3-24-2016

Conversion of Landfill Gas to Liquid Hydrocarbon Fuels: Design and Feasibility Study

Ryan Alexander Kent

Scholar Commons Citation

Kent, Ryan Alexander, "Conversion of Landfill Gas to Liquid Hydrocarbon Fuels: Design and Feasibility Study" (2016). *Graduate Theses and Dissertations.*
http://scholarcommons.usf.edu/etd/6102

Conversion of Landfill Gas to Liquid Hydrocarbon Fuels: Design and Feasibility Study

by

Ryan A. Kent

A thesis submitted in partial fulfillment
of the requirements for the degree of
Master of Science in Chemical Engineering
Department of Chemical and Biomedical Engineering
College of Engineering
University of South Florida

Co-Major Professor: John Kuhn, Ph.D.
Co-Major Professor: Babu Joseph, Ph.D.
Aydin Sunol, Ph.D.

Date of Approval:
March 24, 2016

Keywords: Tri-reforming, Fischer Tropsch Synthesis, Gas to Liquids, Landfill Gas (LFG) to Liquid Fuels, Landfill Gas Recovery, Recovering Energy from Waste Digestion, Waste to Energy

Dedication

I would like to thank all those who assisted in this process, those who fostered my feelings of everlasting enthusiasm and interest in everything around me, and most of all show my greatest appreciation to those who helped me become me. I would like to dedicate all of my work to the five people who enabled me make it out of college, not just one time, but a second time as well. I could compose a thesis on how these people have been there to push me forward. To my parents, Paul and Linda Kent, who supported everything I have ever done and have through one way or another helped me down the path to completing my dreams. To Dr. John Kuhn, who changed my life by taking a gamble on an inexperienced and overly enthusiastic college sophomore. My sister Sarah Kent, who was the person I could tell everything to, especially when I didn't want to. Yolanda Daza, who was my best friend, the person who cared about me, and the entity that grounded me when I needed it most.

Thank You

Acknowledgments

I would like to point out the help I received from a number of people while working on this project. I would like to thank Trash 2 Cash – Energy LLC and VentureWell (Formerly NCIIA) for their financial support into the research. Thanks to the Hinkley Center for Solid and Hazardous Waste Management under the Florida Department of Environmental Protection, and Florida Energy Systems Consortium (FESC), for conference funding and opportunities to present research and learn from the industry. I would like to acknowledge help from Yolanda Daza, and all of the USF Heterogeneous Catalysis Group, for their everyday assistance in my work.

It is necessary to mention all the amazing professors and mentors I had while at USF, from the strange to the devoted and inspiring. The people I had opportunities to meet and work with affirm that I made the best choice when I entered as a beardless freshman.

This project originated out of the University of South Florida's Chemical & Biomedical Engineering Senior design project and as such I would like to send out my greatest appreciation to Kirk Jaunich, Tyler Schimdt, and Zachary Kerbo for their assistance in the original designing of this facility. Many hours were spent plunging into details which laid the groundwork for future developments on this project.

Table of Contents

List of Tables

List of Figures

Abstract

This paper will discuss the conversion of gas produced from biomass into liquid fuel through the combination of naturally occurring processes, which occur in landfills and anaerobic digesters, and a gas-to-liquids (GTL) facility. Landfills and anaerobic digesters produce gases (LFG) that can be converted into syngas via a Tri-reforming process and then synthesized into man-made hydrocarbon mixtures using Fischer-Tropsch synthesis. Further processing allows for the separation into liquid hydrocarbon fuels such as diesel and gasoline, as well as other middle distillate fuels. Conversion of landfill gas into liquid fuels increases their energy density, ease of storage, and open market potential as a common "drop in" fuel. These steps not only allow for profitable avenues for landfill operators but potential methods to decrease greenhouse gas emissions. The objective of this paper is to present a preliminary design of an innovative facility which processes contaminated biogases and produces a valuable product. An economic analysis is performed to show feasibility for a facility under base case scenario. A sensitivity analysis is performed to show the effect of different cost scenarios on the breakeven price of fuel produced. Market scenarios are also presented in order to further analyze situations where certain product portions cannot be sold or facility downtime is increased. This facility is then compared to traditional mitigation options, such as flaring and electricity generation, to assess the effect each option has on cost, energy efficiency, and emissions reduction.

1 Introduction

1.1 Background for Research

An increase in the use of renewable energy and fuels has been occurring over the past few decades as energy demand rises along with the global temperature. The global energy demand is ever increasing with the world's rapidly growing population. According to International Energy Outlook 2013, the current global energy consumption will increase by about 56 percent in the next three decades, from 524 quadrillion Btu to about 820 quadrillion Btu [1]. As of 2014, the United States oil reserves was estimated to be around 39 billion barrels with the world oil reserves amounting to about 1656 billion barrel [2, 3]. In the same year, the US was consuming approximately 19.1 million barrels per day, or ~20% of global consumption, which amounted to 92 million barrels per day [4, 5]. At current rates and reserves, the United States would consume its own reserves in almost 6 years, and the world's reserves would be depleted in just over 50 years. Countries around the world have set forward a number of standards and policies in order to increase the use of renewable sources and limit the use of fossil fuels, thus decreasing their carbon emissions [6]. This includes a large push for the use of renewable fuels which reduce the impact of man on the environment [7]

Although moving the electric grid to renewable sources can fix many energy demand issues, it is additionally important that energy dense transportation fuels be renewably generated for the future of energy development. Limitations in renewable fuels are commonly seen in sources which are temporary, small-scale, or relatively expensive to traditional energy counterparts. Modern day renewable fuels will require diverse feedstocks and depend on a wide array of technologies to fill the gaps left by fossil fuels [8]. A major portion of global transportation services are operated using liquid hydrocarbon fuels, currently derived from fossil fuels, which are presently a plentiful, but nonrenewable, resource. A minimal impact switch could be achieved if the fuel used is not altered, but instead the method of producing that fuel is changed to renewable processes. Generation of diesel fuel and gasoline can occur by gas-to-liquid processes (GTL), which involve catalytic conversion of CH_4, CO_2, and H_2O into syngas (mix of H_2 and CO). This syngas can then be synthesized into artificially produced fuels. Natural gas, whose major component is CH_4 (methane), is a prime candidate for use in these processes. In the United States, the estimated amount of recoverable natural gas was 67 trillion cubic meters as of 2012. Use of advanced gas recovery techniques, such as hydraulic fracturing has increased the amount of viable gas locked inside the US [9]. While widely available, natural gas is a sequestered carbon source and its use contributes substantially towards an increase in greenhouse gas emissions. Due to the variable cost of natural gas, and large general expense of the catalysts used in these processes, these are still considered generally unfeasible projects [10]. In this case liquid fuels are produced by converting one fossil fuel into another while using a great deal of money and energy.

Since CH_4 is also a greenhouse gas which is commonly produced by humans, a number of issues can be solved simultaneously by using anthropogenic sources of methane for production of fuels. A quarter of this methane comes from municipal landfills, which act as large bioreactors. Using landfill-produced methane will cause the reduction of greenhouse gas emissions from the landfill, and the subsequent emissions from the burning of fossil fuels.

1.2 Biomass

Biomass is a composition of matter derived from living organisms or their byproducts. Different forms of biomass include agricultural crops, agricultural or bio-product residues, algae, and municipal or animal wastes [11]. Energy from biomass is the earliest form of energy harvesting, and originated with the combustion of wood, plant matter, and waste products in early human history. Although the first uses of biomass were simple, it still remains a major supplier of energy at 15-25% of the world total. Since the advent of the industrial revolution and discovery of a wide variety of fossil fuels, many industrialized countries have turned away from biomass use. Many less developed countries still use biomass as a primary source of energy; from cooking and heating uses, to waste processing. [12]

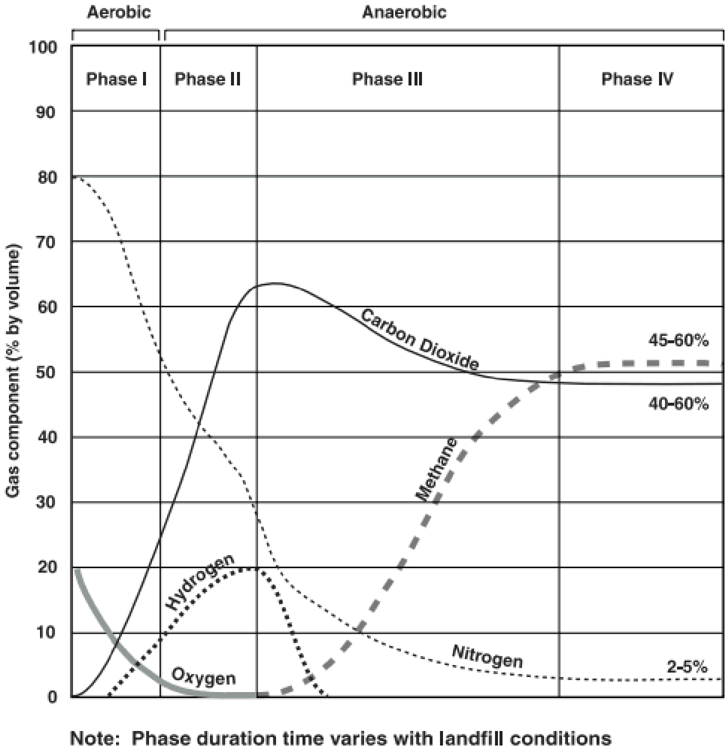
The fossil fuels consumed by the world originate from sequestered biomass and carbon sources. These sources, with the addition of heat and pressure, are converted into significantly more energy dense substances over time. With increasing interest in renewable fuels from developed nations, humans have sought process methods in order to decrease the time it takes to produce fuel from millions of years to immediate production. [13, 14]. Biomass that is converted into liquids for either transportation use or energy storage are termed biofuels. Early methods of ethanol production were the fermentation of simple sugars from crops. Use of food crops has led to debates about using food for the production of fuel. The largest problems surrounding food crops for biofuel is the immense tracts of land required to produce the needed amounts of fuel, and the substantial energy required to produce these crops. These limit the capacity that agriculture crops can contribute to reducing world fuels needs. In order to reduce competition between fuel requirements and food needs, advancements in biofuel technologies lead to the use of waste products which do not compete with current food sources [15-17]. Some of these advances have led to crops which are directly meant for the production of biofuels and bio-products. Initial developments in the field of biofuel production led to large-scale processing of cellulosic ethanol. These include many fast growing grasses, woody biomass tree farms, and bagasse [18]. Besides the use of fuel crops, other sources use inedible cellulosic portions of biomass or food crops which typically come from agricultural and forestry discards [19, 20]. Issues with using these types of waste, as with many biomass products, is their seasonal availability and lack of proximity to production and distribution sites.

In order to reduce the cost of using biomass and increase the efficiency of use, low-value lignocellulosic biomass (woody biomass) must be converted into useable portions. Currently, the difficulty of converting lignin into valuable materials by using biological methods has left this conversion financially unfeasible. This has led to advanced gasification methods which force production of anaerobic gases through thermochemical breakdown of organic chemicals. However, this process by which syngas is produced is energy intensive and varies widely with the composition of the feedstock [21, 22]. Biological processes streamline this conversion by producing decomposition gases with a wide variety of conditions, while maintaining a fairly consistent product composition. Maintenance of bioreactors at these conditions is cost intensive as they must be large, sealed from oxygen permeation, and be constantly tended [23]. Natural examples of bioreactors exist such as swamps, while manmade landfills exhibit all of the required prerequisites for a bioreactor.

1.3 Landfills

In the United States, landfills remain a primary method for the disposal of Municipal Solid Waste (MSW). As of 2013, the U.S. generated 254 million tons of MSW per year, discarding 167 million tons of this waste into landfills [24]. Approximately 38-53% of this waste is biodegradable, and is capable of being broken down by microorganisms to produce methane, carbon dioxide (CO_2) and water using natural processes such as acetogenesis and methanogenesis. [25] This landfill gas (LFG), released by natural anaerobic digestion of MSW, is approximately 54% CH_4

and 46% CO_2 on a dry basis [26]. Landfill microorganism cultures work symbiotically. When MSW is initially landfilled and the cell is closed, the waste undergoes aerobic digestion producing carbon dioxide until the oxygen within the waste and landfill cell is consumed. After ~1 year, enough oxygen is consumed that anaerobic digestion of waste begins, and bacteria begin to produce methane. A diagram showing this process is shown in Figure 1.



Note: Phase duration time varies with landfill conditions

Source: EPA 1997

Figure 1: Landfill Gas Phases as a Function of Time after Placement

ATSDR 2008. Chapter 2: Landfill Gas Basics. Figure 2-1, pp. 5-6. http://www.atsdr.cdc.gov/HAC/landfill/PDFs/Landfill_2001_ch2mod.pdf. Image is public domain through the EPA.

A landfill's gas flow rate typically changes depending on the composition of the waste (i.e. each landfill is unique), as well as the time of year and the amount of rain received. A significant amount of planning goes into renovation and creation of new landfills, adding to their appeal for coexisting projects [27]. The lifetime, or the time period in which gas is produced in viable amounts, typically ranges from 10 to 15 years. However, this can be altered with methods of enhanced gas recovery and changing the minimum viable flow rate [28].

Currently the Environmental Protection Agency (EPA) is keeping track of 636 operational LFG projects which generate 1978 MW of electricity and 305 MMSCFD of gas for other uses. There are 440 candidates' landfills which are planned to add 885 MW electricity and 490 MMSCFD onto current capacity. [29] Of the currently operating usage capacity the average US landfill generates approximately 1970 SCFM. (See Appendix A)

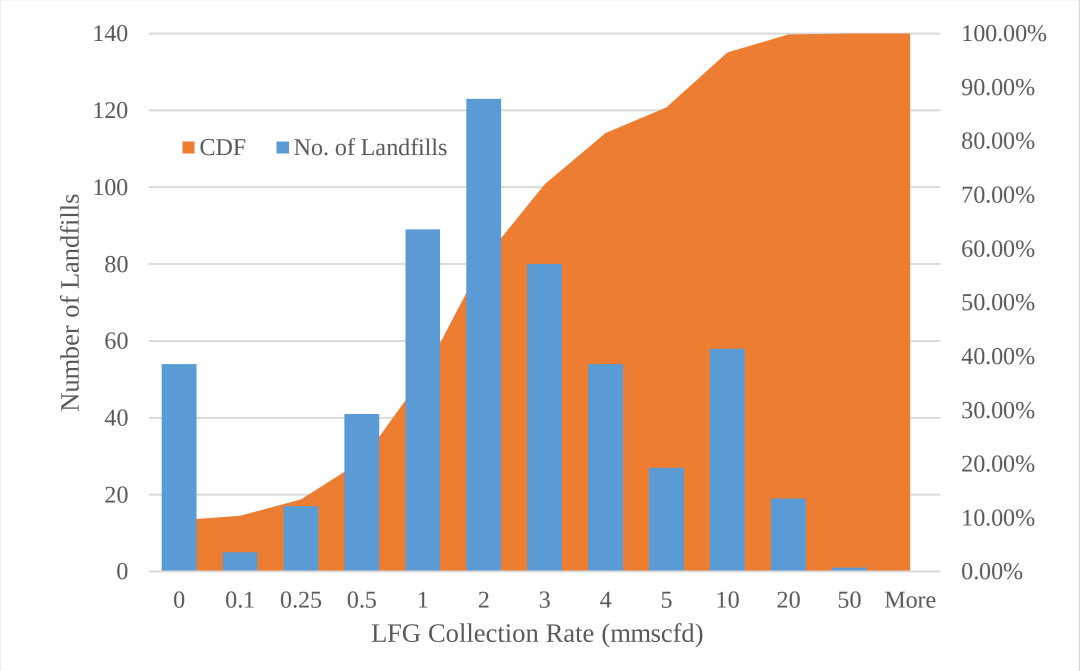


Figure 2: National/State LFG Gas Production Distribution

Figure 2, above, is the distribution of unique, operational landfills which report LFG collection rates. The distribution of LFG output is generated from EPA Landfill Methane Outreach Project data on for voluntary gas reporting.

1.3.1 Landfill Gas (LFG)

LFG can contain upwards of 50% methane by mass, having an energy content of 450-600 BTU/ft^3. Due to high energy content of landfill gases, efforts are made to capture and use it as a resource. Many methods exist to collect and utilize LFG as well as increase LFG evolution rate from the landfill. These include the use of landfill runoff, commonly called leachate, as a way of wetting the MSW and increasing gas production rate. The EPA currently regulates that all new landfills mitigate the emission of methane and other hazardous contaminants such as hydrogen sulfide. In addition to the two major gas components (CO_2 and CH_4), LFG contains siloxanes, halides, volatile organic compounds (VOCs), and a large variety of contaminants considered Non-Methane Organic Compounds (NMOCs), which vary depending on waste composition [30, 31]. In our study 2 model contaminants will be used. Hydrogen sulfide (H_2S) is detrimental to metal catalysts and is present in relatively high concentrations. Silica compounds called siloxanes, which will be used to model larger, non-reactive molecular contaminants are also present in LFG. Siloxanes are a group of manmade organic compounds which contain silicon, methyl, and oxide groups. There are 8 common siloxane compounds which have become a significant issue in modern landfills due to the increased use of silica compounds in household products [32]. In this study, octamethylcyclotetrasiloxane is used as the model siloxane due to its prevalence in LFG representing more than 50% of measured siloxanes [33]. Combustion of siloxanes produces silicon dioxide which forms an abrasive deposit layer on vital engine parts and machine parts. This buildup causes a reduction in heat conduction, part lubrication, and changes combustion chamber geometry. Silicon dioxide can also deactivate catalytic converters, leading to higher exhaust emissions. Acceptable siloxane concentrations range from 0.03-28 mg/m^3 in electrical turbines or gas engines depending on the manufacturer specifications [34]. Some techniques developed for siloxane removal include the use of activated carbon, silica gel, or alumina absorbents. Activated carbon has limited regeneration capabilities while silica and alumina absorbents can be regenerated for subsequent reuse by heating. Other absorption methods, which use liquid stripping agents, can be used to remove siloxanes. These include polyethylene glycol, dimethyl ethers, or water; although they are not very effective methods and incur significant operating costs [35]. Unlike siloxanes, H_2S is reactive and can be removed by a multitude of methods which include ionic liquids, solid scavengers, and catalytic reactors [36-38]. Selection of sulfur removal technologies is greatly dependent on volume of product to be processed and recovery of percentage of sulfur to be removed. For medium scale facilities with varying concentrations of sulfur, reacting a saturated gas stream with an iron solid scavenger has been deemed a solution. Iron solid scavengers are Iron oxide beds which turn H_2S into water removing the sulfur and creating iron pyrite (FeS_2) [39, 40]

Table 1: Reported Components of Landfill Gas

Major Components		**Trace Components**
Component	**Mass %**	
CH4	45-60	**Halides, toluene, acrylonitrile, benzene, dichloroethane, dichloroethylene, dichloromethane, carbonyl sulfide, ethyl benzene, hexane, methyl ethyl ketone, tetrachloroethylene, trichloroethylene, vinyl chloride, and xylenes.**
CO2	40-55	
N2	2-5	
O2	0.1-1	
NMOC	0.01-0.6	
H2S	0-1	

A common method for disposal of LFG is gas flaring. Other methods which are currently available include power generation and production of compressed or liquefied natural gas (CNG/LNG). Issues with current options for using LFG include having a relatively low $/BTU of the product, and its allowance for a significant margin of economic potential when generating common liquid transportation fuels. By utilizing technologies that capture methane from MSW landfills for extended uses rather than just mitigation of hazards, landfill operators can reduce or even remove the cost for compliance. Many local governments across the US are achieving energy, environmental, health, and economic benefits in addition to meeting emissions standards. LFG can be used in various forms for energy including electricity, boiler fuel, steam, alternate vehicle fuel, and pipeline quality gas.

