1,140,352	Aspen Icarus
A300.CLSEP04	1		Venturi Liquid Collection Tank	$74,500	2007	$74,500	$74,500	$224,990	$224,990	Aspen Icarus
A300					**Subtotal**	**$10,384,200**	**$11,123,329**	**$27,076,368**	**$29,308,536**	
A400.A400COND.CDHEAT01	1		Syngas Heater	$60,500	2007	$60,500	$60,500	$182,710	$182,710	Aspen Icarus
A400.A400COND.CDHEAT02	1		Syngas Pre-heater Furnace	$1,949,500	2007	$1,949,500	$1,949,500	$5,887,490	$5,887,490	Aspen Icarus
A400.A400COND.CDHEAT03	1		Reformed Syngas Waste Heat Boiler	$396,600	2007	$396,600	$396,600	$1,197,732	$1,197,732	Aspen Icarus
A400.A400COND.CDHEAT04	1		Syngas Cooler #2	$41,200	2007	$41,200	$41,200	$124,424	$124,424	Aspen Icarus
A400.A400COND.CDREAC01	1		Steam Methane Reformer	$1,650,800	2007	$1,650,800	$1,650,800	$4,985,416	$4,985,416	Aspen Icarus
A400.A400COND.CDREAC02	1		Water Gas Shift Reactor	$136,600	2007	$136,600	$136,600	$412,532	$412,532	Aspen Icarus
A400.A400COND.CDSEP01	2		ZnO Sulfur Removal Beds	$46,400	2007	$92,800	$92,800	$280,256	$280,256	Aspen Icarus
A400.FSCOMP01	2		Booster Syngas Compressor	$921,600	2007	$1,843,200	$1,843,200	$2,211,840	$2,211,840	Aspen Icarus
A400.FSCOMP02	1		Recycle Syngas Booster Compressor	$725,400	2007	$725,400	$725,400	$870,480	$870,480	Aspen Icarus
A400.FSCOMP03	1		PSA Booster Compressor	$1,482,100	2007	$1,482,100	$1,482,100	$1,778,520	$1,778,520	Aspen Icarus
A400.FSDRUM01	1		PSA Knock-out	$1,482,100	2007	$1,482,100	$1,482,100	$4,475,942	$4,475,942	Aspen Icarus
A400.FSHEAT03	1		Syngas Cooler	$165,200	2007	$165,200	$165,200	$498,904	$498,904	Aspen Icarus
A400.FSHEAT04	1		Recycle Syngas Pre-heater	$24,300	2007	$24,300	$24,300	$73,386	$73,386	Aspen Icarus
A400.FSREAC01	1		Fischer-Tropsch Reactor	$7,303,889	2003	$7,303,889	$9,545,928	$26,294,000	$34,365,342	Larson et al. 2005
A400.FSSEP02	12		Pressure Swing Absorption Unit	$30,500	2007	$366,000	$366,000	$1,105,320	$1,105,320	Aspen Icarus
A400.FSSEP03	1		FT knock-out Column	$72,100	2007	$72,100	$72,100	$217,742	$217,742	Aspen Icarus
A400.FSSEP04	1		Water Separator	$39,200	2007	$39,200	$39,200	$118,384	$118,384	Aspen Icarus
A400					**Subtotal**	**$17,792,289**	**$20,034,328**	**$50,596,694**	**$58,668,036**	
A500.HYREAC01	1		Hydrocracking/Hydrotreating Unit	$7,927,152	2007	$7,927,152	$7,927,152	$23,940,000	$23,940,000	Robinson & Dolbear 2007
A500.HYTANK01	1		Gasoline 30-day Storage Tank	$646,300	2007	$646,300	$646,300	$1,951,826	$1,951,826	Aspen Icarus
A500.HYTANK02	1		Diesel 30-day Storage Tank	$1,200,700	2007	$1,200,700	$1,200,700	$3,626,114	$3,626,114	Aspen Icarus
A500					**Subtotal**	**$9,774,152**	**$9,774,152**	**$29,517,940**	**$29,517,940**	

Table 29. Detailed equipment list for Areas 600 and 700 of LT scenario

Equipment Number	Number Required	Number Spares	Equipment Name	Original Equip Cost (per unit) in Base Year	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaled Uninstalled Cost in 2007$	Installed Cost Base Year	Installed Cost in 2007$	Cost Source
A600.COMBB	1		Combustion Turbine - Electric Generator	$18,607,700	2007	$18,607,700	$18,607,700	$22,329,240	$22,329,240	Aspen Icarus
A600.CWPUMP	1	1	Cooling Water Pump	$5,900	2007	$11,800	$11,800	$35,636	$35,636	Aspen Icarus
A600.ECON1_HRSG	1		Heat Recovery Steam Generator	$202,200	2007	$202,200	$202,200	$610,644	$610,644	Aspen Icarus
A600.HPPUMP	1	1	High Pressure Steam Pump	$266,700	2007	$533,400	$533,400	$1,610,868	$1,610,868	Aspen Icarus
A600.HPSEP	1		High Pressure Steam/Water Separation	$107,400	2007	$107,400	$107,400	$324,348	$324,348	Aspen Icarus
A600.LPEXP_ELECGEN	1		Combined Steam Turbine - Electric Gen.	$5,056,300	2007	$5,056,300	$5,056,300	$6,067,560	$6,067,560	Aspen Icarus
A600.LPSEP	1		Low Pressure Water/Steam Separation	$108,800	2007	$108,800	$108,800	$328,576	$328,576	Aspen Icarus
A600.O2COMP	1		Air Compressor	$6,331,200	2007	$6,331,200	$6,331,200	$7,597,440	$7,597,440	Aspen Icarus
A600					**Subtotal**	**$30,958,800**	**$30,958,800**	**$38,904,312**	**$38,904,312**	
A700.COMP1	2		Air Compressor	$3,119,600	2007	$6,239,200	$6,239,200	$7,487,040	$7,487,040	Aspen Icarus
A700.COOLER	1		Air Cooler	$24,300	2007	$24,300	$24,300	$73,386	$73,386	Aspen Icarus
A700.GOXCLR-1	1		Oxygen Compressor Cooler	$23,300	2007	$23,300	$23,300	$70,366	$70,366	Aspen Icarus
A700.GOXCLR-2	1		Oxygen Compressor Cooler	$23,000	2007	$23,000	$23,000	$69,460	$69,460	Aspen Icarus
A700.GOXCMP-1	2		Oxygen Compressor	$1,514,700	2007	$3,029,400	$3,029,400	$3,635,280	$3,635,280	Aspen Icarus
A700.HIGH-P.cond	1		High Pressure Column Condenser	$20,300	2007	$20,300	$20,300	$61,306	$61,306	Aspen Icarus
A700.HIGH-P.cond acc	1		High Pressure Column Condenser Accumulator	$36,300	2007	$36,300	$36,300	$109,626	$109,626	Aspen Icarus
A700.HIGH-P.reflux pump	1	1	High Pressure Column Reflux Pump	$14,300	2007	$28,600	$28,600	$34,320	$34,320	Aspen Icarus
A700.HIGH-P.tower	1		High Pressure Column Tower	$279,900	2007	$279,900	$279,900	$335,880	$335,880	Aspen Icarus
A700.INTRC1	1		Air Compressor Intercooler	$338,300	2007	$338,300	$338,300	$405,960	$405,960	Aspen Icarus
A700.INTRC2	1		Air Compressor Intercooler	$304,500	2007	$304,500	$304,500	$919,590	$919,590	Aspen Icarus
A700.INTRC3	1		Air Compressor Intercooler	$222,500	2007	$222,500	$222,500	$671,950	$671,950	Aspen Icarus
A700.LOW-P.reb	2		Low Pressure Column Reboiler	$19,600	2007	$39,200	$39,200	$118,384	$118,384	Aspen Icarus
A700.LOW-P.tower	1		Low Pressure Column Tower	$1,538,900	2007	$1,538,900	$1,538,900	$4,647,478	$4,647,478	Aspen Icarus
A700.TSA	1		Water Knock-out Drum	$30,100	2007	$30,100	$30,100	$90,902	$90,902	Aspen Icarus
A700.TURB-1	2		Gas Expander	$89,200	2007	$178,400	$178,400	$538,768	$538,768	Aspen Icarus
A700.WK01	1		Water Knock-out Drum	$64,800	2007	$64,800	$64,800	$195,696	$195,696	Aspen Icarus
A700					**Subtotal**	**$12,421,000**	**$12,421,000**	**$19,465,392**	**$19,465,392**	
				Total		$101,552,251	$107,503,567	$210,831,043	$226,719,704	
				Total (with BOP)		**$113,738,521**	**$120,403,995**	**$236,130,768**	**$253,926,068**	

B.4 Discounted Cash Flow

B.4.1 High Temperature Scenario

Table 30. Discounted cash flow sheet for construction period and years 1-8 of HT scenario

DCFROR Worksheet												
Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$50,890,395	$316,115,651	$168,595,014								
Working Capital				$79,028,913								
Loan Payment					$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
GGE (Gallon of Gasoline Equiv.) Sales					$133,364,635	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513
Diesel Sales					$0	$0	$0	$0	$0	$0	$0	$0
By-Product Credit					$4,173,208	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277
Plant Performance					1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Annual Sales					$137,537,843	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791
Annual Manufacturing Cost												
Raw Materials					$44,894,145	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594
SWGS catalysts					$114,621	$0	$0	$114,621	$0	$0	$114,621	$0
Steam reforming catalysts					$0	$0	$0	$0	$0	$0	$0	$0
ZnO					$424,410	$0	$0	$424,410	$0	$0	$424,410	$0
Pressure Swing Adsorption Packing					$497,135	$0	$0	$497,135	$0	$0	$497,135	$0
FT catalysts					$7,686,720	$0	$0	$7,686,720	$0	$0	$7,686,720	$0
Other Variable Costs					$11,727,856	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264
Fixed Operating Costs					$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785
Total Product Cost					$79,690,672	$79,056,643	$79,056,643	$87,779,529	$79,056,643	$79,056,643	$87,779,529	$79,056,643
Annual Depreciation												
General Plant												
DDB					$128,361,546	$91,686,818	$65,490,585	$46,778,989	$33,413,564	$23,866,831	$17,047,737	
SL					$64,180,773	$53,483,977	$45,843,409	$40,931,615	$38,982,491	$38,982,491	$38,982,491	
Remaining Value					$320,903,864	$229,217,046	$163,726,461	$116,947,472	$83,533,909	$59,667,078	$42,619,341	
Actual					$128,361,546	$91,686,818	$65,490,585	$46,778,989	$38,982,491	$38,982,491	$38,982,491	
Steam Plant												
DDB					$5,819,551	$5,383,084	$4,979,353	$4,605,902	$4,260,459	$3,940,925	$3,645,355	$3,371,954
SL					$3,879,700	$3,777,603	$3,688,410	$3,612,472	$3,550,382	$3,503,044	$3,471,767	$3,458,414
Remaining Value					$71,774,458	$66,391,374	$61,412,021	$56,806,119	$52,545,660	$48,604,736	$44,959,380	$41,587,427
Actual					$5,819,551	$5,383,084	$4,979,353	$4,605,902	$4,260,459	$3,940,925	$3,645,355	$3,458,414
Net Revenue					($76,333,925)	$7,257,245	$33,857,210	$44,219,371	$61,084,198	$61,403,732	$52,976,416	$100,868,733
Losses Forward						($76,333,925)	($69,076,681)	($35,219,471)	$0	$0	$0	$0
Taxable Income					($76,333,925)	($69,076,681)	($35,219,471)	$8,999,900	$61,084,198	$61,403,732	$52,976,416	$100,868,733
Income Tax					$0	$0	$0	$3,509,961	$23,822,837	$23,947,455	$20,660,802	$39,338,806
Annual Cash Income					$57,847,171	$104,327,147	$104,327,147	$92,094,301	$80,504,310	$80,379,692	$74,943,459	$64,988,341
Discount Factor		1.21	1.1	1	0.909090909	0.826446281	0.751314801	0.683013455	0.620921323	0.56447393	0.513158118	0.46650738
Annual Present Value	$645,181,377				$52,588,337	$86,220,783	$78,382,530	$62,901,646	$49,986,843	$45,372,241	$38,457,845	$30,317,541
Total Capital Investment + Interest		$61,577,378	$347,727,216	$247,623,927								
Net Present Worth				$0								

Table 31. Discounted cash flow sheet for years 9-20 of HT scenario

DCFROR Worksheet												
Year	9	10	11	12	13	14	15	16	17	18	19	20
Fixed Capital Investment												
Working Capital												($79,028,913)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
GGE (Gallon of Gasoline Equiv.) Sales	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513	$177,819,513
Diesel Sales	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
By-Product Credit	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277	$5,564,277
Plant Performance	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Annual Sales	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791	$183,383,791
Annual Manufacturing Cost												
Raw Materials	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594
SWGS catalysts	$0	$114,621	$0	$0	$114,621	$0	$0	$114,621	$0	$0	$114,621	$0
Steam reforming catalysts	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
ZnO	$0	$424,410	$0	$0	$424,410	$0	$0	$424,410	$0	$0	$424,410	$0
Pressure Swing Adsorption Packing	$0	$497,135	$0	$0	$497,135	$0	$0	$497,135	$0	$0	$497,135	$0
FT catalysts	$0	$7,686,720	$0	$0	$7,686,720	$0	$0	$7,686,720	$0	$0	$7,686,720	$0
Other Variable Costs	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264	$13,403,264
Fixed Operating Costs	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785	$14,345,785
Total Product Cost	$79,056,643	$87,779,529	$79,056,643	$79,056,643	$87,779,529	$79,056,643	$79,056,643	$87,779,529	$79,056,643	$79,056,643	$87,779,529	$79,056,643
Annual Depreciation												
General Plant												
DDB												
SL												
Remaining Value												
Actual												
Steam Plant												
DDB	$3,119,057	$2,885,128	$2,668,743	$2,468,587	$2,283,443	$2,112,185	$1,953,771	$1,807,238	$1,671,696	$1,546,318	$1,430,344	$1,323,069
SL	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414
Remaining Value	$38,468,370	$35,583,242	$32,914,499	$30,445,912	$28,162,468	$26,050,283	$24,096,512	$22,289,273	$20,617,578	$19,071,260	$17,640,915	$16,317,847
Actual	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414	$3,458,414
Net Revenue	$100,868,733	$92,145,848	$100,868,733	$100,868,733	$92,145,848	$100,868,733	$100,868,733	$92,145,848	$100,868,733	$100,868,733	$92,145,848	$100,868,733
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$100,868,733	$92,145,848	$100,868,733	$100,868,733	$92,145,848	$100,868,733	$100,868,733	$92,145,848	$100,868,733	$100,868,733	$92,145,848	$100,868,733
Income Tax	$39,338,806	$35,936,881	$39,338,806	$39,338,806	$35,936,881	$39,338,806	$39,338,806	$35,936,881	$39,338,806	$39,338,806	$35,936,881	$39,338,806
Annual Cash Income	$64,988,341	$59,667,381	$64,988,341	$64,988,341	$59,667,381	$64,988,341	$64,988,341	$59,667,381	$64,988,341	$64,988,341	$59,667,381	$64,988,341
Discount Factor	0.424097618	0.385543289	0.350493899	0.318630818	0.28966438	0.263331254	0.239392049	0.217629136	0.197844669	0.17985879	0.163507991	0.148643628
Annual Present Value	$27,561,401	$23,004,358	$22,778,017	$20,707,288	$17,283,515	$17,113,461	$15,557,692	$12,985,361	$12,857,597	$11,688,724	$9,756,094	$9,660,103
Total Capital Investment + Interest												($11,747,144.32)
Net Present Worth												

B.4.2 Low Temperature Scenario

Table 32. Discounted cash flow sheet for construction period and years 1-8 of LT scenario

DCFROR Worksheet

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$41,888,460	$259,981,649	$138,656,880								
Working Capital				$64,995,412								
Loan Payment					$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
GGE (Gallon of Gasoline Equiv.) Sales					$117,025,289	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719
Diesel Sales					$0	$0	$0	$0	$0	$0	$0	$0
By-Product Credit					$4,945,498	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998
Plant Performance					1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Annual Sales					$121,970,788	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717
Annual Manufacturing Cost												
Raw Materials					$44,894,145	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594
WGS catalysts					$104,732	$0	$0	$104,732	$0	$0	$104,732	$0
Steam reforming catalysts					$103,412	$0	$0	$103,412	$0	$0	$103,412	$0
ZnO					$424,410	$0	$0	$424,410	$0	$0	$424,410	$0
Pressure Swing Adsorption Packing					$497,135	$0	$0	$497,135	$0	$0	$497,135	$0
FT catalysts					$6,127,680	$0	$0	$6,127,680	$0	$0	$6,127,680	$0
Other Variable Costs					$11,238,097	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539
Fixed Operating Costs					$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834
Total Product Cost					$75,794,444	$76,555,967	$76,555,967	$83,813,336	$76,555,967	$76,555,967	$83,813,336	$76,555,967
Annual Depreciation												
General Plant												
DDB					$104,833,121	$74,880,801	$53,486,286	$38,204,490	$27,288,922	$19,492,087	$13,922,919	
SL					$52,416,561	$43,680,467	$37,440,400	$33,428,929	$31,837,075	$31,837,075	$31,837,075	
Remaining Value					$262,082,803	$187,202,002	$133,715,716	$95,511,226	$68,222,304	$48,730,217	$34,807,298	
Actual					$104,833,121	$74,880,801	$53,486,286	$38,204,490	$31,837,075	$31,837,075	$31,837,075	
Steam Plant												
DDB					$4,979,012	$4,605,586	$4,260,167	$3,940,654	$3,645,105	$3,371,722	$3,118,843	$2,884,930
SL					$3,319,341	$3,231,990	$3,155,679	$3,090,709	$3,037,588	$2,997,087	$2,970,327	$2,958,903
Remaining Value					$61,407,813	$56,802,227	$52,542,060	$48,601,405	$44,956,300	$41,584,577	$38,465,734	$35,580,804
Actual					$4,979,012	$4,605,586	$4,260,167	$3,940,654	$3,645,105	$3,371,722	$3,118,843	$2,958,903
Net Revenue					($63,635,790)	$6,585,363	$28,325,297	$36,669,236	$50,589,570	$50,862,953	$43,858,462	$83,112,848
Losses Forward						($63,635,790)	($57,050,426)	($28,725,129)	$0	$0	$0	$0
Taxable Income					($63,635,790)	($57,050,426)	($28,725,129)	$7,944,107	$50,589,570	$50,862,953	$43,858,462	$83,112,848
Income Tax					$0	$0	$0	$3,098,202	$19,729,932	$19,836,551	$17,104,800	$32,414,011
Annual Cash Income					$46,176,343	$86,071,750	$86,071,750	$75,716,179	$66,341,818	$66,235,199	$61,709,581	$53,657,740
Discount Factor		1.21	1.1	1	0.909090909	0.826446281	0.751314801	0.683013455	0.620921323	0.56447393	0.513158118	0.46650738
Annual Present Value	$530,655,988				$41,978,494	$71,133,678	$64,666,980	$51,715,169	$41,193,049	$37,388,043	$31,666,772	$25,031,732
Total Capital Investment + Interest		$50,685,036	$285,979,814	$203,652,292								
Net Present Worth				$0								

Table 33. Discounted cash flow sheet for years 9-20 of LT scenario

DCFROR Worksheet

Year	9	10	11	12	13	14	15	16	17	18	19	20
Fixed Capital Investment												
Working Capital												($64,995,412)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
GGE (Gallon of Gasoline Equiv.) Sales	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719	$156,033,719
Diesel Sales	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
By-Product Credit	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998	$6,593,998
Plant Performance	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Annual Sales	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717	$162,627,717
Annual Manufacturing Cost												
Raw Materials	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594	$51,307,594
WGS catalysts	$0	$104,732	$0	$0	$104,732	$0	$0	$104,732	$0	$0	$104,732	$0
Steam reforming catalysts	$0	$103,412	$0	$0	$103,412	$0	$0	$103,412	$0	$0	$103,412	$0
ZnO	$0	$424,410	$0	$0	$424,410	$0	$0	$424,410	$0	$0	$424,410	$0
Pressure Swing Adsorption Packing	$0	$497,135	$0	$0	$497,135	$0	$0	$497,135	$0	$0	$497,135	$0
FT catalysts	$0	$6,127,680	$0	$0	$6,127,680	$0	$0	$6,127,680	$0	$0	$6,127,680	$0
Other Variable Costs	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539	$12,843,539
Fixed Operating Costs	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834	$12,404,834
Total Product Cost	$76,555,967	$83,813,336	$76,555,967	$76,555,967	$83,813,336	$76,555,967	$76,555,967	$83,813,336	$76,555,967	$76,555,967	$83,813,336	$76,555,967
Annual Depreciation												
General Plant												
DDB												
SL												
Remaining Value												
Actual												
Steam Plant												
DDB	$2,668,560	$2,468,418	$2,283,287	$2,112,040	$1,953,637	$1,807,115	$1,671,581	$1,546,212	$1,430,246	$1,322,978	$1,223,755	$1,131,973
SL	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903
Remaining Value	$32,912,244	$30,443,825	$28,160,538	$26,048,498	$24,094,861	$22,287,746	$20,616,165	$19,069,953	$17,639,706	$16,316,728	$15,092,974	$13,961,001
Actual	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903	$2,958,903
Net Revenue	$83,112,848	$75,855,478	$83,112,848	$83,112,848	$75,855,478	$83,112,848	$83,112,848	$75,855,478	$83,112,848	$83,112,848	$75,855,478	$83,112,848
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$83,112,848	$75,855,478	$83,112,848	$83,112,848	$75,855,478	$83,112,848	$83,112,848	$75,855,478	$83,112,848	$83,112,848	$75,855,478	$83,112,848
Income Tax	$32,414,011	$29,583,637	$32,414,011	$32,414,011	$29,583,637	$32,414,011	$32,414,011	$29,583,637	$32,414,011	$32,414,011	$29,583,637	$32,414,011
Annual Cash Income	$53,657,740	$49,230,744	$53,657,740	$53,657,740	$49,230,744	$53,657,740	$53,657,740	$49,230,744	$53,657,740	$53,657,740	$49,230,744	$53,657,740
Discount Factor	0.424097618	0.385543289	0.350493899	0.318630818	0.28966438	0.263331254	0.239392049	0.217629136	0.197844669	0.17985879	0.163507991	0.148643628
Annual Present Value	$22,756,120	$18,980,583	$18,806,710	$17,097,009	$14,260,393	$14,129,760	$12,845,236	$10,714,044	$10,615,898	$9,650,816	$8,049,620	$7,975,881
Total Capital Investment + Interest												($9,661,153.89)
Net Present Worth												

B.5 Pioneer Plant Analysis Details

Variables used in determining pioneer plant performance (equation 17).

NEWSTEPS (0+): The feedstock handling area was chosen as a new step because of the large scale which has not been demonstrated with biomass. The gasifier and solids feeding are also included as a new step because a pressurized biomass feeding system has not been demonstrated at a commercial scale except for limited campaigns.

BALEQS (0 to 100): The mass and energy balances cannot be validated with current plant data, so a value of zero is chosen.

WASTE (0 to 5): Waste streams for gasification include scrubber sludge, black water, gasifier slag, fly ash, and sulfur. The scrubber sludge and black water requires chemical treatment and the sulfur requires special handling. A mid-range value of is 2.5 chosen.

SOLIDS (0 or 1): Solids are present, therefore a value of 1 is used.

Variables used in determining pioneer plant cost growth (equation 16).

PCTNEW (0 to 100%): The percentage cost of the gasifier, solids pressurizing, and solids feeding out of the total purchased equipment cost.

IMPURITIES (0 to 5): There are two major recycle streams in the gasification process, and there is the possibility of inert component buildup. There is also a potential for equipment corrosion due to sulfur components, hydrogen chloride, and hydrogen, so a value of 4 is assigned.

COMPLEXITY (0+): There are 9 continuously linked steps in the gasification process. These include feedstock handling, solids feeding, gasification, amine scrubbing, sour water-gas-shift, pressure swing adsorption, Fischer-Tropsch synthesis, hydroprocessing, and air separation.

INCLUSIVENESS (0 to 100): Land costs and startup costs are considered in the TCI, however, they have not been rigorously investigated. A value of 33% is used.

PROJECT DEFINITION (2 to 8): The gasification platform is considered to be in the study design stage so a value of 7 was assigned.

Table 34. Pioneer plant analysis parameters and factors

Parameter	Baseline	Optimistic	Pessimistic	Range
NEWSTEPS	2	1	3	0+
BALEQS	0	0	0	0-100
WASTE	4	3	5	0-5
SOLIDS	1	1	1	0 or 1
Plant Perf.	**38.18**	**49.93**	**22.31**	**0-100**
PCTNEW	19 (9)[a]	10 (5)[a]	25 (20)[a]	0-100
IMPURITIES	4	3	5	0-5
COMPLEX	9	6	12	0+
INCLUSIV.	33	50	0	0-100
PROJ. DEF.	7	6	8	2-8
Cost Growth(HT)	**0.47**	**0.63**	**0.30**	**0-1**
Cost Growth(LT)	**0.50**	**0.65**	**0.31**	**0-1**

(a) value in parentheses is value chosen for LT scenario

APPENDIX C. SCENARIO MODELING DETAILS

C.1 Property Method

The model setup includes a particle size distribution in order to better estimate the solids simulation in the grinding and cyclone operations. It operates globally with the Redlich-Kwong-Soave with Boston-Mathias modification (RKS-BM) property method which is recommended for medium temperature refining and gas processing operations including combustion and gasification. During acid gas absorption and stripping another property method, ELECNRTL, is used for more accurate simulation. The solids handling such as in the pretreatment area and cyclones, the SOLIDS property method is used.

C.2 Stream/Block Nomenclature

All streams and blocks within the model follow a specific alphanumeric notation with the purpose of clarity and consistency across scenarios and across platforms. Each area within the model (e.g. Area 200 gasification) has a two letter abbreviation (e.g. gasification is GS). These abbreviations are used for naming both streams as well as blocks. In addition to purposes mentioned above the notation is descriptive (e.g. the notation REAC describes a block as a reactor). Another example is SGAS which describes a stream that contains syngas. ASPEN Plus limits block and stream names to be eight characters.

Figure 18 shows the pattern of notation for a syngas stream in the gasification area:

Area		Number		Description			
G	S	0	1	S	G	A	S

Figure 18. Stream nomenclature used in model

Similarly, the notation for the first reactor block in the gasification area is shown in Figure 19.

Area		Description				Number	
G	S	R	E	A	C	0	1

Figure 19. Block nomenclature used in model

Table 35 contains the abbreviations for areas, unit operation block descriptions, and stream descriptions.

Table 35. Detailed description of stream and block nomenclature

Area	Description	Name	Block	Name	Stream	Name
Plant	All Areas	**PL**	Reactor	**REAC**	Biomass	**BMAS**
A100	Pretreatment	**PR**	Mixer	**MIX**	Steam	**STM**
A100CHOP	Chopping	**CH**	Heat Mixer	**QMX**	Flue gas	**FLUE**
A100DRY	Drying	**DR**	Work Mixer	**WMX**	Syngas	**SGAS**
A100GRIN	Grinding	**GR**	Splitter	**SPL**	Ash	**ASH**
A200	Gasification	**GS**	Separator	**SEP**	Carbon dioxide	**CO2**
A200CYC	Cyclones	**CY**	Cyclones	**CYC**	Air	**AIR**
A200COMB	Combustion	**CB**	Flash Drum	**DRUM**	Hydrogen	**HYD**
A300	Syngas Cleaning	**CL**	Column	**COL**	FT products	**FT**
A300AGR	Acid Gas Removal	**AG**	Distillation	**DIST**	Water	**WAT**
A300SUL	Sulfur Recovery	**SU**	Grinder	**GRIN**	Oxygen	**OX**
A400	Fuel Synthesis	**FS**	Dryer	**DRY**	Sulfur	**SUL**
A400COND	Syngas Conditioning	**CD**	Heater	**HEAT**	Fuel	**FUEL**
A400MTG	Methanol to Gasoline	**MG**	Heat Exchanger	**HX**	Tar	**TAR**
A500	Hydrocracking	**HY**	Tank/Hopper	**TANK**	Char	**CHAR**
A500	Fuel Separation	**SE**	Pump	**PMP**	Acid Gas	**AG**
A600	Power Generation	**PG**	Compressor	**COMP**	Lean MEA soln.	**MEAL**
A700	Air Separation Unit		Turbine	**TURB**	Rich MEA soln.	**MEAR**
					Light gases	**LGAS**
					Nitrogen	**NTGN**

A special notation is used for heat and work streams. In the case that the first reactor in the gasification area includes a heat stream leaving the unit, it follows the nomenclature shown in Figure 20.

Q or W		Area		Block Description			Number
Q	-	G	S	R	E	A	1

Figure 20. Heat and work stream nomenclature used in model

The Q or W sets the stream apart as a heat or work stream. The block description is limited to three characters and number is limited to one character.

C.3 Aspen Plus™ Calculator Block Descriptions

C.3.1 High Temperature scenario

AIRCOMB

This block calculates the nitrogen that accompanies the oxygen in the air inlet for the combustion of unconverted syngas. Molar nitrogen flow (in kmol/hr) is calculated as follows:

$$\dot{M}_{N2} = \left(\frac{0.79}{0.21}\right) \cdot \dot{M}_{O2} \qquad \text{(eqn. 20)}$$

where $\dot{M}_{O2}$ is molar flow of oxygen in kmol/hr.

AMINE

This block calculates the mole flow of monoethanolamine (MEA) needed for the required acid gas removal (CO2 and H2S) arriving from syngas quench and FT unconverted syngas recycle stream. The MEA is able to capture 0.35 moles acid gas per mole MEA. Additionally, the MEA is diluted as explained in *DILUTH2O*.
Molar MEA flow (in kmol/hr) is calculated by

$$\dot{M}_{MEA} = (\dot{M}_{CO2,syn} + \dot{M}_{CO2,rec} + \dot{M}_{H2S,syn})/0.35 \qquad \text{(eqn. 21)}$$

where $\dot{M}_{CO2,syn}$ is molar flow of CO2 from the syngas after the syngas quench, $\dot{M}_{CO2,rec}$ is the molar flow of CO2 from the unconverted syngas recycle after the FT synthesis, and $\dot{M}_{H2S,syn}$ is the molar flow rate of H2S from the syngas quench.

Since the MEA solution in the amine absorption unit is to be 20 wt% concentrated with water, the flow of water must be calculated.
Mole flow of water is calculated as

$$\dot{M}_{H2O} = \frac{\dot{M}_{MEA} * MW_{MEA}/0.20}{MW_{H2O}} \qquad \text{(eqn. 22)}$$

BIOELEM

Because the high temperature gasifier is modeled at equilibrium, the simulation software requires that all components in the input are located in the conventional stream. Therefore, this block splits the biomass into the following compounds based on its ultimate analysis: carbon, hydrogen, oxygen, sulfur, nitrogen, and ash. Water in the biomass is not affected because it is already a conventional component. Biomass in the exit stream is set to zero.

FTDISTR

This calculator block calculates an alpha chain growth parameter using the equation by Song et al. (2004) for cobalt catalyst. Inlet and outlet streams are defined and calculated. FT products include paraffins from C1 through C20. FT waxes are paraffins at C30.

FT reaction is as follows:

$$CO + 2.1 * H_2 \; --> \; -(CH_2)- \; + H_2O \qquad \text{(eqn. 23)}$$

Section 100 sets the CO conversion
Section 200 calculates the reaction extent (in lbmol) based on an alpha value of 0.9

-------------Section 100------------------------

Percent conversion of CO is calculated as follows below and then the molar amount of converted CO (COCONV) is calculated knowing the molar amount of CO entering (COIN).

