NATIONAL ENERGY TECHNOLOGY LABORATORY

Analysis of Natural Gas-to Liquid Transportation Fuels via Fischer-Tropsch

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Acronyms and Abbreviations

AGR	Acid gas recovery
API	American Petroleum Institute
ASU	Air separation unit
ATR	Autothermal Reformer
BACT	Best available control technology
BB	Bituminous Baseline Report
bbl	Barrels
bbl_{ECO}	Equivalent crude oil barrels
bbl_{EPD}	Equivalent petroleum diesel barrels
bbl_{FTL}	Fischer-Tropsch barrels
BBR	Bituminous baseline report
BOP	Balance of plant
bpd	Barrels per day
BTS	Bureau of Transportation Statistics
Btu/kWhr	British thermal units per kilowatt hour
Btu/lb	British thermal units per pound
CCS	Carbon capture and sequestration
CEPCI	Chemical Engineering Plant Cost Index
CF	Capacity Factor
CFB	Circulating fluid bed
CM	Construction Management
CO	Carbon monoxide
Co	Cobalt
CO_2	Carbon dioxide
CO_2e	Carbon dioxide equivalents
COE	Cost of electricity
COP	Cost of production
COP’	Crude oil price
COS	Carbonyl sulfide
CTL	Coal-to-liquids
DOE	Department of Energy
ECO	Equivalent crude oil
eGRID	Emissions & Generation Resource Integrated Database
EIA	Energy Information Administration
EPA	Environmental Protection Agency
EPD	Equivalent petroleum diesel
ESPA	Energy Sector Planning & Analysis
Fe	Iron
FERC	Federal Energy Regulatory Commission
FG	Fuel gas
F-T	Fischer-Tropsch
ft	Foot, Feet
ft/s	Feet per second
GHG	Greenhouse gas
GHGEV	Greenhouse gas emission value
GJ/hr	Gigajoules per hour
GPM	Gallons per minute
GTL	Gas-to-liquid
GWP	Global warming potential
H_2	Hydrogen
HHV	Higher heating value
H.O.	Home Office
HP	High pressure
hp	Horsepower
HRSG	Heat recovery steam generator
HTFT	High temperature Fischer-Tropsch
I&C	Instrumentation & Controls
IGCC	Integrated gasification combined cycle
IPCC	Intergovernmental Panel on Climate Change
IRR	Internal rate of return
IRROE	Internal Rate of Return on Equity
ISO	International Standards Organization
kg/hr	Kilograms per hour
kJ/kWhr	Kilojoules per kilowatt hour
kJ/kg	Kilojoules per kilogram
kV	Kilovolt
kW, kWe	Kilowatt electric
kWth	Kilowatt thermal
lb/hr	Pounds per hour
lbmol/hr	Pound moles per hour
LCA	Life cycle analysis
LCI	Life cycle inventory
LCIA	Life cycle impact assessment
LHV	Lower heating value
LTFT	Low temperature Fischer-Tropsch
m^3/min	Cubic meters per minute
m/s	meters/second
MEA	Monoethanolamine
MMbpd	Million barrels per day
MMBTU	Million British thermal units
MMBtu/hr	Million British thermal units per hour
MOVES	Motor Vehicle Emission Simulator
MPa	Megapascal
MW	Megawatt

N/A	Not applicable
NETL	National Energy Technology Laboratory
NG	Natural Gas
NGL	Natural gas liquids
NGCC	Natural gasification combined cycle
NMVOC	Non-methane volatile organic compounds
NSPS	New Source Performance Standards
O-H	Over head
O&M	Operation and maintenance
OECD	Organization for Economic Co-operation and Development
O.J.	Operator jobs
O/U/E	Offsites, utilities, and electrical
PDGP	Petroleum derived gasoline price
PM	Particulate matter
POTW	Publicly owned treatment works
PSA	Pressure swing adsorber
psia	Pound per square inch absolute
PT	Product transport
RMA	Raw material acquisition
RMT	Raw material transport
SGP	Shell Gasification Process
SMR	Steam methane reformer
Str.	Structures
Syngas	Synthesis gas
TASC	Total as spent cost
TDS	Total dissolved solids
TG	Turbine generator
TOC	Total operating cost
TPC	Total plant cost
TS&M	Transport, storage, and monitoring
TSS	Total suspended solids
U.S.	United States
USDW	Underground sources of drinking water
WGS	Water gas shift
°C	Degrees Celsius
°F	Degrees Fahrenheit

Executive Summary

Domestic, Fischer-Tropsch-based gas-to-liquids (GTL) technology provides options for using the United States' rapidly expanding supplies of recoverable domestic natural gas. Specifically, GTL technology significantly upgrades the economic value of the natural gas resource by providing access to the fungible transportation fuels market. By increasing the supply of secure domestic transportation fuels without relying on additional imports of oil, GTL technology is an effective tool for increasing the country's energy security. Additionally, GTL could mitigate some environmental concerns by displacing higher-sulfur fuels derived from petroleum with essentially sulfur-free fuels. If carbon capture and sequestration is incorporated in the facility and tighter emissions controls are applied to gas extraction then GTL produces liquid transportation fuels that have life cycle greenhouse gases which are similar to those emitted by fuels produced from conventional petroleum refineries today in the U.S., while contributing less than one percent to the cost of fuel production.

Commercial development of GTL has matured recently with two commercial GTL projects coming online during the last decade in Qatar. Both projects were built on the solid operating experience of a pioneer plant, in Malaysia, and commercial operation, in Mossel Bay, South Africa. An additional commercial project is in the process of coming online in Nigeria. Significant interest has been expressed in applying developing GTL technology to the increasing supplies of domestic natural gas. Specifically, Sasol has announced its plan to build an 8-10 billon dollar GTL complex in Calcasieu Parish, LA.[1] The intent to commercialize small-scale modular GTL in the Marcellus Shale region has recently been announced for Karns City, PA.[2] This augments speculation that additional GTL is being considered in the Northeast to offset closed refining assets.