1.4 Emissions and Regulations

Currently the EPA regulates the release of harmful gases such as H_2S and VOC's into the air from landfills; under 40 CFR part 60. Regulations are based on opening date of a landfill as they are "grandfathered in" to emissions regulations. Landfills that have accepted waste after November 1987 are subject to emission regulation under 40 CFR 60.33. Any landfill which generates NMOC's and a capacity of over 2.5 million cubic meters must mitigate their hazardous emissions under the Clean Air Act (40 CFR part 60 subpart WWW) [41], which is typically done through initiation of a flaring project. Combustion of dangerous compounds and gases is preferred, however it generates the increased levels of CO_2. The ability to use potential emissions as an offset for further greenhouse gas emission from fossil fuel holds promise, as 22.38 lbs of CO_2 are generated from burning 1 gallon of diesel fuel and 19.68 lbs from 1 gallon gasoline [42]. Carbon dioxide emissions are preferred over methane, as it has over 25 times the global warming potential as carbon dioxide on a weight basis [43]. Landfills represent a significant portion of US greenhouse gas emissions generating 103 million metric tons or 17.2% (equivalent tons CO_2) as of 2012 [44]. Since a 60% of CH4 emissions come from human activities, and its effect is considered short-lived at 12 years, a reduction in CH4 emissions is considered a swift and effective method of reducing greenhouse gas emissions [45]. An LFG energy plant can reduce methane emissions from a MSW landfill by 60 to 90 percent depending on the efficiency of capture [46]. Even installation of LFG mitigation technologies after landfill closure can reduce lifetime landfill emissions by 20% or more

[47]. The annual emission reductions of a typical 3 MW electricity generation project using LFG is about 34,700 metric tons of carbon equivalents per year - the environmental equivalent of CO2 emissions from about 296,000 barrels of oil [48]. The annual methane and carbon dioxide emission reductions of a typical direct-use LFG to energy project, assuming 1,000 scfm, is 135,750 metric tons of carbon equivalent per year; the environmental equivalent of the carbon emissions from 15.3 million gallons of gasoline consumed [49].

1.5 Current Landfill Gas Usage Methods

1.5.1 Flaring

Flaring is the most commonly used option for facilities which are small or produce a limited amount of LFG. Currently, flares are used in cases in which too little gas flows through the system to support an alternate use project. Even projects which use LFG for other purposes must maintain a flare for safety and emissions concerns [50]. This option is used to directly mitigate the release of harmful gases such as H_2S, CH_4, and other VOCs by combustion. Issues that arise when using a flaring system include the fact that it incurs cost with no benefit other than meeting emissions and safety regulations [51]. In landfills which produce significant quantities of LFG, this resource could be used for energy instead of destruction by flaring. This leads to a potentially valuable resource being wasted, when it could be profitable and also further mitigate CO_2 emissions by better use of the gas.

1.5.2 Direct Burn for Electricity or Heat

The use of LFG for electricity generation is well known [52]. Currently, 70% of facilities use LFG for this purpose [53]. The landfill gas is combusted in order to run reciprocating engines directly or turbines for electricity production. The process and machinery used varies greatly with a number of factors that include LFG flow rate and distance from residential or industrial areas, which can conclude whether heat generation is a profitable option. In some cases this method encompasses “mass burn”, where the MSW is burned directly and the ash is landfilled after incineration. This method is economically predictable and commonly used. However, in the cases of LFG use for generation, profits are generally not as high as expected and are usually only feasible for public entities due to other socioeconomic benefits [54]. Some of these benefits are included in mass burn facilities, where increased land value leads to incineration of MSW. Use of LFG for heat in industrial facilities reduces the required load of natural gas and can lead to significant savings by using LFG to power co-fired boilers [53].

1.5.3 LNG/CNG Generation

This method has grown in popularity for use in recent years. LFG is cleaned, purified of CO_2 and then either compressed or liquefied for use in commercial vehicles or for sale on the open market. This has become an interesting use for landfill gas, as natural gas is considered a cleaner burning fuel. The use of natural gas is attractive due to production of fuel on-site, with many landfill operators converting trucks to run on natural gas [54, 55]. Issues with this method include the high pressures which must be reached in order to separate CO_2 from CH_4. Compressor and utility cost, as well as transportation and storage, are major hurdles in price competition with a much cheaper source of natural gas from fossil deposits.

1.6 Fischer-Tropsch Synthesis (FTS)

The FTS reaction is well studied and involves the conversion of syngas (H_2 and CO) into long chain hydrocarbons. This is primarily due to its appeal in generation of long chain hydrocarbons through the reaction of simple gases. Research began in 1902 with the discovery of catalytic activity in transition metals, such as Co, causing CO hydrogenation into CH4 [56]. This was followed in 1925 by the processes namesake Franz Fischer and Hans Tropsch, who invented the process by which Co metal catalyst could be used to generate hydrocarbon liquids [57]. Fischer Tropsch (FT) synthesis is a process in which hydrocarbons are polymerized and joined stepwise on the surface of the catalyst, as shown in equation (1).

$$nCO + (2n+1)H_2 \rightarrow C_nH_{(2n+2)} + nH_2O \tag{1}$$

Research has been ongoing since the advent of FT synthesis, leading to an appropriation of research in GTL processes. Co and Fe are well known catalysts for FT synthesis and are extensively compared in a number of papers [58, 59]. When used for FT synthesis, Co catalysts have greater activity, selectivity to hydrocarbon production, and lower reaction temperature required for reaction onset. These attributes make Co a preferred catalyst for use in lower temperature GTL reactors [57]. Supports usually include Al_2O_3 or SiO_2, however a large number of factors exist based on the support and catalyst morphology. In many cases supports act as shaping agents to produce a larger portion of desired long chain hydrocarbon products [60-62]. Common drawbacks of using Co catalysts are their high H_2/CO reactant ratios, which are required in order to avoid deactivation of the metallic catalyst by coking. Co catalysts typically operate at a ratio of 2:1 H_2/CO or greater to drive the reaction, while Fe catalysts operate much lower at 0.6-1 H_2/CO due to affinity for the Water-Gas Shift reaction. In order to increase H_2/CO ratios in the feed gas, development of methods which can produce higher quantities of H_2, typically input as water steam. Other common practices include operating at elevated pressures in order to overcome equilibrium limitations of the reactions, as well as enhancement of other properties.

Common issues experienced when using FT synthesis catalysts include the deactivation of the metallic catalyst by contamination, or as a detrimental effect of reaction equilibrium conditions. Contamination of the catalyst by sulfur, heavy waxes and tars, or inorganic compounds can reduce active catalyst sites, or cause unfavorable generation of alcohols or other undesired products.

Effects of the reaction equilibrium are consistently a problem as carbon formation, excess steam buildup, or transportation limitation of products or reactive intermediates can also cause reactor fouling. Because the FT reaction is exothermic, low pressure steam can be generated to offset the energy requirements. These gas-phase reactors have large active volumes which can lead to heat buildup, and reactor hotspots which can lead to runaway reactions and undesirable kinetics. Small changes in temperature through FT reactors can cause relatively large changes in product composition. Recent interest in FT synthesis is due to use of bio-derived renewable fuels for the generation of liquid transportation fuels. The cost associated with generation of these fuels is generally dependent on the cost of the biomass feedstock. Feedstocks that are considered waste or a nuisance can be an easy way to reduce the cost associated with acquiring feedstock. Past research from other group members has looked at the use of a number of biomass options [63, 64].

Advances in FT catalysts are aimed heavily at generating more robust catalysts that can produce more desirable products. Work shown in Gardezi et al. has led to the creation of a Co eggshell catalyst, which reduces mass transfer issues and homongenizes the reactor temperature [65, 66]. The use of this catalyst reduces the formation of heavier tar products, while increasing the conversion of syngas into desirable products which are larger than CH_4.

1.7 Methane Tri-Reforming

Tri-reforming, or Methane Tri-Reforming (MTR) is the combination of the dry reforming of methane, steam reforming of methane, and the partial oxidation of methane (POM), which proceeds according to the following reactions:

Dry Reforming (CO_2 Reforming)

$$CH_4 + CO_2 \Leftrightarrow 2CO + 2H_2 \quad \Delta\mathrm{H} = 247.3\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{2}$$

Steam Reforming (SR)

$$CH_4 + H_2O \Leftrightarrow CO + 3H_2 \quad \Delta\mathrm{H} = 206.3\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{3}$$

Partial Oxidation of Methane (POM)

$$CH_4 + \frac{1}{2}O_2 \Leftrightarrow CO + 2H_2 \quad \Delta\mathrm{H} = -35.6\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{4}$$

Other reactions which occur, either as intermediaries or side reactions, are the WGS Reaction and combustion of methane. Other equations below deal with the formation of coke by decomposition of methane or by Boudouards equilibrium.

Water-Gas Shift

$$CO + H_2O \Leftrightarrow CO_2 + H_2 \quad \Delta\mathrm{H} = -41.09\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{5}$$

Combustion of Methane (POM)

$$CH_4 + 2O_2 \Leftrightarrow CO_2 + 2H_2O \quad \Delta\mathrm{H} = -880\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{6}$$

$$CH_4 + 1.5O_2 \Leftrightarrow CO + 2H_2O \quad \Delta\mathrm{H} = -520\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{7}$$

Thermal Decomposition of Methane

$$CH_4 \Leftrightarrow C_{(s)} + \ 2H_2 \quad \Delta\mathrm{H} = 74.9\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{8}$$

Boudouards Reaction

$$2CO \Leftrightarrow CO_2 + C_{(s)} \quad \Delta\mathrm{H} = -172.2\,\frac{\mathrm{kJ}}{\mathrm{mol}} \tag{9}$$

MTR has a number of benefits for the production of syngas. Reactions (2), (3), and (4), all produce syngas with a varying ratio of H_2/CO. The use of steam reforming is primarily to increase the H_2 available for the reactor by addition of reactant steam.[67] The balance between steam and dry reforming generates the required ratio for input into the FT reactor [68]. POM is used in order to generate heat in-situ that can be used to increase energy efficiency by offsetting required energy inputs. MTR uses 45.8% less energy and produces 92.8% less CO_2 compared to dry reforming. When compared to steam reforming, MTR uses 19.7% less energy and produces 67.5% less CO_2 [69-72]

Typically, MTR catalysts involve Ni or rare metal catalysts on a high surface area support that has high oxygen availability, referred to as oxygen storage capacity (OSC). Although the best activity in MTR, and other variations of reforming reactions, occurs on rare metal catalysts such as Pt, Rh, and Ru, the use of Ni is purely an economical choice [73]. Since coking and deactivation by contamination are a likely occurrence, a less expensive catalyst is chosen. Use of CeO_2 is due to its high OSC and is typically used with Ni [74-78]. ZrO_2 is used in conjunction with CeO_2 as it has be shown to increase dispersion of Ni metal particles on the surface, reduce support and active metal sintering, and stability at high temperatures [79-84]. The $(Ce, Zr)O_2$ solid solution has been effective in reducing sintering, coke formation, and increased H_2/CO product ratios.[85-87]

Although generally explained as catalyst resistance to coke formation, many explanations of MTR involve carbon buildup as an important factor in reaction equilibrium. Carbon restructuring of active Ni species is important to the cleavage process of removing oxygen from support species [88-90]. However, excess buildup of carbon species causes surface coverage and deactivation of the catalyst, as well as carbon whisker formation [68, 91, 92]. The effect of POM mechanisms is vital to oxygen exchange from the catalyst surface and from the support [79, 93]. Magnesium is added as a surface co-catalyst as it reduces carbon buildup on the catalyst surface as well as reducing the degree of sintering by the Ni surface catalyst.

Previous work from group members has focused on using a specific 60/40 (Ce, Zr)O_3 support with an 8% by mass Ni and 8% Mg. This work looked into optimum run conditions which are used throughout this study and design [94].

1.8 Proposed Method: LFG to Liquids (LFGTL)

Liquid fuels from LFG are energy dense, can be easily stored, and have a place in the existing marketplace as a drop in replacement for fossil fuels. The process of generating liquid fuels from waste gases such as LFG could be very beneficial in locations or situations where access to traditional fossil fuels are not available. Typically these countries import the majority of their energy in the form of fossil fuels, further adding to global emissions.

A process that would turn waste gas generated by landfills into a high energy content liquid transportation fuel would be advantageous. Converting LFG into diesel fuel would increase its $/BTU and increase its overall energy density and ability to be stored. Diesel fuel is commonly used by the transportation industry and for commercial vehicles [95]. Advantages of diesel production over other transportation fuels is its suitability for generation by Fischer Tropsch synthesis (FTS) and the fact that its composition is straight chain hydrocarbons, which FTS produces prominently [96]. Using diesel fuel over LNG/CNG is primarily due to diesel's ability to be a "drop in fuel", where infrastructure for its sale, distribution, and use are already in wide practice. This reduces extra and hidden costs associated with fuel production. This can be seen in LNG/CNG, which have a comparable cost to traditional fuels, yet are still only used in discrete areas [97].

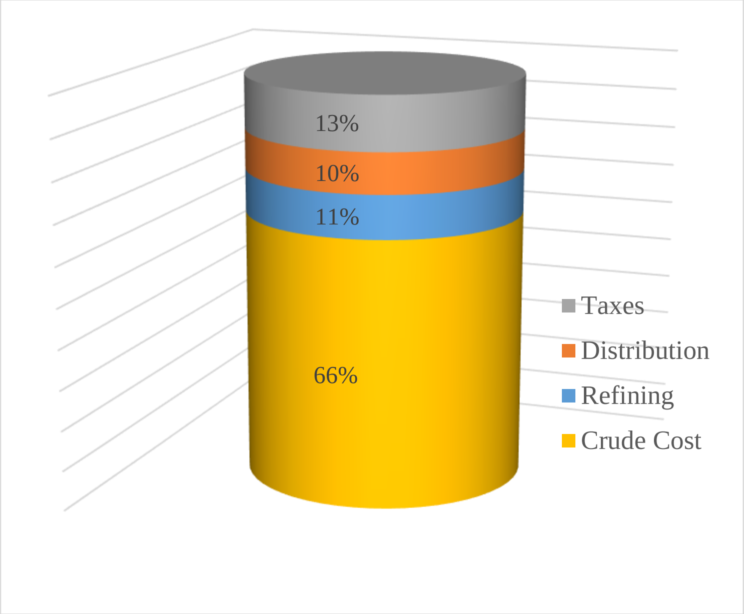


Figure 3: Cost Breakdown for a Gallon of Gasoline

Using fuels which are generated outside of the refinery system incur different costs than those that are. Figure 3 above, is calculated from EIA pricing data in Appendix H, showing the breakdown in the cost for a typical gallon of gasoline. The production of diesel for in-house uses can eliminate refining costs and can still compete, as 10% or more of costs can be reduced.

The composition of diesel fuel required for sale on open market is determined by a host of organizations. These include the EPA, which regulates additives and emissions [50]. Methods of testing performance and specifications are maintained by ASTM standards, Section 5, for testing color, density, viscosity, flash point, and a large number of other properties. FTS produces mostly saturated hydrocarbons, at 98% of total mass, while most diesel produced from fossil fuels can contain 25-75% olefins and aromatics and have less than 50% saturates [98]. This leads to FT fuels having a larger cetane index in comparison.

2 Modeling and Design

2.1 Parameters

The following parameters, listed in Table 2 and shown visually in Figure 4, were used as a model for landfill gas composition in the designed facility. It is important to note that this value was calculated by determining the dry gas composition, then saturating it with water at atmospheric pressure and 40 C. This was meant to simulate conditions in a landfill, where biologic processes heat the gases and leachate processing maintains dew point.

Table 2: Model Composition Used

Component	% Composition
CH_4	55
CO_2	42
N_2	1.7
O_2	~0.4
H_2S	0.07 (700 ppm)
Siloxanes	0.00009 (0.9 ppm)

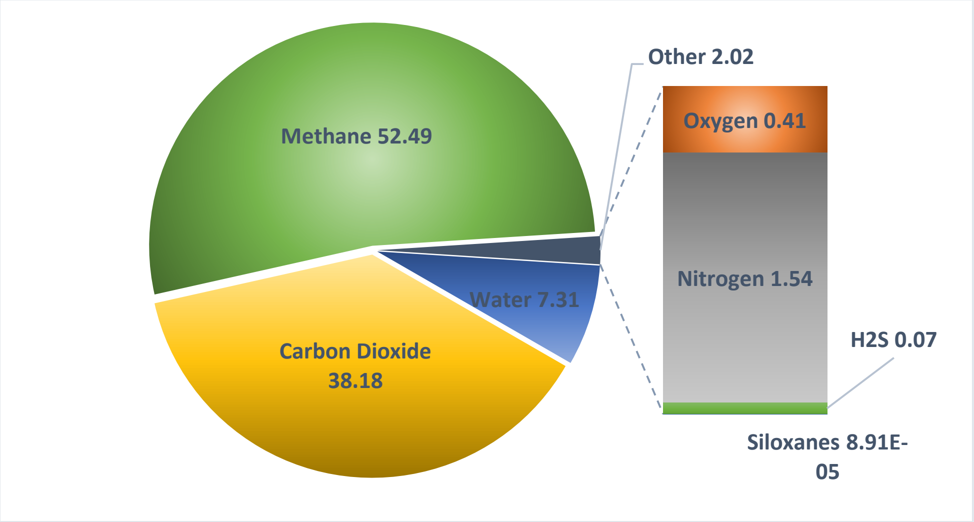


Figure 4: Model LFG used in Paper (% Composition of Saturated Gas)

After processing of the gas through the reactor, the limiting factor in production of fuel, the facility was designed with the following order of priorities. 1) The diesel fuel produced was of sufficient quality, 2) reduction of heating and cooling needs by heat integration of the facility, 3) elimination of heating costs by burning fractions of remaining fuel. An important heuristic that was used throughout the design of the facility was to decrease the outside energy required to run this facility. Overall, it was decided that barring additional plant installations, such as adding solar heating or photovoltaics, the majority of energy could be derived by burning undesired products in the furnace.