PERCEN = 40
CONV=PERCEN/100.0
COCONV=COIN*CONV

------------Section 200--------------------------
R1, R2, R3, etc. represent the molar reaction extent (lbmol/hr) that is utilized in the FT reactor for each reaction (i.e. CO + 3*H2 → CH4 + H2O, 2*CO + 5*H2 → C2H6 + 2*H2O, etc.). The coefficients of each reaction extent are calculated by solving a set of 21 equations shown below and as described in section 5 of this appendix

Table 36. Reaction extent equations for each alkane hydrocarbon

Alkane Component	Equation
C1	$R1 = COCONV * 0.01$
C2	$R2 = COCONV * 0.018/2$
C3	$R3 = COCONV * 0.0243/3$
C4	$R4 = COCONV * 0.02916/4$
C5	$R5 = COCONV * 0.0328/5$
C6	$R6 = COCONV * 0.03543/6$
C7	$R7 = COCONV * 0.0372/7$
C8	$R8 = COCONV * 0.03826/8$
C9	$R9 = COCONV * 0.03874/9$
C10	$R10 = COCONV * 0.03874/10$
C11	$R11 = COCONV * 0.03835/11$
C12	$R12 = COCONV * 0.03766/12$
C13	$R13 = COCONV * 0.03672/13$
C14	$R14 = COCONV * 0.03559/14$
C15	$R15 = COCONV * 0.03432/15$
C16	$R16 = COCONV * 0.03294/16$
C17	$R17 = COCONV * 0.0315/17$
C18	$R18 = COCONV * 0.03002/18$
C19	$R19 = COCONV * 0.02852/19$
C20	$R20 = COCONV * 0.02702/20$

C30	$R21 = COCONV * 0.36473/30$

GRIND

This block calculates the power requirement (in kW) for grinding the biomass from the chop size of 15 mm to final size of 1 mm. This power requirement data is found in Mani, et al. and for 12% exiting moisture. The correlation was changed from a polynomial (quadratic) regression, which Mani, et al. used, to a power regression because the power regression more accurately matched the data. S_{cut} is the final grind size in the units of millimeters.

$$P_{grind} = \left(28.76 \cdot {S_{cut}}^{-0.81}\right) \cdot \dot{m}_{biomass} \qquad \text{(eqn. 24)}$$

HRSG

This calc block totals the heat transfer areas of all the heat exchangers in A600 Power Generation for use in Aspen ICARUS costing of a heat recovery steam generator which is estimated as a waste heat boiler.

HUMIDITY

This block sets humidity of the air entering the Air Separation Unit.

HV-101, HV-203, HV-445

This block calculates the lower and higher heating values of the following streams: biomass, syngas, and FT products.

LOCKHOP

This block calculates the CO2 required for pressurizing the lock hopper. Higman et al. reports 0.09 kg of pressurization gas is required per kg of biomass.

$$XCO2 \;=\; 0.09 * BIOMAS \qquad \text{(eqn. 25)}$$

MEATEMP

This block sets the temperature of the incoming monoethanolamine solution entering the absorber column in the AGR area.

MOISTURE

This block sets the moisture content of the entering biomass to the preprocessing area and sets the biomass moisture content exiting the biomass dryer. Also, the steam loop flow rate for drying the biomass is set at 9 times the amount of moisture removed during the drying process.

Moisture content (% wet basis) of entering biomass feed, $XMOIS1 \;=\; 25$. Inlet mass flow of moisture, $WATERI$, is computed.

$$WATERI = FEED * XMOIS1/100/(1 - XMOIS1/100) \quad \text{(eqn. 26)}$$

Moisture content (% wet basis) of biomass exiting the dryer, XMOIS2 = 10. Mass flow of moisture, WATERO, is computed.

$$\mathrm{WATERO} = \mathrm{FEED} * \mathrm{XMOIS2}/100/(1 - \mathrm{XMOIS2}/100) \quad \text{(eqn. 27)}$$

Specify steam required to remove moisture, STEAMI.

$$\mathrm{STEAMI} = 9 * (\mathrm{WATERI} - \mathrm{WATERO}) \quad \text{(eqn. 28)}$$

O2COMB

Oxygen is required to combust the char and syngas that provides the energy necessary for drying the biomass. A system of stoichiometric combustion reactions are setup to sum all the oxygen required to fully combust the unconverted syngas purge from the FT synthesis outlet. The reactions are as follows in Table 37:

Table 37. Combustion reactions to determine required oxygen

Component	**Reaction**
CO	$CO + 0.5 \cdot O2 \rightarrow CO2$
H2	$H2 + 0.5 \cdot O2 \rightarrow H2O$
CH4	$CH4 + 2O2 \rightarrow 2H2O + CO2$
C2H6	$C2H6 + 3.5 \cdot O2 \rightarrow 3H2O + 2CO2$
C2H4	$C2H4 + 3O2 \rightarrow 2H2O + 2CO2$
C2H2	$C2H2 + 2.5O2 \rightarrow H2O + CO2$
C3H8	$C3H8 + 5O2 \rightarrow 4H2O + 3CO2$
C4H10	$C4H10 + 6.5O2 \rightarrow 5H2O + 4CO2$
C5H12	$C5H12 + 8O2 \rightarrow 6H2O + 5CO2$
C6H14	$C6H14 + 9.5O2 \rightarrow 7H2O + 6CO2$
C7H16	$C7H16 + 11O2 \rightarrow 8H2O + 7CO2$
C8H18	$C8H18 + 12.5O2 \rightarrow 9H2O + 8CO2$
C9H20	$C9H20 + 14O2 \rightarrow 10H2O + 9CO2$
Tar	$C14H10(tar) + 16.5O2 \rightarrow 5H2O + 14CO2$
H2S	$H2S + 1.5O2 \rightarrow H2O + SO2$
NH3	$NH3 + 1.75O2 \rightarrow 1.5H2O + NO2$

The molar flow rate of oxygen entering the combustor is summed and multiplied by factor of 1.25 in order to combust with 25% excess air as shown in equation below.

$$\dot{M}_{O2,in} = 1.25 \cdot (\dot{M}_{CHAR,in} + 0.5\dot{M}_{CO,in} + 0.5\dot{M}_{H2,in} + 2\dot{M}_{CH4,in} + 3.5\dot{M}_{C2H6,in} + 3\dot{M}_{C2H4,in} + 2.5\dot{M}_{C2H2,in} + 5\dot{M}_{C3H8,in} + 6.5\dot{M}_{C4H10,in} + 8\dot{M}_{C5H12,in} + 9.5\dot{M}_{C6H14,in} + 11\dot{M}_{C7H16,in} + 12.5\dot{M}_{C8H18,in} + 14\dot{M}_{C9H20,in} + 16.5\dot{M}_{TAR,in} + 1.5\dot{M}_{H2S,in} + 1.75\dot{M}_{NH3,in}) \quad \text{(eqn. 29)}$$

O2TURB

This block calculates the molar flow rate of air (oxygen and nitrogen) required to combust syngas obtained from FT synthesis and the fuel gas obtained from Area 500 in the gas turbine of Area 600. A excess 25% air is assumed. The calculations are similar to the methodology in ***O2COMB***.

OXYSET

This block sets the entering oxygen at 0.35 lb oxygen per lb dry biomass into the gasifier.

$$\dot{m}_{O2,gas} = 0.35/100 \cdot \dot{m}_{biomass} \qquad \text{(eqn. 30)}$$

SWGSSTM

This block sets the steam flow into the sour water-gas-shift reactor to be at a ratio of 3:1 water to carbon monoxide. This ratio ensures enough water-gas-shift activity occurs within the reactor.

$$\dot{m}_{STM,addition} = 3.0 \cdot \dot{m}_{CO} - \dot{m}_{H2O} \qquad \text{(eqn. 31)}$$

C.3.2 Low Temperature scenario

AMINE

This block calculates the mole flow of monoethanolamine (MEA) needed for the required acid gas removal (CO2 and H2S) arriving from syngas quench and FT unconverted syngas recycle stream. The MEA is able to capture 0.35 moles acid gas per mole MEA. Additionally, the MEA is diluted as explained in *DILUTH2O*.
Molar MEA flow (in kmol/hr) is calculated by

$$\dot{M}_{MEA} = (\dot{M}_{CO2,syn} + \dot{M}_{CO2,rec} + \dot{M}_{H2S,syn})/0.35 \qquad \text{(eqn. 32)}$$

where $\dot{M}_{CO2,syn}$ is molar flow of CO2 from the syngas after the syngas quench, $\dot{M}_{CO2,rec}$ is the molar flow of CO2 from the unconverted syngas recycle after the FT synthesis, and $\dot{M}_{H2S,syn}$ is the molar flow rate of H2S from the syngas quench.

Since the MEA solution in the amine absorption unit is to be 20 wt% concentrated with water, the flow of water must be calculated.
Mole flow of water is calculated as

$$\dot{M}_{H2O} = \frac{\dot{M}_{MEA} * MW_{MEA}/0.20}{MW_{H2O}} \qquad \text{(eqn. 33)}$$

BIOELEM

Same as for the HT scenario

DILUTH2O

This block sets the MEA solution to be 20% concentrated with water.

FTDISTR

Same as High Temperature scenario

GASYIELD

The following model describes how the fluidized bed gasifier keeps an elemental mass balance. Experiments performed at Iowa State University provide the initial gasifier product distribution and the model adjusts the yields of those experiments in order to balance carbon, hydrogen, sulfur, nitrogen, oxygen and ash.

The approach taken to balance each element across the gasifier is by "floating" a component of each element. The "floating" component for element carbon is the char. All sulfur and nitrogen not found in the char is assumed to form hydrogen sulfide and ammonia, respectively. Therefore, sulfur and nitrogen balance. Next, elemental hydrogen is adjusted in the model by either converting diatomic hydrogen to steam or decomposing steam to diatomic hydrogen. Oxygen balance is more complex. Since gasification operates at fuel rich conditions, diatomic oxygen should not present in the syngas leaving the gasifier. Therefore, diatomic oxygen cannot be the "floating" component. Instead, oxygen is balanced by adjusting the carbon monoxide or carbon dioxide in the exiting syngas. Since there is one oxygen difference between those two components, the oxygen can be adjusted to help close the balance.

Carbon balance follows the flow chart shown in Figure 21. If there is less gaseous carbon out than total carbon in, then the difference is made up of char carbon, CCARB. Char is assumed to be comprised of 68% carbon with the rest as H, O, N, and S. Ash is considered apart from the char and is considered inert in the model. Since the char is now fixed, the only pathway for sulfur and nitrogen to take is to form hydrogen sulfide and ammonia. Therefore, the sulfur and nitrogen balance.

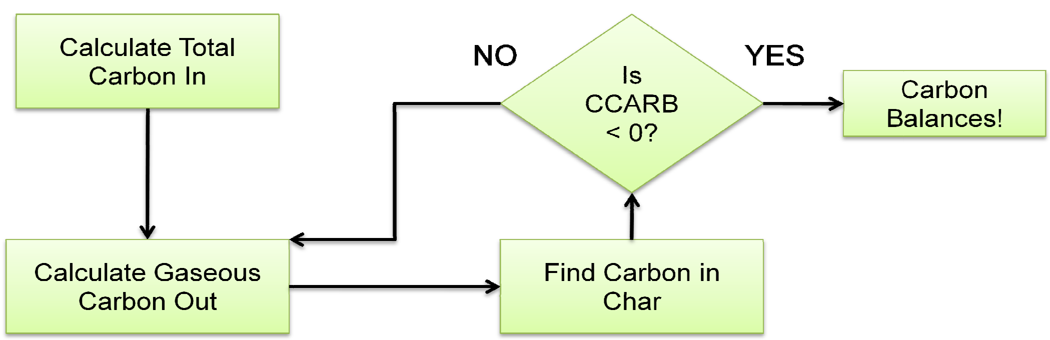


Figure 21. Decision diagram for carbon balance

Next, as show in Figure 22, hydrogen is balanced. Knowing hydrogen in the char and in gaseous products, the hydrogen required (HREQD) is calculated as the sum of those two components. If the hydrogen required is less than hydrogen available (HAVAIL), made up of hydrogen in steam, biomass moisture, and in the biomass itself (THYD), then there is enough hydrogen available to balance. To balance hydrogen, the product yield swings towards either steam or diatomic hydrogen.

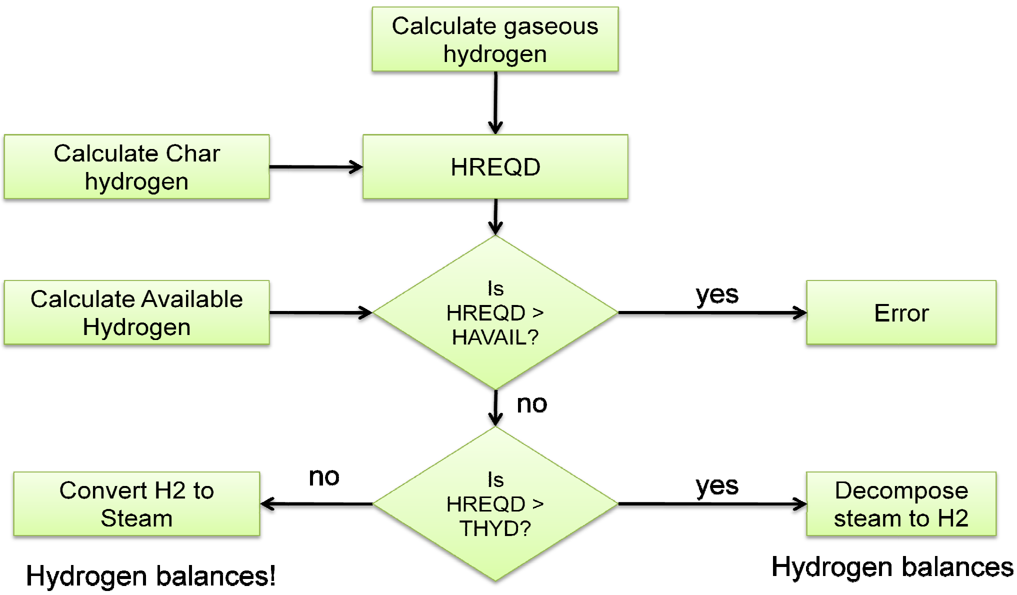


Figure 22. Decision diagram for hydrogen balance

The only element left to balance is oxygen which is accomplished by forcing creation of carbon monoxide or creation of carbon dioxide as shown in Figure 23. The required oxygen (OREQD), made up of oxygen in char and oxygen in syngas, is checked against the available oxygen found in the entering oxygen, steam, and biomass. If there is more oxygen available than required, then the option is to move the excess oxygen to CO_2 by decreasing CO. If there is still oxygen present when CO is decreased to zero, then the yields need to be adjusted since excess oxygen is still present. If there is an oxygen deficit (OREQD > OAVAIL), then CO is increased and CO_2 is decreased. After that, if there is still an oxygen deficit, then insufficient oxygen is present and yields need to be adjusted. When all these steps are completed and no errors generated, there is an elemental mass balance across the gasifier.

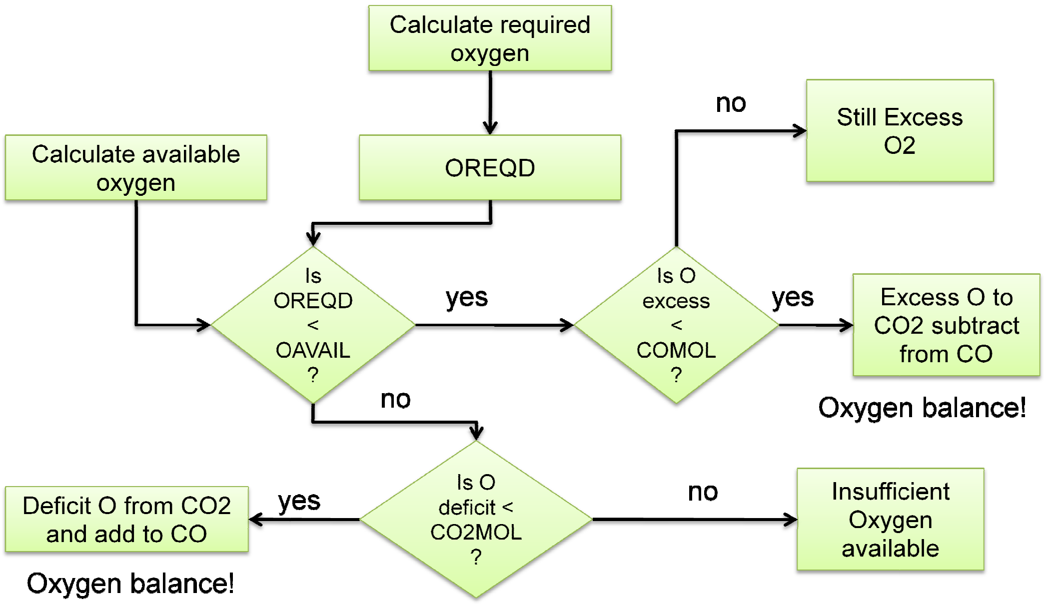


Figure 23. Decision diagram for oxygen balance

GRIND

This block calculates the power requirement (in kW) for grinding the biomass from the chop size of 12 mm to final size of 6 mm. This power requirement data is found in Mani, et al. and for 12% exiting moisture. The correlation has changed from a polynomial regression (which Mani, et al. used) to a power regression because the power regression fit the data better. is in millimeters.

(eqn. 34)

HUMIDITY

This block sets humidity of the air entering the Air Separation Unit.

HV-101, HV-203, HV-445

This block calculates the lower and higher heating values of the following streams: biomass, syngas, and FT products.

MOISTURE

This block is the same as found in the HT scenario.

O2COMB

This block is the same as found in the HT scenario.

O2TURB

This block is the same as found in the HT scenario.

OXYSET

This block sets the entering oxidizing agents, oxygen and steam, into the gasifier. A linear correlation with temperature, T_{gas} (in Fahrenheit), adapted from Bain for oxygen is used because as oxygen increases in the gasifier the temperature increases. Mass flow of oxygen, $\dot{m}_{O2,gas}$, is in percentage of dry feedstock.

$$\dot{m}_{O2,gas} = (-11.567 + 0.02375 \cdot T_{gas})/100 \cdot \dot{m}_{biomass} \qquad \text{(eqn. 35)}$$

The steam feed rate is set at 0.66 lb steam per lb oxygen.

$$\dot{m}_{steam,gas} = 0.66 \cdot \dot{m}_{O2,gas} \qquad \text{(eqn. 36)}$$

Since 95% purity oxygen is produced in the Air Separation Unit, argon mass flow is set at 5% of molar oxygen flow.

$$\dot{m}_{argon} = 0.05 \cdot \left(\frac{\dot{m}_{O2,gas}}{MW_{O2}}\right) * MW_{Ar} \qquad \text{(eqn. 1)}$$

C.4 Aspen Plus™ Design Specifications

C.4.1 High Temperature Scenario

DS-1

The exiting temperature of air in the heat exchanger used to pre-cool the air entering the cryogenic distillation column is varied until a net duty of zero is observed.

FSSPL02

This design specification varies the fraction of unconverted syngas that is piped to area 200 for the combustion of syngas. The syngas, in turn, provides the heat required to dry the biomass.

H2SPLIT

This design spec calculates the required hydrogen that needs to be reserved by the PSA unit for use in Area 500: Hydrocracking. A typical yield from hydrocracking is shown in the table below. Since the FT products are be hydrogen deficient relative to the final blend, then make-up hydrogen is required. The syngas purge amount going to the pressure swing adsorption (PSA) unit is varied so that the calculated delivered hydrogen matches the required hydrogen to Area 500. Without showing the detailed calculations, the basic steps are first calculating the carbon and hydrogen content in the FT product stream. The carbon mass flow is the same as that of the final blend stream flow. Using the blend fractions in Table 38, the amount of hydrogen is calculated in the final blend and the difference in hydrogen is determined. The difference is multiplied by 1.1 to obtain the delivered hydrogen mass flow rate to hydrocracking area.

Table 38. Hydroprocessing product blend

Component	Mass Fraction
Fuel Gas (methane)	0.034
LPG (propane)	0.088
Gasoline (n-octane)	0.261
Diesel (n-hexadecane)	0.617

O2-101, O2-203, O2-445

These design specifications vary the amount of oxygen inlet to the Heating Value blocks (HV-101, HV-203, HV-445) so as to be stoichiometric in the combustion of the duplicate stream.

O2-SULF

This design specification varies the amount of oxygen into the LO-CAT oxidizer unit to fully oxidize the H2S into solid sulfur.

SGSTEMP

The temperature of operation in the sour water-gas-shift reactor is varied until the exiting equilibrium molar ratio of H_2/CO is just above the optimal FT ratio (2.1). A small amount of hydrogen is captured in the PSA unit bringing that ratio down to the optimum for FT synthesis.

C.4.2 Low Temperature scenario

DS-1

This design specification is the same as HT scenario.

H2SPLIT

This design specification is the same as HT scenario.

O2-101, O2-203, O2-445

These design specifications are the same as in the HT scenario.

O2-SULF

This design specification is the same as HT scenario.

STMRECOV

Heat can be recovered from the combustion of syngas and char. This specification varies the steam flow rate (stream 280) to bring the combustion flue gas (stream 252) down to 200 C via heat exchanging.

WGSTEMP

The temperature of operation in the water-gas-shift reactor is varied until the exiting equilibrium molar ratio of H_2/CO is just above the optimal FT ratio (2.1). A small amount of hydrogen is captured in the PSA unit bringing that ratio down to the optimum for FT synthesis.

C.5 Detailed Calculations

ASPEN Model Calculations and Notes

Outline

- Plant Input
- Plant Output
- Carbon Efficiency to Fuels
- Energy Content
- FT Reaction Conversion Solver
- Equipment Sizing
    - Dryer
    - Lock hoppers
    - Slag/Char Collection
    - PSA Unit
    - Fuel Storage
- LT Gasifier Cost
- FT Reactor Cost
- Acid Gas Removal Area Cost
- A500 Hydroprocessing Area Cost
- Reactors and Catalysts
- Natural Gas Utility Usage

Defining Units

$MJ := 10^{6} J$

$MMcf := 10^{6} ft^{3}$

$kPa := 10^{3} \cdot Pa$

$Cp := 100 poise$

$\rho_{water} := 1000 \frac{kg}{m^{3}}$

$MW_{H2O} := 18.02 \frac{gm}{mol}$

$kmol := 1000 mol$

$MMBTU := 10^{6} BTU$

$lbmol := \frac{kmol}{2.2}$

$HHV_{stover} := 17.65 \frac{MJ}{kg}$

$bbl := 42 gal$

$HHV_{stover} = 7.588 \times 10^{3} \frac{BTU}{lb}$

$\rho_{gas} := 737.22 \frac{kg}{m^{3}}$

$100 \frac{kg}{m^{3}} = 6.243 \frac{lb}{ft^{3}}$

$\rho_{diesel} := 840 \frac{kg}{m^{3}}$

$therm := 100000 BTU$

$MMgal := 10^{6} gal$

$dekatherm := 10 therm$

$kJ := 1000 J$

$P_{ref} \equiv 1 atm$

$bpsd := \frac{42 gal}{day}$

$T_{ref} \equiv 298 K$

$PJ := 10^{15} J$

$GJ := 10^{9} J$

Plant Input

Biomass

$m_{dot\_biomass} := 2000 \frac{tonne}{day}$ Availability $:= 310 day$ Load $:= 7446 hr$

Elemental Composition

Carbon	$Frac_{C\_biomass} \equiv 0.4728$	$MW_C := 12.01 \frac{gm}{mol}$
Oxygen	$Frac_{O\_biomass} \equiv 0.4063$	$MW_O := 16. \frac{gm}{mol}$
Hydrogen	$Frac_{H\_biomass} \equiv 0.0506$	$MW_H := 1.01 \frac{gm}{mol}$
Sulfur	$Frac_{S\_biomass} \equiv 0.0022$	$MW_S := 32.07 \frac{gm}{mol}$
Nitrogen	$Frac_{N\_biomass} \equiv 0.008$	$MW_N := 14.01 \frac{gm}{mol}$
Ash	$Frac_{A\_biomass} \equiv 0.0600$	

Elemental Mass Flow

$m_{dot\_C\_in} := m_{dot\_biomass} \cdot Frac_{C\_biomass}$ $\qquad m_{dot\_C\_in} = 945.6 \frac{tonne}{day}$

$m_{dot\_O\_in} := m_{dot\_biomass} \cdot Frac_{O\_biomass}$ $\qquad m_{dot\_O\_in} = 812.6 \frac{tonne}{day}$

$m_{dot\_H\_in} := m_{dot\_biomass} \cdot Frac_{H\_biomass}$ $\qquad m_{dot\_H\_in} = 101.2 \frac{tonne}{day}$

$m_{dot\_S\_in} := m_{dot\_biomass} \cdot Frac_{S\_biomass}$ $\qquad m_{dot\_S\_in} = 4.4 \frac{tonne}{day}$

$m_{dot\_N\_in} := m_{dot\_biomass} \cdot Frac_{N\_biomass}$ $\qquad m_{dot\_N\_in} = 16 \frac{tonne}{day}$

$m_{dot\_A\_in} := m_{dot\_biomass} \cdot Frac_{A\_biomass}$ $\qquad m_{dot\_A\_in} = 120 \frac{tonne}{day}$

Elemental Mole Flow

$$n_{dot\_C\_in} := \frac{m_{dot\_C\_in}}{MW_C} \qquad n_{dot\_C\_in} = 911.278\frac{mol}{s}$$

$$n_{dot\_O\_in} := \frac{m_{dot\_O\_in}}{MW_O} \qquad n_{dot\_O\_in} = 587.818\frac{mol}{s}$$

$$n_{dot\_H\_in} := \frac{m_{dot\_H\_in}}{MW_H} \qquad n_{dot\_H\_in} = 1160\frac{mol}{s}$$

$$n_{dot\_S\_in} := \frac{m_{dot\_S\_in}}{MW_S} \qquad n_{dot\_S\_in} = 1.588\frac{mol}{s}$$

$$n_{dot\_N\_in} := \frac{m_{dot\_N\_in}}{MW_N} \qquad n_{dot\_N\_in} = 13.218\frac{mol}{s}$$

Biomass Moisture

$moist_{in} := 0.25 \qquad moist_{dried} := 0.10$

$$m_{dot\_moist\_in} := \frac{moist_{in} \cdot m_{dot\_biomass}}{1 - moist_{in}} \qquad m_{dot\_moist\_in} = 666.667\frac{tonne}{day}$$

$$m_{dot\_moist\_dried} := \frac{moist_{dried} \cdot m_{dot\_biomass}}{1 - moist_{dried}} \qquad m_{dot\_moist\_dried} = 222.222\frac{tonne}{day}$$

$\rho_{bulk\_stover} := 100\frac{kg}{m^3}$ Source: *Kaliyan and Morey, 2005* for 0.66-0.8 mm sized particles

HT Gasifier Steam/Oxygen addition Source: *Probstein and Hicks, 2006*

Stoichiometric/thermoneutral requirement for synthesis gas according to following equation:

1.34C + 0.34 O2 + H2O --> 0.34CO2 + CO + H2

Oxygen to Carbon: 0.25
Steam to Carbon : 0.75

$$m_{dot\_O2\_in} := 0.35 \cdot m_{dot\_biomass} \qquad m_{dot\_O2\_in} = 700\,\frac{tonne}{day}$$

$$n_{dot\_O2\_in} := \frac{m_{dot\_O2\_in}}{2 \cdot MW_{O}} \qquad n_{dot\_O2\_in} = 253.183\,\frac{mol}{s}$$

$$Ratio_{O2\_to\_C} := \frac{n_{dot\_O2\_in}}{n_{dot\_C\_in}} \qquad Ratio_{O2\_to\_C} = 0.278$$

Steam addition ratio is then three times that of Oxygen minus the moisture in the biomass

$$n_{dot\_H2O\_in} := 3 \cdot Ratio_{O2\_to\_C} \cdot n_{dot\_C\_in} - \frac{m_{dot\_moist\_dried}}{MW_{H2O}} \qquad n_{dot\_H2O\_in} = 616.817\,\frac{mol}{s}$$

$$m_{dot\_H2O\_in} := n_{dot\_H2O\_in} \cdot MW_{H2C} \qquad m_{dot\_H2O\_in} = 960\,\frac{tonne}{day}$$

Plant Output

HT Fuel production

$m_{dot\_gasHT} := 112.78 \frac{tonne}{day}$ $\qquad$ $m_{dot\_dieselHT} := 266.5 \frac{tonne}{day}$

$v_{dot\_gasHT} := \frac{m_{dot\_gasHT}}{\rho_{gas}}$ $\qquad$ $v_{dot\_dieselHT} := \frac{m_{dot\_dieselHT}}{\rho_{diesel}}$

$v_{dot\_gasHT} = 40413 \frac{gal}{day}$ $\qquad$ $v_{dot\_dieselHT} = 83812 \frac{gal}{day}$

$v_{dot\_gasHT} = 962 \frac{bbl}{day}$ $\qquad$ $v_{dot\_dieselHT} = 1996 \frac{bbl}{day}$

$v_{dot\_gasoline\_per\_year} := v_{dot\_gasHT} \cdot Load$ $\qquad$ $v_{dot\_gasoline\_per\_year} = 12.538 MMgal$

$v_{dot\_diesel\_per\_year} := v_{dot\_dieselHT} \cdot Load$ $\qquad$ $v_{dot\_diesel\_per\_year} = 26.003 MMgal$

LT Fuel production

$m_{dot\_gasLT} := 87.12 \frac{tonne}{day}$ $\qquad$ $m_{dot\_dieselLT} := 205.86 \frac{tonne}{day}$

$v_{dot\_gasLT} := \frac{m_{dot\_gasLT}}{\rho_{gas}}$ $\qquad$ $v_{dot\_dieselLT} := \frac{m_{dot\_dieselLT}}{\rho_{diesel}}$

$v_{dot\_gasLT} = 31218 \frac{gal}{day}$ $\qquad$ $v_{dot\_dieselLT} = 64741 \frac{gal}{day}$

$v_{dot\_gasLT} = 743 \frac{bbl}{day}$ $\qquad$ $v_{dot\_dieselLT} = 1541 \frac{bbl}{day}$

$v_{dot\_gasoline\_per\_yearLT} := v_{dot\_gasLT} \cdot Load$ $\qquad$ $v_{dot\_gasoline\_per\_yearLT} = 9.685 MMgal$

$v_{dot\_diesel\_per\_yearLT} := v_{dot\_dieselLT} \cdot Load$ $\qquad$ $v_{dot\_diesel\_per\_yearLT} = 20.086 MMgal$

Carbon Efficiency to Fuels

HT scenario

Gasoline Carbon

$m_{dot\_gasHT} = 112.78\frac{tonne}{day}$ $\quad Frac_{C\_gasoline} := \frac{8\cdot 12.01}{8\cdot 12.01 + 18\cdot 1.01}$ $\quad Frac_{C\_gasoline} = 0.841$

$m_{dot\_C\_gasHT} := Frac_{C\_gasoline}\cdot m_{dot\_gasHT}$ $\quad m_{dot\_C\_gasHT} = 94.835\frac{tonne}{day}$

Diesel Carbon

$m_{dot\_dieselHT} = 266.5\frac{tonne}{day}$

$Frac_{C\_diesel} := \frac{16\cdot 12.01}{16\cdot 12.01 + 34\cdot 1.01}$ $\quad Frac_{C\_diesel} = 0.848$

$m_{dot\_C\_dieselHT} := Frac_{C\_diesel}\cdot m_{dot\_dieselHT}$ $\quad m_{dot\_C\_dieselHT} = 226.096\frac{tonne}{day}$

$m_{dot\_C\_outHT} := m_{dot\_C\_gasHT} + m_{dot\_C\_dieselHT}$

$m_{dot\_C\_outHT} = 320.931\frac{tonne}{day}$ $\quad C\_effHT := \frac{m_{dot\_C\_outHT}}{m_{dot\_C\_in}}$ $\quad C\_effHT = 0.339$

LT scenario

Gasoline Carbon

$$m_{dot\_gasLT} = 87.12\frac{tonne}{day}$$

$$m_{dot\_C\_gasLT} := Frac_{C\_gasoline} \cdot m_{dot\_gasLT} \qquad m_{dot\_C\_gasLT} = 73.258\frac{tonne}{day}$$

Diesel Carbon

$$m_{dot\_dieselLT} = 205.86\frac{tonne}{day}$$

$$m_{dot\_C\_dieselLT} := Frac_{C\_diesel} \cdot m_{dot\_dieselLT} \qquad m_{dot\_C\_dieselLT} = 174.649\frac{tonne}{day}$$

$$m_{dot\_C\_outLT} := m_{dot\_C\_gasLT} + m_{dot\_C\_dieselLT}$$

$$m_{dot\_C\_outLT} = 247.908\frac{tonne}{day} \qquad C\_effLT := \frac{m_{dot\_C\_outLT}}{m_{dot\_C\_in}} \qquad C\_effLT = 0.262$$

Energy Content

This section aquires the energy content (on a LHV basis) from the Aspen data and converts to megawatts for use in developing an energy balance

Biomass

$E_{biomass} := 1400313\frac{MJ}{hr}$ $E_{biomass} = 388.976MW$

Fuel

$E_{fuelHT} := 695598\frac{MJ}{hr}$ $E_{fuelHT} = 193.222MW$

$E_{fuelLT} := 539292\frac{MJ}{hr}$ $E_{fuelLT} = 149.803MW$

Char/Tar

$E_{char\_LT} := 87792\frac{MJ}{hr}$ $E_{char\_LT} = 24.387MW$

$E_{tar\_LT} := 16980\frac{MJ}{hr}$ $E_{tar\_LT} = 4.717MW$

Raw Syngas

$E_{rawsyngas\_HT} := 1230712\frac{MJ}{hr}$ $E_{rawsyngas\_HT} = 341.864MW$

$E_{rawsyngas\_LT} := 964054\frac{MJ}{hr}$ $E_{rawsyngas\_LT} = 267.793MW$

Energy loss across the gasifier

Energy lost across the gasifier is calculated as difference in energy in the biomass and energy in the raw syngas and char (only in LT scenario)

$E_{gasifierloss\_HT} := E_{biomass} - E_{rawsyngas\_HT}$ $E_{gasifierloss\_HT} = 47.111MW$

$E_{gasifierloss\_LT} := E_{biomass} - E_{rawsyngas\_LT} - E_{char\_LT}$ $E_{gasifierloss\_LT} = 96.796MW$

Unconverted Syngas used in A600 Power Generation

$E_{syngasA600\_HT} := 129332 \frac{MJ}{hr}$ $E_{syngasA600\_HT} = 35.926MW$

$E_{syngasA600\_LT} := 109708 \frac{MJ}{hr}$ $E_{syngasA600\_LT} = 30.474MW$

Fuel Gas from A500 used in A600 Power Generation

$E_{fuelgas\_HT} := 104114 \frac{MJ}{hr}$ $E_{fuelgas\_HT} = 28.921MW$

$E_{fuelgas\_LT} := 80718 \frac{MJ}{hr}$ $E_{fuelgas\_LT} = 22.422MW$

Fischer-Tropsch product

$E_{FTliquids\_HT} := 782894 \frac{MJ}{hr}$ $E_{FTliquids\_HT} = 217.471MW$

$E_{FTliquids\_LT} := 606801 \frac{MJ}{hr}$ $E_{FTliquids\_LT} = 168.556MW$

Electricity Generated

$E_{elecgenOUT\_HT} := 48.55MW$ $E_{elecgenOUT\_LT} := 40.73MW$

Net Electricity (exported)

$E_{elecnet\_HT} := 13.8MW$ $E_{elecnet\_LT} := 16.3MW$

Power Generation loss

The loss is the difference between electric generation out and the gas energy in

$$E_{A600losses\_HT} := E_{syngasA600\_HT} + E_{fuelgas\_HT} - E_{elecgenOUT\_HT}$$

$$E_{A600losses\_HT} = 16.296MW$$

$$E_{A600losses\_LT} := E_{syngasA600\_LT} + E_{fuelgas\_LT} - E_{elecgenOUT\_LT}$$

$$E_{A600losses\_LT} = 12.166MW$$

Loss across FT reactor

$E_{FTreactorlosses\_HT} := 226737\frac{MJ}{hr}$ $E_{FTreactorlosses\_HT} = 62.983MW$

$E_{FTreactorlosses\_LT} := 175128\frac{MJ}{hr}$ $E_{FTreactorlosses\_LT} = 48.647MW$

Unconverted Syngas used for biomass drying

Only in HT scenario

$E_{biomass\_drying\_HT} := 24663\frac{MJ}{hr}$ $E_{biomass\_drying\_HT} = 6.851MW$

Fischer-Tropsch Reaction Conversion Solver

This section solves for the reaction fractional conversion for each reaction in the Fischer-Tropsch reactor. A set of equations is developed and solved. The resulting ε values (ε1 - ε30) are used directly in the Aspen Plus conversion reactor block. The reactions in the reactor block are defined as molar extent.