In 1998, the National Energy Technology Laboratory (NETL)[a, 4] established a techno-economic baseline for GTL through a detailed systems analysis. The purpose of this study is to update that baseline based on the continuing evolution of GTL technology; assess the technical, economic, and environmental advantages and challenges associated with constructing a GTL facility in the U.S.; and provide guidance for research and development targets that would facilitate GTL development in order to provide domestic energy security to the U.S. liquid transportation fuels market.

This study models a GTL system that nominally produces 50,000 bbl/day of fuels fungible in the refined product infrastructure without further refining steps. Specifically, the system produces 15,460 bbl/day of finished motor gasoline and 34,543 bbl/day of low-density diesel fuel. The feedstocks to the system are 423,745 MMBTU/day (1863 GJ/hr) of processed natural gas and 499 tons/day of n-butane. The butane input can potentially be sourced from the natural gas liquids recovered during the processing of the natural gas; however, natural gas processing and its integration with the GTL facility is outside of the scope of this system study.

The GTL system considered in this study converts natural gas to synthesis gas in a catalytic reforming section consisting of a pre-reformer and an autothermal reformer (ATR). A small slip stream of natural gas is also sent to a catalytic steam methane reformer to produce hydrogen for product upgrading and to supply a means to augment the hydrogen content of the main synthesis gas stream during operational upsets. The synthesis gas is fed to a low-temperature, slurry-bed Fischer-

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[a] These studies were sponsored by the Pittsburgh Energy Technology Center and the Federal Energy Technology Center, which became the National Energy Technology Laboratory.

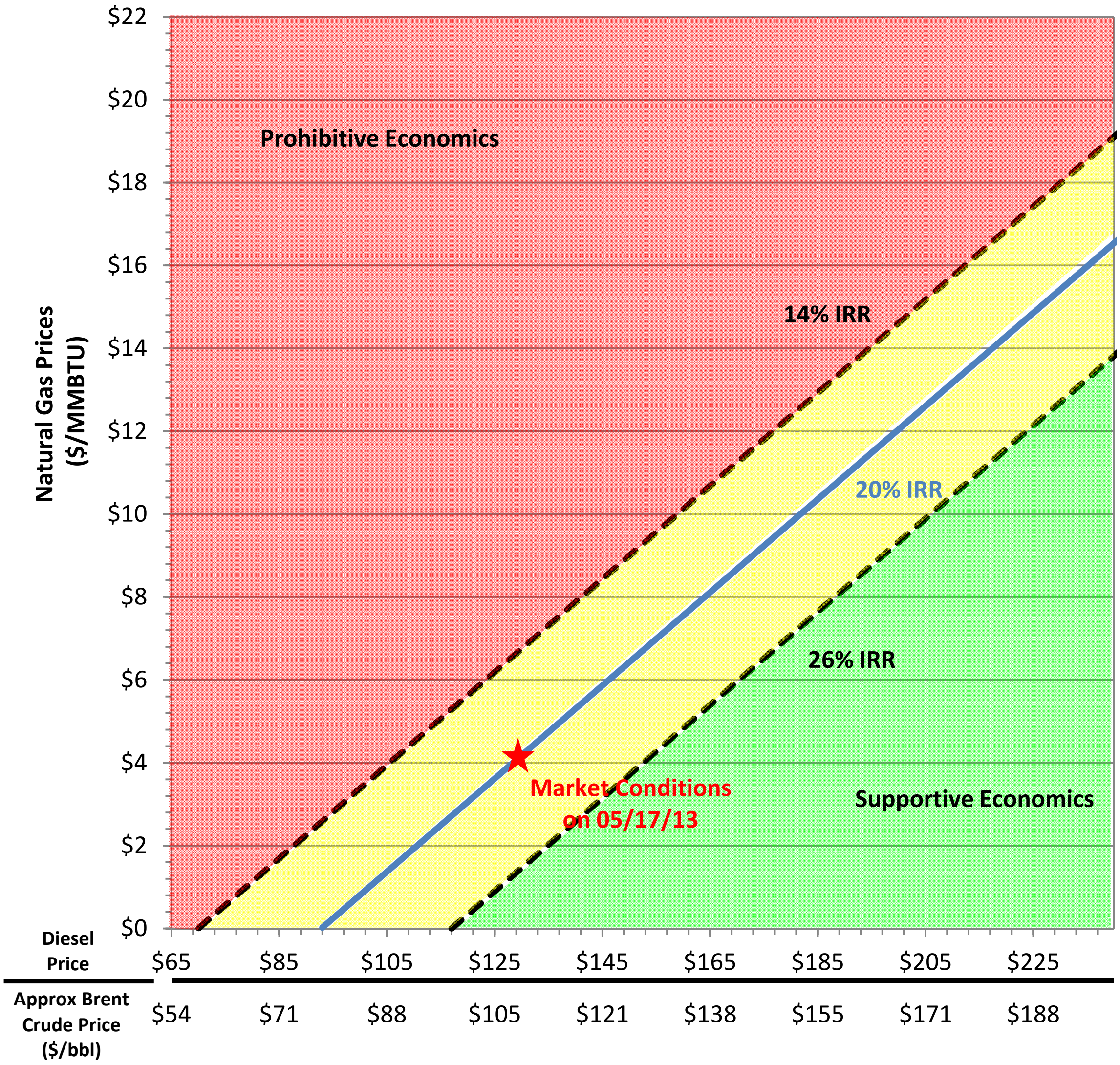
Tropsch reactor that employs a cobalt-based catalyst to reductively oligmerize carbon monoxide into Fischer-Tropsch liquids.

The liquid hydrocarbons are recovered from the vapor overhead of the Fischer-Tropsch reactor. The remaining vapor overhead has oxygenates and residual water removed and the carbon dioxide content captured and sequestered. The remaining vapor product is recycled into the natural gas reforming section. The liquid wax product is separated into appropriate fractions to feed the various product upgrading steps. The wax fraction is hydrocracked into a light hydrocarbon stream, a hydrotreated naphtha stream that is reformed, and a hydrotreated middle distillate stream. The middle distillate fraction is also hydrotreated and combined with the analogous stream from the wax hydrocracker to form a low-density diesel fuel. The naphtha fraction is hydrotreated and reformed, and pooled with isomerized and alkylated light products to form finished motor gasoline. A small amount of butane is imported into the system and isomerized and subsequently alkylated in order to achieve motor gasoline specification. Energy released from the system is recovered as electricity via a steam turbine.