2.2 Assumptions

The basis flow rate of LFG for this paper will be 2500 SCFM, with constant composition and no change in contaminant concentration. This basis will be considered the minimum viable flow rate, or the flow rate which is maintained over the 15 year life span of the facility. This flow rate represents only a small portion of MSW facilities. However, it was chosen as economy of scale for FTS systems is exponentially more viable at larger scales. Pretreatment systems were assumed to be 100% effective until saturation of the media. It is also assumed that there is no trend or alteration in flow rate due to the use of a minimum viable flow rate. Although contamination concentration does not change, it is assumed that all contaminants can or will cause damage to machinery and equipment and therefore must be removed during pretreatment to increase the longevity of the facility and equipment. To correct for this, compressors and pumps were priced with added spares and an increase in maintenance costs was included to resemble an increase in maintenance schedule. In modeling siloxanes, all siloxanes were grouped as a model siloxane, octamethylcyclotetrasiloxane. This is a simplification as this species is dominant in siloxane overall composition, representing greater than 50% of total siloxanes in LFG [99]. Pressure drop throughout the facility was not taken into account, as the placement of facility equipment and their final designs were not calculated. Prices of transportation fuels can change dramatically over a period of a few years and these changes are assumed to be generally consistent or on the rise, therefore profitability at a lower cost would mean profitability in the future.

2.3 Limitations

This is a preliminary analysis of the technology and its feasibility of use. Therefore, a number of simplifications limit the scope and accuracy of the analysis. Some limitations of this analysis are a simplifications of very complex factors, such as changes in LFG flow rate throughout the day or throughout the year according to temperature and rainfall. Not all factors are taken into consideration when looking at the sensitivity of the feasibility analysis, such as equipment inefficiencies when gas flow rate changes or when in non-optimal conditions. Surge capacity is also not taken into account. Taxes, subsidies, and grants were not taken into account as they are different from state to state and year to year. These cannot be dependable factors when looking into the feasibility of a facility which is not limited to a certain state. It is not known how many times an iron-scavenger bed can be regenerated before needing to be replaced. In this study the impact of regeneration is not taken into account. The life span and stability of the catalysts used is not fully understood and could be a significant factor in plant uptime. Simplified kinetics are used and do not fully assess the effect of process upsets which could alter the product distribution.

2.4 Process Design

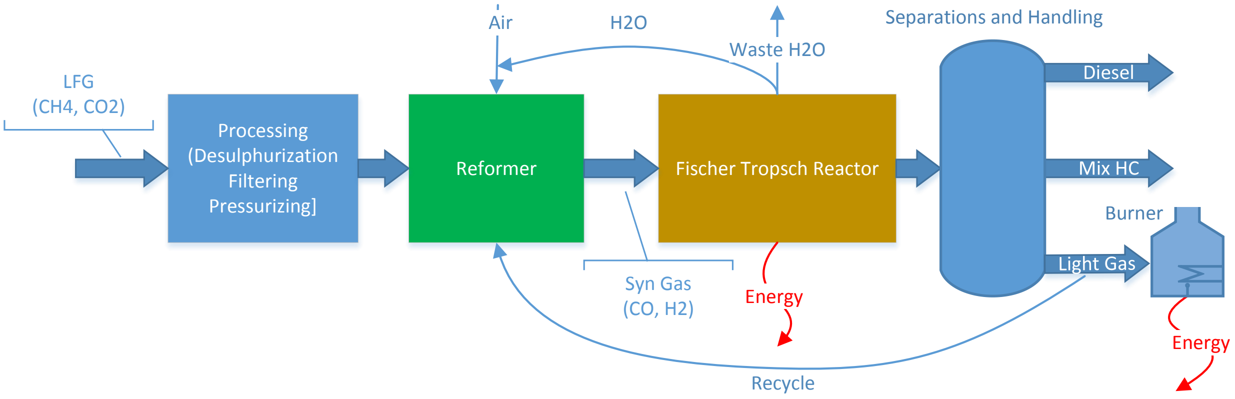


Figure 5: Process Flow Chart

There are 4 major portions of this plant:

1. Contaminant Removal and Preprocessing
2. Tri-Reforming of Methane and Carbon Dioxide into Syngas
3. Fischer Tropsch Synthesis of Hydrocarbon Fuels from Syngas
4. Separation and Upgrading of Liquid Fuel Cut

The gas is first purified and compressed through a series of compressors and absorbent beds. A conversion of the reactant gases into syngas occurs in the MTR reactor, followed by synthesis into long chain hydrocarbons in the FTS reactor. This gas is then separated by distillation or flash column into a multitude of gas components. All hydrocarbons that are considered out of range of middle distillate fuels are removed and sent to be combusted in order to offset the required energy input of the facility. Due to this inclusion, all energy used to run the MTR and distillation towers comes from burning undesired hydrocarbons produced in the facility.

A recycle stream could be added in after the FTS in order to recover unreacted CO_2 and CH4. However this separation is costly and is not specific to the components that need to be separated. Another recycle could be added during the separation stages to return a portion of the unused fuel cut to the MTR. A recycle stream was not implemented due to a dramatic increase in energy required to reform heavy species. Instead, most heavy distillate cuts were kept in the fuel product and were balanced with the addition of lighter cuts. The absence of a recycle significantly cuts operating and capital costs and increases process simplicity.

2.5 Plant Design

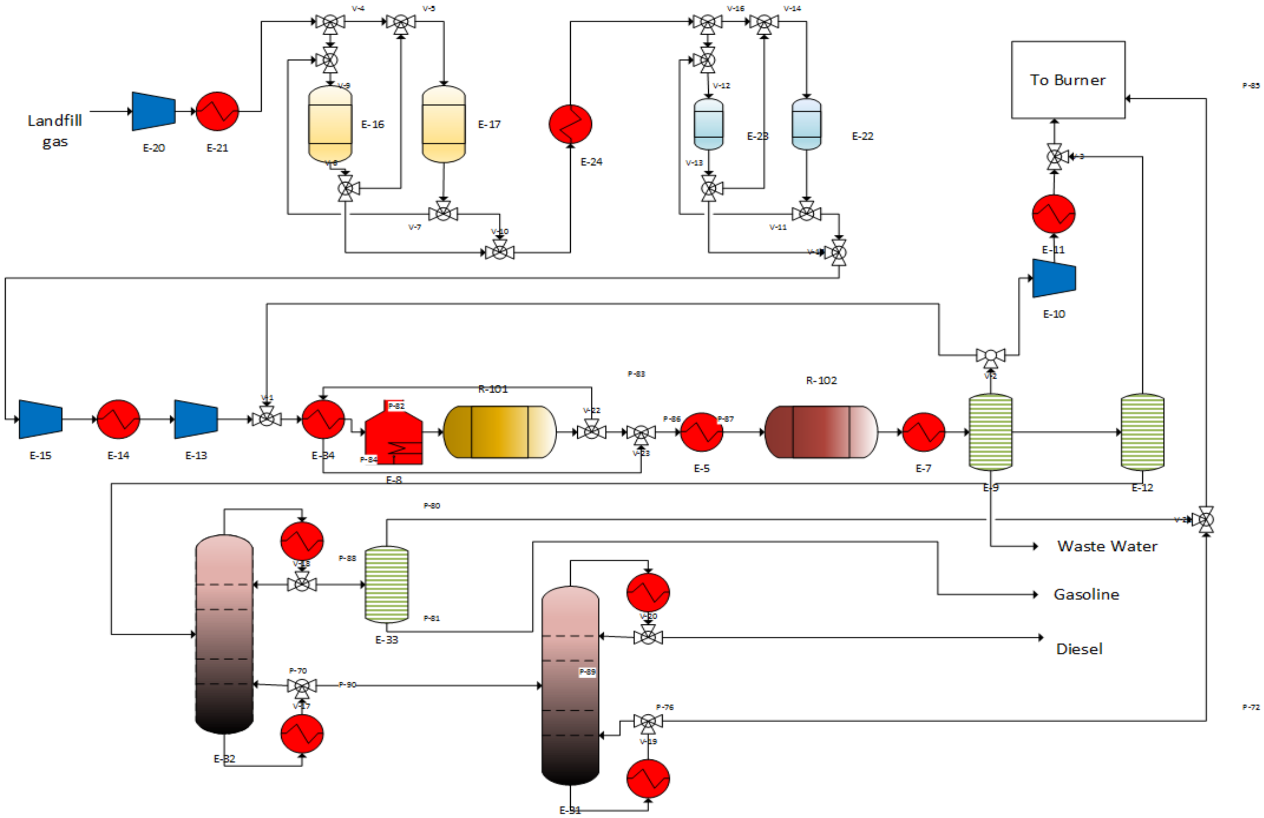


Figure 6: Process Flow Diagram

2.6 Contaminant Removal and Pre-Processing

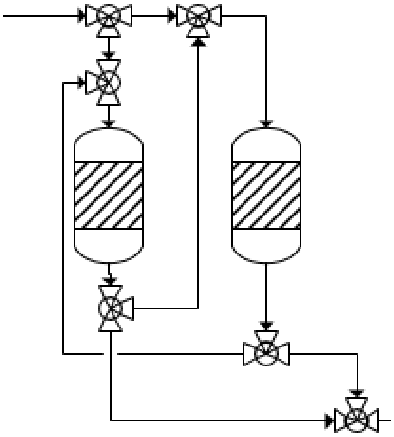


Figure 7: Contaminant Removal Bed Setup

Pre-processing contains all the necessary systems to prepare the incoming gas for the reactors, which includes contaminant removal and a series of compressors. All compressors work at ~3:1 compression ratio and include saturated liquid removal after inter-stage air cooling. The reason for contaminant removal is the fact that the final fuel product cannot contain contaminants which may negatively impact or cause damage to equipment or to the environment upon combustion and the catalytic process used can be stunted, or its effective lifetime reduced, by presence of catalytic poisons such as H_2S and siloxane gases. Removal of these contaminants before any further steps helps to simplify removal, which is easier in the gas phase. This involves a two stage system in order to remove H_2S in an iron based solid scavenger and the remainder of the contaminants on an activated carbon filter bed. Each of the two systems' stages use a lag-lead setup to reduce downtime and optimize contaminate removal. The possibility that the contaminant removal beds can be regenerated a few times before needing to be replaced provides advantages over other adsorption methods.

Incoming LFG is compressed by a single compressor to reduce the effect of contaminant damage to the subsequent compressors while increasing the efficiency of the contaminant removal system. This compressor brings the gas from 1 bar to 3 bar which feeds into the Iron based solid sulfur scavenger unit. It is designed to decrease H_2S from 700 ppm to below 5 ppm. Entering LFG must be saturated with water in order to work properly so that iron oxide is converted into Iron pyrite and sulfur is effectively removed. Industry uses determine that 1 g of iron scavenger (Sulfa-treat©) is required to remove 0.01 g of H_2S. The gas is then cooled before entering the activated carbon molecular sieve where siloxanes and other large NMOCs are removed. It requires 1 g of high surface area, acid washed, activated carbon per 0.04 g of siloxanes removed when less than 1 ppm. After contaminant removal, oxygen is added and the resultant gas is sent to a series of two compressors to bring the gas to 21 bar. A recycle stream of light hydrocarbon gases is mixed in and the gas is preheated by the outgoing reformer product stream.

2.7 Tri-Reforming Process

While steam reforming and dry reforming allow for tuning of the desired H_2:CO product ratio, partial oxidation of methane allows for an exothermic reaction which works in tandem with the first two reactions to reduce the energy required to run the reactor. This reactor is run at 800 C and 21 bar so it is assumed that there is a 99% single pass conversion of methane. A $Ce_{0.6}Zr_{0.4}$ w/ 8% Ni 8% Mg catalyst was chosen because previous literature studies have shown that it reduces coking and reduces sintering. This reactor was operated at 21,00 h-1 GSV in accordance with Walker et al. [94]. The MTR can also intake a portion of light hydrocarbon gases (C2 – C6) that can be recycled later in the process and reformed again. Steam is mixed with the incoming treated LFG to 20% by mole water and then preheated by the exit flow of the reformer, which simultaneously heats the input stream while cooling the output.

2.8 Fischer Tropsch Synthesis

The catalyst used is a specialized silica eggshell catalyst that allows for a higher residence times which lead to higher conversion without the common problem of hydrocarbons C25+ or higher being created [66]. The reactor operates at 21 bar and 230 C and converts 75% of syngas into a hydrocarbon stream. The desired products are hydrocarbons between C8 and C16. The specific catalyst used limits the formation of tar, alcohol, or unwanted formation of methane.

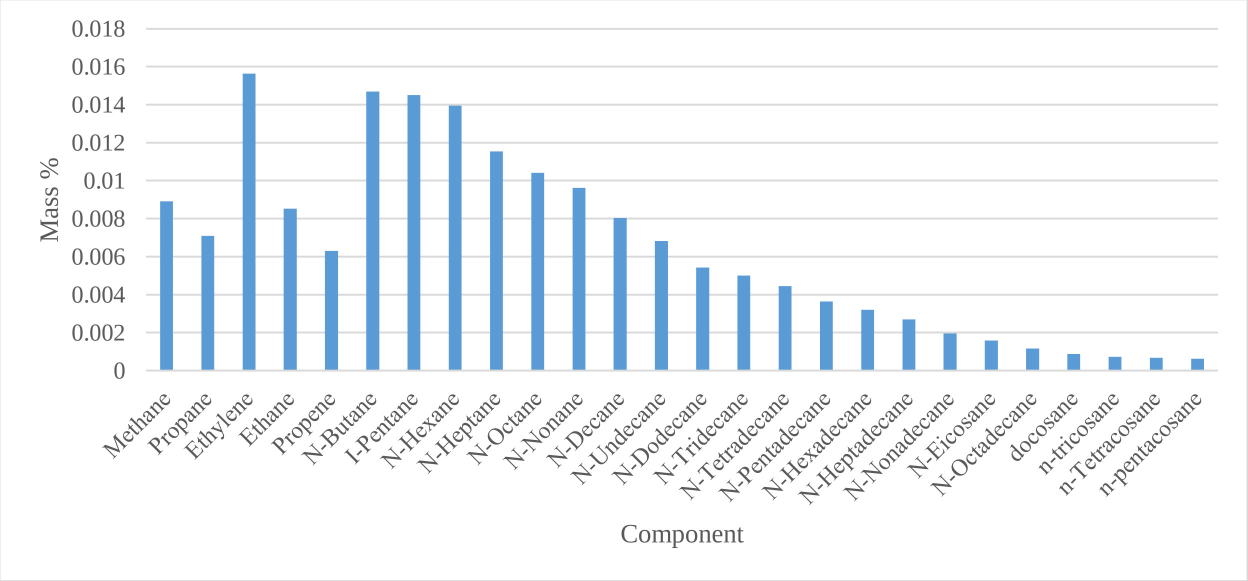


Figure 8: FTS Reactor Product Composition

2.9 Separation and Upgrading

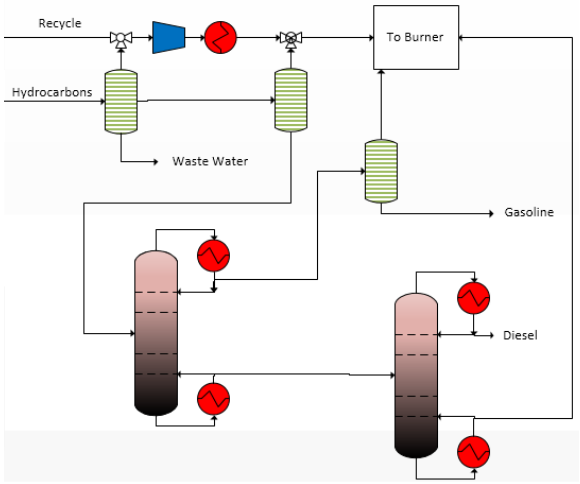


Figure 9: Separation Subsystem Diagram

The fuel separation process includes a 3-phase separator at a reduced temperature while pressure is held constant for the separation of light gases below C6, heavier hydrocarbons above C6, and water. The light gases are sent to be combusted in order to generate heat for the reformer and other equipment throughout the plant. The water is removed and sent back into the reformer for higher pressure steam. The remaining heavy hydrocarbon stream is flashed to remove remaining light gases then sent to a series of 2 distillation towers and 1 packed tower. Although many cuts of fuel can be achieved (Diesel Fuel, Jet Fuel, High Octane Petrol), the distillation system is tuned to optimize the production of diesel with some remainder of a gasoline precursor being created. To obtain this product we run our heavy liquid hydrocarbon stream into our first distillation tower where we separate 90% of the components that are lighter than C9 from those that are equal to or heavier than C9. This gives us a consistent flash point and energy content with that of commercial diesel. The heavier components are then sent through a second distillation tower that separates anything higher than C17 from our product diesel. The end result is the production of ~500kg/hr of diesel and 170 kg/hr of gasoline precursor. This equates to 240 gal/hr of diesel and 90 gal/hr of gasoline precursor. This is a conversion of ~15 wt% of LFG into fuel product. The diesel is considered ready for market sale, however it reaches a better price point when sold on site to the landfill operator as taxes, marketing, and distribution costs are forgone.