Depending on the alpha chain growth probability, the reactor forms different product composition.

Step 1: choose the expected alpha chain growth value

$$\alpha_{FT} := 0.9$$

Step 2: using the α_{FT} chain growth, the mole fraction of each hydrocarbon in the FT product is calculated.

$$M1 := \alpha_{FT}^{1-1}\cdot\left(1-\alpha_{FT}\right) \qquad M1 = 0.1 \qquad M11 := \alpha_{FT}^{11-1}\cdot\left(1-\alpha_{FT}\right) \qquad M11 = 0.035$$

$$M2 := \alpha_{FT}^{2-1}\cdot\left(1-\alpha_{FT}\right) \qquad M2 = 0.09 \qquad M12 := \alpha_{FT}^{12-1}\cdot\left(1-\alpha_{FT}\right) \qquad M12 = 0.031$$

$$M3 := \alpha_{FT}^{3-1}\cdot\left(1-\alpha_{FT}\right) \qquad M3 = 0.081 \qquad M13 := \alpha_{FT}^{13-1}\cdot\left(1-\alpha_{FT}\right) \qquad M13 = 0.028$$

$$M4 := \alpha_{FT}^{4-1}\cdot\left(1-\alpha_{FT}\right) \qquad M4 = 0.073 \qquad M14 := \alpha_{FT}^{14-1}\cdot\left(1-\alpha_{FT}\right) \qquad M14 = 0.025$$

$$M5 := \alpha_{FT}^{5-1}\cdot\left(1-\alpha_{FT}\right) \qquad M5 = 0.066 \qquad M15 := \alpha_{FT}^{15-1}\cdot\left(1-\alpha_{FT}\right) \qquad M15 = 0.023$$

$$M6 := \alpha_{FT}^{6-1}\cdot\left(1-\alpha_{FT}\right) \qquad M6 = 0.059 \qquad M16 := \alpha_{FT}^{16-1}\cdot\left(1-\alpha_{FT}\right) \qquad M16 = 0.021$$

$$M7 := \alpha_{FT}^{7-1}\cdot\left(1-\alpha_{FT}\right) \qquad M7 = 0.053 \qquad M17 := \alpha_{FT}^{17-1}\cdot\left(1-\alpha_{FT}\right) \qquad M17 = 0.019$$

$$M8 := \alpha_{FT}^{8-1}\cdot\left(1-\alpha_{FT}\right) \qquad M8 = 0.048 \qquad M18 := \alpha_{FT}^{18-1}\cdot\left(1-\alpha_{FT}\right) \qquad M18 = 0.017$$

$$M9 := \alpha_{FT}^{9-1}\cdot\left(1-\alpha_{FT}\right) \qquad M9 = 0.043 \qquad M19 := \alpha_{FT}^{19-1}\cdot\left(1-\alpha_{FT}\right) \qquad M19 = 0.015$$

$$M10 := \alpha_{FT}^{10-1}\cdot\left(1-\alpha_{FT}\right) \qquad M10 = 0.039 \qquad M20 := \alpha_{FT}^{20-1}\cdot\left(1-\alpha_{FT}\right) \qquad M20 = 0.014$$

All hydrocarbons greater than C20 make up the balance and modeled using C30.

$$M30 := 1 - \begin{pmatrix} M1 + M2 + M3 + M4 + M5 + M6 + M7 + M8 + M9 + M10 \ldots \\ + M11 + M12 + M13 + M14 + M15 + M16 + M17 + M18 + M19 + M20 \end{pmatrix}$$

$$M30 = 0.122$$

Step 3: Setup a series of equations to solve along with guess values (required for Mathcad)

For a nominal 1000 moles of CO input, the expected CO output is 600 moles since 40% is converted.

Known $CO_{out} := 600$ $CO_{in} := 1000$ <---------------- 40% conversion of CO

Guess $\varepsilon 1 := 20$ $\varepsilon 2 := 20$ $\varepsilon 3 := 20$ $\varepsilon 4 := 20$ $\varepsilon 5 := 20$

$\varepsilon 6 := 20$ $\varepsilon 7 := 20$ $\varepsilon 8 := 20$ $\varepsilon 9 := 20$ $\varepsilon 10 := 20$

$\varepsilon 11 := 20$ $\varepsilon 12 := 20$ $\varepsilon 13 := 20$ $\varepsilon 14 := 20$ $\varepsilon 15 := 20$

$\varepsilon 16 := 20$ $\varepsilon 17 := 20$ $\varepsilon 18 := 20$ $\varepsilon 19 := 20$ $\varepsilon 20 := 20$

$\varepsilon 30 := 20$ $D := 0.1$ <----- This value to be varied until COconv is equal to desired.

A nominal 400 moles of CO are converted in the FT reactor. The sum of the exiting amount of moles in the FT product distribution will not be 400, since moles are not conserved. Mass is conserved, however. Therefore, the variable "D" represents a factor that adjusts all the conversions (ε1, ε2, etc.).

The resulting value of D is 0.1 meaning that 40 moles of FT products exit the reactor.

Given

$$D = \left(\varepsilon 1 + \frac{1}{2}\varepsilon 2 + \frac{1}{3}\varepsilon 3 + \frac{1}{4}\varepsilon 4 + \frac{1}{5}\varepsilon 5 + \frac{1}{6}\varepsilon 6 + \frac{1}{7}\varepsilon 7 + \frac{1}{8}\varepsilon 8 + \frac{1}{9}\varepsilon 9 + \frac{1}{10}\varepsilon 10 + \frac{1}{11}\varepsilon 11 + \frac{1}{12}\varepsilon 12 \ldots \\ + \frac{1}{13}\varepsilon 13 + \frac{1}{14}\varepsilon 14 + \frac{1}{15}\varepsilon 15 + \frac{1}{16}\varepsilon 16 + \frac{1}{17}\varepsilon 17 + \frac{1}{18}\varepsilon 18 + \frac{1}{19}\varepsilon 19 + \frac{1}{20}\varepsilon 20 + \frac{1}{30}\varepsilon 30\right)$$

$$M1 = \frac{\varepsilon 1}{D} \quad M2 = \frac{\frac{1}{2}\varepsilon 2}{D} \quad M3 = \frac{\frac{1}{3}\varepsilon 3}{D} \quad M4 = \frac{\frac{1}{4}\varepsilon 4}{D} \quad M5 = \frac{\frac{1}{5}\varepsilon 5}{D} \quad M6 = \frac{\frac{1}{6}\varepsilon 6}{D}$$

$$M7 = \frac{\frac{1}{7}\varepsilon 7}{D} \quad M8 = \frac{\frac{1}{8}\varepsilon 8}{D} \quad M9 = \frac{\frac{1}{9}\varepsilon 9}{D} \quad M10 = \frac{\frac{1}{10}\varepsilon 10}{D} \quad M11 = \frac{\frac{1}{11}\varepsilon 11}{D} \quad M12 = \frac{\frac{1}{12}\varepsilon 12}{D}$$

$$M13 = \frac{\frac{1}{13}\varepsilon 13}{D} \quad M14 = \frac{\frac{1}{14}\varepsilon 14}{D} \quad M15 = \frac{\frac{1}{15}\varepsilon 15}{D} \quad M16 = \frac{\frac{1}{16}\varepsilon 16}{D} \quad M17 = \frac{\frac{1}{17}\varepsilon 17}{D} \quad M18 = \frac{\frac{1}{18}\varepsilon 18}{D}$$

$$M19 = \frac{\frac{1}{19}\varepsilon 19}{D} \quad M20 = \frac{\frac{1}{20}\varepsilon 20}{D} \quad M30 = \frac{\frac{1}{30}\varepsilon 30}{D}$$

$$Solve := Find(\varepsilon 1, \varepsilon 2, \varepsilon 3, \varepsilon 4, \varepsilon 5, \varepsilon 6, \varepsilon 7, \varepsilon 8, \varepsilon 9, \varepsilon 10, \varepsilon 11, \varepsilon 12, \varepsilon 13, \varepsilon 14, \varepsilon 15, \varepsilon 16, \varepsilon 17, \varepsilon 18, \varepsilon 19, \varepsilon 20, \varepsilon 30)$$

Solve =

	0
0	0.01
1	0.018

$e1 := Solve_0$	$e1 = 0.01$	$e11 := Solve_{10}$	$e11 = 0.03835$	$D = 0.1$
$e2 := Solve_1$	$e2 = 0.018$	$e12 := Solve_{11}$	$e12 = 0.03766$	
$e3 := Solve_2$	$e3 = 0.0243$	$e13 := Solve_{12}$	$e13 = 0.03672$	
$e4 := Solve_3$	$e4 = 0.02916$	$e14 := Solve_{13}$	$e14 = 0.03559$	
$e5 := Solve_4$	$e5 = 0.0328$	$e15 := Solve_{14}$	$e15 = 0.03432$	
$e6 := Solve_5$	$e6 = 0.03543$	$e16 := Solve_{15}$	$e16 = 0.03294$	
$e7 := Solve_6$	$e7 = 0.0372$	$e17 := Solve_{16}$	$e17 = 0.0315$	
$e8 := Solve_7$	$e8 = 0.03826$	$e18 := Solve_{17}$	$e18 = 0.03002$	
$e9 := Solve_8$	$e9 = 0.03874$	$e19 := Solve_{18}$	$e19 = 0.02852$	
$e10 := Solve_9$	$e10 = 0.03874$	$e20 := Solve_{19}$	$e20 = 0.02702$	
		$e30 := Solve_{20}$	$e30 = 0.36473$	

Step 4: The guess value of D is varied until the sum of all reaction conversions ($\varepsilon 1$, $\varepsilon 2$, etc.) sum to 1.0 as seen below. This means that all 400 moles of CO are converted as expected.

$$CO_{conv} := e1 + e2 + e3 + e4 + e5 + e6 + e7 + e8 + e9 + e10 + e11 + e12 + e13 + e14 + e15 + e16 \ldots + e17 + e18 + e19 + e20 + e30$$

$$CO_{conv} = 1$$

Step 5: Each value for ε is imported into Aspen Plus

Equipment Sizing

Rotary Dryer Source: *Process Engineering Economics* by James Couper, 2003

Typical rpm of rotary dryers $rpm_{dryer} := 4$

Typical product of $rpm \cdot diameter(feet)$ equals 15-25. Assume value of 25 for larger end

$$D_{dryer} := \frac{25ft}{rpm_{dryer}} \qquad D_{dryer} = 6.25ft$$

Typical residence times are 5-90 minutes and holdup of solids is 7-8%. Assume 5 minutes and 8%.

$$t_{res} := 5min \qquad holdup := 0.08$$

Typical exit gas temperature is 10-20℃ above the entering solids.

Feed rate into plant is 2000 ton/day with bulk density of stover equal to 100kg/m^3. Water density is accounted for as well.

$$m_{dot\_feed} := 2000\frac{tonne}{day} \qquad m_{dot\_moist\_in} = 666.667\frac{tonne}{day}$$

$$\rho_{bulk\_stover} = 100\frac{kg}{m^3} \qquad \rho_{water} = 1000\frac{kg}{m^3}$$

Volume of solids in dryer $V_{solids} := \left(\frac{m_{dot\_feed}}{\rho_{bulk\_stover}} + \frac{m_{dot\_moist\_in}}{\rho_{water}}\right) \cdot t_{res}$ $V_{solids} = 71.759m^3$

Volume of solids and steam $V_{dryer\_total} := \frac{V_{solids}}{holdup}$ $V_{dryer\_total} = 896.991m^3$

Length of theoretical dryer $length_{dryer} := \frac{V_{dryer\_total}}{D_{dryer}^{2} \cdot \frac{\pi}{4}}$ $length_{dryer} = 314.708m$

Surface area of theoretical dryer $A_{surf\_dryer} := length_{dryer} \cdot \pi \cdot D_{dryer}$ $A_{surf\_dryer} = 1883.4m^2$

Max surface area as reported by Aspen Icarus is 185 m^2, therefore approximately 10 dryers are required.

Feed throughput in each dryer (used for Icarus input) $\frac{m_{dot\_feed} + m_{dot\_moist\_in}}{10} = 24495.8\frac{lb}{hr}$

Lock hopper System

Source: *CE IGCC Repowering Project Bins and Lockhoppers,* Combustion Eng. 1993

note: this report's feedstock is coal

Assumptions from report
-A receiving bin is situated before the lockhopper with a 40 minute residence time
-design pressure is for 50 psia.
-Cycle time for lockhopper system is designed for 10 minutes resulting in approximately 50,000 cycles per year
-Storage volume for lockhopper and feed bin is assumed to be 10 minutes
-Approximate lockhopper and feed bin vessel thickness is 1.5 inches and design pressure is for 450 psia
-Volume is theoretical + 33%

Residence Time

biomass receiving bin $t_{res\_rbin} := 40min$ $\varepsilon_{void} := 25\%$

biomass lockhopper $t_{res\_lock} := 10min$

biomass feed bin $t_{res\_fbin} := 10min$

$$m_{dot\_feed\_lock} := m_{dot\_feed} + m_{dot\_moist\_dried}$$

$$m_{dot\_feed\_lock} = 2222\frac{tonne}{day}$$

Density of feed $\rho_{stover\_10\%moist} := \dfrac{\rho_{bulk\_stover} \cdot 2000 + \rho_{water} \cdot 222}{2222}$

$$\rho_{stover\_10\%moist} = 189.919\frac{kg}{m^3}$$

HT Scenario Lockhopper system (1 train)

Volume of biomass receiving bin $V_{r\_bin} := \dfrac{t_{res\_rbin} \cdot m_{dot\_feed\_lock}}{\rho_{stover\_10\%moist}} \cdot \dfrac{1}{1 - \varepsilon_{void}}$ $V_{r\_bin} = 433m^3$

Volume of biomass lockhopper $V_{lock} := \dfrac{t_{res\_lock} \cdot m_{dot\_feed\_lock}}{\rho_{stover\_10\%moist}} \cdot \dfrac{1}{1 - \varepsilon_{void}}$ $V_{lock} = 108m^3$

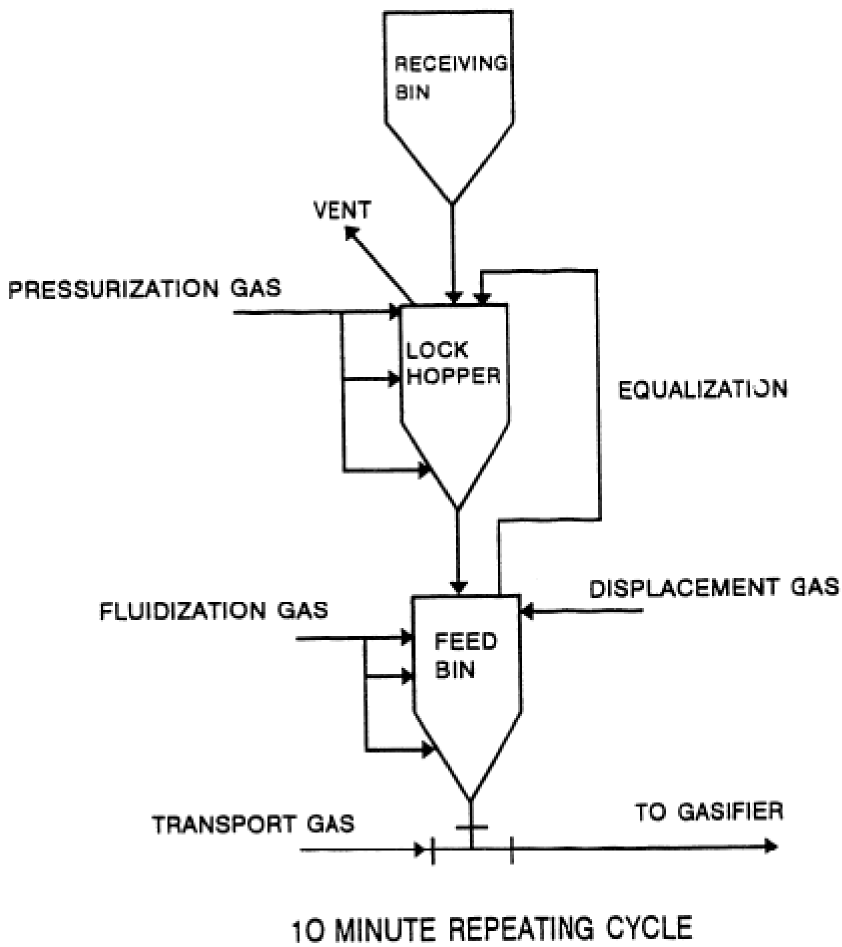
Volume of biomass feed bin $V_{f\_bin} := \dfrac{t_{res\_fbin} \cdot m_{dot\_feed\_lock}}{\rho_{stover\_10\%moist}} \cdot \dfrac{1}{1 - \varepsilon_{void}}$ $V_{f\_bin} = 108m^3$

Low Temperature Lockhopper System (7 trains)

Volume of biomass receiving bin $V_{r\_binLT} := \frac{V_{r\_bin}}{7}$ $V_{r\_binLT} = 61.909m^3$

Volume of biomass lockhopper $V_{lockLT} := \frac{V_{lock}}{7}$ $V_{lockLT} = 15.477m^3$

Volume of biomass feed bin $V_{f\_binLT} := \frac{V_{f\_bin}}{7}$ $V_{f\_binLT} = 15.477m^3$



Source: Combustion Engineering 1993

Lockhopper Power Consumption

Source: *Techno-Economic Analysis of Hydrogen Production by Gasification of Biomass* by Lau et al. [2002]

Specific Power of lockhopper, kW/tonne/day $SP_{lock} := 0.082 \frac{kW}{\frac{tonne}{day}}$

Biomass inlet to gasifier $m_{dot\_gasifier} := m_{dot\_biomass} + m_{dot\_moist\_dried}$

$Power_{lock} := SP_{lock} \cdot m_{dot\_gasifier}$ $Power_{lock} = 182.222kW$

Fly Ash Collection Storage Tank (assume 7 days storage)

$\rho_{ash} := 700\frac{kg}{m^3}$ (assumed) $\qquad m_{dot\_ash} := 5.88\frac{tonne}{day}$

$V_{tank} := \frac{m_{dot\_ash}}{\rho_{ash}}\cdot 7day \qquad V_{tank} = 58.8m^3 \qquad V_{tank} = 2.077\times 10^3\,ft^3$

Slag Separation drum (5 minute residence time, 20% volume)

$\rho_{slag} := 2700\frac{kg}{m^3} \qquad m_{dot\_slag} := 114\frac{tonne}{day} \qquad \varepsilon_{void\_slag} := 0.8$

$V_{drum} := \frac{m_{dot\_slag}}{\rho_{slag}}\cdot 5min\cdot\frac{1}{1-\varepsilon_{void\_slag}} \qquad V_{drum} = 0.733m^3$

Slag collection Storage tank (7 days storage)

$V_{slag\_storage} := \frac{m_{dot\_slag}}{\rho_{slag}}\cdot 7day \qquad V_{slag\_storage} = 295.6m^3$

Char collection storage bin (1-day residence time, 80% volume)

$\rho_{char} := 2700\frac{kg}{m^3} \qquad m_{dot\_char} := 214\frac{tonne}{day} \qquad \varepsilon_{void\_char} := 0.2$ assume 20% voidage

$V_{chardrum} := \frac{m_{dot\_char}}{\rho_{char}}\cdot 1day\cdot\frac{1}{1-\varepsilon_{void\_char}} \qquad V_{chardrum} = 99.074m^3$

Note: the resulting volumes are used to assist in costing using Aspen Icarus

Pressure Swing Adsorption Unit Sizing

References in parentheses are given at the end of this section.

$\mathrm{Pi} := 3.1415$ $\quad \mathrm{nm} := 10^{-9} \cdot \mathrm{m}$

The adsorbtion unit is 1/3 molesieve and 2/3 Activated Carbon

Molsieve 13X $\quad$ (a) $\mathrm{BulkDens} := 43 \cdot \dfrac{\mathrm{lb}}{\mathrm{ft}^3}$ $\quad$ (b) $\mathrm{SA} := 1320 \dfrac{\mathrm{m}^2}{\mathrm{gm}}$ $\quad$ (b) $\mathrm{PoreVol} := 0.51 \cdot \dfrac{\mathrm{cm}^3}{\mathrm{gm}}$

Determine dry volumetric flow rate of the syngas stream at atmospheric pressure and 25 deg C

$$\mathrm{VolFlowRate} := (167 - 1) \cdot \frac{\mathrm{kmol}}{\mathrm{hr}} \cdot 22.414 \frac{\mathrm{m}^3}{\mathrm{kmol}} \cdot \frac{14.696\,\mathrm{psi}}{400 \cdot \mathrm{psi}} \cdot \frac{(273.15 + 25) \cdot \mathrm{K}}{273.15\,\mathrm{K}}$$

$$\mathrm{VolFlowRate} = 149.211 \frac{\mathrm{m}^3}{\mathrm{hr}}$$

Mole fraction of components that are adsorbed

$\mathrm{CO} := 23$ $\quad \mathrm{CO2} := 1$ $\quad \mathrm{CH4} := 1$

Actual Flow rate of components adsorbed

$$\mathrm{FlowRateAds} := \mathrm{VolFlowRate} \cdot \frac{\mathrm{CO} + \mathrm{CO2} + \mathrm{CH4}}{100}$$

$$\mathrm{FlowRateAds} = 37.303 \frac{\mathrm{m}^3}{\mathrm{hr}}$$

Adsorbent Capacity

$$\mathrm{AdsCap} := 0.34 \cdot \frac{\mathrm{ft}^3}{\mathrm{lb}} \cdot \frac{14.696\,\mathrm{psi}}{400 \cdot \mathrm{psi}} \cdot \frac{(273.15 + 25) \cdot \mathrm{K}}{273.15\,\mathrm{K}}$$

(d) SCF/lb corrected for P and T to actual cm3/gm; PSA occurs at ambient temperature

$$\mathrm{AdsCap} = 0.851 \frac{\mathrm{cm}^3}{\mathrm{gm}}$$

Mass of molsieve required

$\mathrm{CycleTime} := 5 \cdot \mathrm{min}$

$$\mathrm{MolSieveMass} := \frac{\mathrm{FlowRateAds} \cdot \mathrm{CycleTime}}{\mathrm{AdsCap}} \qquad \mathrm{MolSieveMass} = 3.652 \times 10^3\,\mathrm{kg}$$

Determine volume and length of molsieve bed and activated carbon bed

$$BedVolume := \frac{MolSieveMass}{BulkDens} \qquad BedVolume = 5.302\,m^3$$

$$Diam := 4 \cdot ft \qquad Diam = 1.219\,m$$ (assumed)

$$Length := \frac{BedVolume}{Pi \cdot Diam^2} \qquad Length = 3.725\,ft$$ (Just molsieve bed)

$$RxtrLength := 3 \cdot Length$$ (bed is 1/3 molsieve, 2/3 activated carbon)

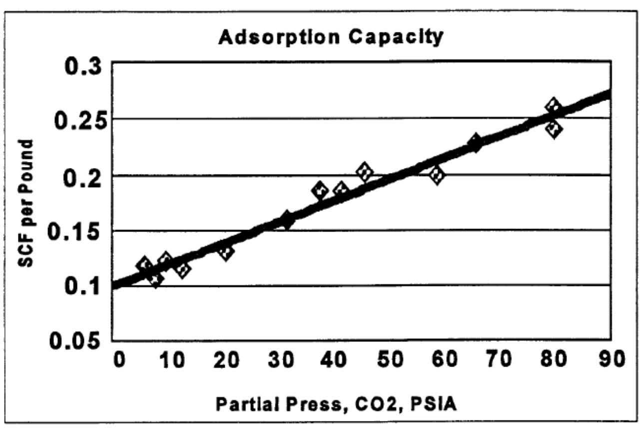
$$RxtrLength = 11.175\,ft \qquad RxtrLength = 3.406\,m$$

$$RxtrVolume := RxtrLength \cdot Diam^2 \cdot 0.25 \cdot \pi \qquad RxtrVolume = 3.977\,m^3$$

(a) http://www.sigmaaldrich.com/Brands/Aldrich/Tech_Bulletins/AL_143/Molecular_Sieves.html

(b) US Pat 6117810

(d) WO/1998/058726 BULK SEPARATION OF CARBON DIOXIDE FROM METHANE USING NATURAL CLINOPTILOLITE --extrapolate to partial pressure of CO2+CH4+N2+CO=32.6%*400 psi



GRAPH C

HT Scenario Fuel Storage

Gasoline Storage Tank (30 days storage)

$$m_{dot\_gasHT} = 112.78\frac{tonne}{day}$$

$$v_{dot\_gasHT} = 4.041\times 10^{4}\ \frac{gal}{day}$$

$$V_{gas\_tankHT} := v_{dot\_gasHT} \cdot 30day$$

$$V_{gas\_tankHT} = 4589m^{3}$$

Diesel Storage Tank (30 days storage)

$$m_{dot\_dieselHT} = 266.5\frac{tonne}{day}$$

$$v_{dot\_dieselHT} = 8.381\times 10^{4}\ \frac{gal}{day}$$

$$V_{diesel\_tankHT} := v_{dot\_dieselHT} \cdot 30day$$

$$V_{diesel\_tankHT} = 9518m^{3}$$

Note: the resulting volumes are used to assist in costing using Aspen Icarus

LT Scenario Fuel Storage

Gasoline Storage Tank (30 days storage)

$$m_{dot\_gasLT} = 87.12\frac{tonne}{day}$$

$$v_{dot\_gasLT} = 3.122\times 10^{4}\ \frac{gal}{day}$$

$$V_{gas\_tankLT} := v_{dot\_gasLT}\cdot 30day$$

$$V_{gas\_tankLT} = 3545m^{3}$$

Diesel Storage Tank (30 days storage)

$$m_{dot\_dieselLT} = 205.86\frac{tonne}{day}$$

$$v_{dot\_dieselLT} = 6.474\times 10^{4}\ \frac{gal}{day}$$

$$V_{diesel\_tankLT} := v_{dot\_dieselLT}\cdot 30day$$

$$V_{diesel\_tankLT} = 7352m^{3}$$

Note: the resulting volumes are used to assist in costing using Aspen Icarus

LT Gasifier Cost

Source: *Larson et al. 2005* in 2003$

$C_{0\_gasifier} := 6.41 \cdot 10^{6}$ $MM $\qquad S_{0\_gasifier} := 41.7 \frac{tonne}{hr}$ $\qquad S_{max} := 120 \frac{tonne}{hr}$ $\qquad f := 0.7$

Biomass throughput of 300 tpd

$S_{gasifierLT} := 300 \frac{ton}{day}$ $\qquad S_{gasifierLT} = 11.34 \frac{tonne}{hr}$

The cost ($MM) of one train at 300 ton per day

$$C_{gasifierLT} := C_{0\_gasifier} \cdot \left(\frac{S_{gasifierLT}}{\frac{tonne}{hr}} \cdot \frac{1}{\frac{S_{0\_gasifier}}{\frac{tonne}{hr}}} \right)^{f}$$

$C_{gasifierLT} = 2.576 \times 10^{6}$ $MM

Since 2205 ton /day we need 7 gasifiers but we can apply the multiple train scaling exponent

$m_{train} := 0.9$

$C_{gasifierLTtrain} := C_{gasifierLT} \cdot 7^{m_{train}}$ $\qquad C_{gasifierLTtrain} = 1.484 \times 10^{7}$ $MM

FT Reactor Costing

Source: *Larson et al. 2005* in 2003$

$C_{FT\_base} := 10.5$ $MM $f_{FT2} := 0.72$ $S_{FT\_base} := 2.52\frac{MMcf}{hr}$

HT Scenario

$M_{dot\_FTHT} := 13829\frac{kmol}{hr}$ $V_{standard\_FTHT} := M_{dot\_FTHT}\cdot 22.4\frac{L}{mol}$

$V_{standard\_FTHT} = 10.939\frac{MMcf}{hr}$

$$C_{FTHT\_reac} := C_{FT\_base}\cdot\left(\frac{V_{standard\_FTHT}}{S_{FT\_base}}\right)^{f_{FT2}}$$

$C_{FTHT\_reac} = 30.217$ Installed cost $MM (assume 3.6 install factor consistent with Peters et al.)