This system analysis provides an updated evaluation of the cost and performance of an approximately 50,000 bbl/day of Fischer-Tropsch liquids GTL facility. Such an evaluation allows for the quantification of the window of viability for a domestic, moderate-scale, commercial GTL facility. This window is illustrated in **Exhibit ES-1** below. Conditions are more favorable in the lower right-hand side of the **Exhibit ES-1**, where market conditions are represented by natural gas and diesel fuel prices, and lines identifying the market conditions provide an internal rate of return of 14, 20, and 26 percent for the GTL system model in this study.

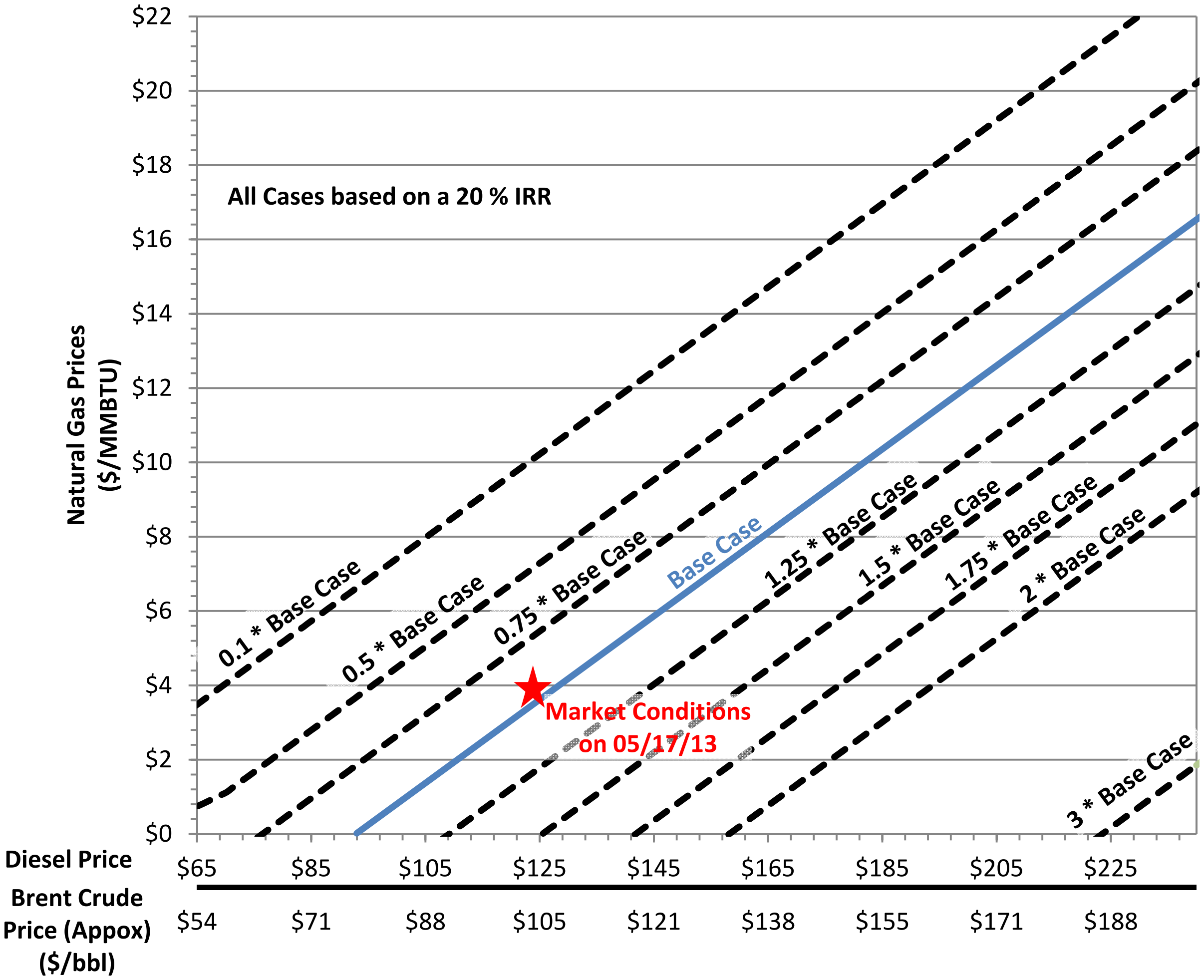
With an estimated total as-spent capital cost of 4.3 billion dollars (3.7 – 5.6 billion dollars) or \$86,188 (\$73,260 - \$112,045) per bbl of daily production of Fischer-Tropsch liquids, such a facility would be commercially viable should the market conditions (liquid fuel and natural gas prices) remain as favorable or better throughout the life of the project than during the middle of May 2013.

Exhibit ES-1: Viability Envelope for GTL Projects



The industry has been slow to accept the risk associated with the high probability that market conditions will regularly occur during the lifetime of a GTL facility which will not be sufficiently profitable to support such high capital investment. Additionally, such mega-projects have a higher risk of capital-cost escalation. **Exhibit ES-2**, organized similar to **Exhibit ES-1** with more favorable market conditions being located in the lower right hand corner of the graph, provides insight on the impact of varying changes in capital costs. Product and feedstock pricing not previously observed would be required to justify a project should realized capital expenditures be twice those anticipated by the estimates of this study – a capital cost escalation which is based on the escalation that occurred at Shell Pearl). Fortunately, the window of viability for GTL widens significantly if capital costs can be reduced by leveraging technology development or creating long-term contracts for natural gas structured to reduce feedstock cost variability.