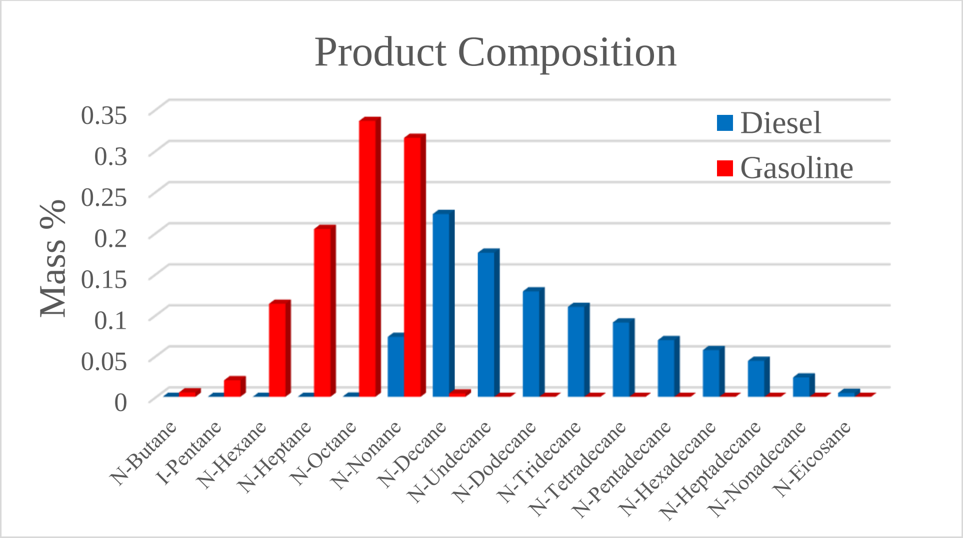


Figure 10: Product Composition of Diesel/Gas Precursor

The gasoline precursor is mostly fuel cuts which can be isomerized through further refining, but must be sold and shipped off site to a third party. The quality of this fuel is greater than typical crude oil feedstocks used by refineries and incurs significantly lower processing costs.

2.10 Minor Systems

A collection of other systems were integrated into the plant that are involved in optimizing the use of heat around the facility. Utility water is used to cool multiple streams. Water utility is taken in, used, and then cooled in an attached evaporation tower where the remaining water is reused. Pumps and other equipment are included with the full system cost of each unit operation. Electrical power is provided by an onsite micro turbine system.

3 Results and Discussion

3.1 Assumptions in Economic Analysis

This economic analysis hinges on the understanding of a few simple heuristics. For one, the plant should be financially feasible without tax breaks, government incentives, or other market modifiers. This is primarily due to the lack of consistency in availability of these programs, which can occur some years but be taken away others. Being dependent upon these factors for feasibility would reduce the overall chance of project stability. The second factor is that as a new technology, this project will incur many unforeseen costs. In order to combat that occurrence, costs associated with experimental pretreatment and reactor technologies were increased to compensate. This was done by taking worst case scenario prices on these systems and increasing the amount of pretreatment material used in cleanup. These changes also include increased maintenance and replacement costs for systems that see a large degree of contamination from LFG. The energy content of LFG used is 557 BTU/ft^3, this is representative of methane content at 55%. This does not include the effect of water content on LFG energy content as this water is removed throughout the process.

3.2 Product Pricing

The final product has the same flash point, cetane number, and energy density of diesel. Although all of these variables are consistent with diesel, the composition of the process product varies slightly compared to commercial diesel. Since the purpose of the diesel is to be used on site and avoid the costs associated with commercial distribution, selling the diesel very close to market value is justified. Synthetically generated diesel is traditionally of a higher quality when compared to fossil fuel diesel because the process produces long chain hydrocarbons which are favorable in diesel. However, this increase in quality does not increase the value of the product, but rather just the approachable market. The second product of this process, the mixed hydrocarbons or "gasoline precursor," also differs from traditional commercial composition. The gasoline precursor that is produced still needs to be sent to a refinery for isomerization and final finishing in order to be sold on the open market.

In order to find a price for both diesel and gasoline, a comparison between their prices must be made. In an open market sale the price ratio of diesel to gasoline is ~1.126, taking into account all seasons from 2009 to 2014 based on EIA pricing data (Look at Appendix H and Figure 3 for gasoline cost breakdown). Although this does not mean that this facility can always match these prices, it indicates an open market scenario that the facility must meet in order to maintain feasibility. The costs of distribution are also subtracted (~10%) because this is not a step that sale on site would require. All other systems are similar between the production of fossil fuel and synthetic diesel. Breakeven prices of diesel and gasoline were ~$2.73 Diesel and ~$2.18 gasoline.

Table 3: Liquid Fuel Production Volume per Year

Component	**US Gallons Produced**
Diesel	2,021,760
Gasoline precursor	758,160

Table 4: Fuel Revenue per Year

Component	**Revenue per year**
Diesel	$5,519,000
Gasoline precursor	$1,653,000
TOTAL	**$7,180,000**

3.3 Equipment Costing

The equipment is going to be exposed to a large amount of contamination, which can vary depending on what is in the landfill and the waste that is entombed. This leads to degradation of facility equipment. The equipment cost was purposely increased by 50% so that it would offset the replacement costs of the compressors, particularly the compressor prior to pretreatment. The reactors were priced as if they were shell-in-tube heat exchangers. This is primarily due to temperature control and cost being the most known cost at a larger scale. The 3-phase separator, pretreatment beds, and flash vessels were sized and costed by hand.

In the proposed system, ChemCad is used for plant design and calculation of thermodynamics, phases, and streams. Capcost is used to cost pieces of equipment.

3.4 Land Costing

It is assumed that the cost of purchasing land was negligible as the facility would be built on a landfill. This land is generally considered undesirable or already owned by the customer or operator. Due to many LFG agreements, there is usually an acknowledgment that the plant would take over the cost and duty of greenhouse gas emission as well as dealing with subsequent emissions fines from the landfill operator. This would also be assumed to be negligible due to the use of green energy credits and carbon offsets. There is also possibility of the facility being owned by the landfill itself, in which the individual cost would be null.

3.5 Capital Investment

A significant capital investment is required to operate this facility. Approximately 40% of all capital costs go to the construction of the two main reactors. Construction of extra pumps, compressors, and drivers were added in order to decrease the chance of facility downtime.

Table 5: Facility Cost Breakdown

Unit Name	**Cost**
Heat Exchangers	$622,000
Compressors	$3,988,000
Drivers	$440,000
Towers	$525,000
Reforming Reactor	$1,300,000
FTS Reactor	$2,500,000
Iron Packed Beds	$408,000
Carbon Beds	$84,000
Flash Vessels	$311,000
Total Bare Module	**$9,578,000**
Total Install Cost	**$11,250,000**

3.6 Operating Costs

It is common practice to purchase LFG from the landfill operator at 30 to 50 percent below the average monthly cost of natural gas on the indexes such as NYMEX [53]. Using LFG at 45% of NG prices and considering NG at $3 per MMBTU, LFG is 1.35 $/MMBTU. A detailed method of LFG costing is described in Appendix B. The final cost of LFG is 0.84 Million USD per year and is a recurring operating cost.

Many of the values chosen for operating costs represent a less than optimal scenario. Maintenance cost of the facility is raised slightly as contamination and variability will inevitably cause increased downtime if equipment is not maintained. A breakdown of operating costs in shown in Table 6. Utilities are primarily the cost of electricity and cooling water for removal of low quality waste heat.

Table 6: Operating Costs

Parameter		Cost
Maintenance	5.5% of FCI	$619,000
Labor	7 operators	$1,139,000
Materials and Utilities		$3,174,000
Utilities		$1,268,000
Materials		$457,000
Clean up		$16,000
LFG Purchase	45% of NYMEX	$840,000
Total		**$4,932,000**

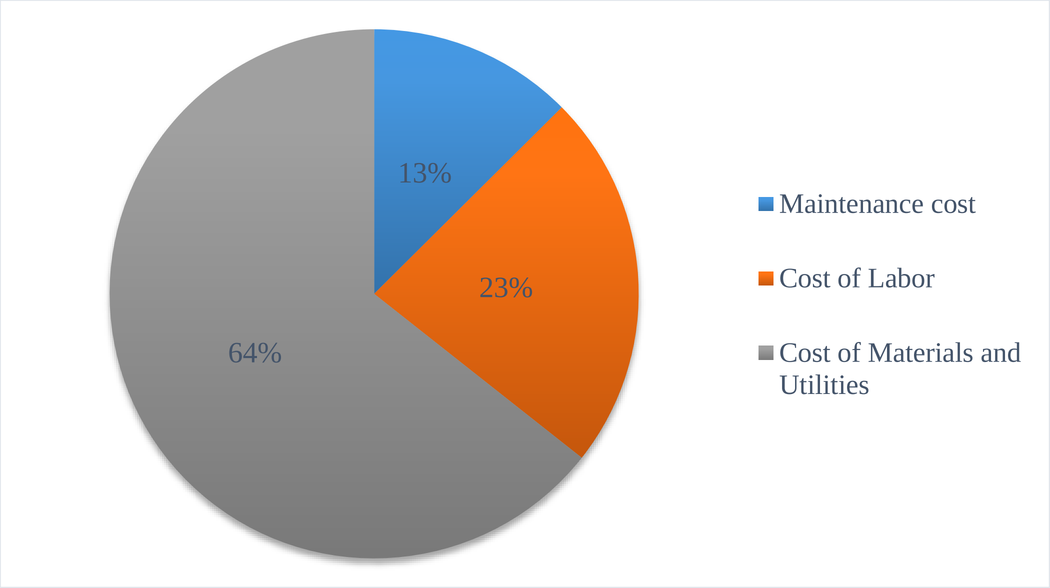


Figure 11: Operating Costs Distribution

Table 7: Utilities Pricing

Utility	**30◦ C cooling water**
Price per unit	$14.8 / 1000 m3
Price/year	$ 1,267,635.80

It is assumed all heating fuel that will be needed will come from the processing of non-usable gas portions. The cost of pretreatment is based on industry pricing for the adsorbent media with the large degree of sulfur contamination adding significant costs. The prices for the Sulfa-rite and acid washed activated carbon are based on industry specifications from Merichem and Cabot Corp [100].

Table 8: Raw Materials Cost

Raw material	**Sulfa-Rite©**	**Acid washed A.C.**
cost per lb	$ 0.50	$ 1.50
lb / yr	880,175	11,070
Price/year	$440,000	$16,600

The total combined cost for pre-treatment materials is 456,600 $/yr. Disposing of this material, as well as waste water generate in the process is 16,522 $/yr

Table 9: Waste Removal

Waste	**Spent packed column media**	**Hazardous Waste water**
cost per unit	$36 per ton	$56/1000 m^3
Amount removed	445.6 tons	8,575 m^3/yr
Price/year	$16,042	$480

3.7 Economic Analysis

The following tables show the breakdown of the facility economics over its 15 years lifespan. In this simulation tax was left at a normal 30% bracket, even though a possibility of being in a decreased bracket exist, as this project is a green energy project.

Table 10: Feasibility Analysis Parameters

Parameter	**Value**
Plant Life	15 years
Tax Rate	30%
Minimum Attractive Interest	15%
Depreciation Method	MACRS
Depreciation Life	9 Years

A theoretical sale price of $3 per gallon for diesel and $2.40 per gallon gasoline is used in order to create a situation where a profitable sale price occurs. Table 11 shows the breakdown of this scenario, where the plant will generate $2.94 million dollars in worth and be paid off in under 9.5 years.

Table 11: Plant Financial Analysis

$3 Diesel, $2.40 gas		
Fixed Capital Investment	11.25	Million $
Total Capital Investment	12.5	Million $
Operating and Manufacturing Cost	4.93	Million $/yr
Revenue	7.88	Million $/yr
Yearly Profit	~2	Million $/yr
Feasibility Results		
Discounted Payback Time	9.5	Years
Discounted Cash Flow Rate of Return (DCFROR)	19.7%	
Net Present Worth (NPW)	**$2.94**	**Million**

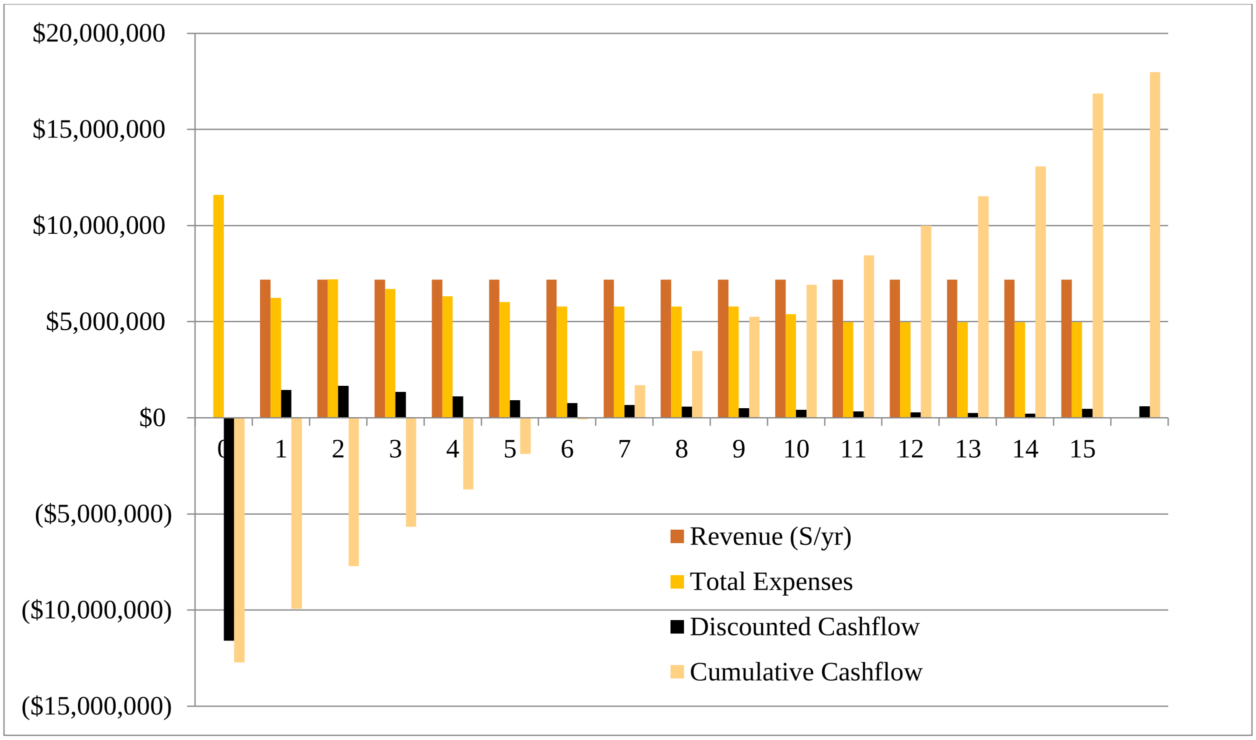


Figure 12: Financial Flow Chart ($3 Diesel, $2.40 Gas)

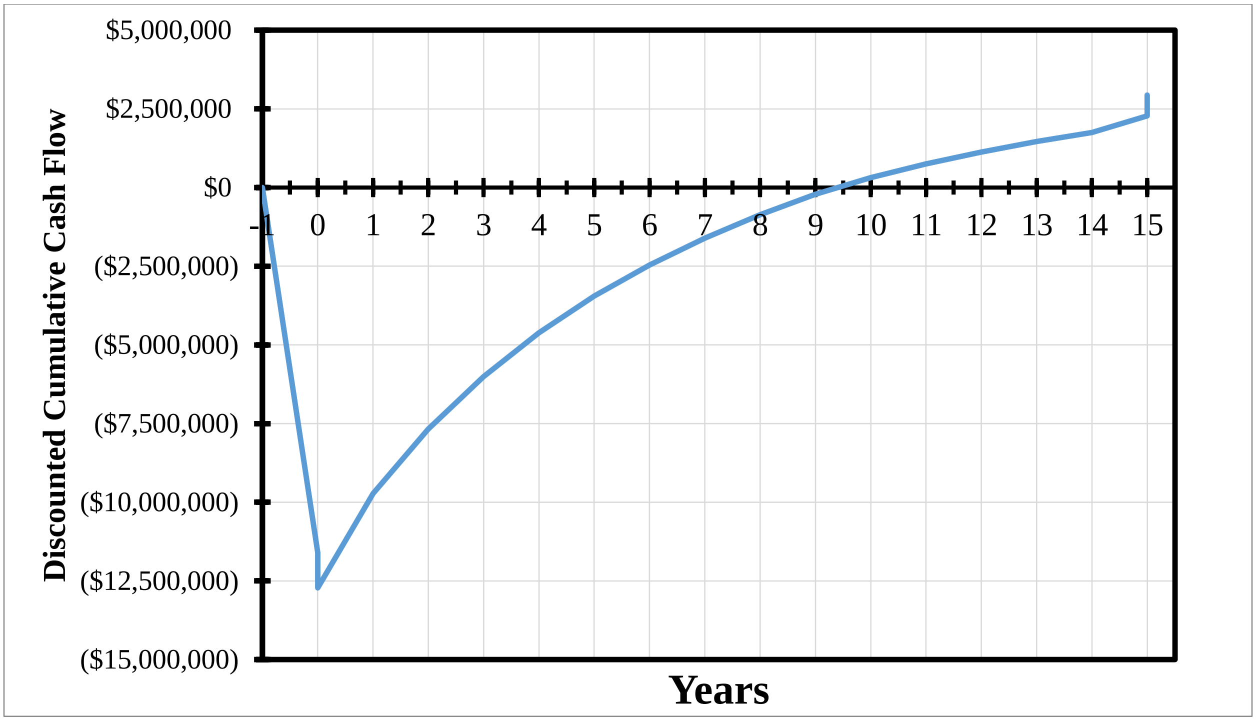


Figure 13: Discounted Cumulative Cash Flow ($3 Diesel, $2.40 Gas)

3.8 Sensitivity Analysis

A sensitivity analysis was performed based on multiple parameters to determine the major components necessary for feasibility of the facility. The analysis was performed so that the breakeven cost of diesel could be found (NPW=0). The base case scenario provided a cost of diesel at $2.73 and gasoline at $2.18. Gas flow rate was the largest factor, as it is assumed that the facility can handle gas flow fluctuations and the production rate was changed by the same percentage (assuming that all of values remained the same and only fuel production rate was affected). A general swing in price of equipment was included at 20% in order to account for final design parameters. Contamination alterations were made (changing the O&M of the preprocessing) knowing that the incoming gas could contain a variable degree of contamination. Since the base case scenario assumed a near worst case contamination concentration, the majority of the range was in decreasing the requirements of contamination reduction beds as contamination was reduced. As a change in the concentration of contaminants directly impacts the amount of bed material used, no other impact was taken into account.