LT Scenario

$M_{dot\_FTLT} := 11400\frac{kmol}{hr}$ $V_{standard\_FTLT} := M_{dot\_FTLT}\cdot 22.4\frac{L}{mol}$

$V_{standard\_FTLT} = 9.018\frac{MMcf}{hr}$

$$C_{FTLT\_reac} := C_{FT\_base}\cdot\left(\frac{V_{standard\_FTLT}}{S_{FT\_base}}\right)^{f_{FT2}}$$

$C_{FTLT\_reac} = 26.294$ Installed cost $MM (assume 3.6 install factor consistent with Peters et al.)

Acid Gas Removal Area Cost

Source: Phillips et al. 2007 in 2005$

Calculated by adding the input syngas streams to the absorber column

$S_{AGR\_base} := 332910\frac{lb}{hr}$ $\qquad f_{AGR} := 0.65$ $\qquad C_{AGR\_base} := 544650$

HT scenario

$S_{AGR\_HT} := (2965 + 2308)\cdot\frac{tonne}{day}$ $\qquad S_{AGR\_HT} = 484374\frac{lb}{hr}$

$$C_{AGR\_HT} := C_{AGR\_base}\cdot\left(\frac{S_{AGR\_HT}}{\frac{lb}{hr}}\cdot\frac{1}{\frac{S_{AGR\_base}}{\frac{lb}{hr}}}\right)^{f_{AGR}}$$

$C_{AGR\_HT} = 6949808$

LT scenario

$S_{AGR\_LT} := (2070 + 2190)\frac{tonne}{day}$ $\qquad S_{AGR\_LT} = 391321\frac{lb}{hr}$

$$C_{AGR\_LT} := C_{AGR\_base}\cdot\left(\frac{S_{AGR\_LT}}{\frac{lb}{hr}}\cdot\frac{1}{\frac{S_{AGR\_base}}{\frac{lb}{hr}}}\right)^{f_{AGR}}$$

$C_{AGR\_LT} = 6049946$

A500 Hydroprocessing area cost

Source: Robinson et al. 2007 in 2007$

Note: "bpsd" is barrels per standard day

$AreaCost_0 := \frac{4000}{bpsd}$ $S_{0\_HY} := 25000\,bpsd$ $f_{HY} := 0.65$ (assumed)

$C_{0\_HY} := AreaCost_0 \cdot S_{0\_HY}$ $C_{0\_HY} = 100000000$

HT scenario

$m_{dot\_FTL\_HT} := 428\frac{tonne}{day}$ $\rho_{FTL} := 750\frac{kg}{m^3}$ (from ASPEN model)

$v_{dot\_FTL\_HT} := \frac{m_{dot\_FTL\_HT}}{\rho_{FTL}}$ $v_{dot\_FTL\_HT} = 3.589\times 10^3\,bpsd$

$$C_{HY\_HT} := C_{0\_HY}\cdot\left(\frac{v_{dot\_FTL\_HT}}{bpsd}\cdot\frac{1}{\frac{S_{0\_HY}}{bpsd}}\right)^{f_{HY}}$$ $C_{HY\_HT} = 2.832\times 10^7$

Power required for A500

$Power_{per\_bpsd} := \frac{15kW\cdot hr}{bpsd\cdot day}$

$Power_{areaHT} := Power_{per\_bpsd}\cdot v_{dot\_FTL\_HT}$ $Power_{areaHT} = 2.243MW$

LT Scenario

$m_{dot\_FTL\_LT} := 330.42\frac{tonne}{day}$ (from ASPEN model)

$v_{dot\_FTL\_LT} := \frac{m_{dot\_FTL\_LT}}{\rho_{FTL}}$ $v_{dot\_FTL\_LT} = 2.771\times 10^3\,bpsd$

$$C_{HY\_LT} := C_{0\_HY}\cdot\left(\frac{v_{dot\_FTL\_LT}}{bpsd}\cdot\frac{1}{\frac{S_{0\_HY}}{bpsd}}\right)^{f_{HY}}$$ $C_{HY\_LT} = 2.394\times 10^7$

Power required for A500

$Power_{areaLT} := Power_{per\_bpsd}\cdot v_{dot\_FTL\_LT}$ $Power_{areaLT} = 1.732MW$

Reactors and Catalysts

Fischer-Tropsch reactor and cobalt catalyst

FT reactor volume

Using gas hourly space velocity and actual volumetric flow rate, the volume of the reactor is determined

$$GHSV = \frac{v_0}{V}$$

$GHSV_{FT} := 100 \cdot hr^{-1}$ (assumed)

$v_{rate\_actHT} := 6.298 \frac{m^3}{s}$ (from ASPEN model)

$V_{FTHT} := \frac{v_{rate\_actHT}}{GHSV_{FT}}$ $V_{FTHT} = 226.728 m^3$ $V_{FTHT} = 8.007 \times 10^3 ft^3$

$v_{rate\_actLT} := 5.021 \frac{m^3}{s}$

$V_{FTLT} := \frac{v_{rate\_actLT}}{GHSV_{FT}}$ $V_{FTLT} = 180.756 m^3$ $V_{FTLT} = 6.383 \times 10^3 ft^3$

FT catalyst cost

$Co_{cost} := \frac{15}{lb}$ (assumed) $\rho_{Co} := 64 \frac{lb}{ft^3}$ (assumed)

$Co_{vol\_cost} := Co_{cost} \cdot \rho_{Co}$ $Co_{vol\_cost} = 960 \frac{1}{ft^3}$

Replacement cost of cobalt catalyst

$Co_{total\_costHT} := Co_{vol\_cost} \cdot V_{FTHT}$ $Co_{total\_costHT} = 7.687 \times 10^6$

$Co_{total\_costLT} := Co_{vol\_cost} \cdot V_{FTLT}$ $Co_{total\_costLT} = 6.128 \times 10^6$

Water Gas Shift reactor and catalyst

Sour WGS reactor volume (HT scenario)

Using gas hourly space velocity and actual volumetric flow rate, the volume of the reactor is determined

$$GHSV = \frac{\nu_0}{V}$$

$$GHSV_{WGS} := 1000 \cdot hr^{-1}$$ (assumed)

$$\nu_{rate\_actSWGS} := 2.008 \frac{m^3}{s}$$ (from ASPEN)

$$V_{SWGS} := \frac{\nu_{rate\_actSWGS}}{GHSV_{WGS}} \qquad V_{SWGS} = 7.229 m^3 \qquad V_{SWGS} = 255.283 ft^3$$

WGS reactor volume (LT scenario)

$$\nu_{rate\_actWGS} := 1.834 \frac{m^3}{s}$$ (from ASPEN)

$$V_{WGS} := \frac{\nu_{rate\_actWGS}}{GHSV_{WGS}} \qquad V_{WGS} = 6.602 m^3 \qquad V_{WGS} = 233.162 ft^3$$

WGS and SWGS Catalyst Cost

$$CatCost_{WGS} := \frac{8}{lb} \qquad \rho_{cat\_WGS} := 56 \frac{lb}{ft^3} \qquad \rho_{cat\_WGS} = 897.034 \frac{kg}{m^3}$$

$$CatCost_{vol\_WGS} := CatCost_{WGS} \cdot \rho_{cat\_WGS} \qquad CatCost_{vol\_WGS} = 448 \frac{1}{ft^3}$$

Replacement cost of WGS catalyst

$$TotalCatCost_{SWGS} := CatCost_{vol\_WGS} \cdot V_{SWGS} \qquad TotalCatCost_{SWGS} = 114367$$

$$TotalCatCost_{WGS} := CatCost_{vol\_WGS} \cdot V_{WGS} \qquad TotalCatCost_{WGS} = 104456$$

Steam Methane Reformer reactor and catalyst (LT scenario)

$GHSV_{SMR} := 2600 hr^{-1}$ (assumed)

$\nu_{rate\_actSMR} := 7.082 \frac{m^3}{s}$ (from Aspen model)

$V_{SMR} := \frac{\nu_{rate\_actSMR}}{GHSV_{SMR}}$ $V_{SMR} = 9.806 m^3$ $V_{SMR} = 346.29 ft^3$

SMR Catalyst Cost

$CatCost_{SMR} := \frac{4.67}{lb}$ $\rho_{cat\_SMR} := 64 \frac{lb}{ft^3}$ $\rho_{cat\_SMR} = 1.025 \times 10^3 \frac{kg}{m^3}$

$CatCost_{vol\_SMR} := CatCost_{SMR} \cdot \rho_{cat\_SMR}$ $CatCost_{vol\_SMR} = 298.88 \frac{1}{ft^3}$

Replacement cost of SMR catalyst

$TotalCatCost_{SMR} := CatCost_{vol\_SMR} \cdot V_{SMR}$ $TotalCatCost_{SMR} = 103499$

Natural Gas utility consumption

Annual natural gas requirement at 5% of yearly operating hours

Natural gas properties

Price (Source: Energy Information Administration)

$HHV_{ng} := 54\frac{MJ}{kg}$ $MW_{ng} := 16.04\frac{gm}{mol}$ $\rho_{ng} := 22.4\frac{L}{mol}$

$Cost_{ng} := \frac{6.4}{1000ft^3}$

$Cost_{ng}\cdot\frac{\rho_{ng}}{MW_{ng}} = 286.335\frac{1}{ton}$

HT scenario

$P_{required\_plantHT} := 32.813MW$ (from Aspen model, includes power required for gas turbine air compressor)

$Eff_{ng\_to\_power} := 0.35$ (assumed)

$$m_{dot\_ngHT} := \frac{\frac{P_{required\_plantHT}}{Eff_{ng\_to\_power}}}{HHV_{ng}} \qquad m_{dot\_ngHT} = 1.378\times 10^{4}\frac{lb}{hr}$$

Annual natural gas requirement

$$M_{ngHT} := m_{dot\_ngHT}\cdot Availability\cdot 0.05 \qquad M_{ngHT} = 2563ton$$

Average flowrate of natural gas

$$m_{dot\_ng\_5\%HT} := \frac{M_{ngHT}}{8760hr} \qquad m_{dot\_ng\_5\%HT} = 585.14\frac{lb}{hr}$$

LT scenario

$P_{required\_plantLT} := 24.3MW$ (from Aspen model, includes power required for gas turbine air compressor)

$$m_{dot\_ngLT} := \frac{\frac{P_{required\_plantLT}}{Eff_{ng\_to\_power}}}{HHV_{ng}} \qquad m_{dot\_ngLT} = 1.02 \times 10^{4} \frac{lb}{hr}$$

Annual natural gas requirement

$$M_{ngLT} := m_{dot\_ngLT} \cdot Availability \cdot 0.05 \qquad \boxed{M_{ngLT} = 1898ton}$$

Average flowrate of natural gas

$$m_{dot\_ng\_5\%LT} := \frac{M_{ngLT}}{8760hr} \qquad m_{dot\_ng\_5\%LT} = 433.331 \frac{lb}{hr}$$