Exhibit ES-2: Impact of Variation of Capital Costs on Project Viability



Increases in catalyst losses, on the order of magnitude often observed in the application of multiphase catalytic reactors, can significantly erode the profitability of a GTL facility. The internal rate of return drops by 0.5, 1.4, and 3 percentage points as catalyst losses increase 4-, 10-, and 20-fold, respectively. Catalyst activity gains will barely improve the internal rate of return of the facility. Increasing the liquids selectivity to 97 percent, while difficult, can add 2.7 percentage points to the internal rate of return of a GTL facility. Consequently, further development of reactor and catalyst technology should focus on improving the effective catalyst lifetime (improved attrition resistance and deactivation prevention) and selectivity.

The life cycle analysis (LCA) results of this analysis are limited to greenhouse gas (GHG) emissions, expressed as carbon dioxide equivalents (CO_2e) using the Intergovernmental Panel on Climate Change (IPCC) 2007 100-yr global warming potentials. Scenarios for diesel and gasoline were assessed, using 1 MJ of combusted fuel as the functional unit (or basis of comparison). The GTL facility produces multiple products, so this analysis uses co-product displacement to express GHG results for one product at a time.[b] The LCA also accounts for near-term improvement to upstream natural gas emissions. Results were generated for current practices in the natural gas industry and for improved performance once the approved New Source Performance Standards (NSPS) for the Oil and Gas Sector are fully implemented. The combination of two functional units (diesel and gasoline) and two upstream natural gas practices (current and NSPS) yield four scenarios for the LCA GHG results. The GHG results for these scenarios are shown in **Exhibit ES-3**.

The life cycle GHG emissions for GTL diesel and gasoline when based on current practices in the natural gas industry are 90.6 g CO_2e/MJ and 89.4 g CO_2e/MJ, respectively. If the natural gas extraction and processing sector complies with NSPS, the upstream GHG emissions from natural gas are reduced by 23 percent. With the implementation of NSPS standards, the life cycle values for GTL diesel and gasoline decrease to 85.3 g CO_2e/MJ (5.9 percent reduction) and 77.0 g CO_2e/MJ (13.9 percent reduction), respectively. In **Exhibit ES-3**, these values are compared to the NETL petroleum baseline values for petroleum based fuels, which are 90.0 and 91.3 g CO_2e/MJ for diesel and gasoline, respectively. The expected LCA GHG emissions for GTL diesel are 0.6 percent higher than the baseline without NSPS, and 5.3 percent lower with the implementation of NSPS. For GTL gasoline, the LCA GHG emissions are 2.1 percent lower than the baseline without NSPS and 15.7 percent lower with the implementation of NSPS. The NSPS regulations will reduce the venting rates from well development and maintenance, increase the flaring rates at natural gas wells, and reduce compressor emissions at natural gas processing plants; this will lead to an overall reduction in the GHG emissions from natural gas extraction and processing.

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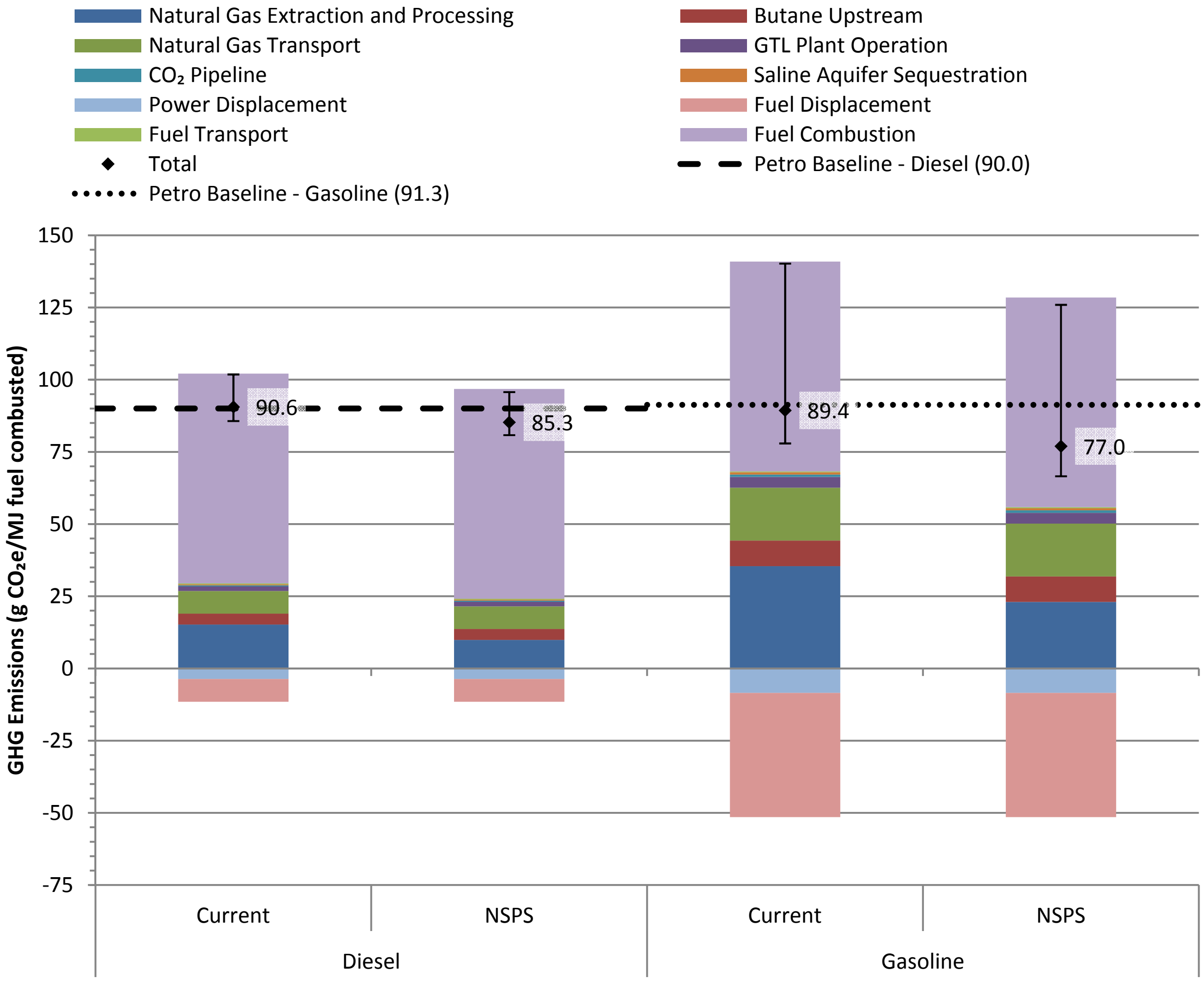
[b] The objective of LCA is to assign ownership of environmental burdens to a single product or function. The GTL facility produces three products (diesel, gasoline, and electricity), making it necessary to use a co-product management method to assign burdens to the functional unit. Displacement and allocation are two options for co-product management. Displacement expands system boundaries to envelop alternative pathways that are displaced by co-products, thus isolating one product of interest. Allocation uses a physical or economic relationship as a basis for splitting burdens among co-products. This analysis uses displacement for co-product management. Allocation is not effective in this analysis because it requires comparison of two forms of energy (electricity and liquid fuel) that are not used for similar applications and do not have the same degree of usefulness.