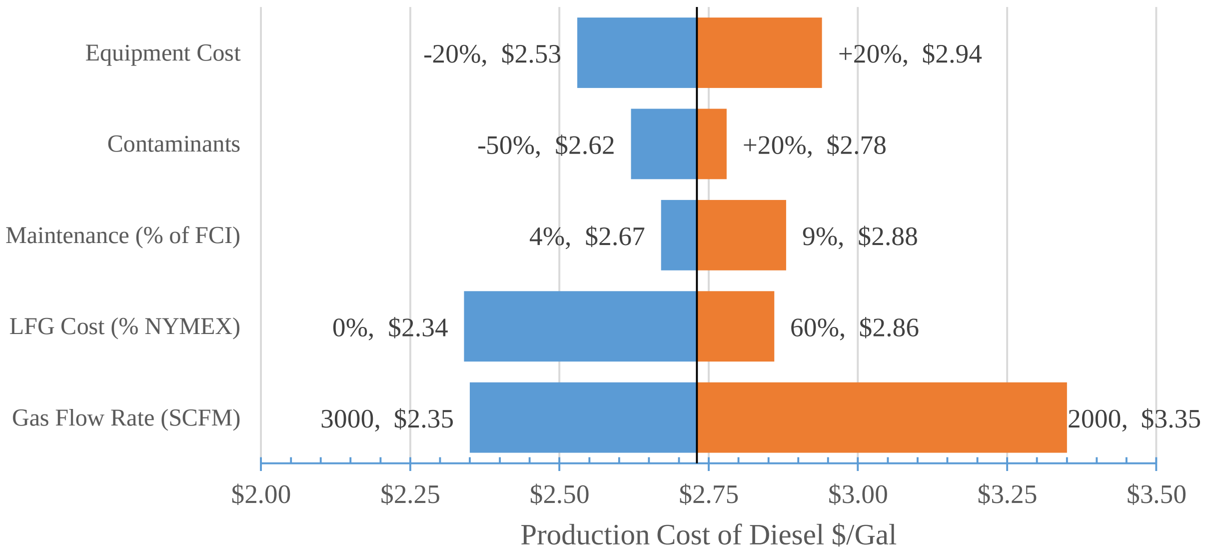


Figure 14: Economic Sensitivity Factors

It is important to note that the price of both diesel and gasoline is projected to rise over the life of this plant. Therefore, using the previously mentioned prices is rather pessimistic, but is done so purposely to offset some unforeseen costs.

3.9 Product Factors

It is important to note three factors when generating renewable liquid fuels. The first is determining the conversion efficiency when compared to the maximum theoretical conversion of LFG into a model hydrocarbon. The second is the amount of CO_2 sequestered using the fuel. The third is the amount of CO_2 not generated by the use of the renewable fuel over the traditional fossil source. Results of using the EPA LFG Project Calculator, located in Appendix J, for a 2500 SCFM LFG project show a reduction of 13,896 tons of CH_4 emissions per year.

As carbon chains grow the fuel can be represented as nCH_2 where the traditional formula denotes C_nH_{2n+2}. At C14, tetradecane, where the tradition notation shows 14 carbon and 30 hydrogen atoms, the simplified version denotes 2 less hydrogen atoms (only 7% off).

$$14CO + 28H_2 \rightarrow 14CH_2 + 14H_2O \tag{10}$$

Since the FTS reactor converts 75% of syngas in useful hydrocarbons, this is assumed to be entirely C14, and the MTR converts methane into syngas at 99% the maximum efficiency for fuel generation is 74%.

A carbon balance, depicted in Table 12 below, was performed to show the ability of the system to mitigate carbon emissions by sequestering it in fuels. The LFGTL system removes 30% of the carbon which would emitted had it all been flared. A maximum of 55% of the carbon can be removed and utilized if the removal process is 100% efficient. This limitation is based on the fact that energy is required to convert CO_2, without outside energy input methane must be used for this purpose. Although this process uses CO_2 as a reactant, which is helpful as removal isn't required, it produces more CO_2 than it takes in as some energy from methane is used to generate the higher energy fuels.

Table 12: Carbon Balance

	LFG Input	**Diesel Product**	**Gasoline Product**	**Flue Gas**
Moles Carbon /hr	185376	40789	14450	130005
% Carbon		22%	7.8%	70%

At breakeven the base case scenario prices of $2.73 for diesel and $2.18 for gasoline, an approximate $/MMBTU can be generated. This calculation can be found in Appendix D. Increases in the products value are seen as a rise in the cost per BTU. Fuels with higher demand and overall worth cost more per BTU. Converting landfill gas to diesel increases the $/BTU by 15 times; 1.35 $/MMBTU LFG to 19.71 $/MMBTU Diesel.

Table 13: BTU/$ of Conventional and Product Fuels

Fuel	**Specification**	**$/MMBTU**
Diesel ULS #2	2.73 $/gal	19.71
Conventional Gasoline	2.18 $/gal	18.1
Landfill Gas	45% of NYMEX	1.35
Electricity (Turbine)	0.065 $/kWh 13,000 BTU/kWh	5

Another important factor that should be shown from this project is the energy storage capability of the facility. In simpler terms, this would equate to the amount of energy that is converted into fuel compared against the amount of energy used to produce the fuel. Table 14 shows this efficiency by a balance on LFG input and fuel energy output. Further calculations for these numbers can be found in Appendix E.

Table 14: Energy Storage Efficiency of LFGTL

	MMBTU/year
LFG Energy Input	703,825
Fuel Energy Output	371,266
Energy Storage	53%

A breakdown of process energy is provided using the heat consumption of the reactors and the generation. Of the 83 MMBTU/hr which enters the system, 53% is converted into useable fuels, while the rest leaves the plant as thermal energy. A breakdown of these energy flows can be found in Appendix E. The MTR reactor requires ~40 MMBTU to preheat and react the incoming treated LFG. Approximately 18 MMBTU/hr of heat is required raise the incoming LFG to the 800C reacting temperature. The remaining 22 MMBTU/hr is required to convert the reactants into syngas. 15 MMBTU /hr of the heat required to raise the temperature can be obtained by preheating the treated LFG with the hot syngas leaving the MTR reactor; more cannot be used due to pinch point limitations. The remaining 25 MMBTU/hr required by the MTR reactor is provided by the combustion of waste hydrocarbons from the furnace. 75% of the furnaces total usable thermal energy, at ~34 MMBTU/hr (85% of 40 due to parasitic loss), is sent to the reactor, while the remaining portion is used to run reboilers throughout the plant. Without the use of heat integration and reuse through the plant, an additional 40 MMBTU/hr of heating would be required and 80 MMBTU/hr of cooling. If heat removed from the FTS, equaling ~20 MMBTU/hr, is converted into electricity at 13,000 BTU/kWh, it would generate greater than 1 MW of electricity which could power the remainder of the facility. Over the course of the entire conversion, 47% of energy is lost as heat through cooling.

4 Market Scenario

4.1 Only Diesel Fuel Can Be Sold

In a scenario where the gasoline precursor is considered unsellable or unusable, as no third party will purchase the product or because more energy is required to heat the furnaces, its sale price is reduced to $0 per gallon. In this event no other production factor is changed as it is most likely that the product will be burned, regardless of if the extra energy is required. Although this does hold promise for the use of electric steam turbines or other energy recovery methods, they are not taken into account here. In order to maintain an NPW of $0 over plant lifetime the price of diesel must be sold at a minimum of $3.55 per gallon in order to remain feasible at 15% interest.

4.2 Equipment Downtime is Increased

In a scenario where the facility has increased downtime, and is operating less than the previously identified 351 days per year, profitability decreases. Even though increased maintenance scheduling and equipment backups were priced for accessory equipment, failure in one of the reactors or distillation columns would cause product loss. In the event the reactor catalyst needs to be swapped out greater than once per year, the system would require should down for extended periods on multiple occasions. FTS reactors using this catalyst typically take a significant amount of time to reach reaction steady state. This leaves days of lower than expected production.

The number of working days is decreased from 351 to 330, or the addition of 3 weeks of downtime. This effect is calculated as the following effects on extra days 1) No profit from downtime days, 2) 75% reduction in operating costs (heating, cooling, and electricity). As LFG purchasing is under contractual agreement, it must still be purchased on non-operating days. In order to maintain an NPW of $0 over plant lifetime the price of diesel must be sold at a minimum of $3.76 per gallon in order to remain feasible at 15% interest. If gasoline is also sold the price of diesel will be $2.90 per gallon and gasoline will be $2.32 a gallon.

4.3 Required Capital Investment Ceiling

In this scenario the price of diesel and gasoline are locked according to a current day sale cost, and the purchase price of landfill gas is locked to natural gas projections. The price per gallon of diesel projected in 2016 and 2017 is $2.22 and $2.58 according to Energy Information Association outlooks (As of February 9th, 2016). Since purchase and sale are known locked variables, the cost of capital investment will be altered to find the highest cost of facility construction to maintain feasibility. The effect of changing capital cost is far reaching, it is essentially altering the plant itself and therefore changes the majority of factors which go it feasibility analysis. The only factors which were co-calculated with this analysis were those which have the largest effect on feasibility outcome and are not speculative. Therefore depreciation, maintenance cost, and salvage value are the other variables which are iterated. Variables such as operators, costs associated with contamination, and minor costs were omitted. The following table shows the results of changing fuel price on initial plant investment cost.

Table 15: Capital Ceiling Feasibility

Sale Price	Capital Investment Ceiling
$2.22 for Diesel $1.77 for Gasoline	$5.60 Million
$2.58 for Diesel $2.06 for Gasoline	$9.57 Million

5 LFG Usage Option Comparison

The following section will compare the use of LFG for liquid fuel production versus other options, such as flaring and electricity generation. A flaring system is installed on every landfill and is considered a preexisting cost when approaching installation of another LFG usage option. In order to generate 2500 SCFM of gas flow rate, 8.33 Million tons of waste must be in place at the landfill. It is assumed that there is 1 collection well per acre, and each well can collect 10-30 SCFM. This mean that between 83-250 wells are required.

5.1 Flaring

The price of a flaring system alone is produced to show the economic impact of LFG mitigation without use. The addition of another system will not affect the initial install cost of a flare, however, it will greatly reduce or even remove the operating costs associated with its discontinued use. Table 16 shows the financial breakdown for a flaring system which generates no revenue and only incurs annual costs. Flaring systems do use electricity in their blower and gas control systems.

Table 16: Flaring System Financial Analysis

Flaring System		
Total Capital Investment	2.8	Million $
Operating and Manufacturing Cost	0.5	Million $/yr
Revenue	NONE	Million $/yr
Yearly Profit	-0.6	Million $/yr
Feasibility Results		
Net Present Worth (NPW) (15% Rate)	**-4.76**	**Million $**

5.2 Electricity Generation

A number of different methods exist for the combustion and use of LFG for energy or heat generation. Projects that include mass burn have additional advantages, such as the reduction of overall mass landfilled, which are not easily quantifiable and comparable to other options. Electricity generation can proceed by reciprocating engine, standard or micro turbine, or Combined Heat and Power (CHP). Two of these options will be presented, both of which use turbines and depend on whether waste heat is utilized in addition to generating electricity. Turbines will be used instead of reciprocating engines as generating capacities of greater than 3 MW cannot be done without the use of multiple reciprocating engines. Electricity is priced at current national industrial purchase price, at 0.065 $/kWh. The discount rate is maintained at 15%.

The following options show the difference between using waste heat developed in the electrical generation process. At this scale, heat recovery is not expensive and develops a significant net of 2.4 MM USD over not including it. Heat recovery in this instance is comparable to using co-generation in LFGTL and not indicative of heat integration, but the direct sale of thermal energy. Electrical generation is greatly dependent on agreements for the sale cost of electricity to a network, and the ability to sell off thermal energy as steam. These prices do not vary widely, except regionally where electrical distribution prices are different.

Table 17: CHP Turbine Financial Analysis

6.4 MW Capacity and 271,000MM BTU/yr produced		
Total Capital Investment	10.9	Million $
Operating and Manufacturing Cost	0.8	Million $/yr
Revenue	4.4	Million $/yr
Yearly Profit	1.7	Million $/yr
Feasibility Results		
Discounted Payback Time	None	Years
Rate of Return (ROR)	16%	
Net Present Worth (NPW) (15% Rate)	**0.4**	**Million $**

Table 18: Turbine Financial Analysis

6.4 MW Capacity		
Total Capital Investment	9.5	Million $
Operating and Manufacturing Cost	0.8	Million $/yr
Revenue	3.2	Million $/yr
Yearly Profit	0.8	Million $/yr
Feasibility Results		
Discounted Payback Time	None	Years
Rate of Return (ROR)	10%	
Net Present Worth (NPW) (15% Rate)	**-2.0**	**Million $**

The financial return of these facilities may seem low compared to other literature values, this is due to the method by which the system is analyzed. In order to compare the facilities with LFGTL, the facility is paid for 100% up front, does not consider a number of inflationary costs, and mimics the development of this facility by a private entity purchasing the gas from the landfill at 1.35 $/MMBTU.

5.3 Impact on Greenhouse Gas Emissions Comparison

It is important to note that any facility that either destroys or uses the CH_4 in 2500 SCFM of landfill gas will offset 723 Million ft^3 of CH_4 emissions per year, equal to 0.347 $MMTCO_2$. Depending on the use of the LFG, additional equivalent CO_2 emissions may be avoided due to the offset of fossil fuel use. The following table presents the total emissions offset by each facility. CO_2 production by electrical generation is based on 2014 U.S. average of 1.18 lb CO2/kWh.

Table 19: Greenhouse Gas Comparison

	Flaring	**LFGTL**	**CHP Turbine**	**Turbine**
CO2 Emissions Avoided ($MMTCO_2$/yr)	0	0.029	.044	.026

5.4 Facility Comparison Chart

The following is the comparison of all four of the LFG usage options presented in this paper. All units are based on yearly values. Table 21 shows the side by side comparison of these options based on economic, emission, and product values.

Table 20: Comparison Parameters

Flow Rate	2500 SCFM (1.2636 Billion SCF per year)
Methane Content	55%
LFG Energy Content	557 BTU/SCF
Energy Content	703,825 MMBTU per year
Carbon Content	1.5616 Billion moles carbon per year

Price of Electricity Sold	$0.065 per kWh
Price of Diesel Sold	$3.0 per gallon
Price of Gasoline Sold	$2.4 per gallon
Price f Steam Sold	$4.5 per MMBTU

Table 21: LFG Usage Option Comparison

	LFGTL	**CH&P**	**Electric**	**Flaring (Only)**
Product	Diesel Gasoline	Electricity Thermal	Electricity	None
Product Rate (per year)	2,021,760 gal diesel 758,160 gal gasoline	49.36 Million kWh 271,468 MMBTU Steam	49.36 Million kWh	-
Energy Required	-	-	-	2.53 Million kWh
Energy Recovered (MMBTU/yr)	371,266	666,348	394,880	None
Plant Emissions ($MMTCO_2$/yr)	0.0482	.0703	.0703	0.0703
Avoided Emissions Using Product ($MMTCO_2$/yr equiv)	0.0290	.0440	.0260	0
Total Emissions (Plant-Avoided)	0.0192	0.0263	0.0443	0.0703
Capital Investment (Million USD)	12.5	10.5	9.5	2.8
Operating Expenses (Million USD per year)	4.9	0.8	0.8	0.5
Net Present Worth (15% Rate)	2.94	0.40	-2.00	-4.76

6 Conclusion

After a preliminary feasibility analysis it is recommended that an investigation into the process continue to further develop the use of gas-to-liquids technology on naturally occurring biogases. Currently, the feasibility of the project depends on 4 major factors: royalty price paid for LFG ($/MMBTU), the capital cost of the installed facility, facility downtime, and is heavily dependent of the sale price of the diesel and gasoline products. An experimental facility that is based on the price of an economically volatile component incurs a significant amount of risk. A price swing of a few cents can have a dramatic effect on profit margins. The LFGTL facility itself reduces 30% of CO_2 released by options that combust LFG. An additional amount of CO_2 emissions are avoided through the offset of fossil fuel use, however, this process could stand to gain with better use of excess facility heat. When LFGTL is compared to electrical generation, an additional $2-3 million in capital expenditures are required as well as over ~$4 million per year in operating costs. LFGTL also achieves the lowest overall equivalent CO_2 emissions, with 27% lower emissions compared to CH&P. As the price of liquid fuels rise, the generation of liquid fuels will become more attractive than generating electricity.

6.1 Potential Process Improvements

Many options remain for further development of the system. Understanding the pretreatment and contaminant removal process is vital to the longevity and feasibility of this process. Contaminants that are not removed can cause significant damage to capital equipment and catalyst beds, further increasing the cost of the facility. It is advised that further research be done to find a packing media which is selective to molecules larger than CH_4 and CO_2 for easy removal. It was assumed that the catalysts would react unfavorably to any contamination present in LFG. However, investigation of the actual resistance of the two catalysts to the major contaminants found in landfill gases would allow better sizing and construction of the pretreatment process.

Other options remain when looking at processing of the hydrocarbon products. Instead of separating it into diesel and gasoline fractions onsite, the hydrocarbon mixture could be sold as a high quality, light, sweet crude. The option was looked into for sale to local refineries for post processing, however, when the mixture is sold to refineries it must compete with the price of incoming crude and its sale price is significantly reduced. This is mainly due to the generation of the hydrocarbon stream being the most expensive processing step in both operating and capital costs.

Due to the generation of a large amount of low quality waste heat, it may be advantageous to use an air cooling system instead of water cooling using utility. The addition of nitrogen, provided by air into the system, either by landfill leaks or from addition of air for use in reforming process, decreases the efficiency of the process or will increase the likelihood of byproducts being made from nitrogen. A solution to this problem is to use excess waste heat to run turbines, whose energy can be used for electricity generation for the facility or used in compressors for air separation processing. The coupling of this plant with other facilities in cogeneration will, overall, be advantageous. The amount of low quality waste heat is very valuable in water treatment and for other industrial processes. Using additional systems, such as solar heating, can reduce the amount of energy required for heating allowing a higher recycle rate of gas and subsequent generation of more product. In this facility design, water separated from the FTS reactor contains dissolved hydrocarbons and a fee is charged for disposal. However, a more feasible option may be onsite cleanup through aeration of the water and desorption of the dissolved hydrocarbons.

6.2 Market Instability

It is important to mention that the timing of this thesis coincided with a major crash of world oil and fossil fuel prices. At the beginning of my master's degree, Diesel ULS #2 and conventional gasoline hovered between 3 and 4 $/gallon, a level that was generally stable for the previous 3 years. Over the next year, this price would fall to under 2 $/gallon. This dramatic change is price was due to a large dump of oil from OPEC nations, flooding the market with very cheap oil. The price of oil in June of 2014 was over $100 a barrel, and as of writing this paper the price of oil has fallen 80% to around 20 $/barrel.

While the drop in fuel prices has been helpful in some ways to the US economy, it has led to many alternative energy projects, such as this one, losing their footing as being economically feasible options. Due to a large number of variables and risks in developing a facility like this, a large safety margin is usually needed by investors to make alternative energy projects attractive. This project's feasibility competes directly with open sale of transportation fuels, but does not have the same protections that large refiners incur.

This facility holds promise, however, in order for the process to be financially feasible gas prices will need to rise above the breakeven point. Additional increases above this point will grant a larger margin of safety for investing in this system.