APPENDIX D. PROCESS FLOW DIAGRAMS

D.1 High Temperature Scenario

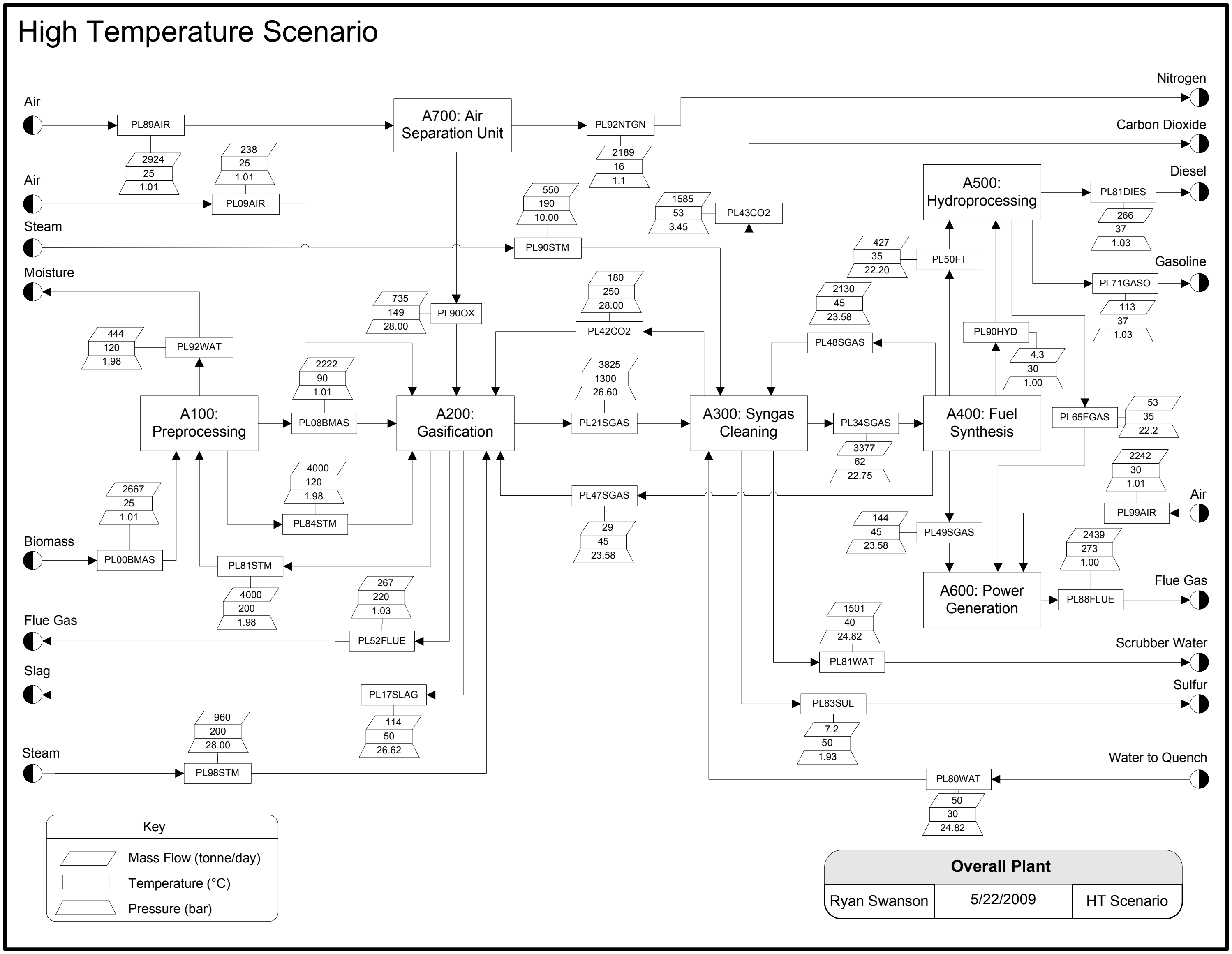


Figure 24. Overall plant area process flow diagram for HT scenario

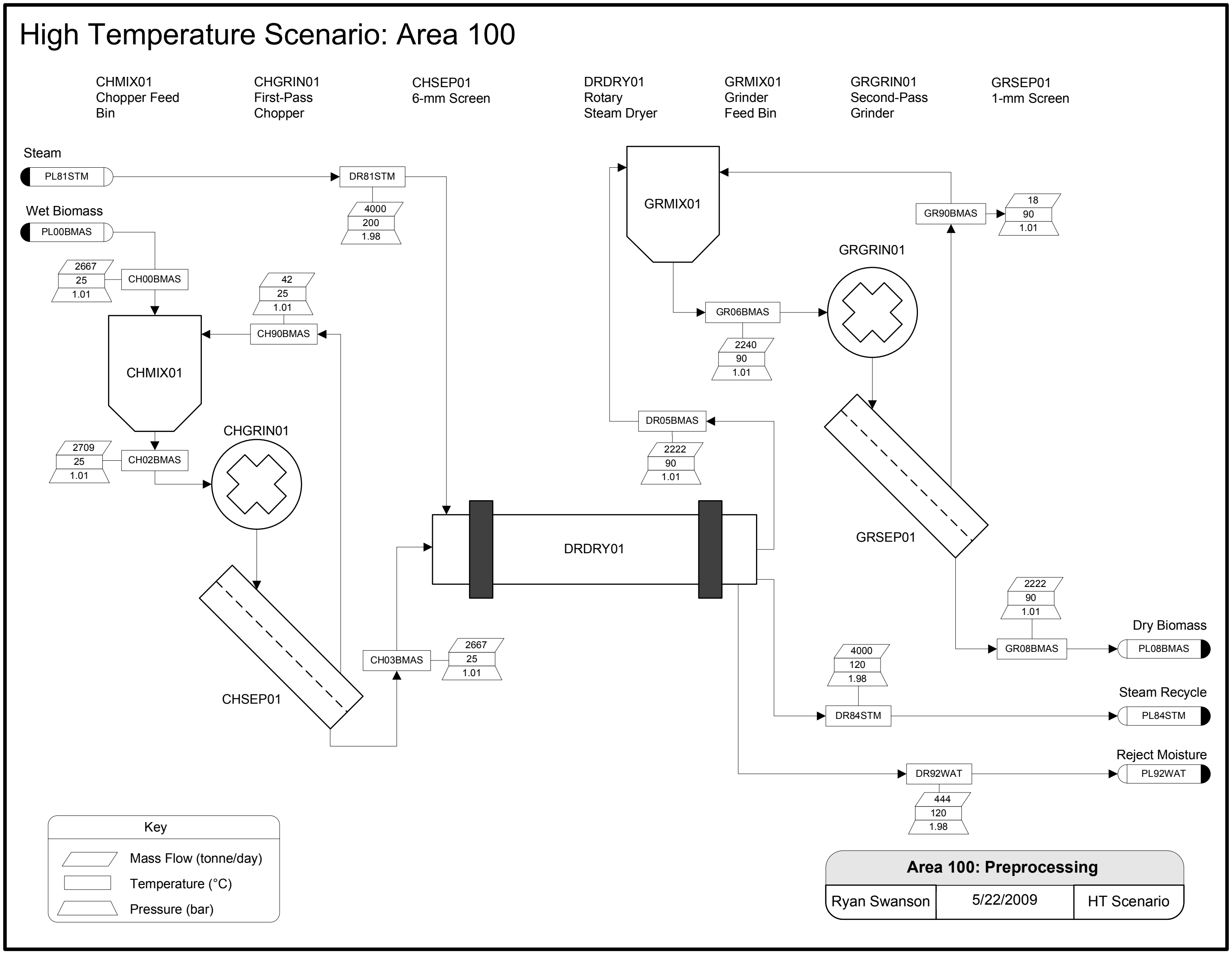


Figure 25. Preprocessing area process flow diagram for HT scenario

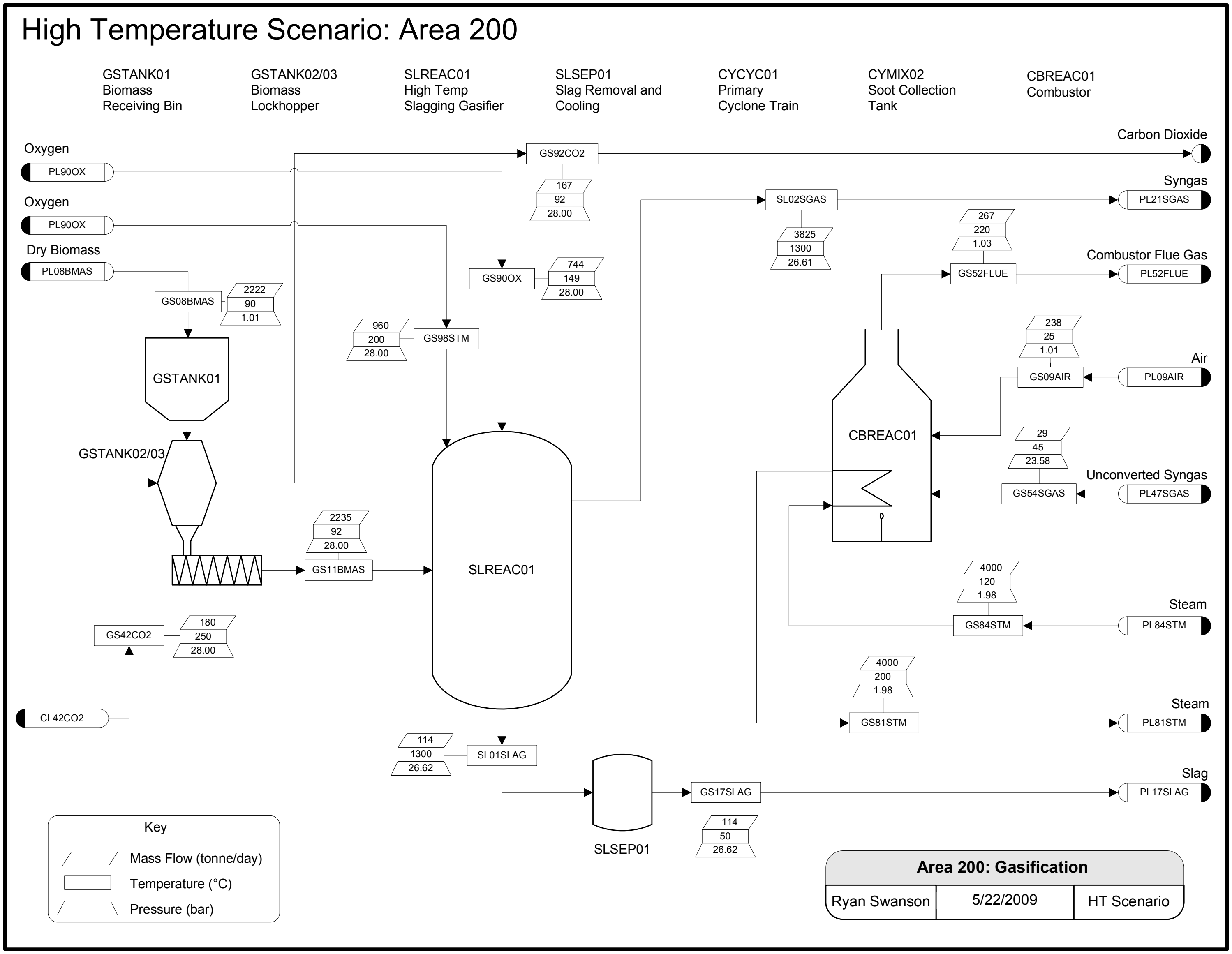


Figure 26. Gasification area process flow diagram for HT scenario

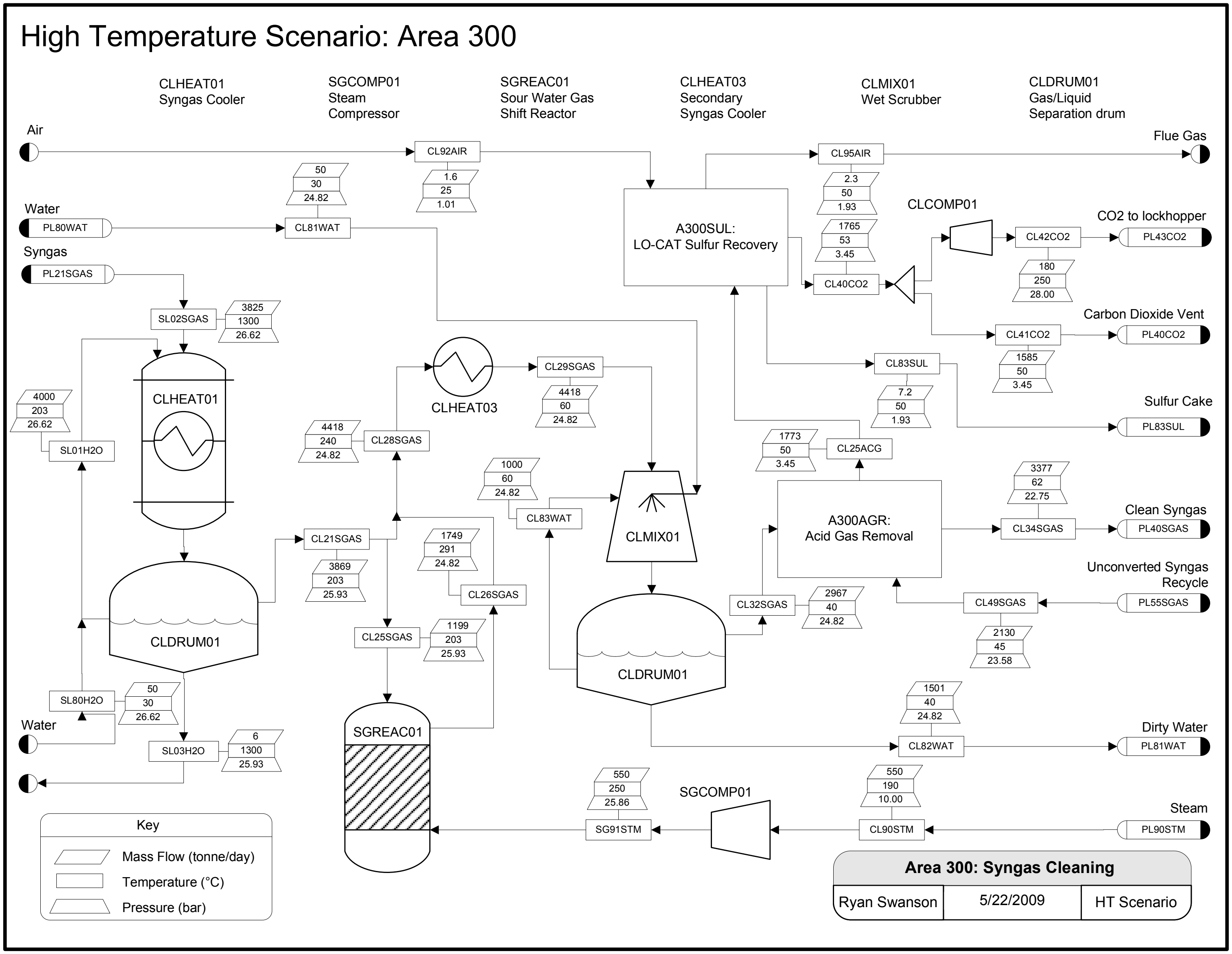


Figure 27. Syngas cleaning area process flow diagram for HT scenario

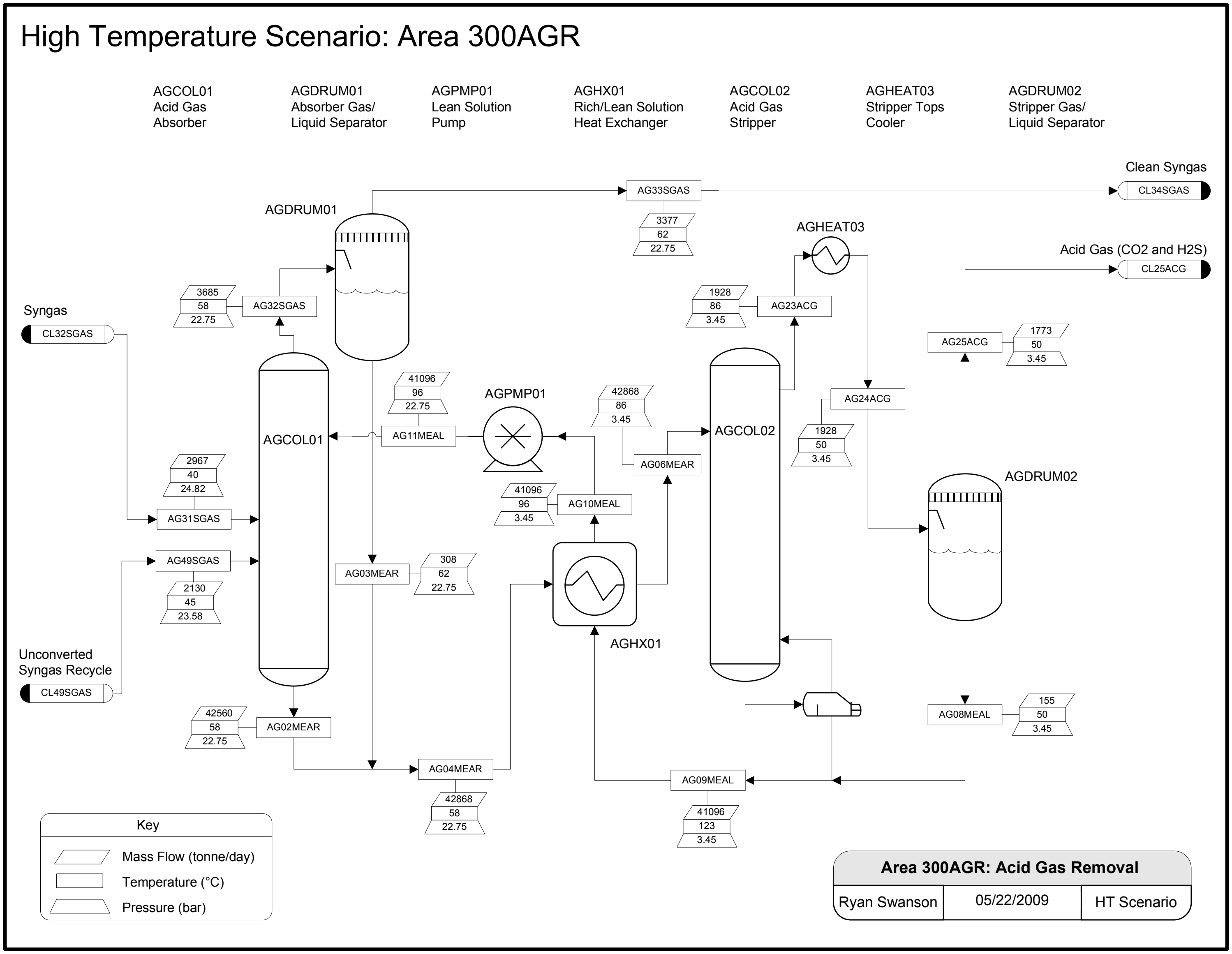


Figure 28. Acid gas removal area process flow diagram for HT scenario

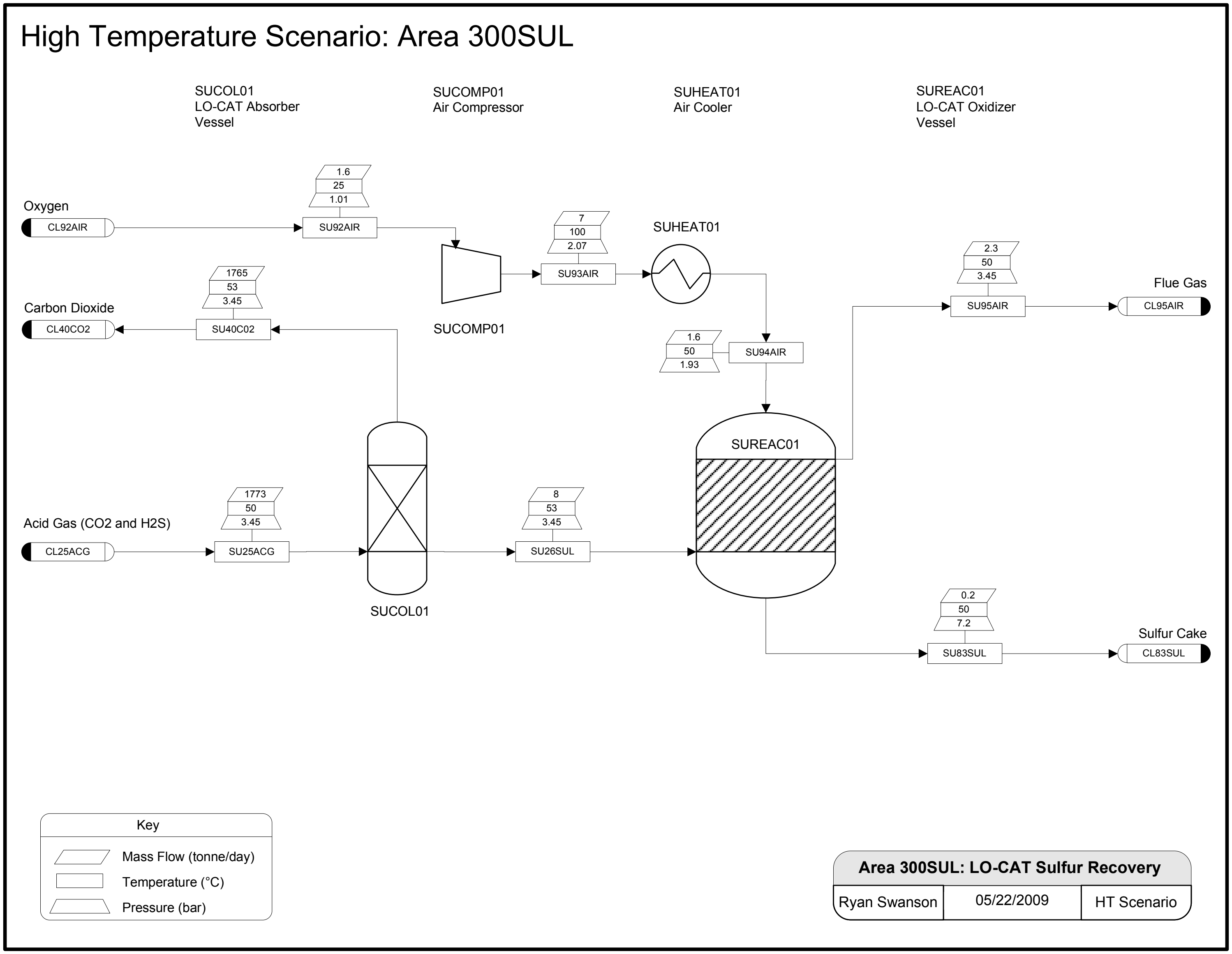


Figure 29. Sulfur recovery area process flow diagram for HT scenario

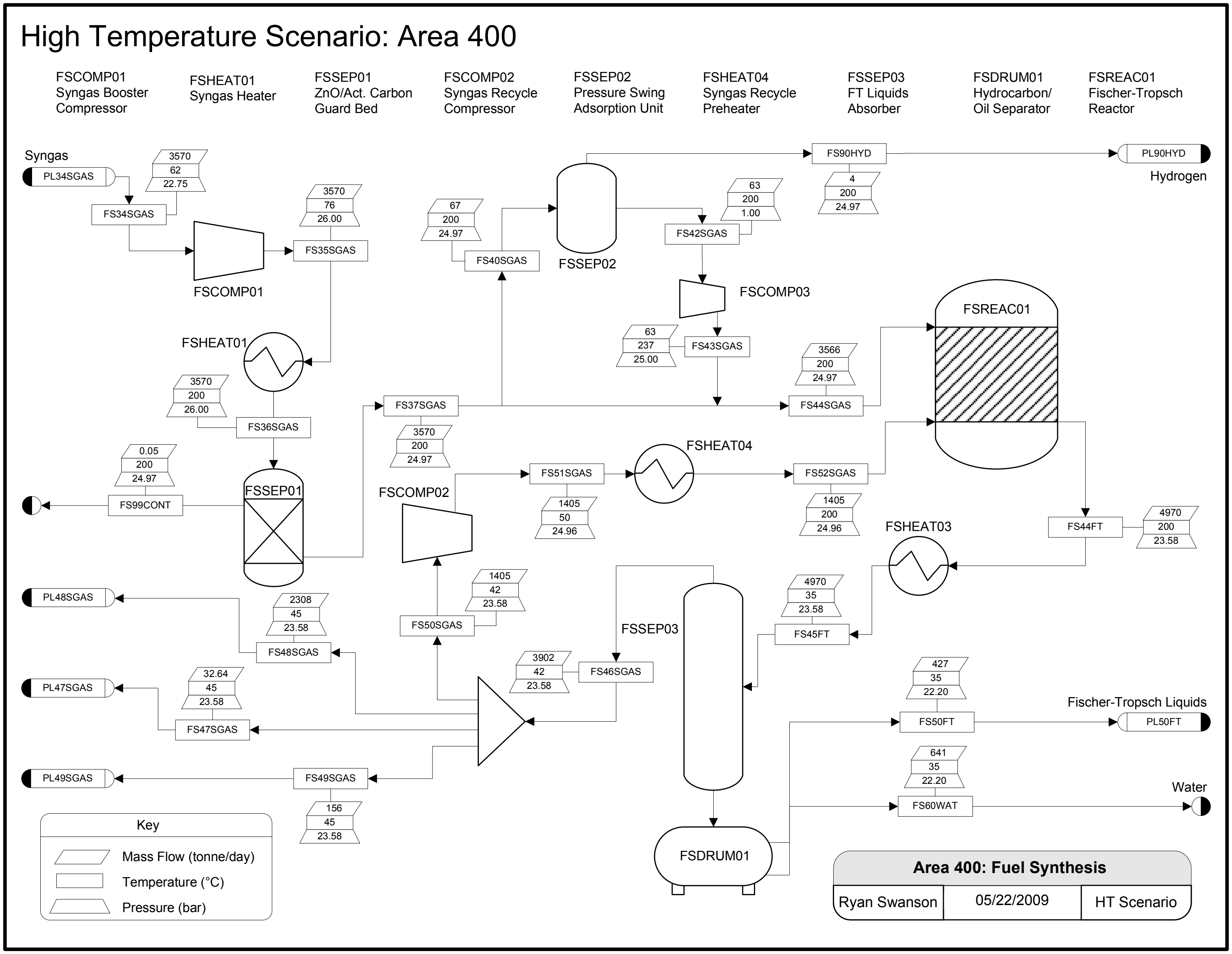


Figure 30. Fuel synthesis area process flow diagram for HT scenario

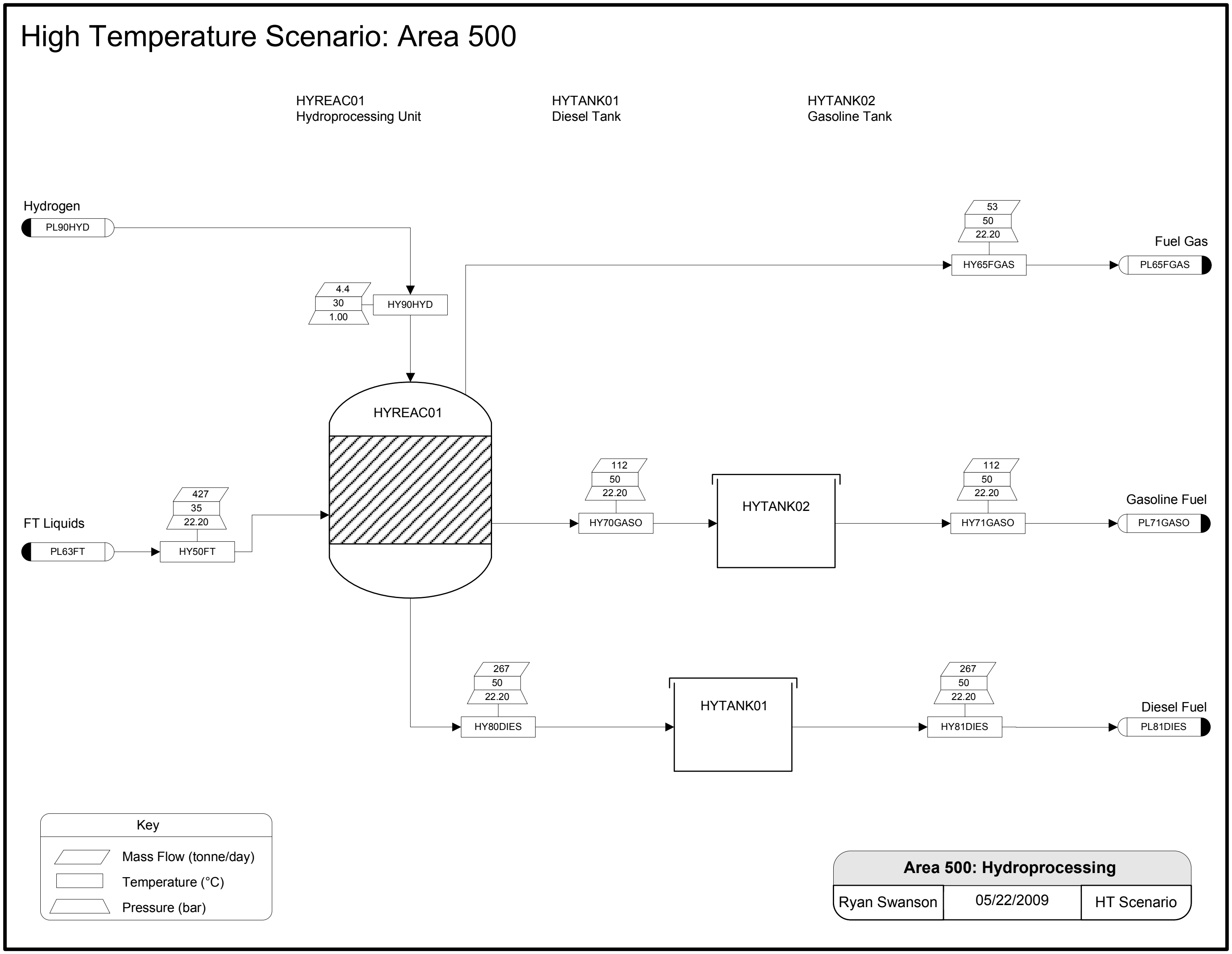


Figure 31. Hydroprocessing area process flow diagram for HT scenario

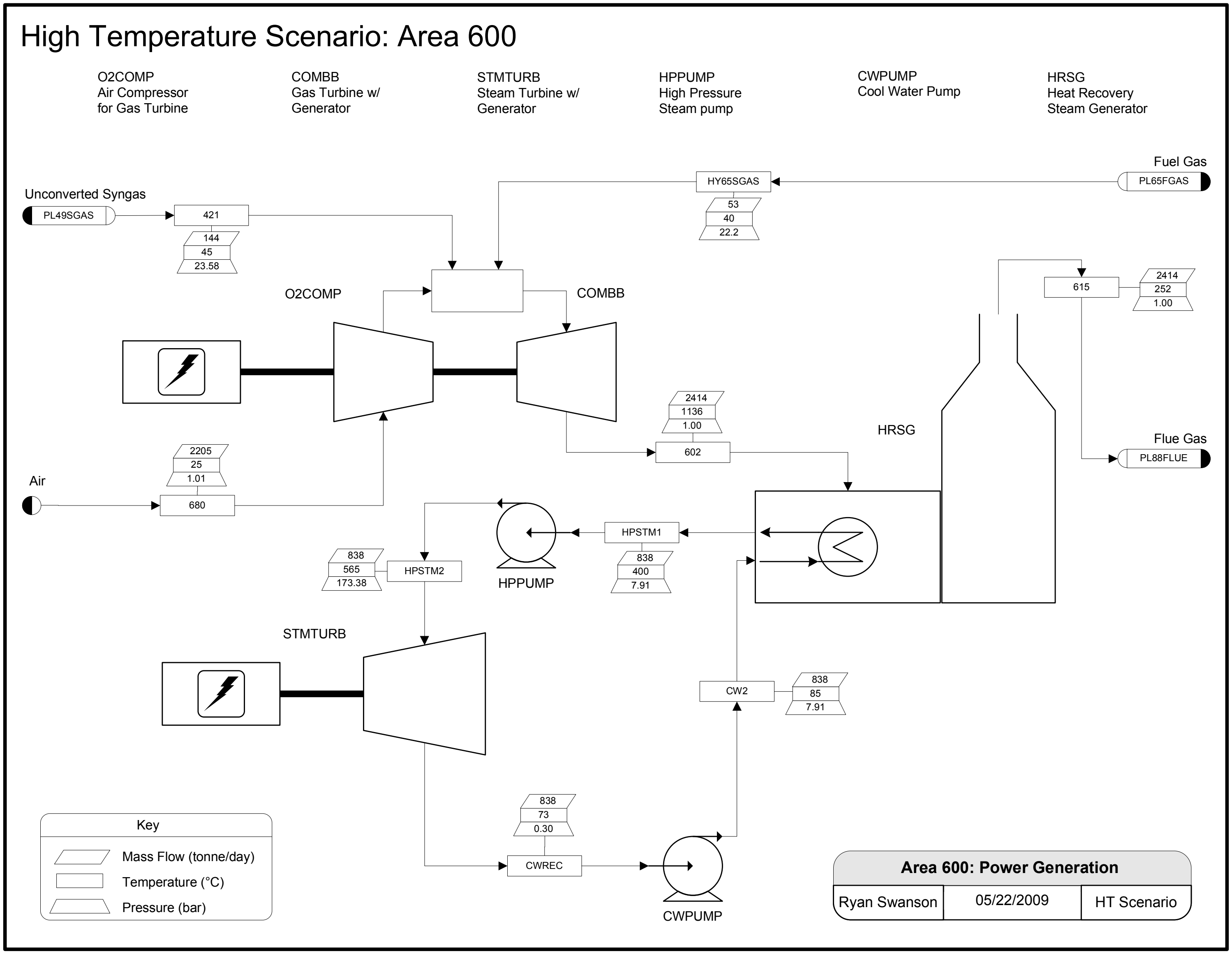


Figure 32. Power generation area process flow diagram for HT scenario

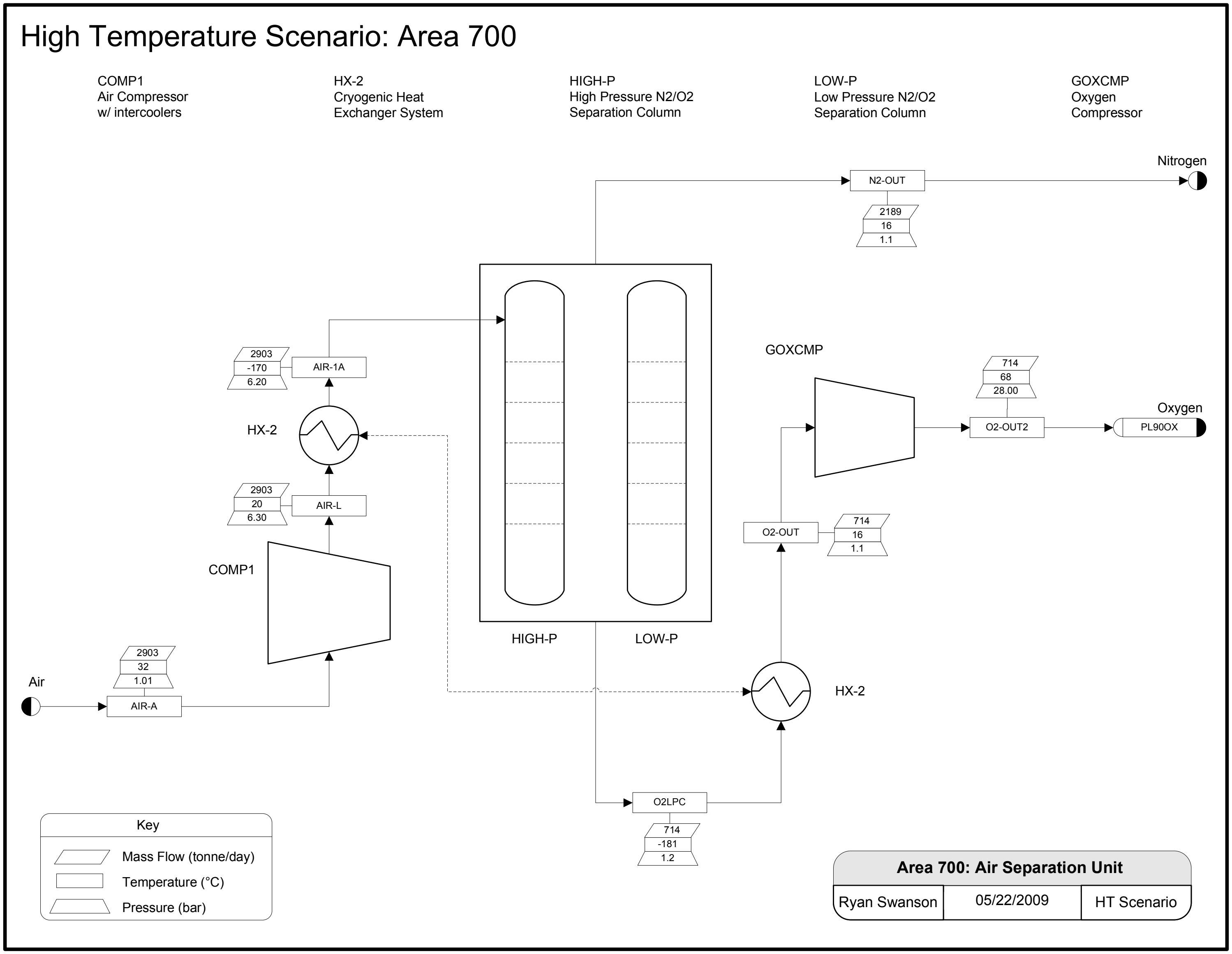


Figure 33. Air separation unit process flow diagram for HT scenario

D.2 Low Temperature Scenario

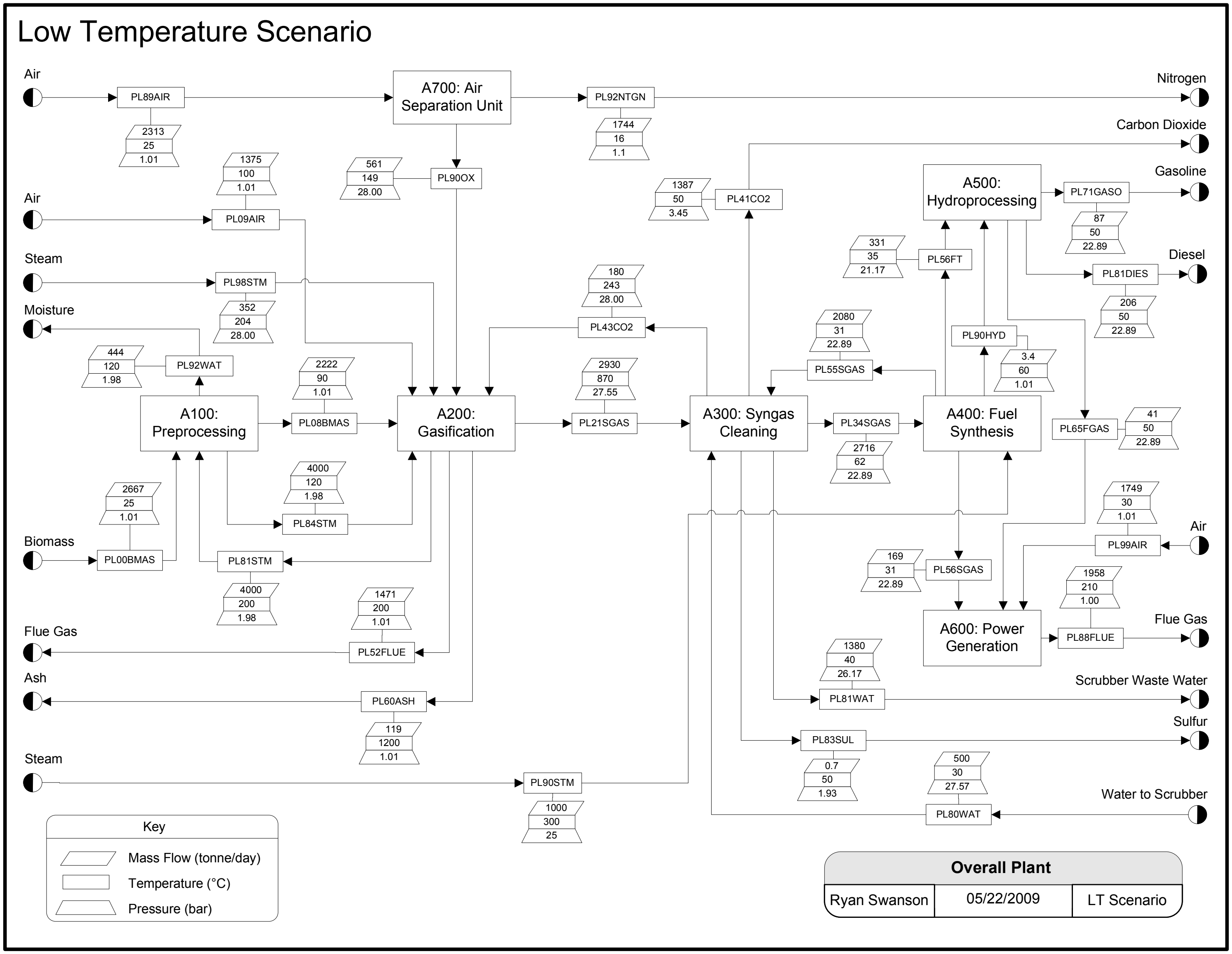


Figure 34. Overall plant area process flow diagram for LT scenario

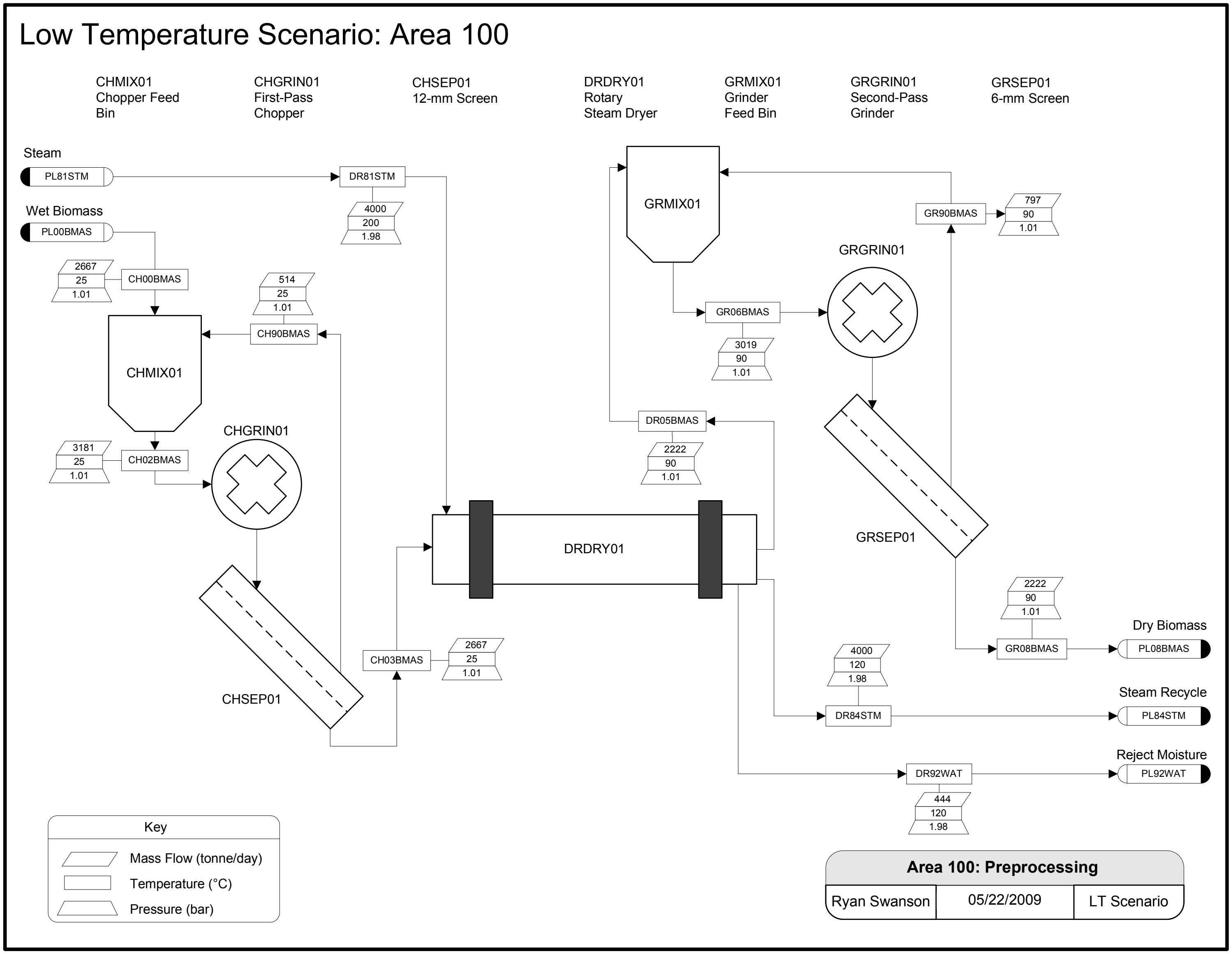


Figure 35. Preprocessing area process flow diagram for LT scenario

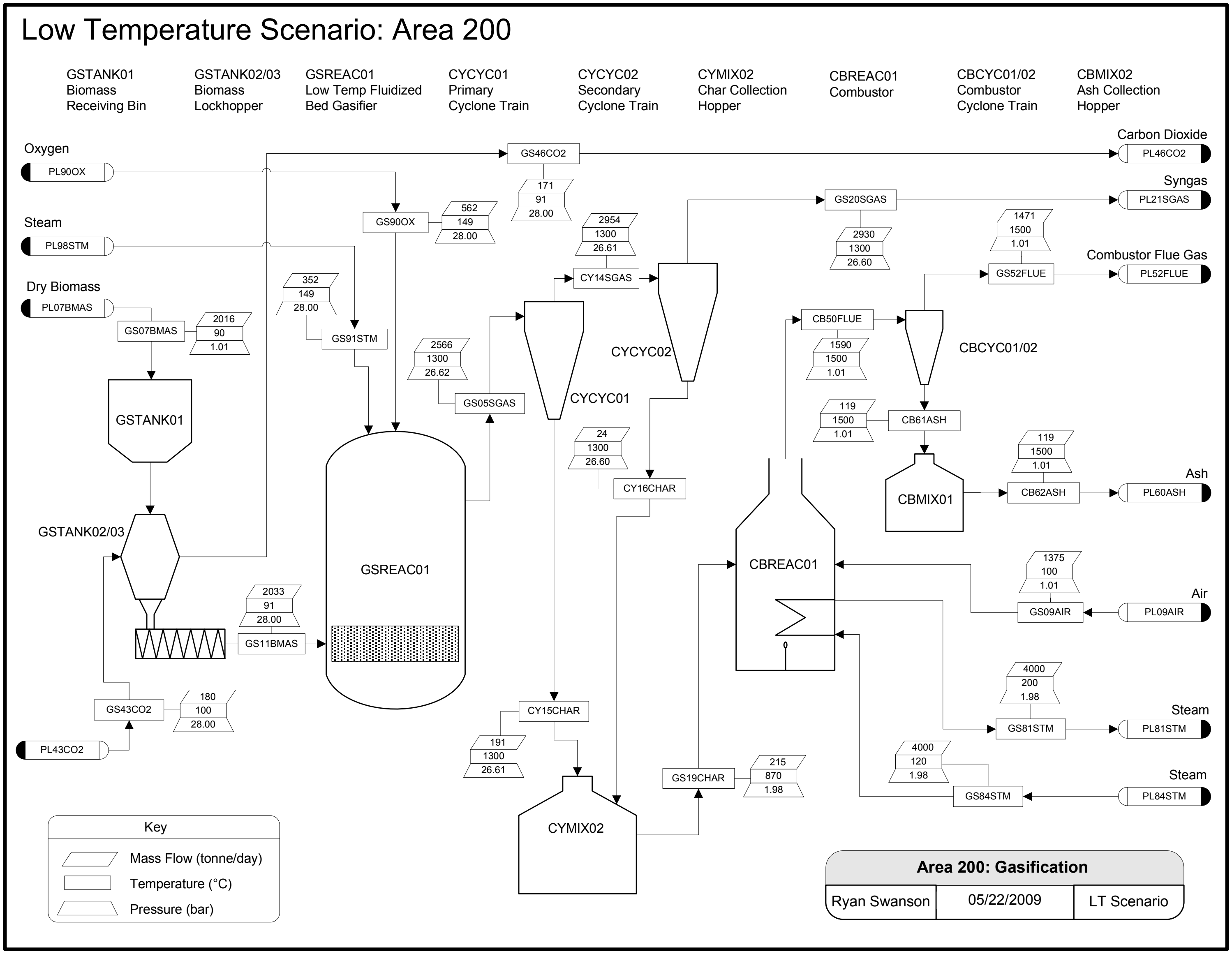


Figure 36. Gasification area process flow diagram for LT scenario

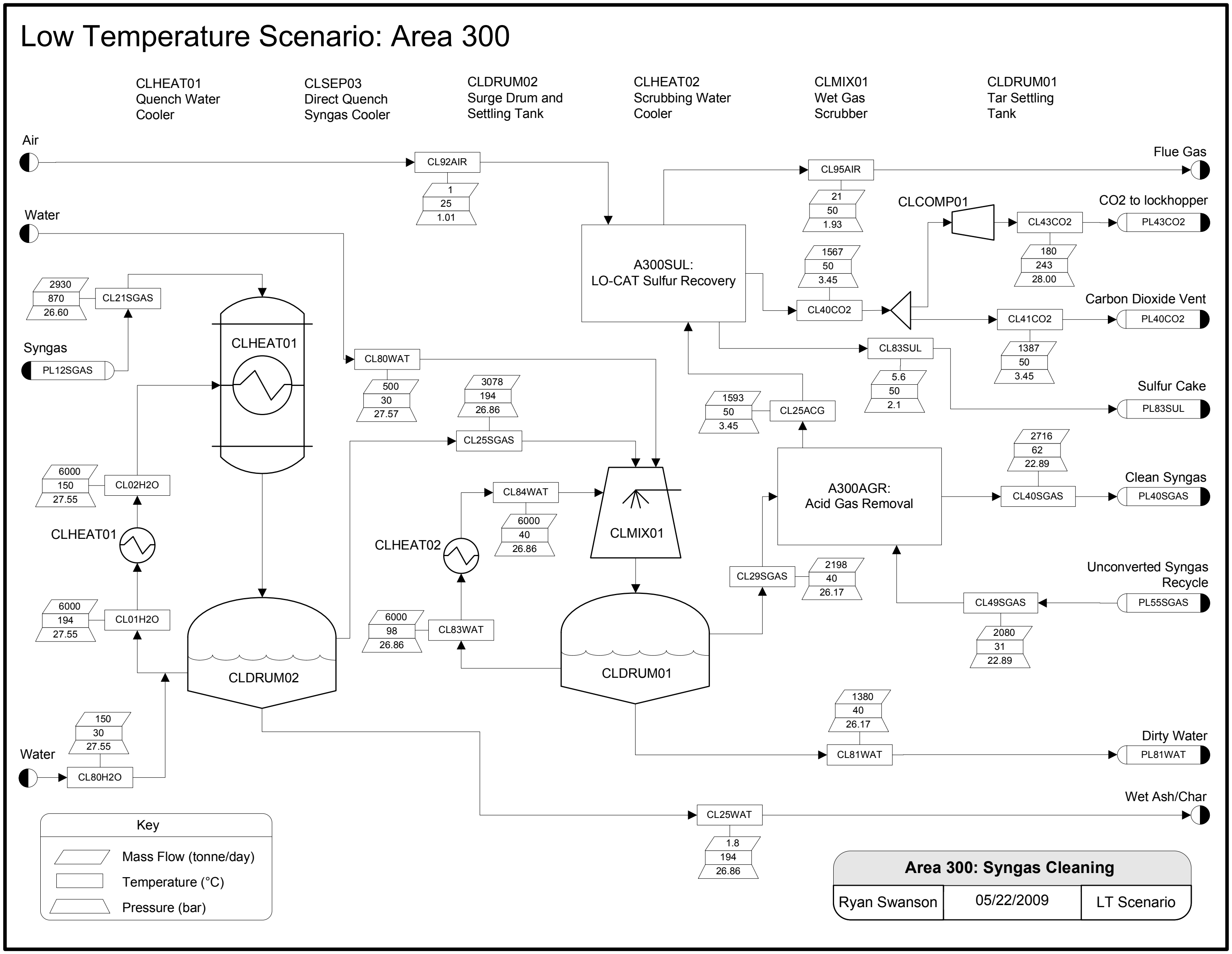


Figure 37. Syngas cleaning area process flow diagram for LT scenario

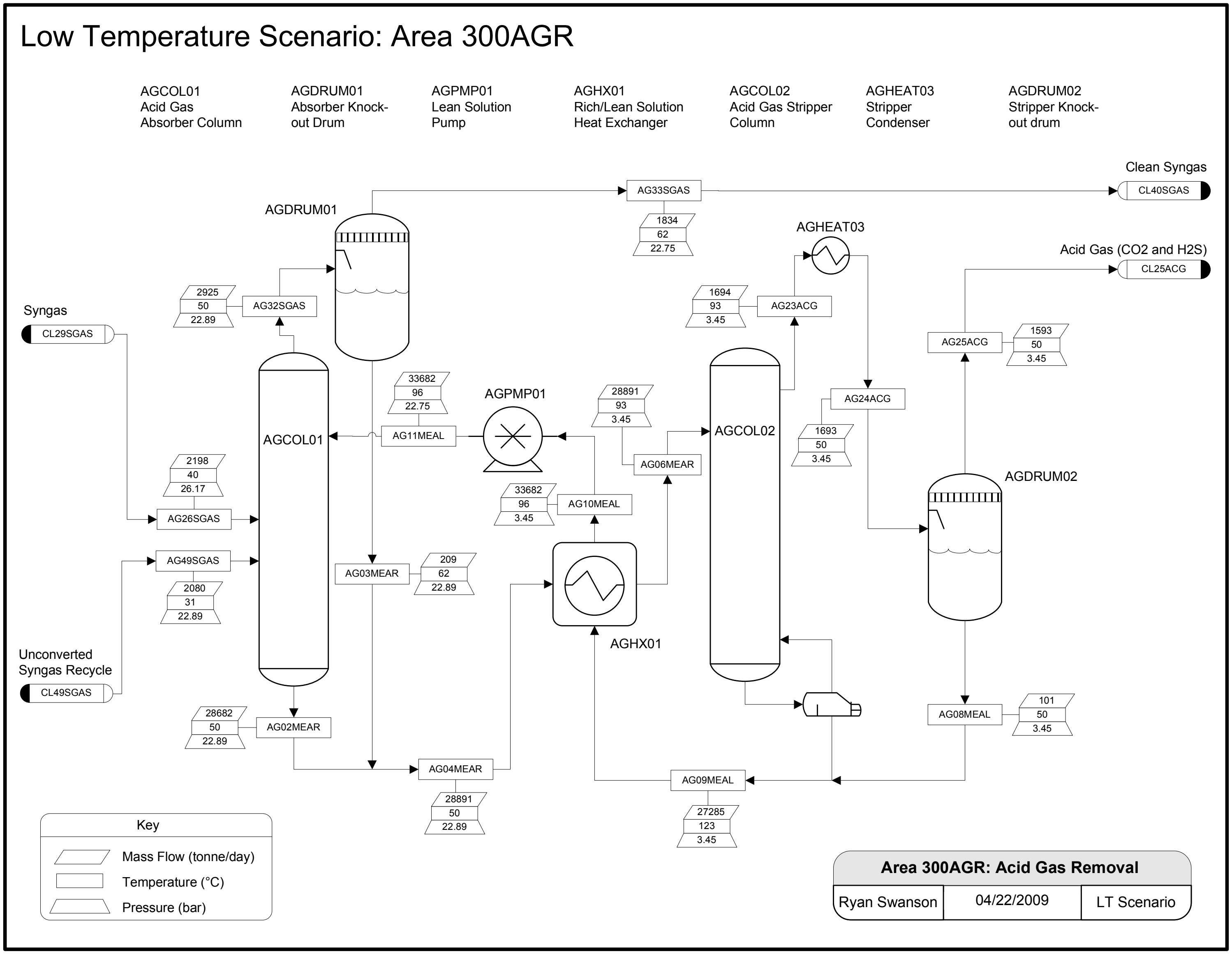


Figure 38. Acid gas removal area process flow diagram for LT scenario

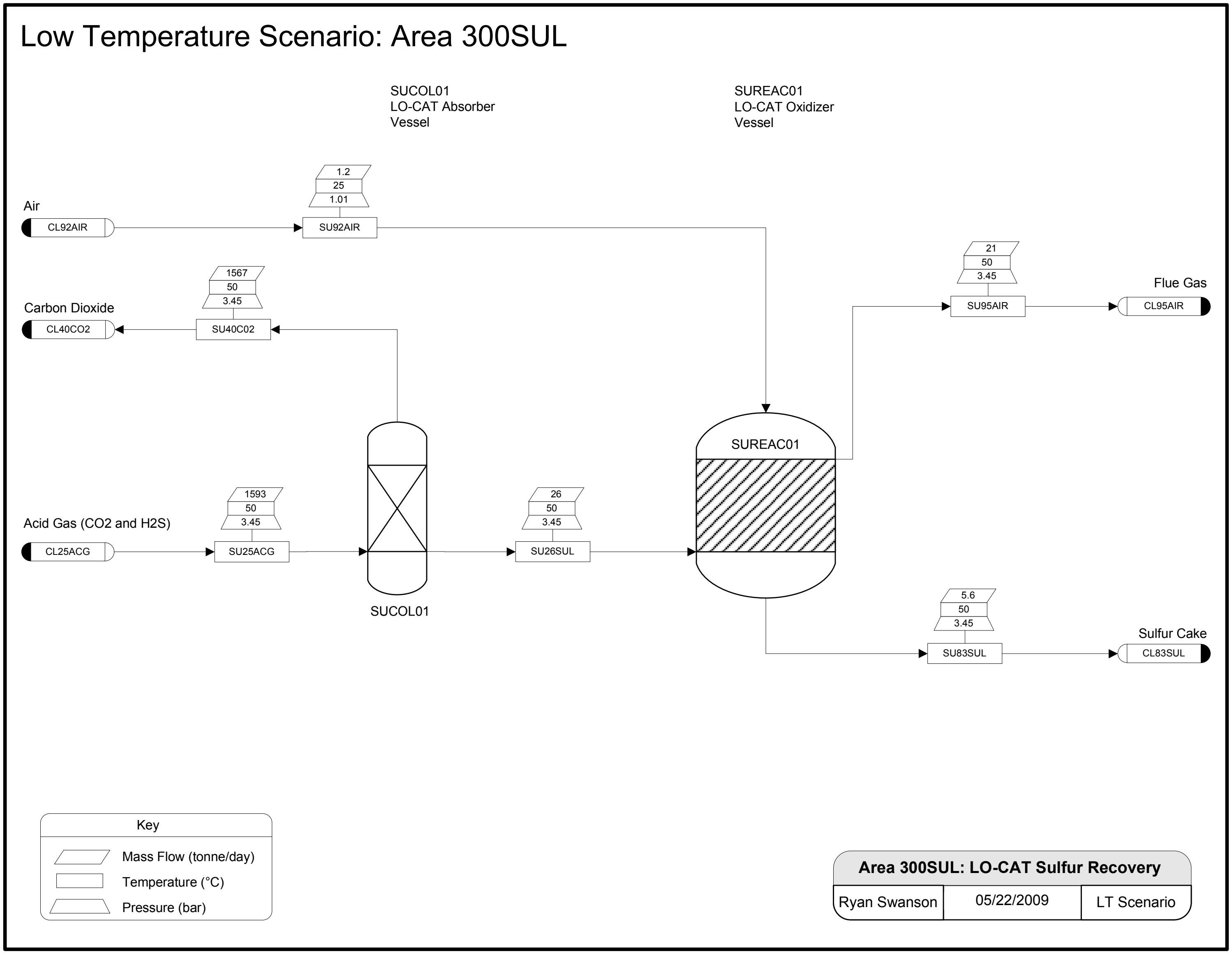


Figure 39. Sulfur recovery process flow diagram for LT scenario

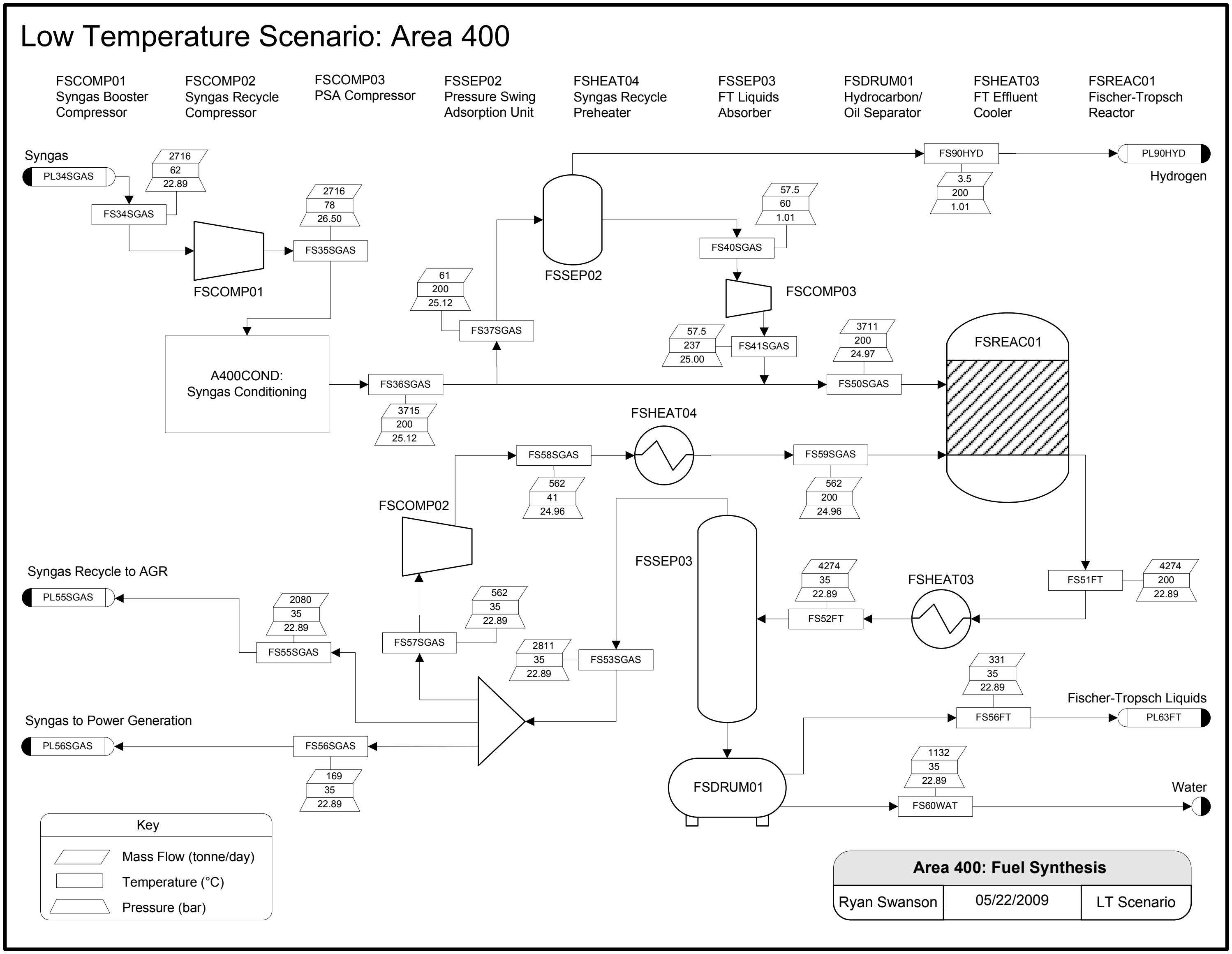


Figure 40. Fuel synthesis area process flow diagram for LT scenario

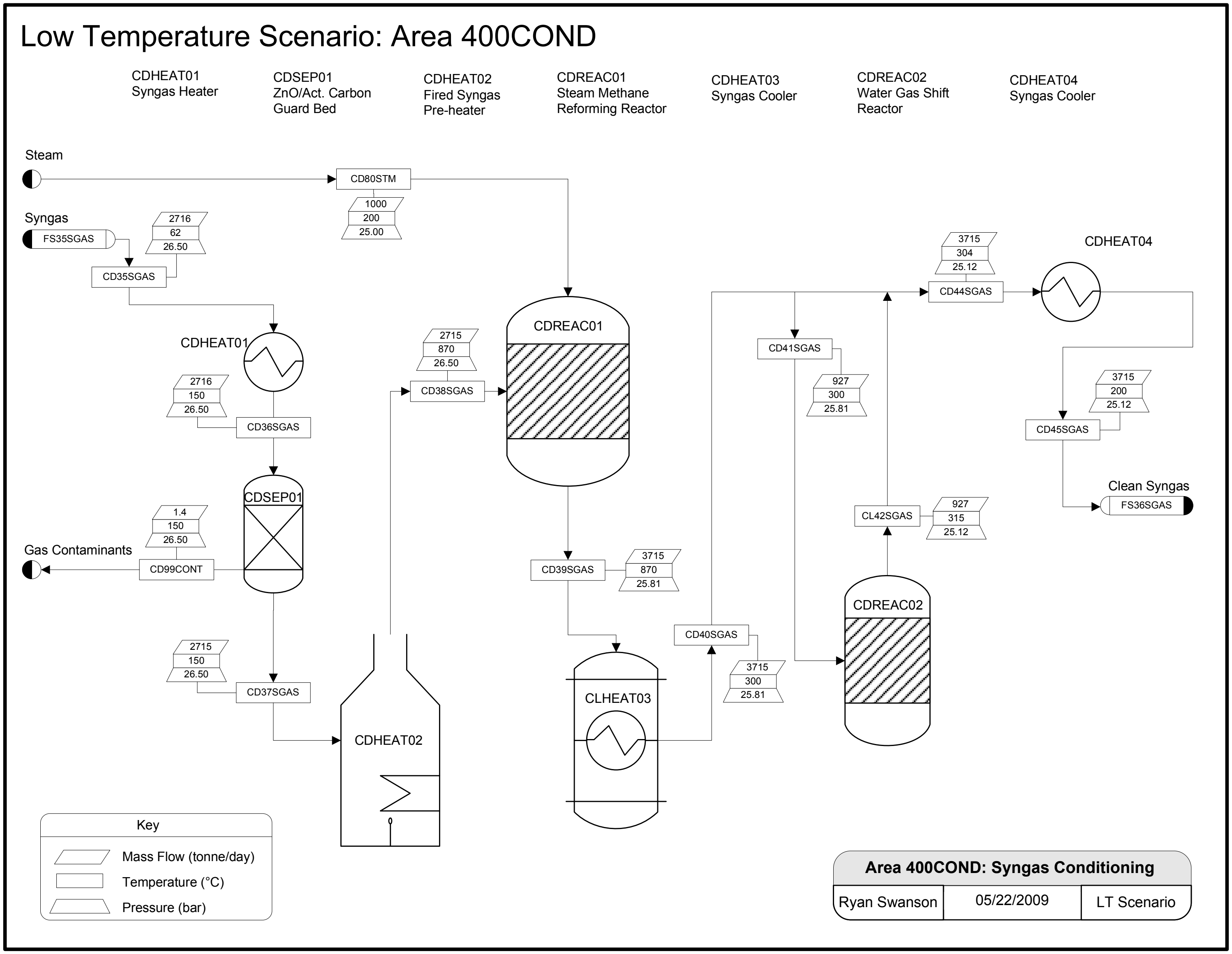


Figure 41. Syngas conditioning area process flow diagram for LT scenario

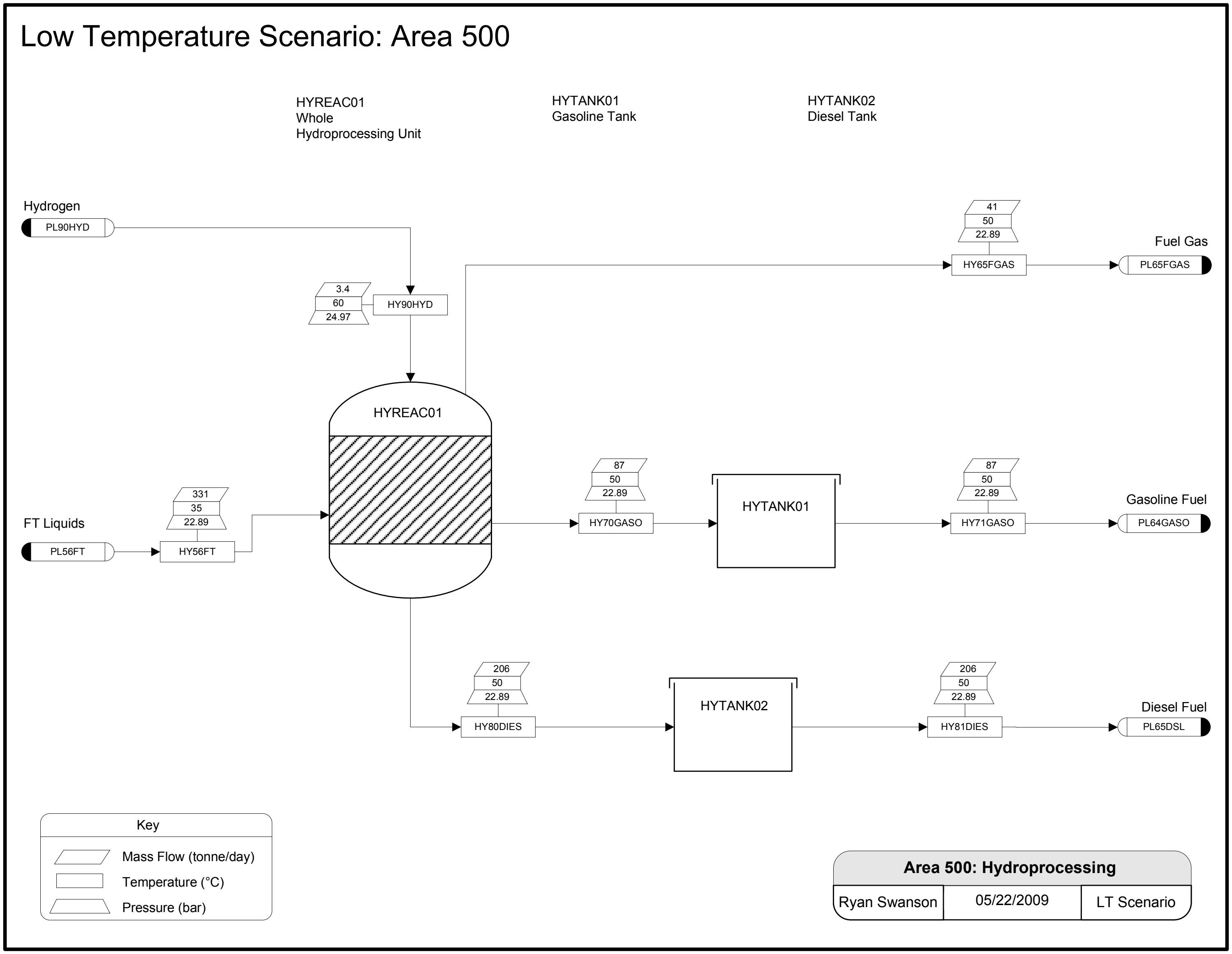


Figure 42. Hydroprocessing area process diagram for LT scenario

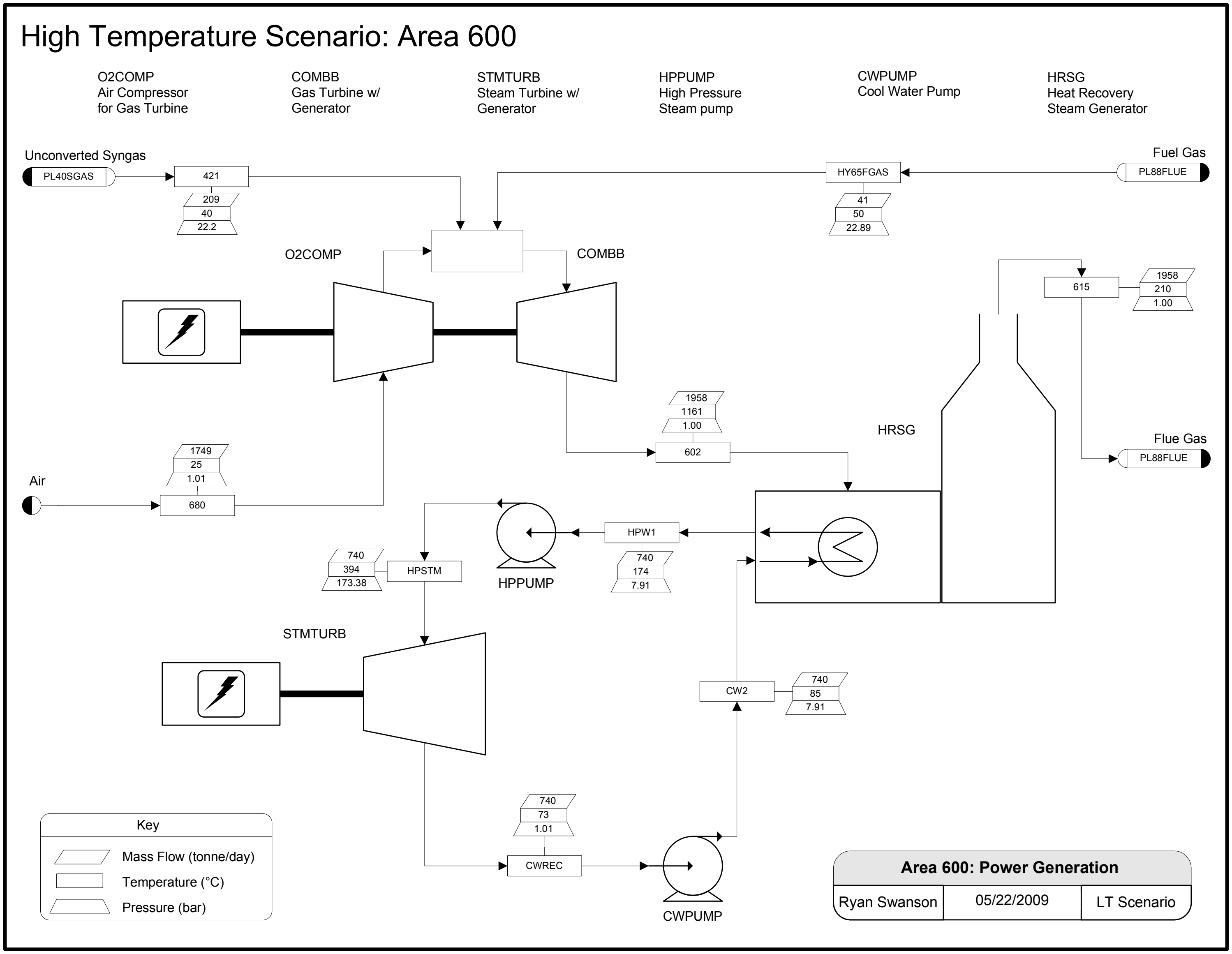


Figure 43. Power generation area process flow diagram for LT scenario

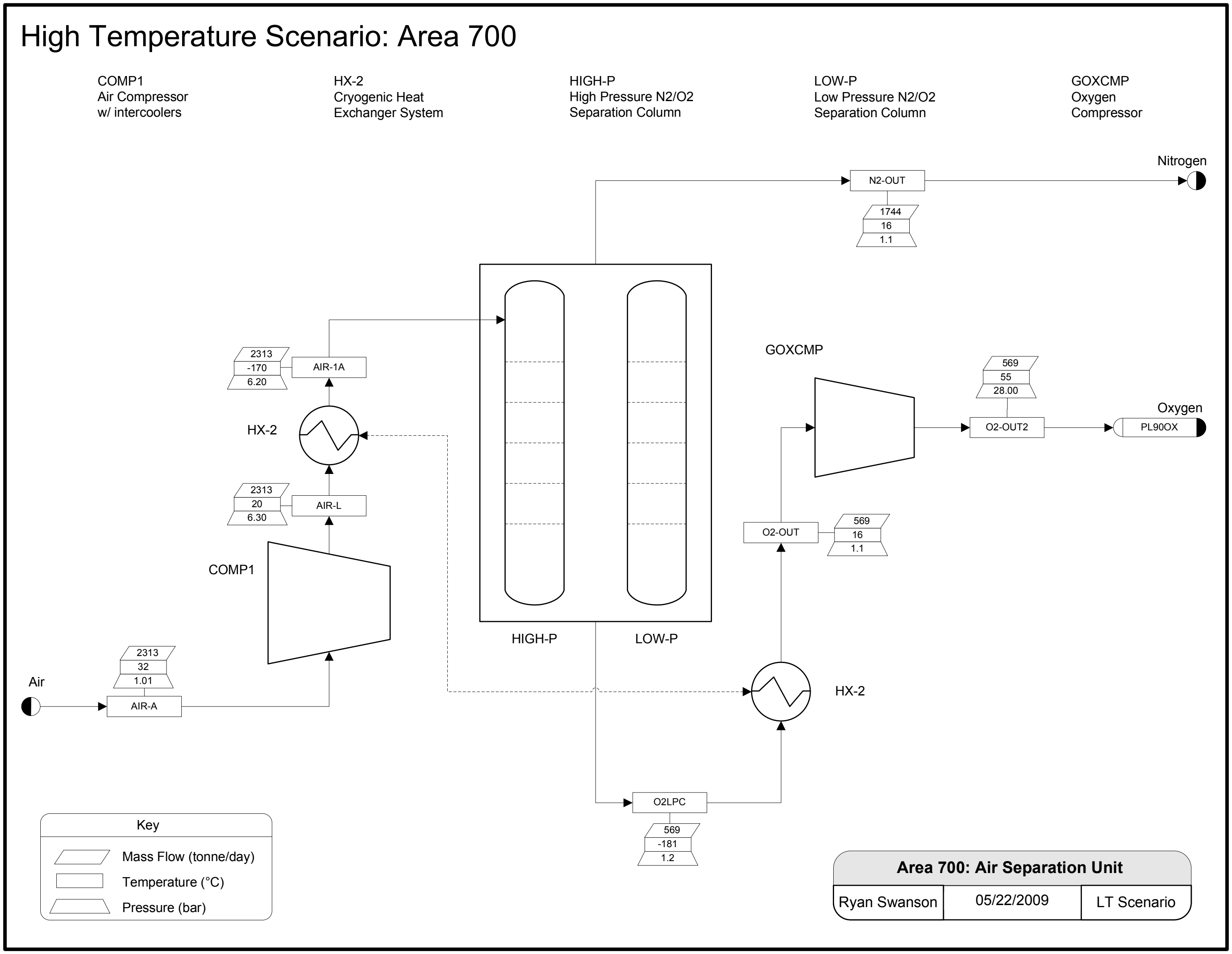


Figure 44. Air separation unit process flow diagram for LT scenario

APPENDIX E. STREAM DATA

E.1 High Temperature Scenario

Table 39. Overall plant stream data for HT scenario

HT Overall Plant	PL00BMAS	PL08BMAS	PL09CAIR	PL17SLAG	PL21SGAS	PL34SGAS	PL42CO2	PL43CO2	PL47SGAS	PL48SGAS	PL49SGAS	PL50FT	PL52FLUE	PL65FGAS	PL71GASO	PL81DIES	PL81STM	PL81WAT	PL83SUL	PL84STM	PL88FLUE	PL89AIR	PL90HYD	PL90OX	PL90STM	PL92NTGN	PL92WAT	PL98STM	PL99AIR
Temperature (C)	25	90	25	50	1300	62	250	53	45	45	45	35	220	35	36	37	200	40	50	120	273	32	30	149	190	-179	120	200	30
Pressure (bar)	1.01	1.01	1.01	26.62	26.62	22.75	28.00	3.45	23.58	23.58	23.58	22.20	1.03	22.20	1.03	1.03	1.98	24.82	1.93	1.98	1.00	1.01	1.00	28.00	10.00	1.20	1.98	28.00	1.01
Vapor Fraction	0.00	0.00	1.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00	1.00	0.00	0.00	1.00	0.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00
Volume Flow** (m^3/sec)	0.43	0.14	2.33	0	11.23	3.45	0.07	3.32	0.02	1.64	0.11	0.01	4.33	0.02	0	0.01	50.53	0.02	0	42.34	45.37	29.07	0.63	0.34	1.30	5.76	4.70	0.02	22.37
Mole Flow** (kmol/hr)	1542	513.97	343.36	0	8192	10089	175.05	1541	71.82	5260	355.46	89.49	393.42	74.55	41.08	48.97	9251	3276	13.46	9251	3595	4177	90.27	957.07	1271	3255	1028	2220	3237