Exhibit ES-3: LCA Greenhouse Gas Results for GTL Diesel and Gasoline



The key challenges of GTL are the risk associated with varying gas and product prices, the lack of sustained effort in its development, and its high capital costs. A robust research and development program, besides driving capital cost reductions, can serve the role of sustaining the deep knowledge base in GTL. The research and development needs should focus on capital cost drivers such as non-cryogenic gas separation (which can lower the need for air separation units by enabling easier purge of inert, fixed gases) and process intensifications such as small-scale modular processing. Smaller scale processing will lower the barriers to entry into the GTL space, allowing more institutions to participate and drive innovation. Additionally, carbon efficiency can be improved by enabling the recovery and upgrading oxygenated co-products (which will also help biofuels development as well). Step-out, non-Fischer-Tropsch-based technologies hold additional potential to change the economics of converting natural gas to liquid transportation fuels, and they would be a useful topic for future analysis.

Research and development can also significantly contribute to the positive environmental impact of applying GTL technology. In most cases, the operation of the GTL facility accounts for a minority of life cycle GHG emissions. However, there are opportunities for improving the emissions from natural gas extraction, processing, and transport. Reducing gas losses in the upstream natural gas sector is a research and development goal that could reduce the life cycle GHG emissions from GTL to values below the petroleum baseline. Furthermore, GTL could be used to solve the environmental problem of stranded natural gas that is currently flared. The translation of GTL concepts to small modular-scale could enable altering the disposition of natural gas currently stranded at the well head, such as currently flared natural gas associated with the development of shale oil in the Bakken formation, to a useful liquid fuel product that augments domestic energy security and lower environmental impact associated with developing shale plays. The combination of thermochemical conversion processes for gas and biomass is another R&D goal that could further improve the value provided by GTL technology. It is also possible that the GTL technology basis could allow for a synergistic pairing of natural gas and biomass feedstocks in circumstances where feedstock supplies are limited. An example of such circumstances include combining natural gas from landfills or a one-off well-site with the practical limits on biomass supply for a military installation in a forward area mitigating the risk of defending supply lines or for making a domestic installation self-sufficient.

Market conditions, environmental concerns, and energy security can drive domestic application of GTL technology. These applications can leverage recent commercial development, but there is considerable need to reduce the risks associated with these projects. Research and development is required across the GTL value chain to unlock the potential for GTL to provide lower carbon-intensity fuels and to provide supply security to military installations. Gas-to-liquids can ease the strains on the refined product supply chain created by the reduction of East Coast refining capacity and can do so based on domestic resources. Additionally, GTL can provide additional sources of high-value fuels to support continued export of refined products from the Gulf Coast to Latin America helping the U.S. balance of trade and further solidify national energy security.

1 Introduction

New and expanding supplies of domestic natural gas due to the rapid development of shale gas plays and the resulting downward pressure on natural gas prices have motivated renewed commercial interest in domestic gas-to-liquid (GTL) operations. Traditional paradigms related to the viability of GTL are open to re-evaluation and challenge, given the current spread between crude oil and natural gas prices. GTL, or the production of liquid fuels from natural gas via the Fischer-Tropsch process, may provide an opportunity to address our economic and national security concerns related to the supply of liquid transportation fuels.

The Fischer-Tropsch process is the reductive oligomerization of carbon monoxide (CO) by hydrogen to form hydrocarbons (liquid hydrocarbons and aliphatic compounds). The feed mixture of CO and hydrogen is commonly referred to as synthesis gas. The reactions are typically catalyzed by iron or cobalt-based active materials.

Fischer-Tropsch technology reaches back to the 1920s and was first applied in areas where either the feedstock was “stranded”[c] or the market was blocked from access to conventional fuels. Recently, Fischer-Tropsch GTL technology has been seen as a means of monetizing stranded gas in places such as Qatar in western Asia. This has spurred pioneer projects in Malaysia (Shell Bintulu) and commercial projects in Nigeria (Escravos) and Qatar (Shell Pearl and Oryx). These projects are typically referred to as megaprojects due to their large size in hopes of capturing sufficient economies of scale and improving their viability. These projects also establish a solid technology platform from which to consider future projects.

The success of these commercial projects, particularly Pearl, and the expansion of recoverable natural gas resources in the U.S. have motivated this updated look at the feasibility of a domestic GTL facility. This system study updates and evaluates the cost and performance of a commercial-scale Fisher-Tropsch GTL facility. National Energy Technology Laboratory (NETL) has sponsored and shared considerable previous detailed work, on GTL[3, 4] and related “X”-to-liquids technologies such as coal-to-liquids[5, 6] and coal and biomass-to-liquids.[7] This previous work provided the foundation for this current system analysis. The scale of this study was set to align to the scale of these recent studies for ease of comparison.