References

1. EIA. *International Energy Outlook*. 2013 [cited 2014; Available from: http://www.eia.gov/forecasts/ieo/world.cfm.

2. EIA, *U.S. Crude Oil and Natural Gas Proved Reserves, 2014*. 2015, U.S. Department of Energy: Washington D.C.

3. Oil & Gas Journal, 2008. **106**(13).

4. EIA, *SHORT-TERM ENERGY OUTLOOK: February 2016*, in *Short Term Energy Outlook*. 2016, Energy Information Association.

5. Whitney, G., C.E. Behrens, and C. Glover, *U.S. Fossil Fuel Resources:Terminology, Reporting, and Summary.* Congressional Research Service, 2010: p. 24.

6. Carley, S., *State renewable energy electricity policies: An empirical evaluation of effectiveness.* Energy Policy, 2009. **37**(8): p. 3071-3081.

7. Schnepf, R. and B.D. Yacobucci. *Renewable fuel standard (RFS): overview and issues*. in *CRS report for congress*. 2013.

8. Lund, H., *Renewable energy strategies for sustainable development.* Energy, 2007. **32**(6): p. 912-919.

9. Proctor, C. *U.S. natural gas reserves at record levels, says School of Mines panel*. 2013 [cited 2014; Available from: http://www.bizjournals.com/denver/blog/earth_to_power/2013/04/us-natural-gas-reserves-hit-record.html?page=all.

10. Lyubovsky, M., S. Roychoudhury, and R. LaPierre, *Catalytic partial "oxidation of methane to syngas" at elevated pressures.* Catalysis Letters, 2005. **99**(3-4): p. 113-117.

11. McKendry, P., *Energy production from biomass (part 1): overview of biomass.* Bioresource technology, 2002. **83**(1): p. 37-46.

12. Andreae, M.O., *Biomass burning: its history, use, and distribution and its impact on environmental quality and global climate.* Global biomass burning: Atmospheric, climatic and biospheric implications, 1991: p. 3-21.

13. Yerkes, R.F., T.H. McCulloh, J. Schoellhamer, and J.G. Vedder, *Geology of the Los Angeles basin, California: an introduction.* US, Geol. Surv., Prof. Pap.;(United States), 1965. **420**.

14. Bertine, K. and E.D. Goldberg, *Fossil fuel combustion and the major sedimentary cycle.* Science, 1971. **173**(3993): p. 233-235.

15. Watanabe, Y., Y. Shimada, A. Sugihara, and Y. Tominaga, *Enzymatic conversion of waste edible oil to biodiesel fuel in a fixed-bed bioreactor.* Journal of the American Oil Chemists' Society, 2001. **78**(7): p. 703-707.

16. Tsai, W.-T., C.-C. Lin, and C.-W. Yeh, *An analysis of biodiesel fuel from waste edible oil in Taiwan.* Renewable and Sustainable Energy Reviews, 2007. **11**(5): p. 838-857.

17. Panda, A.K., R. Singh, and D. Mishra, *Thermolysis of waste plastics to liquid fuel: A suitable method for plastic waste management and manufacture of value added products—A world prospective.* Renewable and Sustainable Energy Reviews, 2010. **14**(1): p. 233-248.

18. Reddy, B.V., S. Ramesh, A.A. Kumar, S. Wani, R. Ortiz, H. Ceballos, and T. Sreedevi, *Bio-fuel crops research for energy security and rural development in developing countries.* Bioenergy Research, 2008. **1**(3-4): p. 248-258.

19. Johnson, J.-F., R. Allmaras, and D. Reicosky, *Estimating source carbon from crop residues, roots and rhizodeposits using the national grain-yield database.* Agronomy journal, 2006. **98**(3): p. 622-636.

20. Puig-Arnavat, M., J.C. Bruno, and A. Coronas, *Review and analysis of biomass gasification models.* Renewable and Sustainable Energy Reviews, 2010. **14**(9): p. 2841-2851.

21. Bahng, M.-K., C. Mukarakate, D.J. Robichaud, and M.R. Nimlos, *Current technologies for analysis of biomass thermochemical processing: a review.* Analytica Chimica Acta, 2009. **651**(2): p. 117-138.

22. Larson, E.D., R.H. Williams, and M.R.L. Leal, *A review of biomass integrated-gasifier/gas turbine combined cycle technology and its application in sugarcane industries, with an analysis for Cuba.* Energy for sustainable development, 2001. **5**(1): p. 54-76.

23. Mitchell, D.A., M. Berovič, and N. Krieger, *Solid-state fermentation bioreactor fundamentals: Introduction and overview*. 2006: Springer.

24. EPA, *Municipal Solid Waste Generation, Recycling, and Disposal in the United States: Facts and Figures for 2013*. 2015.

25. Bryant, M.P., *Microbial Methane Production—Theoretical Aspects1 , 2.* Journal of Animal Science, 1979. **48**(1).

26. Themelis, N.J. and P.A. Ulloa, *Methane generation in landfills.* Renewable Energy, 2007. **32**(7): p. 1243-1257.

27. Townsend, T., A. Byrne, H. Carter, K. Hodoval, and S. Sikora, *ALACHUA COUNTY WASTE COMPOSITION STUDY*. 2010, University of Florida: Alachua County Public Works Department.

28. Barlaz, M.A., J.P. Chanton, and R.B. Green, *Controls on landfill gas collection efficiency: instantaneous and lifetime performance.* Journal of the Air & Waste Management Association, 2009. **59**(12): p. 1399-1404.

29. EPA, *Landfill Methane Outreach Program*. 2014, United States Enviromental Protection Agency.

30. Rasi, S., A. Veijanen, and J. Rintala, *Trace compounds of biogas from different biogas production plants.* Energy, 2007. **32**(8): p. 1375-1380.

31. Piechota, G., B. Igliński, and R. Buczkowski, *Development of measurement techniques for determination main and hazardous components in biogas utilised for energy purposes.* Energy Conversion and Management, 2013. **68**: p. 219-226.

32. Wheless, E. and J. Pierce. *Siloxanes in landfill and digester gas update*. in *Proceedings of the 27th SWANA Landfill Gas Symposium, San Antonio, TX, March*. 2004.

33. Schweigkofler, M. and R. Niessner, *Determination of siloxanes and VOC in landfill gas and sewage gas by canister sampling and GC-MS/AES analysis.* Environmental science & technology, 1999. **33**(20): p. 3680-3685.

34. Wellinger, A. and C. Mandereau, *Rationale for siloxane levels*, BioSurf, Editor. 2013, European Biogas Associastion.

35. Läntelä, J., S. Rasi, J. Lehtinen, and J. Rintala, *Landfill gas upgrading with pilot-scale water scrubber: Performance assessment with absorption water recycling.* Applied Energy, 2012. **92**: p. 307-314.

36. Zhang, S. and Z.C. Zhang, *Novel properties of ionic liquids in selective sulfur removal from fuels at room temperature.* Green Chemistry, 2002. **4**(4): p. 376-379.

37. Blumer, M., *Removal of elemental sulfur from hydrocarbon fractions.* Analytical Chemistry, 1957. **29**(7): p. 1039-1041.

38. Srivastava, V.C., *An evaluation of desulfurization technologies for sulfur removal from liquid fuels.* Rsc Advances, 2012. **2**(3): p. 759-783.

39. Samuels, A., *H sub 2 S removal system shows promise over iron sponge.* Oil and Gas Journal;(USA), 1990. **88**(6).

40. Nagl, G.J., *Removing H2S from gas streams.* Chemical Engineering, 2001. **108**(7): p. 97.

41. CFR, *40 CFR part 60 subpart WWW STANDARDS OF PERFORMANCE FOR MUNICIPAL SOLID WASTE LANDFILLS*, EPA, Editor. 1996.

42. EIA, *Voluntary Reporting of Greenhouse Gases Program Fuel Emission Coefficients*. 2011, U.S. Energy Information Association.

43. IPCC, *2.10.2 Direct Global Warming Potentials*. 2007, Intergovernmental Panel on Climate Change.

44. EPA, *Inventory of U.S. Greenhouse Gas Emissions and Sinks: 1990–2012*, U.S.E.P. Agency, Editor. 2014.

45. EPA. *Methane Emissions*. Emissions 2016 [cited 2016; Available from: http://www3.epa.gov/climatechange/ghgemissions/gases/ch4.html.

46. Spokas, K., J. Bogner, J.P. Chanton, M. Morcet, C. Aran, C. Graff, Y.M. Golvan, and I. Hebe, *Methane mass balance at three landfill sites: what is the efficiency of capture by gas collection systems?* Waste Manag, 2006. **26**(5): p. 516-25.

47. IPCC, *Climate Change 2014: Mitigation of Climate Change*. 2014, Intergovernmental Panel on Climate Change.

48. IPCC, *Chapter 2: Waste Generation, Composition and Managment Data*, in *2006 IPCC Guidelines for National Greenhouse Gas Inventories*, I.P.o.C. Change, Editor. 2006.

49. EPA. *Greenhouse Gas Equivalencies Calculator*. Energy and the Environment 2015 [cited 2016; Available from: https://www.epa.gov/energy/greenhouse-gas-equivalencies-calculator.

50. CFR, *40 FR Chapter 1 79.55 STATE OPERATING PERMIT PROGRAMS*, in *40*, EPA, Editor. 1997.

51. *Chapter 5: Incineration and Open Burning of Waste*, in *2006 IPCC Guidelines for National Greenhouse Gas Inventories*, I.P.o.C. Change, Editor. 2006.

52. Bove, R. and P. Lunghi, *Electric power generation from landfill gas using traditional and innovative technologies.* Energy Conversion and Management, 2006. **47**(11-12): p. 1391-1401.

53. EPA, *EPA LFG Energy Project Handbook*, E.P. Agency, Editor.

54. Jaramillo, P. and H.S. Matthews, *Landfill-gas-to-Energy projects: analysis of net private and social benefits.* Environ Sci Technol, 2005. **39**(19): p. 7365-73.

55. Pilarczyk, E., K.-D. Henning, and K. Knoblauch, *Natural gas from landfill gases.* Resources and conservation, 1987. **14**: p. 283-294.

56. Sabatier, P. and J. Senderens, *Hydrogenation of CO over nickel to produce methane.* J. Soc. Chim. Ind, 1902. **21**: p. 504-506.

57. Bartholomew, C.H. and R.J. Farrauto, *Fundamentals of Industrial Catalytic Processes*. 2nd ed. 2006, Hoboken, NJ: John Wiley & Sons.

58. Van Berge, P. and R. Everson. *Cobalt as an alternative Fischer-Tropsch catalyst to iron for the production of middle distillates*. in *Fuel and Energy Abstracts*. 1997.

59. Raje, A.P., S.J. Liaw, R. Srinivas, and B.H. Davis, *Second row transition metal sulfides for the hydrotreatment of coal-derived naphtha I. Catalyst preparation, characterization and comparison of rate of simultaneous removal of total sulfur, nitrogen and oxygen.* Applied Catalysis A: General, 1997. **150**: p. 297-318.

60. Bezemer, G.L., J.H. Bitter, H.P. Kuipers, H. Oosterbeek, J.E. Holewijn, X. Xu, F. Kapteijn, A.J. van Dillen, and K.P. de Jong, *Cobalt particle size effects in the Fischer-Tropsch reaction studied with carbon nanofiber supported catalysts.* Journal of the American Chemical Society, 2006. **128**(12): p. 3956-3964.

61. Iglesia, E., *Design, synthesis, and use of cobalt-based Fischer-Tropsch synthesis catalysts.* Applied Catalysis A: General, 1997. **161**(1): p. 59-78.

62. Bessell, S., *Support effects in cobalt-based fischer-tropsch catalysis.* Applied Catalysis A: General, 1993. **96**(2): p. 253-268.

63. Gardezi, S.A., B. Joseph, F. Prado, and A. Barbosa, *Thermochemical biomass to liquid (BTL) process: Bench-scale experimental results and projected process economics of a commercial scale process.* Biomass & Bioenergy, 2013. **59**: p. 168-186.

64. Gardezi, S.A.Z., J.T. Wolan, and B. Joseph, *Technical Presentation: An Integrated Approach to the preparation of effective Catalyst for Biomass-to-Liquid (BTL) process* in *Department of Chemical & Biomedical Engineering*. USF Tampa.

65. Gardezi, S.A., L. Landrigan, B. Joseph, and J.T. Wolan, *Synthesis of Tailored Eggshell Cobalt Catalysts for Fischer–Tropsch Synthesis Using Wet Chemistry Techniques.* Industrial & Engineering Chemistry Research, 2012. **51**(4): p. 1703-1712.

66. Gardezi, S.A., J.T. Wolan, and B. Joseph, *Effect of catalyst preparation conditions on the performance of eggshell cobalt/SiO2 catalysts for Fischer-Tropsch synthesis.* Applied Catalysis a-General, 2012. **447**: p. 151-163.

67. Rostrup-Nielsen, J.R., *Catalytic steam reforming*. 1984: Springer.

68. Rostrup-Nielsen, J.R. and J. Sehested, *Hydrogen and Synthesis Gas by Steam- and CO2 Reforming.* Advances in Catalysis, 2002. **47**: p. 65.

69. Sukonket, T., A. Khan, B. Saha, H. Ibrahim, S. Tantayanon, P. Kumar, and R. Idem, *Influence of the catalyst preparation method, surfactant amount, and steam on CO2 reforming of CH4 over 5Ni/Ce0. 6Zr0. 4O2 catalysts.* Energy & Fuels, 2011. **25**(3): p. 864-877.

70. Cho, W.J., T.Y. Song, A. Mitsos, J.T. McKinnon, G.H. Ko, J.E. Tolsma, D. Denholm, and T. Park, *Optimal design and operation of a natural gas tri-reforming reactor for DME synthesis.* Catalysis Today, 2009. **139**(4): p. 261-267.

71. Song, C. and W. Pan, *Tri-reforming of methane: a novel concept for catalytic production of industrially useful synthesis gas with desired H 2/CO ratios.* Catalysis Today, 2004. **98**(4): p. 463-484.

72. Farniaei, M., H. Rahnama, M. Abbasi, and M.R. Rahimpour, *Simultaneous production of two types of synthesis gas by steam and tri-reforming of methane using an integrated thermally coupled reactor: mathematical modeling.* International Journal of Energy Research, 2014. **38**(10): p. 1260-1277.

73. Rostrup-Nielsen, J.R., *Activity of nickel catalysts for steam reforming of hydrocarbons.* Journal of Catalysis, 1973. **31**(2): p. 173-199.

74. Pakulska, M.M., C.M. Grgicak, and J.B. Giorgi, *The effect of metal and support particle size on NiO/CeO2 and NiO/ZrO2 catalyst activity in complete methane oxidation.* Appl. Catal. A: Gen., 2007. **332**: p. 124-129.

75. Goguet, A., F. Meunier, J.P. Breen, R. Burch, M.I. Petch, and A.F. Ghenciu, *Study of the origin of the deactivation of a Pt/CeO2 catalyst during reverse water gas shift (RWGS) reaction.* J. Catal., 2004. **226**: p. 382-392.

76. Kobune, M., S. Sato, and R. Takahashi, *Surface-structure sensitivity of CeO2 for several catalytic reactions.* J. Mol. Catal. A: Chem., 2008. **279**: p. 10-19.

77. Vita, A., G. Cristiano, C. Italiano, L. Pino, and S. Specchia, *Syngas production by methane oxy-steam reforming on Me/CeO2 (Me=Rh, Pt, Ni) catalyst lined on cordierite monoliths.* Applied Catalysis B: Environmental, 2015. **162**: p. 551-563.

78. Zhang, R., Y. Wang, and R.C. Brown, *Steam reforming of tar compounds over Ni/olivine catalysts doped with* CeO_2. Energy Conversion & Management, 2007. **48**: p. 68.

79. Dong, W.-S., K.-W. Jun, H.-S. Roh, Z.-W. Liu, and S.-E. Park, *Comparative study on partial oxidation of methane over Ni/ZrO2, Ni/CeO2 and Ni/Ce–ZrO2 catalysts.* Catalysis Letters, 2002. **78**(1/4): p. 215-222.

80. Itome, M. and A.E. Nelson, *Methane oxidation over M–8YSZ and M–CeO2/8YSZ (M=Ni, Cu, Co, Ag) catalysts.* Catalysis Letters, 2006. **106**(1-2): p. 21.

81. Kumar, P., Y. Sun, and R.O. Idem, *Nickel-Based Ceria, Zirconia, and Ceria–Zirconia Catalytic Systems for Low-Temperature Carbon Dioxide Reforming of Methane.* Energy & Fuels, 2007. **21**(6): p. 3113-3123.

82. Guo, Y. and Z. Dong, *Well-dispersed Ceria-promoted Sulfated Zirconia Supported on Mesoporous Silica.* Catal. Lett., 2007. **119**: p. 120-125.

83. Chen, X.R., Y.H. Ju, and C.Y. Mou, *Direct synthesis of mesoporous sulfated silica-zirconia catalysts with high catalytic activity for biodiesel via esterification.* Journal of Physical Chemistry C, 2007. **111**(50): p. 18731-18737.

84. Dajiang, M., C. Yaoqiang, Z. Junbo, W. Zhenling, M. Di, and G. Maochu, *Catalytic Partial Oxidation of Methane over Ni/Ce02 - ZrO2-Al203.* JOURNAL OF RARE EARTHS, 2007. **25**: p. 311-315.

85. Youn, M.H., J.G. Seo, K.M. Cho, S. Park, D.R. Park, J.C. Jung, and I.K. Song, *Hydrogen production by auto-thermal reforming of ethanol over nickel catalysts supported on Ce-modified mesoporous zirconia: Effect of Ce/Zr molar ratio.* International journal of hydrogen energy, 2008. **33**(19): p. 5052-5059.

86. Laosiripojana, N. and S. Assabumrungrat, *Methane steam reforming over Ni/Ce–ZrO 2 catalyst: Influences of Ce–ZrO 2 support on reactivity, resistance toward carbon formation, and intrinsic reaction kinetics.* Applied Catalysis A: General, 2005. **290**(1): p. 200-211.

87. Noronha, F.B., E.C. Fendley, R.R. Soares, W.E. Alvarez, and D.E. Resasco, *Correlation between catalytic activity and support reducibility in the CO 2 reforming of methane over Pt/Ce x Zr 1− x O 2 catalysts.* Chemical Engineering Journal, 2001. **82**(1): p. 21-31.

88. Somorjai, G.A. and Y. Li, *Introduction to Surface Chemistry and Catalysis*. 2 ed. 2010, New Jersey: John Wiley & Sons, Inc.