Mass Flow (tonnes/day)	2667	2222	237.82	114.00	3825	3377	180.00	1585	29.08	2130	143.95	427.14	266.91	52.77	112.62	266.11	4000	1501	7.20	4000	2439	2903	4.37	743.70	549.62	2189	444.44	960.00	2242
H2O	666.67	222.22	0	0	988.43	45.53	2.50	21.97	0	0	0	0	27.89	0	0	0	0	1348	4.04	0	233.43	0	0	0	549.62	0	444.44	960.00	0
CO	0	0	0	0	1457	1818	0.47	4.15	11.24	823.54	55.65	0	0	0	0	0	0	22.10	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	122.88	288.34	0	0	1.83	134.08	9.06	0	0	0	0	0	0	0	0	0	0	0	4.37	0	0	0	0	0	0
CO2	0	0	0	0	1184	190.01	175.21	1543	2.40	175.74	11.88	0	40.00	0	0	0	0	128.26	0	0	352.24	0	0	0	0	0	0	0	0
O2	0	0	55.97	0	0	0	0	0	0	0	0	0	10.28	0	0	0	0	0	0	0	104.25	672.19	0	700.00	0	0	0	0	527.27
N2	0	0	181.85	0	17.68	0	1.78	15.63	0	0	0	0	181.85	0	0	0	0	0.26	0	0	1715	2194	0	0	0	2189	0	0	1715
CH4	0	0	0	0	0.02	63.41	0.01	0.07	0.87	63.47	4.29	0	0	14.94	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	106.90	0.01	0.10	1.46	107.01	7.23	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C3	0	0	0	0	0	141.45	0.01	0.10	1.93	141.56	9.57	0	0	37.83	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C4	0	0	0	0	0	167.38	0.02	0.15	2.29	167.55	11.32	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	4.50	0.03	0	0	0	0	0	0	0	0	0	0	0	1.02	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0.11	0	0	0	0	0	0	0	0	0	0	0	0	0.10	0	0	0	0	0	0	0	0	0	0	0
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3.16	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4000	0	0	4000	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0.02	0	0	0	0	0	0	0	0.08	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	170.22	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	2.29	0	0	0.03	2.29	0.15	15.49	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C6	0	0	0	0	0	2.46	0	0	0.03	2.46	0.17	16.65	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C7	0	0	0	0	0	1.15	0	0	0.02	1.15	0.08	17.54	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C8	0	0	0	0	0	1.18	0	0	0.02	1.18	0.08	17.99	0	0	112.62	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C9	0	0	0	0	0	1.11	0	0	0.02	1.19	0.08	18.10	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C10	0	0	0	0	0	1.16	0	0	0.02	1.19	0.08	18.12	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C11	0	0	0	0	0	0.83	0	0	0.02	1.16	0.08	17.62	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12	0	0	0	0	0	0.13	0	0	0.01	0.54	0.04	17.25	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C13	0	0	0	0	0	0.02	0	0	0.01	0.52	0.04	16.70	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C14	0	0	0	0	0	0	0	0	0.01	0.50	0.03	16.16	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C15	0	0	0	0	0	0	0	0	0.01	0.49	0.03	15.57	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C16	0	0	0	0	0	0	0	0	0.01	0.47	0.03	14.93	0	0	0	266.11	0	0	0	0	0	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0	0	0	0	14.75	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	14.05	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	13.35	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	12.64	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0.30	1.20	0	0	0.02	1.12	0.08	0	0	0	0	0	0	0.13	0	0	0	0	0	0	0	0	0	0	0
AR	0	0	0	0	43.70	544.37	0	0	6.87	502.93	33.99	0	6.87	0	0	0	0	1.03	0	0	33.99	37.28	0	43.70	0	0.12	0	0	0
BIOMASS	2000	2000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOOT	0	0	0	0	6.00	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SLAG	0	0	0	114.00	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