The selection of specific GTL technologies (and consequently how previous work was adapted and used) is a function of the following study design choices. The first choice is the purpose of the facility, since Fischer-Tropsch can be targeted at fuels, chemicals, blendstocks, and waxes. The current system analysis focuses on the production of liquid transportation fuels from natural gas. Aligning with current market demand, this system analysis targets predominately middle distillate (i.e. diesel) and some lower distillate (i.e., motor gasoline) fuels. This affects the choice of a low-temperature Fischer-Tropsch process rather than a high-temperature process to obtain the desired product distribution. Given that the natural gas feedstock for this analysis is hydrogen-rich, water-gas shift functionality is not required; therefore, a cobalt-based system was selected. Additionally, a cobalt-based system was chosen, because, relative to iron-based systems, it produces less oxygenates and more saturated products,[8] leading to lower hydrogen requirements for product upgrading.

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[c] “Stranded” refers to the situation when the infrastructure (e.g., pipelines) are insufficiently developed and/or the distance to market too long to allow an asset to be brought to market and capture its full value.

Furthermore, cobalt-based catalyst systems are considered to have superior deactivation behavior when compared to iron-based systems[8] and can take advantage of significant, recent commercial Fischer-Tropsch catalyst research and development.[9, 10] A synthesis gas ratio (H_2/CO) of approximately 2 was selected to align with the product ratio of the desire transportation fuel products, dictating the technology choice of converting natural gas to synthesis gas.

The GTL plant assessed in this study is based on a new plant construction that is not co-located or associated via reasonable transportation and access agreements with other existing refinery or chemical upgrading facilities. As a result, the techno-economic analysis includes all components in order to independently make a salable fuels product.

In summary, there is considerable variability in determining what technologies are applied within the context of a GTL facility, based on the intent and project/enterprise specific considerations (summarized below in **Exhibit 2-2**). As such, this system analysis cannot align to all examples of GTL projects. This system analysis targets the domestic production of saleable fuels from natural gas at a moderated commercial scale. Sensitivity analysis and sharing of the detailed cost basis should allow stakeholders to adapt the results of this analysis to support their needs in different circumstances. This system analysis builds on the technical and cost analysis of previous NETL-sponsored GTL studies and provides insights on the economic viability of domestic GTL in the current environment.

2 Study Approach and Method

This system analysis developed a robust process simulation of the GTL facility reflecting current best technology practices for a fuel-oriented facility. The process simulation was developed to set a basis for estimating capital expenditures and operating expenses and revenue. The results of the process simulation were applied in standard NETL cost estimation and life cycle analysis methodologies in order to conduct sensitivity analyses to provide insight on the economic viability of fuels production from GTL technology in the current environment.

2.1 Method

The tools used for evaluating GTL technology, and a general description of GTL technology, are discussed below.

2.1.1 Process Simulation and Modeling

Process simulations were conducted in AspenPlus®, leveraging previous NETL systems studies to form the initial basis of the model. Specifically, models developed for the "Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2"[11] and the "Synergistic Production of Transport Fuels (Diesel, Jet, and Gasoline) from Coal"[5] were used to augment and update the models used for the basis from the "Baseline Design/Economics for Advanced Fischer-Tropsch Technology"4 report. Adjustments were made to the flowsheet (which is described in Section 3) to reflect adjustments in technology and intent (such as upgrading to fuels). This analysis also adds CO_2 compression equipment to allow for carbon capture and storage.

Process simulations were conducted using engineering judgment in a manner consistent with NETL's established Quality Guidelines for Energy System Studies.[12, 13] This system study provides sufficient information needed to estimate capital and operating costs, as well as to conduct subsequent sensitivity analyses to provide insight on a fuel-centered GTL system, but not at the level of detail design associated with a front end-engineering design study. The previous GTL study4 included some elements of detailed engineering and catalyst kinetic models as the basis of the code blocks used in the AspenPlus® modeling. The detailed AspenPlus® blocks were also used within this study to provide a firm basis for the results of the process simulation.

The process simulation results were also used to calculate the life cycle greenhouse gas (GHG) emissions associated with the Fischer-Tropsch diesel fuel and gasoline products. Results are reported based on emissions generated during resource acquisition, fuels production, transport of product, and the end use of the fuel. This was done by leveraging previous work by NETL on modeling the life cycle greenhouse gas footprint of petroleum-based and alternative fuels productions. Details are provided in Section 7.3.

2.1.2 Cost Estimation Method

Capital costs are developed for each major component or subsystem individually, based on the sizing developed in the process simulations. The estimate of capital costs, operating costs, and power production were scaled from previous NETL reports that obtained a cost basis from an engineering firm's cost engineering data and method. No vendor quotes were obtained or detailed engineering was performed as part of this systems analysis, which built on previous system analyses. All costs were escalated to 2011 dollar values by use of the Chemical Engineering Plant Cost Index (CEPCI).[14] The cost estimates reported below are not definitive; having a reasonable (-15 to +30 percent) range of accuracy. Additionally, there are certain project-specific risks and design choices that cannot be covered comprehensively in a generalized systems analysis. The components of the capital cost estimate (e.g., such as bare erected costs, total overnight costs) and relative accuracy are discussed in Section 5.

2.1.3 Sensitivity Analysis

A set of sensitivity analyses, covered in more detail in Section 5.3, was conducted to increase the utility of this system analysis. Such analysis is critical to using the results of this study to provide insight on GTL systems. Furthermore, a sensitivity analysis assures that the framework of the system analysis is well formed, by evaluating if the system responds to adjustments in a rational and consistent manner. Therefore, a variety of key systems parameters sensitivity analyses were conducted.

These sensitivity analyses were done outside of the AspenPlus® process simulations, using Microsoft® Excel. The impact of varying key economic parameters, such as the price of natural gas, fuel products, catalyst losses and capital expenditures, were examined. This allows for a basic mapping of the window of economic viability of a fuels-centered GTL process as a function of feedstock and product prices. Additionally, such analysis provides insight on how the window of viability moves due to shifts in performance (catalyst losses, activity, and selectivity) or capital costs escalation or improvement.