89. Christian Enger, B., R. Lødeng, and A. Holmen, *A review of catalytic partial oxidation of methane to synthesis gas with emphasis on reaction mechanisms over transition metal catalysts.* Applied Catalysis A: General, 2008. **346**(1-2): p. 1-27.

90. Bengaard, H.S., J.K. Nørskov, J. Sehested, B.S. Clausen, L.P. Nielsen, A.M. Molenbroek, and J.R. Rostrup-Nielsen, *Steam Reforming and Graphite Formation on Ni Catalysts.* Journal of Catalysis, 2002. **209**: p. 365.

91. Czekaj, I., F. Loviat, F. Raimondi, J. Wambach, S. Biollaz, and A. Wokaun, *Characterization of surface processes at the Ni-based catalyst during the methanation of biomass-derived synthesis gas: X-ray photoelectron spectroscopy (XPS).* Applied Catalysis A: General, 2007. **329**: p. 68-78.

92. Lu, Y., J.Z. Xue, C.C. Yu, Y. Liu, and S.K. Shen, *Mechanistic investigations on the partial oxidation of methane to synthesis gas over a nickel-on-alumina catalyst.* Applied Catalysis a-General, 1998. **174**(1-2): p. 121-128.

93. Diskin, A.M., R.H. Cunningham, and R.M. Ormerod, *The oxidative chemistry of methane over supported nickel catalysts.* Catal. Tod., 1998. **46**: p. 147-154.

94. Walker, D.M., S.L. Pettit, J.T. Wolan, and J.N. Kuhn, *Synthesis gas production to desired hydrogen to carbon monoxide ratios by tri-reforming of methane using Ni–MgO–(Ce,Zr)O2 catalysts.* Applied Catalysis A: General, 2012. **445-446**: p. 61-68.

95. Rakopoulos, C., K. Antonopoulos, D. Rakopoulos, D. Hountalas, and E. Giakoumis, *Comparative performance and emissions study of a direct injection diesel engine using blends of diesel fuel with vegetable oils or bio-diesels of various origins.* Energy conversion and management, 2006. **47**(18): p. 3272-3287.

96. Demirbas, A., *Progress and recent trends in biofuels.* Progress in energy and combustion science, 2007. **33**(1): p. 1-18.

97. Thomas, C.S., *Transportation options in a carbon-constrained world: Hybrids, plug-in hybrids, biofuels, fuel cell electric vehicles, and battery electric vehicles.* International Journal of Hydrogen Energy, 2009. **34**(23): p. 9279-9296.

98. Erwin, J., *Assay of diesel fuel components properties and performance* in *Symposium on processing and produt seletivityof synthetic fuels*. 1992, American Chemical Society: Washington D.C.

99. Wheless, E. and J. Pierce, *Siloxanes in Landfill and Digester Gas Update*. 2003, SCS Engineers.

100. Heguy, D. and J. Bogner, *Cost-Effective Hydrogen Sulfide Treatment Strategies for Commercial Landfill Gas Recovery*. 2015, Merichem Company.

101. EPA. *Landfill Methane Outreach Project: Operational Landfills Data*. 2016 1/20/2016 [cited 2016; Available from: http://www3.epa.gov/lmop/projects-candidates/operational.html.

102. Turton, R., R.C. Bailie, W.B. Whiting, J.A. Shaeiwitz, and D. Bhattacharyya, *Analysis, Synthesis, and Design of Chemical Processes*. 4th ed. 2013: Pearson.

103. DOE, *Alternative Fuels Comparison Chart*, A.F.D. Center, Editor. 2015, Department of Energy.

104. Prometheus_Energy, *LNG Quick Facts Sheet*. 2015.

105. NIST, *Report of the 78th National Conference on Weights and measures*, in *NIST Special Publication*. 1993. p. pp 322-326.

106. ORNL, *Heating Values of Fuels*. 2011, Oak Ridge National Labs: Biomass Energy Data Book.

107. EIA. *Natural Gas Spot Prices : NYMEX*. 2016 [cited 2016 2/9/2016]; Available from: https://www.eia.gov/dnav/ng/ng_pri_fut_s1_w.htm.

Appendices

Appendix A. EPA LFG Calculation (Average Landfill Size)

Calculations for average LFG production were made using data provided by the EPA Landfill Methane Outreach Project. As of March 2015 the EPA published that 645 projects belong to 595 unique landfill sites. These numbers are reported by the EPA as such

> "The information contained in the LMOP database is compiled from a variety of sources, including annual voluntary submissions by LMOP Partners and industry publications. Due to the limitations of voluntary reporting, LMOP cannot guarantee the accuracy of these data."[101]

All numbers reported units of million standard cubic feet a day (mmscfd) or standard cubic feet per minute (scfm). All of these values are based on unique, operational, and reporting landfills. Landfills that do not provide gas data are not included.

Table A: Landfill Operation Statistics

Number of Unique, Operational Landfills reporting LFG collection	514
Total LFG Collection (mmscfd)	1460
Average LFG Flow Rate (mmscfd)	2.84013
Avg. LFG Flow Rate (scfm)	1972
Avg. LFG Flared (scfm)	483

$$\frac{mscf}{day} * \frac{1\ day}{1440\ min} * \frac{10^6 scf}{1\ mscf} \qquad 1\ mscfd = 694.44\ scfm$$

Appendix B. Calculation of Landfill Gas Price

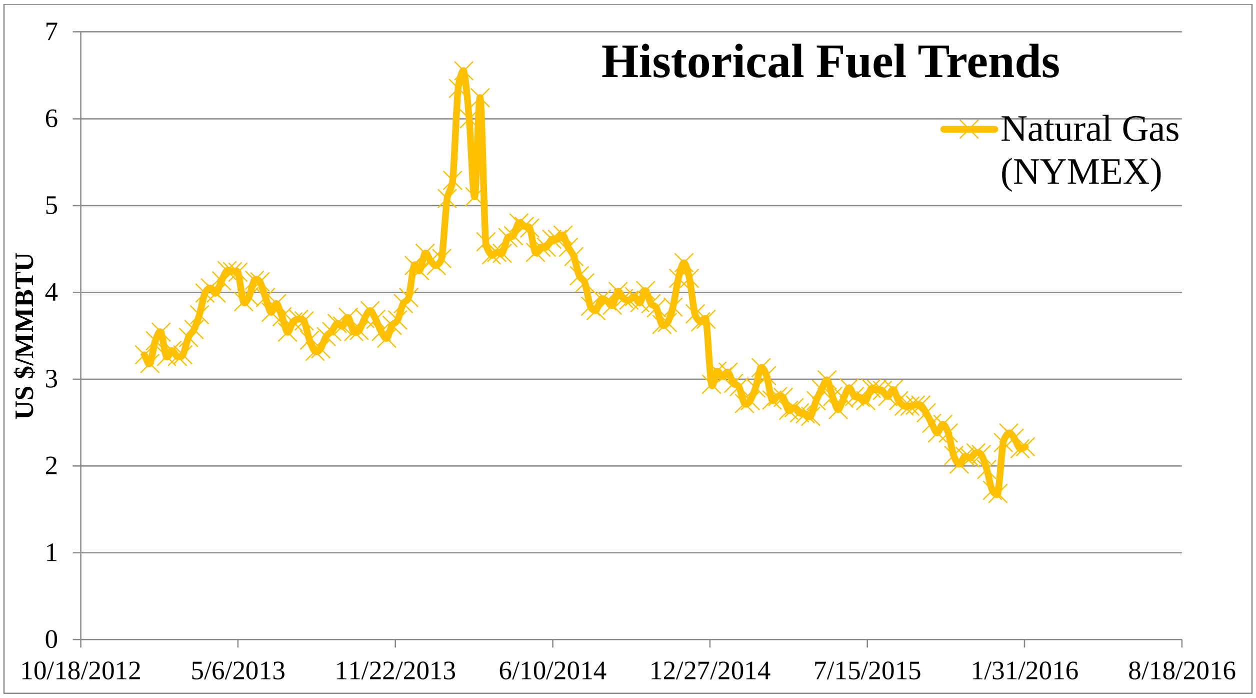


Figure A: Natural Gas Price History

Landfill gas is typically sold at 40-60% the cost of Open market Natural Gas prices. It would seem that a usable number would be 50%, as that is close to the composition of methane in LFG, however 45% was chosen in order to use a scenario in which the LFG was heavily contaminated and would incur additional costs to users.

- Methane content of LFG is 55%.
- Methane heat content is 1,012 Btu/scf methane

$$\frac{1012\ Btu}{scf\ methane} * \frac{0.55\ scf\ methane}{1\ scf\ LFG} * \frac{1\ MMBTU}{10^6\ BTU} = \frac{0.000557\ MMBTU}{scf\ LFG}$$

- At the time of this plant design natural gas was at ~3 \$/MMBTU.
- LFG is 45% of Natural gas cost

$$\frac{3\ \$}{MMBTU\ NG} * \frac{0.45\ \$\ LFG}{1\ \$\ NG} = 1.35\ \frac{\$}{MMBTU\ LFG}$$

$$\frac{1.35\ \$}{MMBTU\ LFG} * \frac{0.000557\ MMBTU}{scf\ LFG} = 0.000752\frac{\$}{scf\ LFG}$$

$$2500\frac{scf}{min} *\ 0.000752\frac{\$}{scf\ LFG} =\ 1.88\frac{\$}{min}\ \ (\$2700\ per\ day)$$

- Using an operating factor of 85% (Net capacity factor, for gas flow variability and parasitic loses)

$$2700\frac{\$}{day} *\ 0.85\ * 365\frac{\text{days}}{\text{yr}} = \sim 0.84\ Million\ USD\ \$$$

Appendix C. Cost of Labor

To estimate the cost of labor the number of operators was calculated using the following equations which can be found in Turton et al. [102]

$$N_{OL}= (6.29 + 31.7P^2 + 0.23N_{NP})^{0.5}$$

P is the number of processing steps involving the handling of particulate solids. N_{NP} is the number of nonparticulate processing steps or N_{NP}= sum (Equipment). N_{OL} is the number of operators per shift.

$$N_{OL}= (6.29 + 31.7(0)^2 + 0.23*(15))^{0.5}$$

$$N_{OL}=3.12 \text{ operators per shift}$$

Due to relativity increased level of automation, in part due to the difficulty of running FTS reactors, the plant is very automated, allowing a decreased cost of typical operators to 50% normal.

$$N_{OL}=1.56 \text{ operators per shift}$$

Assuming the plant is operating 365 days a year with 5 eight hour shifts per week per operator you will need 1095 shifts per year with 245 shifts per operator year.

$$Operators = 1.56 * \frac{1095}{245} = 7 \text{ operators}$$

Labor Cost = (7 operators x \$59,580) = \$417,061

Appendix D. Calculation of Gallon Gasoline Equivalency

The values listed below were compiled from a number of sources on the energy density of common fuels. The energy density of NG is greater per pound, however its density is relativity low, giving a fraction of the energy when its volume is the determining factor.

Table B: Gallon Gasoline Equivalency

Fuel	**Energy Content**	
Low-Sulfur Highway Diesel	138, 490 BTU/gal	18,320 BTU/lb
Conventional Gasoline	120,388 BTU/gal	18,679 BTU/lb
Natural Gas (compressed)	----	20,267 BTU/lb
Natural Gas (Liquefied)	82,644 BTU/gal	20,908 BTU/lb

[103-106]

At the selling prices used in this paper above. The price of natural gas used for this analysis is a long term average of the cost and is not directly reflected by current prices.

Diesel is $2.73 a gallon, and gasoline is $2.1840 per gallon.

$$138490 \frac{BTU}{gal} * \frac{1}{2.73} \frac{gal}{\$} = 49110 \frac{BTU}{\$} \quad (1)$$

Table C: BTU/$

Fuel	BTU/$	$/MMBTU
Low-Sulfur Highway Diesel	50,736	19.71
Conventional Gasoline	55,224	18.1
Natural Gas	333,333	3
Landfill Gas (45% NYMEX)	740,741	1.35

Appendix E. LFGTL Energy Distribution

The energy flow in can be calculated by using 557 BTU/SCFM of LFG.

$$2500\,SCFM * 557\frac{BTU}{SCFM} * \frac{1}{10^6}\frac{MMBTU}{BTU} = 1.39\frac{MMBTU}{min}$$

$$(83.55\frac{MMBTU}{hr})(\,713{,}851\frac{MMBTU}{yr}\,)$$

The energy out can be calculated using BTU/gal numbers from Table B.

$$138490\frac{BTU}{gal\ diesel} * 240\frac{gal\ diesel}{hr} * \frac{1}{10^6}\frac{MMBTU}{BTU} = 33.24\frac{MMBTU}{hr}$$

$$120388\frac{BTU}{gal\ gas} * 90\frac{gal\ gas}{hr} * \frac{1}{10^6}\frac{MMBTU}{BTU} = 10.83\frac{MMBTU}{hr}$$

$$(\,44.07\frac{MMBTU}{hr}\ combined)(\,376{,}576\ \frac{MMBTU}{yr})$$

Table D: Energy Flow in LFGTL

Section	**Type**	**MMBTU/hr**
Landfill Gas	Stream (Energy In)	83.55
Fuel Production	Stream (Energy Out)	-40.115
Heat Leaving	Heat (Energy Out)	-43.4
Furnace Usage	Heat (Generation)	43
Tri-Reformer (MTR Reactor)	Heat (Consumption)	-22.15
Fired Heater	Heat (Consumption)	-18.1
Fischer Tropsch Reactor	Heat (Generation)	23.1
Driver Power	Electricity (Consumption)	-2

Appendix F. Reactor Outputs

Table E: Fischer Tropsch Reactor Output

Chemical	**Mass %**
Hydrogen	0.007606
Water	0.174952
Carbon Monoxide	0.110269
Carbon Dioxide	0.519313
Methane	0.008907
Propane	0.007098
Ethane	0.00852
Propene	0.006302
N-Butane	0.01469
I-Pentane	0.014507
N-Hexane	0.013956
N-Heptane	0.011542
N-Octane	0.010405
N-Nonane	0.009626
N-Decane	0.008035
N-Undecane	0.006831
N-Dodecane	0.005432
N-Tridecane	0.005008
N-Tetradecane	0.004452
N-Pentadecane	0.003638
N-Hexadecane	0.003209
N-Heptadecane	0.002698
N-Nonadecane	0.001954
N-Eicosane	0.001586
N-Octadecane	0.001168
NH3	0
Oxygen	0
Nitrogen	0.018774
Ethylene	0.015628
	0.000981
docosane	0.00088
N-Tricosane	0.000728
N-Tetracosane	0.00068
N-pentacosane	0.000625

Table F: MTR Reactor Output

Chemical	**Mass %**
Hydrogen	0.07921232
Water	0.001850625
Carbon Monoxide	0.5482671
Carbon Dioxide	0.3556202
Methane	0.001786542
Propane	9.999407e-007
Ethane	1.126095e-006
Propene	1.288773e-005
N-Butane	1.149542e-006
I-Pentane	1.67676e-005
N-Hexane	1.157184e-005
N-Heptane	4.705648e-006
N-Octane	1.58357e-006
N-Nonane	5.527469e-007
N-Decane	1.745998e-007
N-Undecane	4.591474e-008
N-Dodecane	1.2804e-008
N-Tridecane	3.271031e-009
N-Tetradecane	1.419595e-009
N-Pentadecane	4.410317e-010
N-Hexadecane	1.224802e-010
N-Heptadecane	6.03382e-011
N-Nonadecane	9.854508e-012
N-Eicosane	2.85634e-012
N-Octadecane	5.804425e-013
NH3	0
Oxygen	0
Nitrogen	0.01320931
Ethylene	2.305658e-006
	1.911166e-013
docosane	7.06271e-014
N-Tricosane	2.084904e-014
N-Tetracosane	5.056259e-015
N-pentacosane	1.740997e-015