Table 40. Preprocessing area stream data for HT scenario

HT A100	CH00BMAS	CH02BMAS	CH03BMAS	CH90BMAS	DR05BMAS	DR81STM	DR84STM	DR92WAT	GR06BMAS	GR08BMAS	GR90BMAS
Temperature (C)	25	25	25	25	90	200	120	120	90	90	90
Pressure (bar)	1.01	1.01	1.01	1.01	1.01	1.98	1.98	1.98	1.01	1.01	1.01
Vapor Fraction	0.00	0.00	0.00	0.00	0.00	1.00	1.00	1.00	0.00	0.00	0.00
Volume Flow** (m^3/sec)	0.43	0.43	0.43	0	0.14	50.53	42.34	4.70	0.14	0.14	0
Mole Flow** (kmol/hr)	1542	1542	1542	0	513.97	9251	9251	1028	513.97	513.97	0
Mass Flow (tonnes/day)	2667	2709	2667	42.20	2222	4000	4000	444.44	2240	2222	17.60
H2O	666.67	666.67	666.67	0	222.22	0	0	444.44	222.22	222.22	0
CO	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0	0	0
C3	0	0	0	0	0	0	0	0	0	0	0
C4	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0
TAR	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	4000	4000	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0	0	0	0
C6	0	0	0	0	0	0	0	0	0	0	0
C7	0	0	0	0	0	0	0	0	0	0	0
C8	0	0	0	0	0	0	0	0	0	0	0
C9	0	0	0	0	0	0	0	0	0	0	0
C10	0	0	0	0	0	0	0	0	0	0	0
C11	0	0	0	0	0	0	0	0	0	0	0
C12	0	0	0	0	0	0	0	0	0	0	0
C13	0	0	0	0	0	0	0	0	0	0	0
C14	0	0	0	0	0	0	0	0	0	0	0
C15	0	0	0	0	0	0	0	0	0	0	0
C16	0	0	0	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	2000	2042	2000	42.20	2000	0	0	0	2018	2000	17.60
ASH	0	0	0	0	0	0	0	0	0	0	0
SOOT	0	0	0	0	0	0	0	0	0	0	0
SLAG	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

Table 41. Gasification area stream data for HT scenario

HT A200	GS02SLAG	GS08BMAS	GS09CAIR	GS11BMAS	GS42CO2	GS52FLUE	GS54SGAS	GS81STM	GS84STM	GS90OX	GS92CO2	GS98STM	S06SGAS	SL01SLAG	SL02SGAS
Temperature (C)	50	90	25	90	250	220	45	200	120	149	92	200	203	1300	1300
Pressure (bar)	26.62	1.01	1.01	1.01	28.00	1.03	23.58	1.98	1.98	28.00	28.00	28.00	25.93	26.62	26.62
Vapor Fraction	0.00	0.00	1.00	0.53	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00	0.00	1.00
Volume Flow** (m^3/sec)	0	0.14	2.33	30.59	0.07	4.33	0.02	50.53	42.34	0.34	0.04	0.02	3.50	0	11.23
Mole Flow** (kmol/hr)	0	513.97	343.36	6974	175.05	393.42	71.82	9251	9251	957.07	158.25	2220	8308	0	8192
Mass Flow (tonnes/day)	114.00	2222	237.82	2222	180.00	266.91	29.08	4000	4000	743.70	166.90	960.00	3869	114.00	3825
H2O	0	222.22	0	222.22	2.50	27.89	0	0	0	0	0	960.00	1038	0	988.43
CO	0	0	0	0	0.47	0	11.24	0	0	0	0.45	0	1457	0	1457
H2	0	0	0	101.20	0	0	1.83	0	0	0	0	0	122.88	0	122.88
CO2	0	0	0	0	175.21	40.00	2.40	0	0	0	166.45	0	1184	0	1184
O2	0	0	55.97	812.60	0	10.28	0	0	0	700.00	0	0	0	0	0
N2	0	0	181.85	16.00	1.78	181.85	0	0	0	0	0	0	17.68	0	17.68
CH4	0	0	0	0	0.01	0	0.87	0	0	0	0	0	0.02	0	0.02
C2H6	0	0	0	0	0.01	0	1.46	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C3	0	0	0	0	0.01	0	1.93	0	0	0	0	0	0	0	0
C4	0	0	0	0	0.02	0	2.29	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	4.50	0	4.50
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0.11	0	0.11
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	4.40	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	945.60	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	4000	4000	0	0	0	0	0	0
SO2	0	0	0	0	0	0.02	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0.03	0	0	0	0	0	0	0	0
C6	0	0	0	0	0	0	0.03	0	0	0	0	0	0	0	0
C7	0	0	0	0	0	0	0.02	0	0	0	0	0	0	0	0
C8	0	0	0	0	0	0	0.02	0	0	0	0	0	0	0	0
C9	0	0	0	0	0	0	0.02	0	0	0	0	0	0	0	0
C10	0	0	0	0	0	0	0.02	0	0	0	0	0	0	0	0
C11	0	0	0	0	0	0	0.02	0	0	0	0	0	0	0	0
C12	0	0	0	0	0	0	0.01	0	0	0	0	0	0	0	0
C13	0	0	0	0	0	0	0.01	0	0	0	0	0	0	0	0
C14	0	0	0	0	0	0	0.01	0	0	0	0	0	0	0	0
C15	0	0	0	0	0	0	0.01	0	0	0	0	0	0	0	0
C16	0	0	0	0	0	0	0.01	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0.02	0	0	0	0	0	0.30	0	0.30
AR	0	0	0	0	0	6.87	6.87	0	0	43.70	0	0	43.70	0	43.70
BIOMASS	0	2000	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	120.00	0	0	0	0	0	0	0	0	0	0	0
SOOT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	6.00
SLAG	114.00	0	0	0	0	0	0	0	0	0	0	0	0	114.00	0

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

Table 42. Syngas cleaning area stream data for HT scenario

HT A300	CL25ACG	CL25SGAS	CL26SGAS	CL28SGAS	CL29SGAS	CL32SGAS	CL40CO2	CL41CO2	CL42CO2	CL49SGAS	CL81WAT	CL82WAT	CL83SUL	CL83WAT	CL90STM	CL92AIR	CL95AIR	GS84STM	SG91STM	SL01H2O	SL02SGAS	SL03H2O	SL80H2O
Temperature (C)	50	203	291	240	60	40	53	53	250	45	30	40	50	60	190	25	50	62	250	203	1300	203	30
Pressure (bar)	3.45	25.93	24.82	24.82	24.82	24.82	3.45	3.45	28.00	23.58	24.82	24.82	1.93	24.82	10.00	1.01	1.93	22.75	25.86	26.62	26.62	25.93	26.62
Vapor Fraction	1.00	1.00	1.00	1.00	0.69	1.00	1.00	1.00	1.00	1.00	0.00	0.00	0.00	0.00	1.00	1.00	0.93	1.00	1.00	0.00	1.00	0.00	0.00
Volume Flow** (m^3/sec)	3.69	1.08	2.01	4.54	2.07	1.86	3.70	0.38	0.07	1.64	0	0.02	0	0.02	1.30	0.01	0.01	3.45	0.54	0.07	11.23	0	0
Mole Flow** (kmol/hr)	1728	2576	3847	9579	9579	6419	1716	175.05	175.05	5260	115.64	3276	13.46	2313	1271	2.07	2.58	10089	1271	9251	8192	0	115.64
Mass Flow (tonnes/day)	1773	1199	1749	4418	4418	2967	1765	180.00	180.00	2130	50.00	1501	7.20	1000	549.62	1.59	2.32	3377	549.62	4000	3825	6.00	50.00
H_2O	27.18	321.91	591.14	1308	1308	9.19	24.46	2.50	2.50	0	50.00	1348	4.04	1000	549.62	0	0.45	45.53	549.62	4000	988.43	0	50.00
CO	4.62	451.69	15.74	1021	1021	999.01	4.62	0.47	0.47	823.54	0	22.10	0	0	0	0	0	1818	0	0	1457	0	0
H_2	0	38.09	69.47	154.26	154.26	154.26	0	0	0	134.08	0	0	0	0	0	0	0	288.34	0	0	122.88	0	0
CO_2	1718	367.08	1052	1869	1869	1741	1718	175.21	175.21	175.74	0	128.26	0	0	0	0	0	190.01	0	0	1184	0	0
O_2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1.59	0.02	0	0	0	0	0	0
N_2	17.41	5.48	5.48	17.68	17.68	17.43	17.41	1.78	1.78	0	0	0.26	0	0	0	0	0	0	0	0	17.68	0	0
CH_4	0.08	0.01	0.01	0.02	0.02	0.02	0.08	0.01	0.01	63.47	0	0	0	0	0	0	0	63.41	0	0	0.02	0	0
C_2H_6	0.11	0	0	0	0	0	0.11	0.01	0.01	107.01	0	0	0	0	0	0	0	106.90	0	0	0	0	0
C_2H_4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C_2H_2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C3	0.11	0	0	0	0	0	0.11	0.01	0.01	141.56	0	0	0	0	0	0	0	141.45	0	0	0	0	0
C4	0.17	0	0	0	0	0	0.17	0.02	0.02	167.55	0	0	0	0	0	0	0	167.38	0	0	0	0	0
H_2S	3.39	1.40	1.40	4.50	4.50	3.48	0	0	0	0	0	1.02	0	0	0	0	0.03	0.03	0	0	4.50	0	0
NH_3	0	0.03	0.03	0.11	0.11	0.01	0	0	0	0	0	0.10	0	0	0	0	0	0	0	0	0.11	0	0
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	3.16	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO_2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO_2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0	0	2.29	0	0	0	0	0	0	0	2.29	0	0	0	0	0
C6	0	0	0	0	0	0	0	0	0	2.46	0	0	0	0	0	0	0	2.46	0	0	0	0	0
C7	0	0	0	0	0	0	0	0	0	1.15	0	0	0	0	0	0	0	1.15	0	0	0	0	0
C8	0	0	0	0	0	0	0	0	0	1.18	0	0	0	0	0	0	0	1.18	0	0	0	0	0
C9	0.08	0	0	0	0	0	0	0	0	1.19	0	0	0	0	0	0	0.08	1.11	0	0	0	0	0
C10	0.03	0	0	0	0	0	0	0	0	1.19	0	0	0	0	0	0	0.03	1.16	0	0	0	0	0
C11	0.23	0	0	0	0	0	0	0	0	1.16	0	0	0	0	0	0	0.23	0.83	0	0	0	0	0
C12	0.13	0	0	0	0	0	0	0	0	0.54	0	0	0	0	0	0	0.13	0.13	0	0	0	0	0
C13	0.04	0	0	0	0	0	0	0	0	0.52	0	0	0	0	0	0	0.04	0.02	0	0	0	0	0
C14	0.01	0	0	0	0	0	0	0	0	0.50	0	0	0	0	0	0	0.01	0	0	0	0	0	0
C15	0	0	0	0	0	0	0	0	0	0.49	0	0	0	0	0	0	0	0	0	0	0	0	0
C16	0	0	0	0	0	0	0	0	0	0.47	0	0	0	0	0	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
COS	0.08	0.09	0.09	0.30	0.30	0.17	0	0	0	1.12	0	0.13	0	0	0	0	0.08	1.20	0	0	0.30	0	0
AR	1.22	13.55	13.55	43.70	43.70	42.66	0	0	0	502.93	0	1.03	0	0	0	0	1.22	544.37	0	0	43.70	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOOT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	6.00	6.00	0
SLAG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

Table 43. Acid gas removal and sulfur recovery areas stream data for HT scenario

HT A300AGR	AG02MEAR	AG03MEAR	AG04MEAR	AG06MEAR	AG08MEAL	AG09MEAL	AG10MEAL	AG11MEAL	AG23ACG	AG24ACG	AG25ACG	AG31SGAS	AG32SGAS	AG33SGAS	AG49SGAS
Temperature (C)	58	62	58	86	50	123	96	96	86	50	50	40	58	62	45
Pressure (bar)	22.75	22.75	22.75	3.45	3.45	3.45	3.45	20.68	3.45	3.45	3.45	24.82	22.75	22.75	23.58
Vapor Fraction	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.98	0.83	1.00	1.00	0.93	1.00	1.00
Volume Flow** (m^3/sec)	0.480	0	0.480	1.590	0	0.480	0.470	0.470	4.820	3.690	3.690	1.860	3.410	3.450	1.640
Mole Flow** (kmol/hr)	84877	699.000	85576	85576	342.110	83848	83848	83848	2070	2070	1728	6419	10788	10089	5260
Mass Flow (tonnes/day)	42560	308.340	42868	42868	155.170	41096	41096	41096	1928	1928	1773	2967	3685	3377	2130
H2O	33956	297.460	34254	34254	144.090	34226	34226	34226	171.270	171.270	27.180	9.190	342.990	45.530	0
CO	0	4.640	4.640	4.640	0.010	0.010	0.010	0.010	4.630	4.630	4.620	999.010	1823	1818	823.540
H2	0	0	0	0	0	0	0	0	0	0	0	154.260	288.340	288.340	134.080
CO2	1725	1.650	1727	1727	8.690	8.690	8.690	8.690	1727	1727	1718	1741	191.660	190.010	175.740
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	17.430	0	17.430	17.430	0.020	0.020	0.020	0.020	17.430	17.430	17.410	17.430	0	0	0
CH4	0	0.080	0.080	0.080	0	0	0	0	0.080	0.080	0.080	0.020	63.490	63.410	63.470
C2H6	0	0.110	0.110	0.110	0	0	0	0	0.110	0.110	0.110	0	107.010	106.900	107.010
C2H4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C3	0	0.110	0.110	0.110	0	0	0	0	0.110	0.110	0.110	0	141.560	141.450	141.560
C4	0	0.170	0.170	0.170	0	0	0	0	0.170	0.170	0.170	0	167.550	167.380	167.550
H2S	3.450	0	3.450	3.450	0.060	0.060	0.060	0.060	3.450	3.450	3.390	3.480	0.030	0.030	0
NH3	0.010	0	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0	0.010	0	0	0
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	6858	0	6858	6858	0	6858	6858	6858	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0	0	0	0	0	2.290	2.290	2.290
C6	0	0	0	0	0	0	0	0	0	0	0	0	2.460	2.460	2.460
C7	0	0	0	0	0	0	0	0	0	0	0	0	1.150	1.150	1.150
C8	0	0	0	0	0	0	0	0	0	0	0	0	1.180	1.180	1.180
C9	0	0.080	0.080	0.080	0.010	0.010	0.010	0.010	0.080	0.080	0.080	0	1.190	1.110	1.190
C10	0	0.030	0.030	0.030	0	0	0	0	0.030	0.030	0.030	0	1.190	1.160	1.190
C11	0	0.330	0.330	0.330	0.100	0.100	0.100	0.100	0.320	0.320	0.230	0	1.160	0.830	1.160
C12	0	0.410	0.410	0.410	0.280	0.280	0.280	0.280	0.410	0.410	0.130	0	0.540	0.130	0.540
C13	0	0.500	0.500	0.500	0.460	0.460	0.460	0.460	0.500	0.500	0.040	0	0.520	0.020	0.520
C14	0	0.500	0.500	0.500	0.490	0.490	0.490	0.490	0.500	0.500	0.010	0	0.500	0	0.500
C15	0	0.490	0.490	0.490	0.490	0.490	0.490	0.490	0.490	0.490	0	0	0.490	0	0.490
C16	0	0.470	0.470	0.470	0.470	0.470	0.470	0.470	0.470	0.470	0	0	0.470	0	0.470
C17	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
COS	0	0.080	0.080	0.080	0	0	0	0	0.080	0.080	0.080	0.170	1.290	1.200	1.120
AR	0	1.220	1.220	1.220	0	0	0	0	1.220	1.220	1.220	42.660	545.590	544.370	502.930
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOOT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SLAG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

A300SUL	SU25ACG	SU26SUL	SU40CO2	SU83SUL	SU92AIR	SU93AIR	SU94AIR	SU95AIR
Temperature (C)	50	53	53	50	25	100	50	50
Pressure (bar)	3.45	3.45	3.45	1.93	1.01	2.07	1.93	1.93
Vapor Fraction	1.00	0.49	1.00	0.00	1.00	1.00	1.00	0.93
Volume Flow** (m^3/sec)	3.690	0.010	3.700	0	0.010	0.010	0.010	0.010
Mole Flow** (kmol/hr)	1728	11.910	1716	13.460	2.070	2.070	2.070	2.580
Mass Flow (tonnes/day)	1773	7.930	1765	7.200	1.590	1.590	1.590	2.320
H2O	27.180	2.720	24.460	4.040	0	0	0	0.450
CO	4.620	0	4.620	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
CO2	1718	0	1718	0	0	0	0	0
O2	0	0	0	0	1.590	1.590	1.590	0.020
N2	17.410	0	17.410	0	0	0	0	0
CH4	0.080	0	0.080	0	0	0	0	0
C2H6	0.110	0	0.110	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0
C3	0.110	0	0.110	0	0	0	0	0
C4	0.170	0	0.170	0	0	0	0	0
H2S	3.390	3.390	0	0	0	0	0	0.030
NH3	0	0	0	0	0	0	0	0
TAR	0	0	0	0	0	0	0	0
SULFUR	0	0	0	3.160	0	0	0	0
CARBON	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0
C6	0	0	0	0	0	0	0	0
C7	0	0	0	0	0	0	0	0
C8	0	0	0	0	0	0	0	0
C9	0.080	0.080	0	0	0	0	0	0.080
C10	0.030	0.030	0	0	0	0	0	0.030
C11	0.230	0.230	0	0	0	0	0	0.230
C12	0.130	0.130	0	0	0	0	0	0.130
C13	0.040	0.040	0	0	0	0	0	0.040
C14	0.010	0.010	0	0	0	0	0	0.010
C15	0	0	0	0	0	0	0	0
C16	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0
COS	0.080	0.080	0	0	0	0	0	0.080
AR	1.220	1.220	0	0	0	0	0	1.220
BIOMASS	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0
SOOT	0	0	0	0	0	0	0	0
SLAG	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

Table 44. Fuel synthesis area stream data for HT scenario

HT A400	FS34SGAS	FS35SGAS	FS36SGAS	FS37SGAS	FS40SGAS	FS42SGAS	FS43SGAS	FS44FT	FS44SGAS	FS45FT	FS45SGAS	FS46FT	FS46SGAS	FS47SGAS	FS48SGAS	FS49SGAS	FS50FT	FS50SGAS	FS51SGAS	FS52SGAS	FS60WAT	FS89HYD	FS90HYD	FS99CONT
Temperature (C)	62	76	200	200	30	30	417	200	202	35	202	43	0	45	45	45	35	45	50	200	35	30	30	200
Pressure (bar)	22.75	26.00	26.00	24.97	1.00	1.00	24.97	23.58	24.97	23.58	24.96	23.58	0.00	23.58	23.58	23.58	22.20	23.58	24.96	24.96	22.20	1.00	1.00	24.97
Vapor Fraction	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.86	1.00	0.00	1.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00
Volume Flow** (m^3/sec)	3.45	3.16	4.29	4.47	1.26	0.63	0.06	4.84	4.45	2.71	5.86	0.02	2.74	0.02	1.64	0.11	0.01	1.00	0.96	1.42	0.01	0.63	0.63	0
Mole Flow** (kmol/hr)	10089	10089	10089	10089	179.78	89.51	89.51	10438	9998	10438	13198	1575	8887	71.82	5260	355.46	89.49	3199	3199	3199	1483	90.27	90.27	0.04
Mass Flow (tonnes/day)	3377	3377	3377	3377	60.18	55.81	55.81	4668	3372	4668	4668	1069	3599	29.08	2130	143.95	427.14	1296	1296	1296	641.34	4.37	4.37	0.03
H2O	45.53	45.53	45.53	45.53	0.81	0.81	0.81	642.07	45.53	642.07	45.53	642.07	0	0	0	0	0	0	0	0	641.34	0	0	0
CO	1818	1818	1818	1818	32.40	32.40	32.40	1391	1818	1391	2319	0	1391	11.24	823.54	55.65	0	500.87	500.87	500.87	0	0	0	0
H2	288.34	288.34	288.34	288.34	5.14	0.77	0.77	225.34	283.97	225.34	365.52	0	226.52	1.83	134.08	9.06	0	81.55	81.55	81.55	0	4.37	4.37	0
CO2	190.01	190.01	190.01	190.01	3.39	3.39	3.39	296.89	190.01	296.89	296.89	0	296.90	2.40	175.74	11.88	0	106.88	106.88	106.88	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	63.41	63.41	63.41	63.41	1.13	1.13	1.13	107.33	63.41	107.33	102.02	0	107.23	0.87	63.47	4.29	0	38.60	38.60	38.60	0	0	0	0
C2H6	106.90	106.90	106.90	106.90	1.91	1.91	1.91	180.95	106.90	180.95	171.99	0	180.79	1.46	107.01	7.23	0	65.08	65.08	65.08	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C3	141.45	141.45	141.45	141.45	2.52	2.52	2.52	239.37	141.45	239.37	227.54	0	239.16	1.93	141.56	9.57	0	86.10	86.10	86.10	0	0	0	0
C4	167.38	167.38	167.38	167.38	2.98	2.98	2.98	283.31	167.38	283.31	269.28	0	283.06	2.29	167.55	11.32	0	101.90	101.90	101.90	0	0	0	0
H2S	0.03	0.03	0.03	0	0	0	0	0	0	0	0	0	0.00	0	0	0	0	0	0	0	0	0	0	0.03
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0.00	0	0	0	0	0	0	0	0	0	0	0
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	170.22	0	170.22	0	170.22	0	0	0	0	170.22	0	0	0	0	0	0	0
C5	2.29	2.29	2.29	2.29	0.04	0.04	0.04	19.36	2.29	19.36	3.69	15.49	3.87	0.03	2.29	0.15	15.49	1.39	1.39	1.39	0	0	0	0
C6	2.46	2.46	2.46	2.46	0.04	0.04	0.04	20.81	2.46	20.81	3.96	16.65	4.16	0.03	2.46	0.17	16.65	1.50	1.50	1.50	0	0	0	0
C7	1.15	1.15	1.15	1.15	0.02	0.02	0.02	19.49	1.15	19.49	1.85	17.54	1.95	0.02	1.15	0.08	17.54	0.70	0.70	0.70	0	0	0	0
C8	1.18	1.18	1.18	1.18	0.02	0.02	0.02	19.99	1.18	19.99	1.90	17.99	2.00	0.02	1.18	0.08	17.99	0.72	0.72	0.72	0	0	0	0
C9	1.11	1.11	1.11	1.11	0.02	0.02	0.02	20.11	1.11	20.11	1.83	18.10	2.01	0.02	1.19	0.08	18.10	0.72	0.72	0.72	0	0	0	0
C10	1.16	1.16	1.16	1.16	0.02	0.02	0.02	20.13	1.16	20.13	1.88	18.12	2.01	0.02	1.19	0.08	18.12	0.72	0.72	0.72	0	0	0	0
C11	0.83	0.83	0.83	0.83	0.01	0.01	0.01	19.58	0.83	19.58	1.54	17.62	1.96	0.02	1.16	0.08	17.62	0.71	0.71	0.71	0	0	0	0
C12	0.13	0.13	0.13	0.13	0	0	0	18.16	0.13	18.16	0.45	17.25	0.91	0.01	0.54	0.04	17.25	0.33	0.33	0.33	0	0	0	0
C13	0.02	0.02	0.02	0.02	0	0	0	17.58	0.02	17.58	0.33	16.70	0.88	0.01	0.52	0.04	16.70	0.32	0.32	0.32	0	0	0	0
C14	0	0	0	0	0	0	0	17.01	0	17.01	0.31	16.16	0.85	0.01	0.50	0.03	16.16	0.31	0.31	0.31	0	0	0	0
C15	0	0	0	0	0	0	0	16.39	0	16.39	0.30	15.57	0.82	0.01	0.49	0.03	15.57	0.29	0.29	0.29	0	0	0	0
C16	0	0	0	0	0	0	0	15.72	0	15.72	0.28	14.93	0.79	0.01	0.47	0.03	14.93	0.28	0.28	0.28	0	0	0	0
C17	0	0	0	0	0	0	0	14.75	0	14.75	0	14.75	0	0	0	0	14.75	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	14.05	0	14.05	0	14.05	0	0	0	0	14.05	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	13.35	0	13.35	0	13.35	0	0	0	0	13.35	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	12.64	0	12.64	0	12.64	0	0	0	0	12.64	0	0	0	0	0	0	0
COS	1.20	1.20	1.20	1.20	0.02	0.02	0.02	1.88	1.20	1.88	1.88	0	1.88	0.02	1.12	0.08	0	0.68	0.68	0.68	0	0	0	0
AR	544.37	544.37	544.37	544.37	9.70	9.70	9.70	850.25	544.37	850.25	850.25	0	849.66	6.87	502.93	33.99	0	305.88	305.88	305.88	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOOT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SLAG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

Table 45. Hydroprocessing, power generation, and air separation areas stream data for HT scenario

HT A500	HY50FT	HY65FGAS	HY70GASO	HY71GASO	HY80DIES	HY81DIES	HY90HYD	HT A600	421	602	615	680	CW2	CWREC	HPSTM1	HPSTM2	HT A700	AIR-1A	AIR-A	AIR-L	N2-OUT	O2LPC	O2-OUT	O2-OUT2
Temperature (C)	35	35	35	36	35	37	30		45	1144	273	30	85	73	170	565		-170	32	20	16	-177	16	68
Pressure (bar)	22.20	22.20	22.20	1.03	22.20	1.03	1.00		23.58	1.00	1.00	1.01	7.91	0.30	7.91	173.38		6.20	1.01	6.30	1.10	1.88	1.10	29.97
Vapor Fraction	0.00	1.00	0.00	0.00	0.00	0.00	1.00		1.00	1.00	1.00	1.00	0.00	0.97	0.00	0.00		1.00	1.00	1.00	1.00	1.00	0.98	1.00
Volume Flow** (m^3/sec)	0.01	0.02	0	0	0.01	0.01	0.63		0.11	117.68	45.37	22.37	0.01	49.13	0.01	0.01		1.33	29.07	4.49	19.73	1.04	5.58	0.24
Mole Flow** (kmol/hr)	89.49	74.55	41.08	41.08	48.97	48.97	90.27		355.46	3595	3595	3237	1939	1939	1939	1939		3968	4177	4177	3255	921.35	921.35	921.35
Mass Flow (tonnes/day)	427.14	52.77	112.62	112.62	266.11	266.11	4.37		143.95	2439	2439	2242	838.24	838.23	838.23	838.23		2758	2903	2903	2189	714.26	714.26	714.26
H2O	0	0	0	0	0	0	0		0	233.43	233.43	0	838.24	838.23	838.23	838.23		0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0		55.65	0	0	0	0	0	0	0		0	0	0	0	0	0	0
H2	0	0	0	0	0	0	4.37		9.06	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0		11.88	352.24	352.24	0	0	0	0	0		0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0		0	104.25	104.25	527.27	0	0	0	0		638.58	672.19	672.19	0	672.19	672.19	672.19
N2	0	0	0	0	0	0	0		0	1715	1715	1715	0	0	0	0		2084	2194	2194	2189	4.91	4.91	4.91
CH4	0	14.94	0	0	0	0	0		4.29	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0		7.23	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C3	0	37.83	0	0	0	0	0		9.57	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C4	0	0	0	0	0	0	0		11.32	0	0	0	0	0	0	0		0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
TAR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0		0	0.08	0.08	0	0	0	0	0		0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
WAXES	170.22	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C5	15.49	0	0	0	0	0	0		0.15	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C6	16.65	0	0	0	0	0	0		0.17	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C7	17.54	0	0	0	0	0	0		0.08	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C8	17.99	0	112.62	112.62	0	0	0		0.08	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C9	18.10	0	0	0	0	0	0		0.08	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C10	18.12	0	0	0	0	0	0		0.08	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C11	17.62	0	0	0	0	0	0		0.08	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C12	17.25	0	0	0	0	0	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C13	16.70	0	0	0	0	0	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C14	16.16	0	0	0	0	0	0		0.03	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C15	15.57	0	0	0	0	0	0		0.03	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C16	14.93	0	0	0	266.11	266.11	0		0.03	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C17	14.75	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C18	14.05	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C19	13.35	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C20	12.64	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0		0.08	0	0	0	0	0	0	0		0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0		33.99	33.99	33.99	0	0	0	0	0		35.42	37.28	37.28	0.12	37.16	37.16	37.16
BIOMASS	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
SOOT	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
SLAG	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

E.2 Low Temperature Scenario

Table 46. Overall area stream data for LT scenario

HT Overall Plant	PL00BMAS	PL08BMAS	PL09CAIR	PL21SGAS	PL34SGAS	PL41CO2	PL43CO2	PL52FLUE	PL55SGAS	PL56FT	PL56SGAS	PL60ASH	PL65FGAS	PL71GASO	PL81DIES	PL81STM	PL81WAT	PL83SUL	PL84STM	PL88FLUE	PL90HYD	PL90OX	PL90STM	PL92WAT	PL98STM	PL89AIR	PL92NTGN	PL99AIR
Temperature (C)	25	90	100	870	62	50	243	200	32	35	32	0	50	50	50	200	40	50	120	344	60	149	300	120	204	32	16	30
Pressure (bar)	1.01	1.01	1.01	27.55	22.89	3.45	28.00	1.00	22.89	22.89	22.89	1.00	22.89	22.89	22.89	1.98	26.86	2.07	1.98	1.00	1.01	22.00	25.00	1.98	22.00	1.01	1.10	1.01
Vapor Fraction	0.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00	0.00	1.00	0.00	0.00	1.00	0.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00
Volume Flow** (m^3/sec)	0.43	0.14	16.89	4.76	2.40	2.84	0.07	22.48	1.50	0.01	0.12	0	0.02	0	0	50.53	0.02	0	42.34	40.84	0.53	0.32	1.15	4.70	0.01	23.17	15.73	17.42
Mole Flow** (kmol/hr)	1542	513.97	1985	4939	7066	1334	170.42	2053	4869	69.40	394.75	0	57.67	31.78	37.88	9251	2804	5.56	9251	2863	69.96	722.81	2313	1028	814.33	3328	2594	2522
Mass Flow (tonnes/day)	2667	2222	1375	2930	2706	1380	180.00	1471	2071	330.42	167.90	118.88	40.83	87.12	205.86	4000	1388	3.10	4000	1955	3.38	561.66	1000	444.44	352.09	2313	1744	1746
H2O	666.67	222.22	0	413.42	30.18	20.50	0	27.16	0	0.09	0	0	0	0	0	0	1060	2.40	0	189.23	0	0	1000	444.44	352.09	0	0	0
CO	0	0	0	797.86	1575	0	0	0	795.10	0	64.47	0	0	0	0	0	15.27	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	47.75	168.38	0	0	0	120.71	0	9.79	0	0	0	0	0	0.07	0	0	0	3.38	0	0	0	0	0	0	0
CO2	0	0	0	1427	167.80	1359	180.00	239.52	559.11	0	45.33	0	0	0	0	0	263.09	0	0	318.35	0	0	0	0	0	0	0	0
O2	0	0	320.19	0	0	0	0	148.99	0	0	0	0	0	0	0	0	0	0	0	80.99	0	528.66	0	0	0	535.65	0	410.78
N2	0	0	1055	0	0	0	0	1055	0	0	0	0	0	0	0	0	0	0	0	1336	0	0	0	0	0	1748	1744	1336
AR	0	0	0	33.00	411.50	0	0	0	380.61	0	30.86	0	0	0	0	0	0.69	0	0	30.86	0	33.00	0	0	0	29.71	0.10	0
CH4	0	0	0	103.81	149.06	0	0	0	52.01	0	4.22	0	11.56	0	0	0	4.58	0	0	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	21.82	28.49	0	0	0	13.55	0	1.10	0	0	0	0	0	5.75	0	0	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	44.96	36.05	0	0	0	0.36	0	0.03	0	0	0	0	0	8.11	0	0	0	0	0	0	0	0	0	0	0
C6H6	0	0	0	4.10	39.50	0	0	0	36.54	0	2.96	0	0	0	0	0	0.98	0	0	0	0	0	0	0	0	0	0	0
C3	0	0	0	0	54.18	0	0	0	58.62	0	4.75	0	29.27	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C4	0	0	0	0	34.31	0	0	0	41.73	0	3.38	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	4.51	0.13	0	0	0	0	0	0	0	0	0	0	0	1.87	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	19.09	1.52	0	0	0	0.28	0	0.02	0	0	0	0	0	17.36	0	0	0	0	0	0	0	0	0	0	0
TAR	0	0	0	10.57	0	0	0	0	0	0	0	0	0	0	0	0	10.57	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4000	0	0	4000	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0.30	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.06	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	131.46	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	2.21	0	0	0	2.21	11.92	0.18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C6	0	0	0	0	2.37	0	0	0	2.37	12.82	0.19	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C7	0	0	0	0	1.11	0	0	0	1.11	13.53	0.09	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C8	0	0	0	0	1.14	0	0	0	1.14	13.88	0.09	0	0	87.12	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C9	0	0	0	0	1.08	0	0	0	1.15	13.95	0.09	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C10	0	0	0	0	1.13	0	0	0	1.15	13.98	0.09	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C11	0	0	0	0	0.50	0	0	0	0.54	13.86	0.04	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12	0	0	0	0	0.44	0	0	0	0.53	13.54	0.04	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C13	0	0	0	0	0.34	0	0	0	0.51	13.10	0.04	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C14	0	0	0	0	0.20	0	0	0	0.49	12.57	0.04	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C15	0	0	0	0	0	0	0	0	0.46	11.93	0.04	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C16	0	0	0	0	0.04	0	0	0	0.45	11.48	0.04	0	0	0	205.86	0	0	0	0	0	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0	0	11.39	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	10.85	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	10.31	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	9.76	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	2000	2000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	1.01	0	0	0	0.01	0	0	0	118.88	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0.81	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char