2.2 System Overview

The GTL system modeled in this study has the objective of producing nominally 50,000 bbl/day of fuels fungible in the refined product infrastructure without further refining steps. Specifically, the system produces 15,460 bbl/day of finished motor gasoline and 34,543 bbl/day of low-density diesel fuel. The feedstocks to the system are 423,745 MMBTU/day (1863 GJ/hr) of processed natural gas and 499 tons/day of n-butane. The butane input can potentially be sourced from the natural gas liquids recovered during the processing of the natural gas; however, natural gas processing and its integration with the GTL facility is outside of the scope of this system study.[d]

Exhibit 2-1 below provides a simplified flow diagram of the system considered for this GTL study. Processing sections will be described in further detail in Section 3, and details of the system (e.g., stream tables) are found in Section 7. The system converts natural gas to synthesis gas in a catalytic reforming section consisting of a pre-reformer and an autothermal reformer (ATR). A small slip stream of natural gas is also sent to a catalytic steam methane reformer to produce hydrogen for product upgrading and to supply a means to augment the hydrogen content of the main synthesis gas stream during operational upsets. The synthesis gas is fed to a low-temperature, slurry-bed Fischer-Tropsch reactor, which employs a cobalt-based catalyst to reductively oligmerize carbon monoxide into Fischer-Tropsch liquids.

The liquid hydrocarbons are recovered from the vapor overhead of the Fischer-Tropsch reactor. The remaining vapor overhead has oxygenates and residual water removed, and the carbon dioxide content captured and stored. The remaining vapor product is recycled into the natural gas reforming section. The liquid wax product is separated into appropriate fractions to feed the various product upgrading steps. The wax fraction is hydrocracked into a light hydrocarbon stream, a hydrotreated naphtha stream that is reformed, and a hydrotreated middle distillate stream. The middle distillate fraction is also hydrotreated and combined with the analogous stream from the wax hydrocracker to form a low-density diesel fuel. The naphtha fraction is hydrotreated and reformed, and pooled with isomerized and alkylated light products to form finished motor gasoline. A small amount of butane is imported into the system and isomerized, and subsequently alkylated in order to achieve motor gasoline specification. Energy released from the system is recovered as electricity via a steam turbine.

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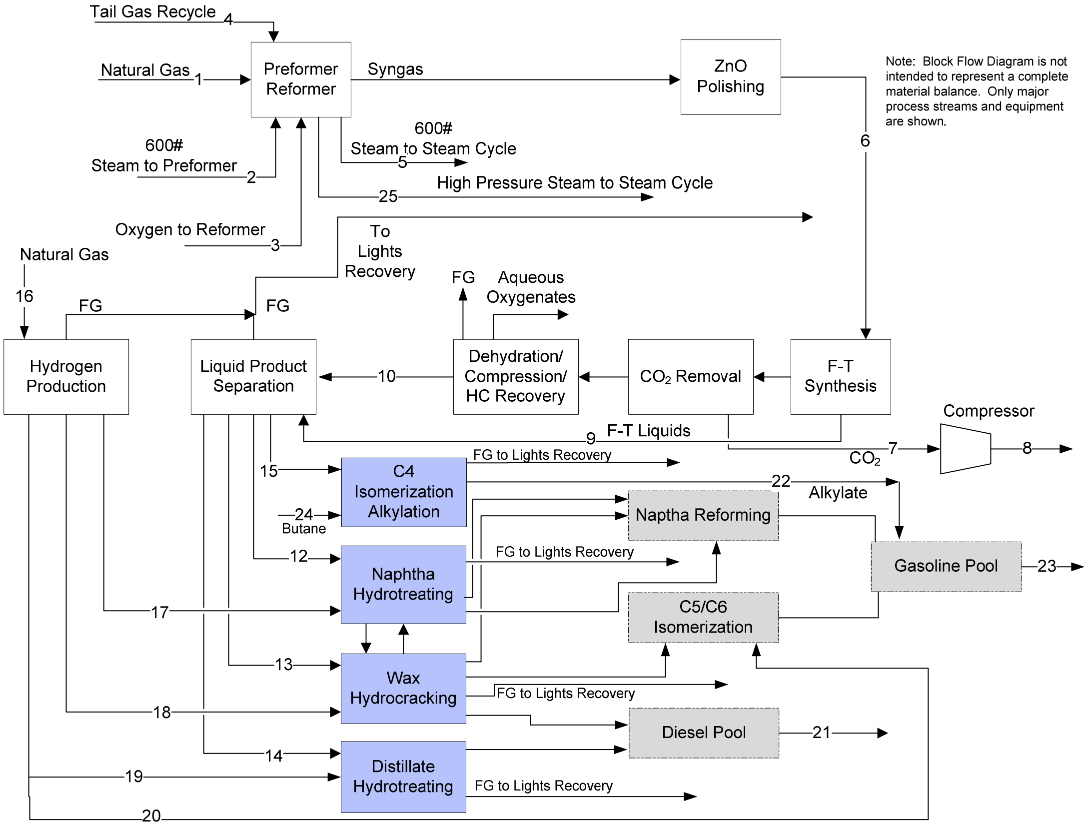
[d] With the exception of the environmental life cycle analysis

Exhibit 2-1: Simplified Flow Diagram of this GTL System



Cross comparisons between different GTL studies and commercial projects are informative but difficult due to differences in objective, scope, and scale. As mentioned in Section 1 above, the intent of the GTL projects (e.g., fuels, blendstocks, chemicals) greatly affects design choices for the GTL facility. **Exhibit 2-2** below provides a summary of GTL studies and commercial projects that can be used to provide context for the results of this systems analysis. **Exhibit 2-3** below provides a geographic representation of GTL projects that have moved forward.