Appendix G. Chemcad Plant Design

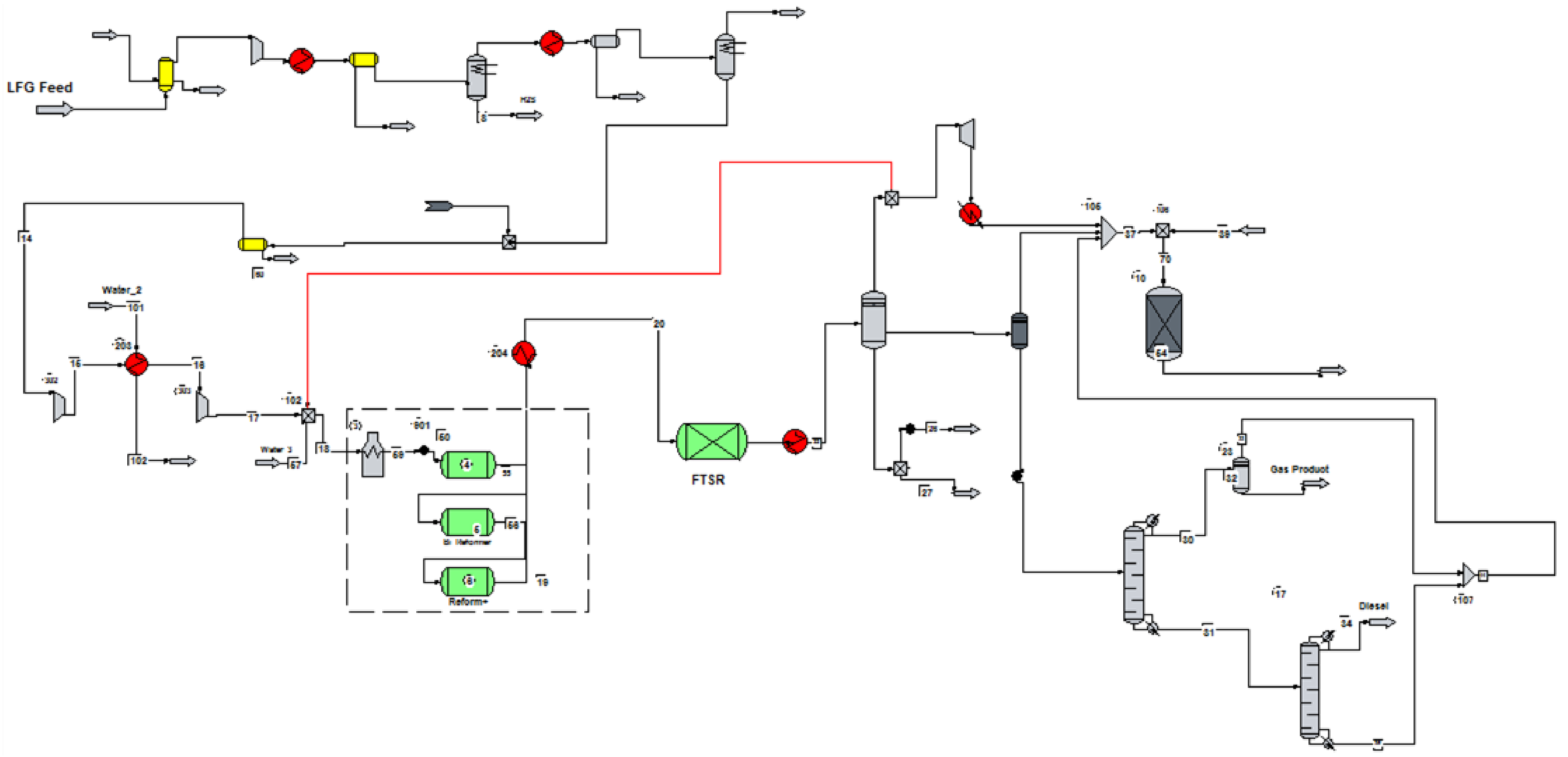


Figure B: ChemCAD CFD

Appendix H. Composite Cost of a Gallon of Gasoline

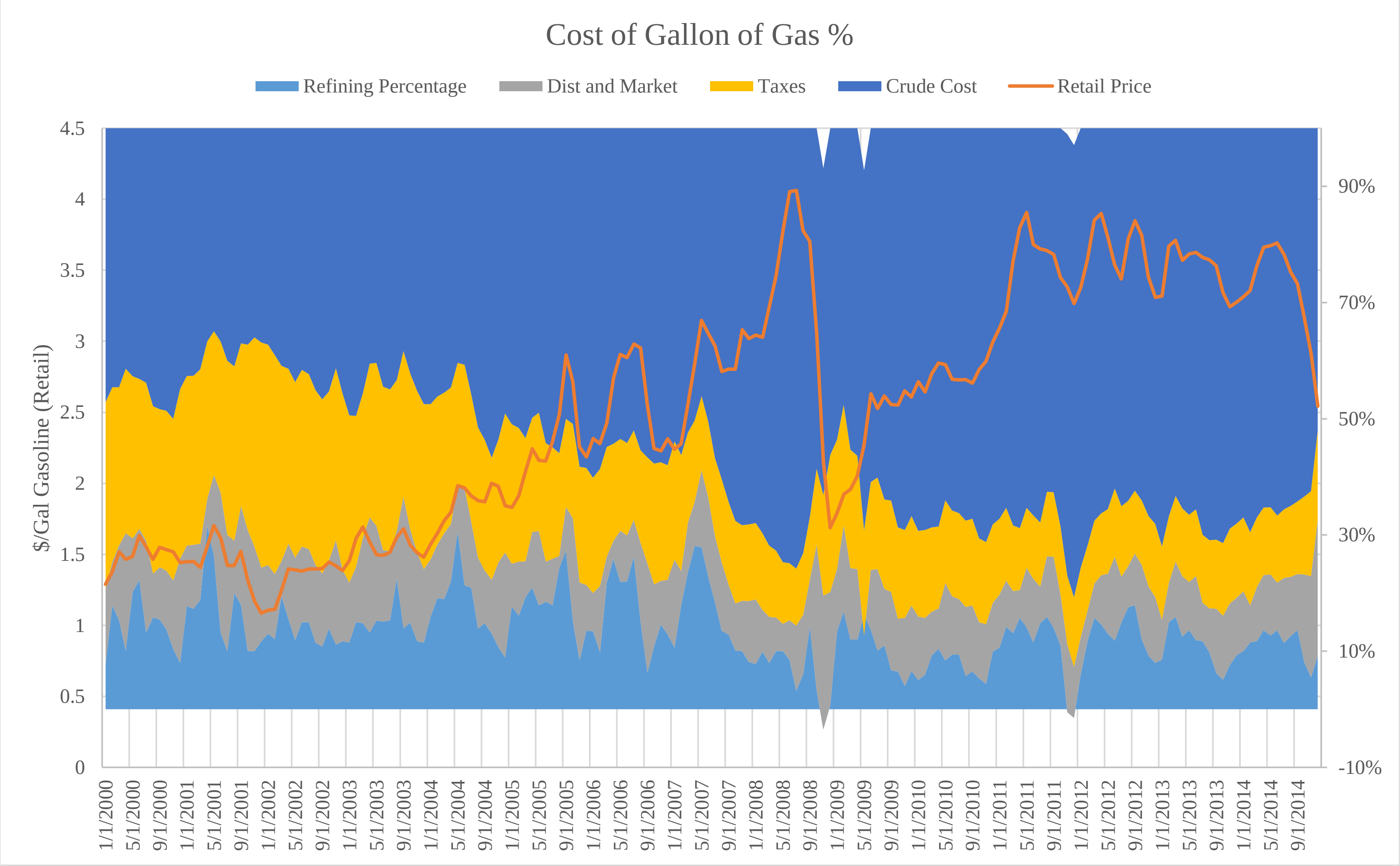


Figure C: Gas Price Breakdown

Appendix I. Fossil Fuels Market Prices

Data is compiled from EIA sources on weekly fuel price updates on a national average[107]

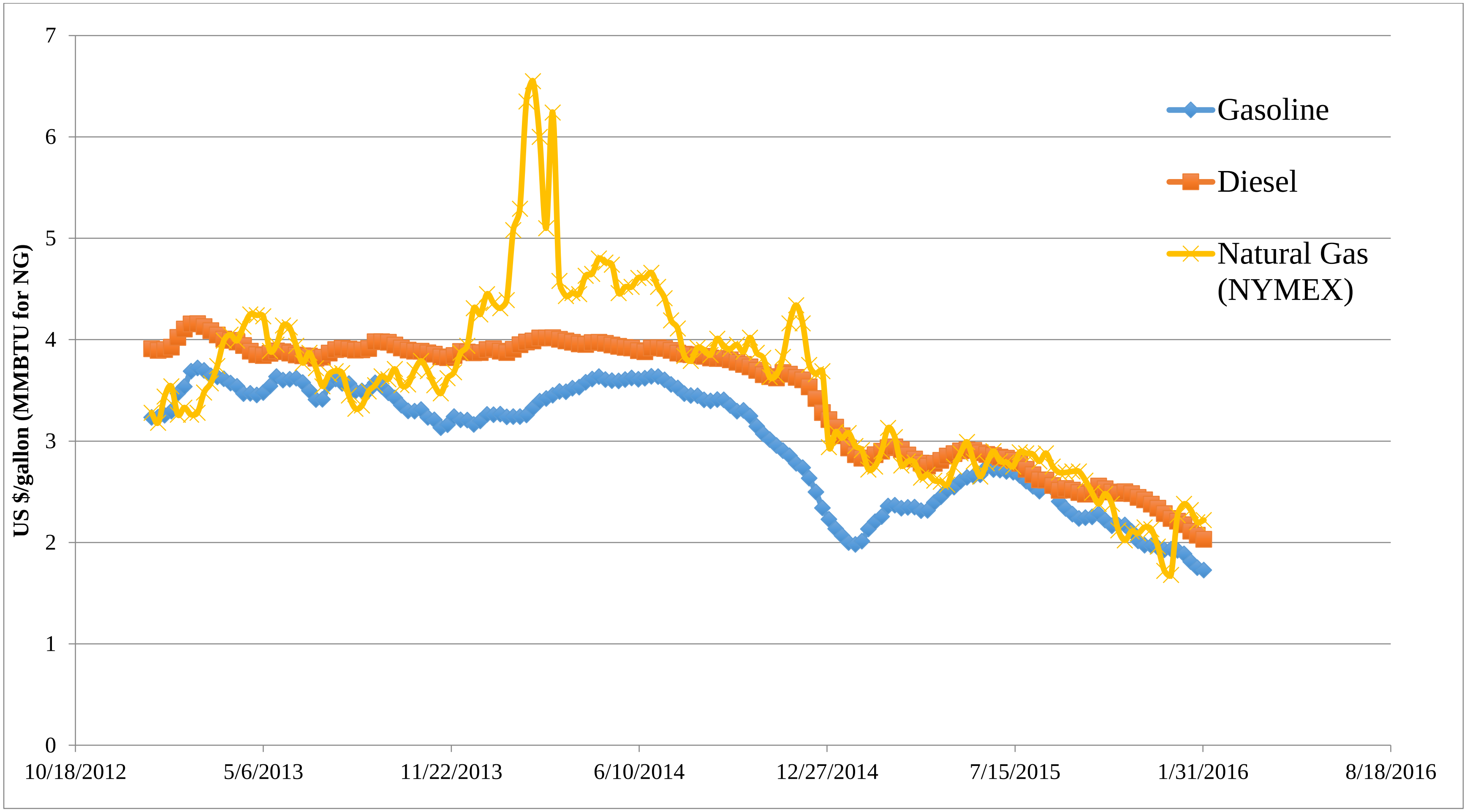
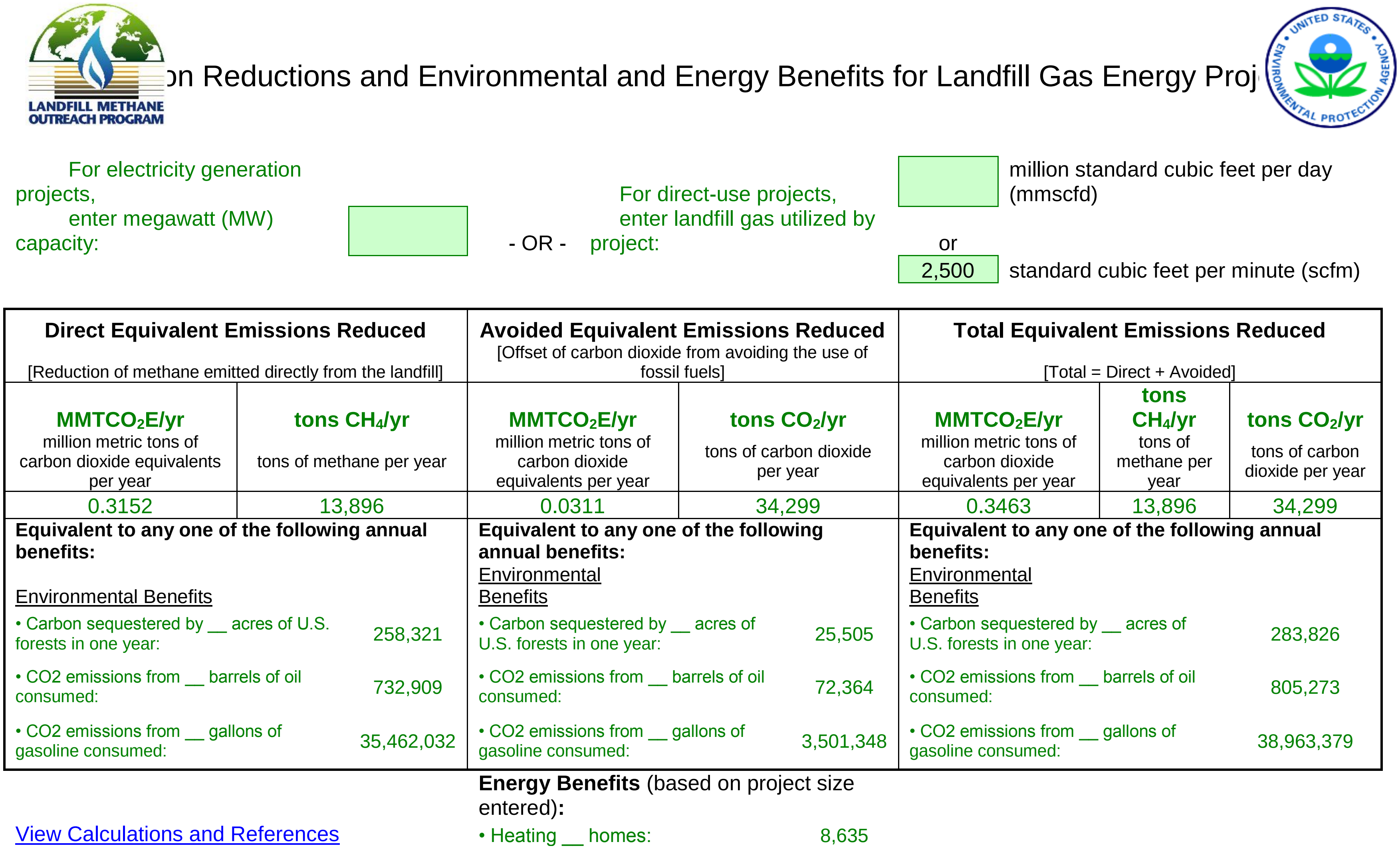


Figure D: Fuel Price History

Appendix J. EPA LFG Project Calculator Results

Table G: EPA Calculator



LANDFILL METHANE OUTREACH PROGRAM

on Reductions and Environmental and Energy Benefits for Landfill Gas Energy Proj

For electricity generation projects, enter megawatt (MW) capacity: - OR - For direct-use projects, enter landfill gas utilized by project:

million standard cubic feet per day (mmscfd)

or

2,500 standard cubic feet per minute (scfm)

Direct Equivalent Emissions Reduced [Reduction of methane emitted directly from the landfill]		**Avoided Equivalent Emissions Reduced** [Offset of carbon dioxide from avoiding the use of fossil fuels]		**Total Equivalent Emissions Reduced** [Total = Direct + Avoided]		
MMTCO$_2$E/yr million metric tons of carbon dioxide equivalents per year	**tons CH$_4$/yr** tons of methane per year	**MMTCO$_2$E/yr** million metric tons of carbon dioxide equivalents per year	**tons CO$_2$/yr** tons of carbon dioxide per year	**MMTCO$_2$E/yr** million metric tons of carbon dioxide equivalents per year	**tons CH$_4$/yr** tons of methane per year	**tons CO$_2$/yr** tons of carbon dioxide per year
0.3152	13,896	0.0311	34,299	0.3463	13,896	34,299
Equivalent to any one of the following annual benefits: Environmental Benefits		**Equivalent to any one of the following annual benefits:** Environmental Benefits		**Equivalent to any one of the following annual benefits:** Environmental Benefits		
• Carbon sequestered by __ acres of U.S. forests in one year:	258,321	• Carbon sequestered by __ acres of U.S. forests in one year:	25,505	• Carbon sequestered by __ acres of U.S. forests in one year:		283,826
• CO2 emissions from __ barrels of oil consumed:	732,909	• CO2 emissions from __ barrels of oil consumed:	72,364	• CO2 emissions from __ barrels of oil consumed:		805,273
• CO2 emissions from __ gallons of gasoline consumed:	35,462,032	• CO2 emissions from __ gallons of gasoline consumed:	3,501,348	• CO2 emissions from __ gallons of gasoline consumed:		38,963,379

Energy Benefits (based on project size entered)**:**

• Heating __ homes: 8,635

View Calculations and References

Appendix K. Electrical Turbine Economics Output

STANDARD TURBINE-GENERATOR SET

Cost Component	Cost (2008$'s)	Cost Unit
Installed cost of gas compression/treatment, turbine/generator, site work, and housings	$2,340(x) - 0.103($x^2$)	$, x = kW capacity [$1,015/kW for ($2,340 - 0.103/kW capacity) < 1,015
Installed cost of electrical interconnect equipment	$250,000	per system
Annual O&M of compression/treatment and turbine/generator (excluding energy)	$0.0144	per kWh generated

Project Component	Quantity
Gross capacity factor (%)	93%
System operating schedule (hours/year)	8,147
Fuel use rate (Btu/kWh generated)	13,000
Parasitic loss efficiency (%)	88%
Landfill gas heat content (Btu/ft^3)	557
Turbine capacity (kW)	6,885

Installed Capital Costs:	2015
Gas Compression/Treatment, Turbine/Generator, Site Work, and Housings:	$9,228,104
Electrical Interconnect Equipment:	$250,000
Total Capital Costs Including Cost Contingency	$9,478,104

CHP TURBINE-GENERATOR SET

Cost Component	Cost (2008$'s)	Cost Unit
Installed cost of gas compression/treatment, turbine/generator, site work, and housings	$\$2,340(x) - 0.103(x^2)$	$, x = kW capacity [$1,370/kW for ($2,340 - 0.103(x)) < $1,370]
Installed cost of heat exchangers	$355(x)	$, x = kW capacity
Installed cost of electrical interconnect equipment	$250,000	per system
Installed cost of gas pipeline	$63	per ft
Installed cost of steam pipelines (assumes 2 lines for supply and return)	$106	per ft of trench
Installed cost of circulation pump	$12,000	per system
Annual O&M of compression/treatment, turbine/generator, and exchangers (excluding energy)	$0.0144	per kWh generated

Project Component	Quantity
Gross capacity factor (%)	93%
System operating schedule (hours/year)	8,147
Fuel use rate (Btu/kWh generated)	13,000
Steam production (Btu/kWh, net)	5,500
Parasitic loss efficiency (%)	88%
Landfill gas heat content (Btu/ft^3)	557
Turbine capacity (kW)	6,885

Installed Capital Costs:	2015
Gas Compression/Treatment, Turbine/Generator, Site Work, and Housings:	$9,228,104
Heat Recovery Exchangers:	$2,444,073
Electrical Interconnect Equipment:	$250,000
Gas Pipeline:	$0
Steam Pipelines and Circulation Pump:	$12,000
Total Capital Costs Including Cost Contingency	$10,934,177

Cap and Flaring System

Cost Component	Cost (2013$'s)
Drilling and pipe crew mobilization	$20,000
Installed cost of vertical gas extraction wells	$4,675
Installed cost of wellheads and pipe gathering system	$17,000
Installed cost of knockout, blower, and flare system	$(x)^{0.61}$ * $4,600
Engineering, permitting, and surveying	$700
Annual O&M for collection (excluding energy)	$2,600
Annual O&M for flare (excluding electricity)	$5,100
Electricity price (depends on type of project)	$0.065

Project Component	Quantity
Average depth of landfill waste (ft)	65
Number of wells (1 well per acre)	100
Number of flares (1 flare per system)	1
Collected landfill gas design flow rate (ft^3/min)	2,941
Electricity usage by blowers (kWh/ft^3)	0.002

Installed Capital Costs:	2015
Mobilization:	$20,000
Extraction Wells:	$467,500
Wellheads and Pipe Gathering System:	$1,700,000
Knockout, Blower, and Flare System:	$600,559
Engineering, Permitting, and Surveying:	$70,000
Total Capital Costs Including Cost Contingency	$2,858,059

Annual Costs:

Year	2016
O&M for Collection	$260,000
O&M for Flare	$5,100
Electricity	$200,965
Total Annual Costs Including Cost Contingency	$466,065