Table 47. Preprocessing area stream data for LT scenario

HT A100	CH00BMAS	CH02BMAS	CH03BMAS	CH90BMAS	DR05BMAS	.DR81STM	.DR84STM	.DR92WAT	GR06BMAS	GR08BMAS	GR90BMAS
Temperature (C)	25	25	25	0	90	200	120	120	90	90	0
Pressure (bar)	1.01	1.01	1.01	0.00	1.01	1.98	1.98	1.98	1.01	1.01	0.00
Vapor Fraction	0.00	0.00	0.00	0.00	0.00	1.00	1.00	1.00	0.00	0.00	0.00
Volume Flow** (m^3/sec)	0.43	0.43	0.43	0	0.14	50.53	42.34	4.70	0.14	0.14	0
Mole Flow** (kmol/hr)	1542	1542	1542	0	513.97	9251	9251	1028	513.97	513.97	0
Mass Flow (tonnes/day)	2667	3181	2667	514.02	2222	4000	4000	444.44	3019	2222	796.81
H2O	666.67	666.67	666.67	0	222.22	0	0	444.44	222.22	222.22	0
CO	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0	0	0
C6H6	0	0	0	0	0	0	0	0	0	0	0
C3	0	0	0	0	0	0	0	0	0	0	0
C4	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0
TAR	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	4000	4000	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0	0	0	0
C6	0	0	0	0	0	0	0	0	0	0	0
C7	0	0	0	0	0	0	0	0	0	0	0
C8	0	0	0	0	0	0	0	0	0	0	0
C9	0	0	0	0	0	0	0	0	0	0	0
C10	0	0	0	0	0	0	0	0	0	0	0
C11	0	0	0	0	0	0	0	0	0	0	0
C12	0	0	0	0	0	0	0	0	0	0	0
C13	0	0	0	0	0	0	0	0	0	0	0
C14	0	0	0	0	0	0	0	0	0	0	0
C15	0	0	0	0	0	0	0	0	0	0	0
C16	0	0	0	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	2000	2514	2000	514.02	2000	0	0	0	2797	2000	796.81
ASH	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char

Table 48. Gasification area stream data for LT scenario

HT A200	CB52FLUE	CB61ASH	CB62ASH	CY14SGAS	CY15CHAR	CY16CHAR	GS05SGAS	GS07SGAS	GS09CAIR	GS11BMAS	GS19CHAR	GS20SGAS	GS43CO2	GS46CO2	GS52FLUE	GS81STM	GS84STM	GS90OX	GS91STM
Temperature (C)	1200	0	0	870	0	0	871	870	100	96	0	870	243	100	200	200	120	149	204
Pressure (bar)	1.00	1.00	1.00	27.57	27.57	27.55	28.00	27.58	1.01	1.01	27.55	27.55	28.00	27.58	1.00	1.98	1.98	22.00	22.00
Vapor Fraction	1.00	0.00	0.00	1.00	0.00	0.00	1.00	1.00	1.00	0.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00
Volume Flow** (m^3/sec)	69.99	0	0	2.38	0	0	4.69	4.76	16.89	0	0	4.76	0.07	0.05	22.48	50.53	42.34	0.32	0.01
Mole Flow** (kmol/hr)	2053	0	0	2470	0	0	4931	4939	1985	513.97	0	4939	170.42	161.90	2053	9251	9251	722.81	814.33
Mass Flow (tonnes/day)	1471	17.82	118.88	1477	95.38	12.09	3136	3145	1375	2222	214.95	2930	180.00	171.00	1471	4000	4000	561.66	352.09
H2O	27.16	0	0	206.71	0	0	413.42	413.42	0	222.22	0	413.42	0	0	27.16	0	0	0	352.09
CO	0	0	0	398.93	0	0	797.86	797.86	0	0	0	797.86	0	0	0	0	0	0	0
H2	0	0	0	23.87	0	0	47.75	47.75	0	0	0	47.75	0	0	0	0	0	0	0
CO2	239.52	0	0	713.66	0	0	1418	1427	0	0	0	1427	180.00	171.00	239.52	0	0	0	0
O2	148.99	0	0	0	0	0	0	0	320.19	0	0	0	0	0	148.99	0	0	528.66	0
N2	1055	0	0	0	0	0	0	0	1055	0	0	0	0	0	1055	0	0	0	0
AR	0	0	0	16.50	0	0	33.00	33.00	0	0	0	33.00	0	0	0	0	0	33.00	0
CH4	0	0	0	51.90	0	0	103.81	103.81	0	0	0	103.81	0	0	0	0	0	0	0
C2H6	0	0	0	10.91	0	0	21.82	21.82	0	0	0	21.82	0	0	0	0	0	0	0
C2H4	0	0	0	22.48	0	0	44.96	44.96	0	0	0	44.96	0	0	0	0	0	0	0
C6H6	0	0	0	2.05	0	0	4.10	4.10	0	0	0	4.10	0	0	0	0	0	0	0
C3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	2.25	0	0	4.51	4.51	0	0	0	4.51	0	0	0	0	0	0	0
NH3	0	0	0	9.54	0	0	19.09	19.09	0	0	0	19.09	0	0	0	0	0	0	0
TAR	0	0	0	5.29	0	0	10.57	10.57	0	0	0	10.57	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4000	4000	0	0
SO2	0.30	0	0	0	0	0	0	0	0	0	0	0	0	0	0.30	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C6	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C7	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C8	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C9	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C10	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C13	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C14	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C15	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C16	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C17	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	2000	0	0	0	0	0	0	0	0	0
ASH	0.01	17.82	118.88	7.19	52.75	6.69	119.90	119.90	0	0	118.89	1.01	0	0	0.01	0	0	0	0
CHAR	0	0	0	5.81	42.63	5.40	96.87	96.87	0	0	96.06	0.81	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char

Table 49. Syngas cleaning area stream data for LT scenario

HT A300	CL01H2O	CL02H2O	CL21SGAS	CL25ACG	CL25SGAS	CL25WAT	CL29SGAS	CL40CO2	CL40SGAS	CL41CO2	CL43CO2	CL49SGAS	CL80H2O	CL80WAT	CL81WAT	CL83SUL	CL83WAT	CL84WAT	CL92AIR	CL95AIR
Temperature (C)	194	150	870	50	194	0	40	50	62	50	243	32	30	30	40	50	98	40	50	50
Pressure (bar)	27.55	27.55	27.55	3.45	27.55	27.55	26.86	3.45	22.89	3.45	28.00	22.89	27.57	27.57	26.86	2.07	27.55	27.55	2.07	2.07
Vapor Fraction	0.00	0.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	1.00
Volume Flow** (m^3/sec)	0.08	0.08	4.76	3.28	2.03	0	0.98	3.21	2.40	2.84	0.07	1.50	0	0.01	0.02	0	0.07	0.07	0	0.10
Mole Flow** (kmol/hr)	13877	13877	4939	1538	5286	0	3638	1505	7066	1334	170.42	4869	346.93	1156	2804	5.56	13877	13877	0.45	27.74
Mass Flow (tonnes/day)	6000	6000	2930	1586	3078	1.82	2190	1560	2706	1380	180.00	2071	150.00	500.00	1388	3.10	6000	6000	0.35	22.89
H2O	6000	6000	413.42	22.78	563.42	0	3.78	20.50	30.18	20.50	0	0	150.00	500.00	1060	2.40	6000	6000	0	0.27
CO	0	0	797.86	3.07	797.86	0	782.59	0	1575	0	0	795.10	0	0	15.27	0	0	0	0	3.07
H2	0	0	47.75	0	47.75	0	47.68	0	168.38	0	0	120.71	0	0	0.07	0	0	0	0	0
CO2	0	0	1427	1539	1427	0	1164	1539	167.80	1359	180.00	559.11	0	0	263.09	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.35	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AR	0	0	33.00	1.42	33.00	0	32.31	0	411.50	0	0	380.61	0	0	0.69	0	0	0	0	1.42
CH4	0	0	103.81	2.18	103.81	0	99.23	0	149.06	0	0	52.01	0	0	4.58	0	0	0	0	2.18
C2H6	0	0	21.82	1.11	21.82	0	16.07	0	28.49	0	0	13.55	0	0	5.75	0	0	0	0	1.11
C2H4	0	0	44.96	1.14	44.96	0	36.85	0	36.05	0	0	0.36	0	0	8.11	0	0	0	0	1.14
C6H6	0	0	4.10	0.15	4.10	0	3.11	0	39.50	0	0	36.54	0	0	0.98	0	0	0	0	0.15
C3	0	0	0	4.29	0	0	0	0	54.18	0	0	58.62	0	0	0	0	0	0	0	4.29
C4	0	0	0	6.73	0	0	0	0	34.31	0	0	41.73	0	0	0	0	0	0	0	6.73
H2S	0	0	4.51	2.48	4.51	0	2.64	0	0.13	0	0	0	0	0	1.87	0	0	0	0	1.73
NH3	0	0	19.09	0.42	19.09	0	1.73	0	1.52	0	0	0.28	0	0	17.36	0	0	0	0	0.42
TAR	0	0	10.57	0	10.57	0	0	0	0	0	0	0	0	0	10.57	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0	2.21	0	0	2.21	0	0	0	0	0	0	0	0
C6	0	0	0	0	0	0	0	0	2.37	0	0	2.37	0	0	0	0	0	0	0	0
C7	0	0	0	0	0	0	0	0	1.11	0	0	1.11	0	0	0	0	0	0	0	0
C8	0	0	0	0	0	0	0	0	1.14	0	0	1.14	0	0	0	0	0	0	0	0
C9	0	0	0	0.07	0	0	0	0	1.08	0	0	1.15	0	0	0	0	0	0	0	0.07
C10	0	0	0	0.02	0	0	0	0	1.13	0	0	1.15	0	0	0	0	0	0	0	0.02
C11	0	0	0	0.03	0	0	0	0	0.50	0	0	0.54	0	0	0	0	0	0	0	0.03
C12	0	0	0	0.07	0	0	0	0	0.44	0	0	0.53	0	0	0	0	0	0	0	0.07
C13	0	0	0	0.10	0	0	0	0	0.34	0	0	0.51	0	0	0	0	0	0	0	0.10
C14	0	0	0	0.09	0	0	0	0	0.20	0	0	0.49	0	0	0	0	0	0	0	0.09
C15	0	0	0	0	0	0	0	0	0	0	0	0.46	0	0	0	0	0	0	0	0
C16	0	0	0	0.02	0	0	0	0	0.04	0	0	0.45	0	0	0	0	0	0	0	0.02
C17	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	1.01	0	0	1.01	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0.81	0	0	0.81	0	0	0	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char

Table 50. Acid gas removal and sulfur recovery areas stream data for LT scenario

HT A300AGR	AG02MEAR	AG03MEAR	AG04MEAR	AG06MEAR	AG08MEAL	AG09MEAL	AG10MEAL	AG11MEAL	AG23ACG	AG24ACG	AG25ACG	AG26SGAS	AG32SGAS	AG33SGAS	AG49SGAS	A300SUL	SU25ACG	SU26SUL	SU40CO2	SU83SUL	SU92AIR	SU95AIR
Temperature (C)	50	62	50	93	50	123	80	80	93	50	50	40	50	62	32		50	50	50	50	50	50
Pressure (bar)	22.89	22.89	22.89	3.45	3.45	3.45	3.45	26.00	3.45	3.45	3.45	26.86	22.89	22.89	22.89		3.45	3.45	3.45	2.07	2.07	2.07
Vapor Fraction	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.88	1.00	1.00	0.93	1.00	1.00		1.00	0.82	1.00	0.00	1.00	1.00
Volume Flow** (m^3/sec)	0.31	0	0.31	0.33	0	0.31	0.30	0.30	4.24	3.28	3.28	0.98	2.30	2.40	1.50		3.28	0.06	3.21	0	0	0.10
Mole Flow** (kmol/hr)	54067	449.42	54516	54516	207.83	52978	52978	52978	1746	1746	1538	3638	7515	7066	4869		1538	33.30	1505	5.56	0.45	27.74
Mass Flow (tonnes/day)	28640	208.71	28848	28848	101.05	27263	27263	27263	1687	1687	1586	2190	2915	2706	2071		1586	25.65	1560	3.10	0.35	22.89
H2O	20923	181.16	21104	21104	82.74	21081	21081	21081	105.52	105.52	22.78	3.78	211.34	30.18	0		22.78	2.28	20.50	2.40	0	0.27
CO	0	3.07	3.07	3.07	0	0	0	0	3.07	3.07	3.07	782.59	1578	1575	795.10		3.07	3.07	0	0	0	3.07
H2	0	0	0	0	0	0	0	0	0	0	0	47.68	168.38	168.38	120.71		0	0	0	0	0	0
CO2	1551	4.53	1556	1556	16.15	16.15	16.15	16.15	1556	1556	1539	1164	172.33	167.80	559.11		1539	0	1539	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0.35	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
AR	0	1.42	1.42	1.42	0	0	0	0	1.42	1.42	1.42	32.31	412.92	411.50	380.61		1.42	1.42	0	0	0	1.42
CH4	0	2.19	2.19	2.19	0.01	0.01	0.01	0.01	2.19	2.19	2.18	99.23	151.24	149.06	52.01		2.18	2.18	0	0	0	2.18
C2H6	0	1.13	1.13	1.13	0.02	0.02	0.02	0.02	1.13	1.13	1.11	16.07	29.62	28.49	13.55		1.11	1.11	0	0	0	1.11
C2H4	0	1.16	1.16	1.16	0.01	0.01	0.01	0.01	1.16	1.16	1.14	36.85	37.21	36.05	0.36		1.14	1.14	0	0	0	1.14
C6H6	0	0.15	0.15	0.15	0	0	0	0	0.15	0.15	0.15	3.11	39.65	39.50	36.54		0.15	0.15	0	0	0	0.15
C3	0	4.44	4.44	4.44	0.15	0.15	0.15	0.15	4.44	4.44	4.29	0	58.62	54.18	58.62		4.29	4.29	0	0	0	4.29
C4	0	7.42	7.42	7.42	0.69	0.69	0.69	0.69	7.41	7.41	6.73	0	41.73	34.31	41.73		6.73	6.73	0	0	0	6.73
H2S	2.51	0.01	2.52	2.52	0.04	0.04	0.04	0.04	2.52	2.52	2.48	2.64	0.13	0.13	0		2.48	2.48	0	0	0	1.73
NH3	0	0.49	0.49	0.49	0.07	0.07	0.07	0.07	0.49	0.49	0.42	1.73	2.01	1.52	0.28		0.42	0.42	0	0	0	0.42
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
MEA	6163	0	6163	6163	0	6163	6163	6163	0	0	0	0	0	0	0		0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
C5	0	0	0	0	0	0	0	0	0	0	0	0	2.21	2.21	2.21		0	0	0	0	0	0
C6	0	0	0	0	0	0	0	0	0	0	0	0	2.37	2.37	2.37		0	0	0	0	0	0
C7	0	0	0	0	0	0	0	0	0	0	0	0	1.11	1.11	1.11		0	0	0	0	0	0
C8	0	0	0	0	0	0	0	0	0	0	0	0	1.14	1.14	1.14		0	0	0	0	0	0
C9	0	0.07	0.07	0.07	0	0	0	0	0.07	0.07	0.07	0	1.15	1.08	1.15		0.07	0.07	0	0	0	0.07
C10	0	0.02	0.02	0.02	0	0	0	0	0.02	0.02	0.02	0	1.15	1.13	1.15		0.02	0.02	0	0	0	0.02
C11	0	0.04	0.04	0.04	0	0	0	0	0.04	0.04	0.03	0	0.54	0.50	0.54		0.03	0.03	0	0	0	0.03
C12	0	0.08	0.08	0.08	0.02	0.02	0.02	0.02	0.08	0.08	0.07	0	0.53	0.44	0.53		0.07	0.07	0	0	0	0.07
C13	0	0.17	0.17	0.17	0.08	0.08	0.08	0.08	0.17	0.17	0.10	0	0.51	0.34	0.51		0.10	0.10	0	0	0	0.10
C14	0	0.29	0.29	0.29	0.20	0.20	0.20	0.20	0.29	0.29	0.09	0	0.49	0.20	0.49		0.09	0.09	0	0	0	0.09
C15	0	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0	0	0.46	0	0.46		0	0	0	0	0	0
C16	0	0.41	0.41	0.41	0.38	0.38	0.38	0.38	0.41	0.41	0.02	0	0.45	0.04	0.45		0.02	0.02	0	0	0	0.02
C17	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char

Table 51. Fuel synthesis area stream data for LT scenario

HT A400	FS34SGAS	FS35SGAS	FS36SGAS	FS37SGAS	FS40SGAS	FS41GAS	FS50SGAS	FS51FT	FS52FT	FS53SGAS	FS54FT	FS55SGAS	FS56FT	FS56SGAS	FS57SGAS	FS58SGAS	FS59SGAS	FS60WAT	FS90HYD
Temperature (C)	62	78	200	200	60	235	200	200	35	32	35	32	35	32	32	41	200	35	60
Pressure (bar)	22.89	26.50	25.12	25.12	1.01	25.00	24.96	22.89	22.89	22.89	22.89	22.89	22.89	22.89	22.89	25.30	24.96	22.89	1.01
Vapor Fraction	1.00	1.00	1.00	1.00	0.88	1.00	1.00	1.00	0.72	1.00	0.00	1.00	0.00	1.00	1.00	1.00	1.00	0.00	1.00
Volume Flow** (m^3/sec)	2.40	2.18	4.44	0.07	0.65	0.05	5.02	4.39	2.08	2.02	0.01	1.50	0.01	0.12	0.41	0.38	0.58	0.01	0.53
Mole Flow** (kmol/hr)	7066	7066	10153	166.97	97.01	97.01	11399	9268	9268	6579	70.21	4869	69.40	394.75	1316	1316	1316	2619	69.96
Mass Flow (tonnes/day)	2706	2706	3705	60.93	57.55	57.55	4261	4261	4261	2798	330.77	2071	330.42	167.90	559.66	559.66	559.66	1132	3.38
H2O	30.18	30.18	671.91	11.05	11.05	11.05	671.91	1133	1133	0	0.43	0	0.09	0	0	0	0	1132	0
CO	1575	1575	1576	25.92	25.92	25.92	1791	1074	1074	1074	0	795.10	0	64.47	214.89	214.89	214.89	0	0
H2	168.38	168.38	242.13	3.98	0.60	0.60	271.37	163.12	163.12	163.12	0	120.71	0	9.79	32.62	32.62	32.62	0	3.38
CO2	167.80	167.80	604.44	9.94	9.94	9.94	755.55	755.55	755.55	755.55	0	559.11	0	45.33	151.11	151.11	151.11	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AR	411.50	411.50	411.50	6.77	6.77	6.77	514.36	514.36	514.36	514.34	0	380.61	0	30.86	102.87	102.87	102.87	0	0
CH4	149.06	149.06	52.13	0.86	0.86	0.86	66.19	70.29	70.29	70.29	0	52.01	0	4.22	14.06	14.06	14.06	0	0
C2H6	28.49	28.49	7.72	0.13	0.13	0.13	11.38	18.30	18.30	18.30	0	13.55	0	1.10	3.66	3.66	3.66	0	0
C2H4	36.05	36.05	0.39	0.01	0.01	0.01	0.49	0.49	0.49	0.49	0	0.36	0	0.03	0.10	0.10	0.10	0	0
C6H6	39.50	39.50	39.50	0.65	0.65	0.65	49.38	49.38	49.38	49.38	0	36.54	0	2.96	9.88	9.88	9.88	0	0
C3	54.18	54.18	54.18	0.89	0.89	0.89	70.02	79.16	79.16	79.22	0	58.62	0	4.75	15.84	15.84	15.84	0	0
C4	34.31	34.31	34.31	0.56	0.56	0.56	45.59	56.43	56.43	56.40	0	41.73	0	3.38	11.28	11.28	11.28	0	0
H2S	0.13	0.13	0.01	0	0	0	0.01	0.01	0.01	0.01	0	0	0	0	0	0	0	0	0
NH3	1.52	1.52	0.30	0	0	0	0.38	0.38	0.38	0.38	0	0.28	0	0.02	0.08	0.08	0.08	0	0
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	131.46	131.46	0	131.46	0	131.46	0	0	0	0	0	0
C5	2.21	2.21	2.21	0.04	0.04	0.04	2.80	14.91	14.91	2.98	11.92	2.21	11.92	0.18	0.60	0.60	0.60	0	0
C6	2.37	2.37	2.37	0.04	0.04	0.04	3.01	16.02	16.02	3.20	12.82	2.37	12.82	0.19	0.64	0.64	0.64	0	0
C7	1.11	1.11	1.11	0.02	0.02	0.02	1.41	15.03	15.03	1.50	13.53	1.11	13.53	0.09	0.30	0.30	0.30	0	0
C8	1.14	1.14	1.14	0.02	0.02	0.02	1.45	15.42	15.42	1.54	13.88	1.14	13.88	0.09	0.31	0.31	0.31	0	0
C9	1.08	1.08	1.08	0.02	0.02	0.02	1.39	15.50	15.50	1.55	13.95	1.15	13.95	0.09	0.31	0.31	0.31	0	0
C10	1.13	1.13	1.13	0.02	0.02	0.02	1.44	15.54	15.54	1.55	13.98	1.15	13.98	0.09	0.31	0.31	0.31	0	0
C11	0.50	0.50	0.50	0.01	0.01	0.01	0.65	14.59	14.59	0.73	13.86	0.54	13.86	0.04	0.15	0.15	0.15	0	0
C12	0.44	0.44	0.44	0.01	0.01	0.01	0.59	14.26	14.26	0.71	13.54	0.53	13.54	0.04	0.14	0.14	0.14	0	0
C13	0.34	0.34	0.34	0.01	0.01	0.01	0.47	13.79	13.79	0.69	13.10	0.51	13.10	0.04	0.14	0.14	0.14	0	0
C14	0.20	0.20	0.20	0	0	0	0.34	13.23	13.23	0.66	12.57	0.49	12.57	0.04	0.13	0.13	0.13	0	0
C15	0	0	0	0	0	0	0.13	12.55	12.55	0.63	11.93	0.46	11.93	0.04	0.13	0.13	0.13	0	0
C16	0.04	0.04	0.04	0	0	0	0.16	12.08	12.08	0.60	11.48	0.45	11.48	0.04	0.12	0.12	0.12	0	0
C17	0	0	0	0	0	0	0	11.39	11.39	0	11.39	0	11.39	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	10.85	10.85	0	10.85	0	10.85	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	10.31	10.31	0	10.31	0	10.31	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	9.76	9.76	0	9.76	0	9.76	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char

Table 52. Syngas conditioning area stream data for LT scenario

HT A400COND	CD35SGAS	CD36SGAS	CD37SGAS	CD38SGAS	CD39SGAS	CD40SGAS	CD41SGAS	CD42SGAS	CD44SGAS	CD45SGAS	CD80STM	CD81STM	CD99CONT
Temperature (C)	78	150	150	870	870	300	300	474	363	200	300	870	150
Pressure (bar)	26.50	26.50	26.50	26.50	25.81	25.81	25.81	25.12	25.12	25.12	25.00	27.00	26.50
Vapor Fraction	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Volume Flow** (m^3/sec)	2.18	2.64	2.64	7.08	10.44	5.24	1.83	2.46	5.98	4.44	1.15	2.26	0
Mole Flow** (kmol/hr)	7066	7066	7063	7063	10153	10153	3554	3554	10153	10153	2313	2313	3.12
Mass Flow (tonnes/day)	2706	2706	2705	2705	3705	3705	1297	1297	3705	3705	1000	1000	1.34
H2O	30.18	30.18	30.18	30.18	850.65	850.65	297.73	118.99	671.91	671.91	1000	1000	0
CO	1575	1575	1575	1575	1854	1854	648.81	370.91	1576	1576	0	0	0
H2	168.38	168.38	168.38	168.38	222.13	222.13	77.75	97.75	242.13	242.13	0	0	0
CO2	167.80	167.80	167.80	167.80	167.80	167.80	58.73	495.37	604.44	604.44	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0
AR	411.50	411.50	411.50	411.50	411.50	411.50	144.02	144.02	411.50	411.50	0	0	0
CH4	149.06	149.06	149.06	149.06	52.13	52.13	18.25	18.25	52.13	52.13	0	0	0
C2H6	28.49	28.49	28.49	28.49	7.72	7.72	2.70	2.70	7.72	7.72	0	0	0
C2H4	36.05	36.05	36.05	36.05	0.39	0.39	0.14	0.14	0.39	0.39	0	0	0
C6H6	39.50	39.50	39.50	39.50	39.50	39.50	13.83	13.83	39.50	39.50	0	0	0
C3	54.18	54.18	54.18	54.18	54.18	54.18	18.96	18.96	54.18	54.18	0	0	0
C4	34.31	34.31	34.31	34.31	34.31	34.31	12.01	12.01	34.31	34.31	0	0	0
H2S	0.13	0.13	0.01	0.01	0.01	0.01	0	0	0.01	0.01	0	0	0.12
NH3	1.52	1.52	0.30	0.30	0.30	0.30	0.11	0.11	0.30	0.30	0	0	1.21
TAR	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0
WAXES	0	0	0	0	0	0	0	0	0	0	0	0	0
C5	2.21	2.21	2.21	2.21	2.21	2.21	0.77	0.77	2.21	2.21	0	0	0
C6	2.37	2.37	2.37	2.37	2.37	2.37	0.83	0.83	2.37	2.37	0	0	0
C7	1.11	1.11	1.11	1.11	1.11	1.11	0.39	0.39	1.11	1.11	0	0	0
C8	1.14	1.14	1.14	1.14	1.14	1.14	0.40	0.40	1.14	1.14	0	0	0
C9	1.08	1.08	1.08	1.08	1.08	1.08	0.38	0.38	1.08	1.08	0	0	0
C10	1.13	1.13	1.13	1.13	1.13	1.13	0.40	0.40	1.13	1.13	0	0	0
C11	0.50	0.50	0.50	0.50	0.50	0.50	0.18	0.18	0.50	0.50	0	0	0
C12	0.44	0.44	0.44	0.44	0.44	0.44	0.16	0.16	0.44	0.44	0	0	0
C13	0.34	0.34	0.34	0.34	0.34	0.34	0.12	0.12	0.34	0.34	0	0	0
C14	0.20	0.20	0.20	0.20	0.20	0.20	0.07	0.07	0.20	0.20	0	0	0
C15	0	0	0	0	0	0	0	0	0	0	0	0	0
C16	0.04	0.04	0.04	0.04	0.04	0.04	0.01	0.01	0.04	0.04	0	0	0
C17	0	0	0	0	0	0	0	0	0	0	0	0	0
C18	0	0	0	0	0	0	0	0	0	0	0	0	0
C19	0	0	0	0	0	0	0	0	0	0	0	0	0
C20	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char

Table 53. Hydroprocessing, power generation, and air separation areas stream data for LT scenario

HT A500	HY56FT	HY65FGAS	HY70GASO	HY71GASO	HY80DIES	HY81DIES	HY90HYD	A600	421	602	615	680	CW2	CWREC	HPW1	HPSTM	A700	AIR-1A	AIR-A	AIR-L	N2-OUT	O2LPC	O2-OUT	O2-OUT2
Temperature (C)	35	50	50	50	50	50	60		32	1161	344	30	85	23	170	393		-170	32	20	16	35	16	55
Pressure (bar)	22.89	22.89	22.89	22.89	22.89	22.89	1.01		22.89	1.00	1.00	1.01	7.91	0.02	7.91	173.38		6.20	1.01	6.30	1.10	23.08	1.10	27.00
Vapor Fraction	0.00	1.00	0.00	0.00	0.00	0.00	1.00		1.00	1.00	1.00	1.00	0.00	0.89	0.00	1.00		1.00	1.00	1.00	1.00	1.00	1.00	1.00
Volume Flow** (m^3/sec)	0.01	0.02	0	0	0	0	0.53		0.12	94.88	40.84	17.42	0.01	442.16	0	0.03		1.06	23.17	3.58	15.73	0.22	4.45	0.21
Mole Flow** (kmol/hr)	69.40	57.67	31.78	31.78	37.88	37.88	69.96		394.75	2863	2863	2522	1709	1709	1709	1709		3162	3328	3328	2594	734.20	734.20	734.20
Mass Flow (tonnes/day)	330.42	40.83	87.12	87.12	205.86	205.86	3.38		167.90	1955	1955	1746	740.10	740.10	740.10	740.10		2198	2313	2313	1744	569.18	569.18	569.18
H2O	0.09	0	0	0	0	0	0		0	189.23	189.23	0	740.10	740.10	740.10	740.10		0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0		64.47	0	0	0	0	0	0	0		0	0	0	0	0	0	0
H2	0	0	0	0	0	0	3.38		9.79	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0		45.33	318.35	318.35	0	0	0	0	0		0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0		0	80.99	80.99	410.78	0	0	0	0		508.87	535.65	535.65	0	535.65	535.65	535.65
N2	0	0	0	0	0	0	0		0	1336	1336	1336	0	0	0	0		1661	1748	1748	1744	3.91	3.91	3.91
AR	0	0	0	0	0	0	0		30.86	30.86	30.86	0	0	0	0	0		28.22	29.71	29.71	0.10	29.61	29.61	29.61
CH4	0	11.56	0	0	0	0	0		4.22	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0		1.10	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0		0.03	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C6H6	0	0	0	0	0	0	0		2.96	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C3	0	29.27	0	0	0	0	0		4.75	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C4	0	0	0	0	0	0	0		3.38	0	0	0	0	0	0	0		0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0		0.02	0	0	0	0	0	0	0		0	0	0	0	0	0	0
TAR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CARBON	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0		0	0.06	0.06	0	0	0	0	0		0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
WAXES	131.46	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C5	11.92	0	0	0	0	0	0		0.18	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C6	12.82	0	0	0	0	0	0		0.19	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C7	13.53	0	0	0	0	0	0		0.09	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C8	13.88	0	87.12	87.12	0	0	0		0.09	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C9	13.95	0	0	0	0	0	0		0.09	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C10	13.98	0	0	0	0	0	0		0.09	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C11	13.86	0	0	0	0	0	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C12	13.54	0	0	0	0	0	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C13	13.10	0	0	0	0	0	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C14	12.57	0	0	0	0	0	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C15	11.93	0	0	0	0	0	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C16	11.48	0	0	0	205.86	205.86	0		0.04	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C17	11.39	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C18	10.85	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C19	10.31	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C20	9.76	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CHAR	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0		0	0	0	0	0	0	0

**Volumetric and Mole flow values do not include biomass, ash, or char