Our current study is represented in the first row of **Exhibit 2-2**. The studies labeled Kramer3 and Bechtel4 were also sponsored by NETL.[e] The study labeled Steynberg refers to an open literature study[3]; this study and the Mossel Bay, Oryx, and Escravos GTL facilities are reported as interpreted by deKlerk.[8] The Shell's Bintulu and Pearl (Shell Qatar) projects are presented as their operations are reflected in the open literature sources.[8, 15, 9]

The design basis refers to a reasonable interpretation of the intent of the GTL facility. Capacity is provided, including the production of natural gas liquids (labeled NGL in the table) as this is relevant to understanding the scope and capital requirements. The type of Fischer-Tropsch synthesis technology employed is designated by catalyst-type iron (Fe) or cobalt (Co), temperature regime high

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[e] These studies were sponsored by the Pittsburgh Energy Technology Center and the Federal Energy Technology Center, which became the National Energy Technology Laboratory.

(HTFT, lighter product distribution) and low (LTFT, heavier product distribution), and reactor type (fixed bed, slurry bed, and circulating fluid bed [CFB]). The potential component of a GTL facility are broken out by production (of raw natural gas), transport (of natural gas to the gas processing plant), (natural) gas processing (for removal of natural gas liquids and contaminants), synthesis gas production, whether or not it has a recycle loop, product recovery, product upgrading, co-production of electricity, and carbon capture and sequestration (CCS). Black square markers indicate the presence of a given element in a study or facility to facilitate fair comparisons between them. Half-full circles indicate situations where either option was explored or when an element sometimes is considered and sometimes not considered part of a facility.

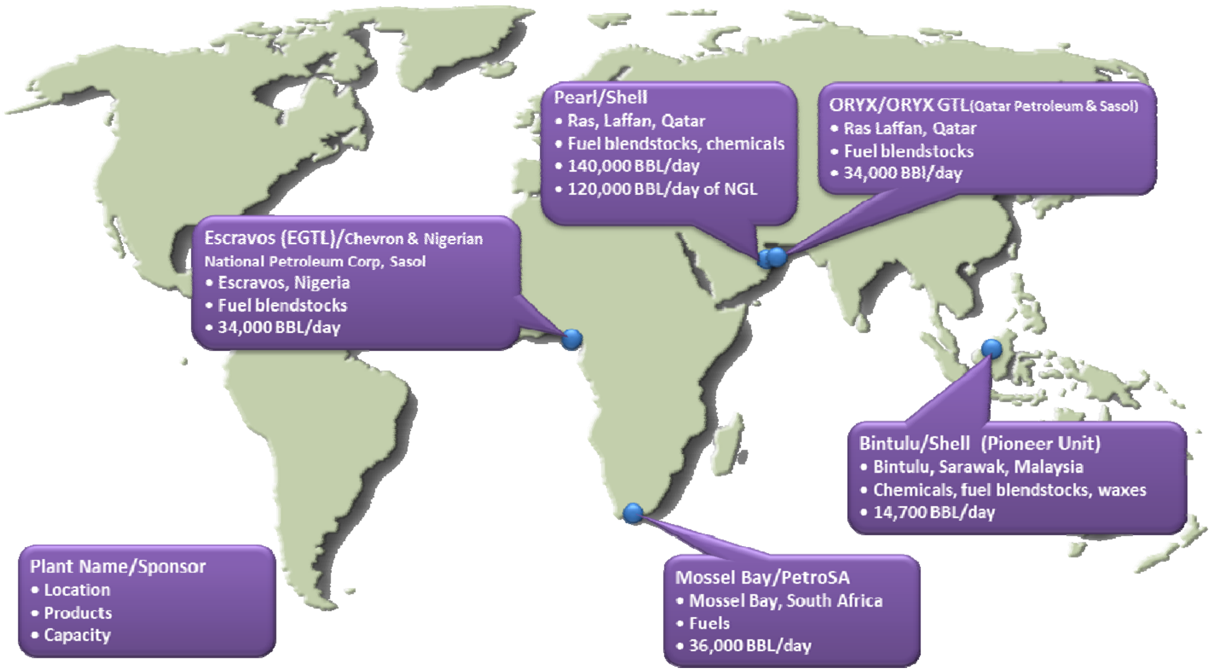
Exhibit 2-2: Overview of GTL Projects and Studies

Study/ Project	Design Basis	Capacity (000 BBL/day)	Fischer-Tropsch Synthesis	Production	Transport	Gas Processing	Synthesis Gas Production	Recycle Loop	Product Recovery	Product Upgrading	Electrical Power Generation	CCS
Current NETL	Fuels	50.0	Co-LTFT Slurry Bed				■	■	■	■	■	■
Bechtel (1998)	Fuels	8.8	Co-LTFT Slurry Bed				■	◐	■	■	■	
Kramer (1997)		43	Co-LTFT Slurry Bed				■	■	■	■	■	
Steynberg (2004)**		33					■		■	Limited		
Mossel Bay	Fuels	24.5+ 11.5 NGL	Fe-HTFT CFB + Co-LTFT Slurry Bed			■	■	■	■	■	■	
Shell Qatar	Fuel Blendstock, Chemicals	140+ 120 NGL	Co-LTFT Fixed Bed	◐	■	■	■	■	■	■ Chemicals	■	
Shell Bintulu	Fuel Blendstock, Chemicals, Waxes	14.7	Co-LTFT Fixed Bed			■	■	■	■	■ Chemicals	■	
Oryx/ Escravos	Fuel Blendstock	34	Co-LTFT Slurry Bed			■ Minor	■	■	■	Limited	■	

***de Klerk interpretation*

These systems will be used to provide context to the results of this current analysis in Section 5. The salient feature of **Exhibit 2-2** and **Exhibit 2-3** is that GTL facilities, while sharing some core elements, have wide variation, and comparisons are not always straight forward.

Exhibit 2-3: Geographic Overview of GTL Projects



3 System Description

A description of the functional sections of the GTL system is provided below. Note that this system is equipped for carbon dioxide capture.

3.1 Synthesis Gas Production

Exhibit 3-1 provides a simplified flow diagram of the synthesis gas production area. A fresh natural gas feed, after natural gas processing, is combined with recycled gas from the Fischer-Tropsch process and sent to a pre-reformer. The intent of the pre-reformer is to convert the non-methane hydrocarbons in the fresh feed and recycle gas into synthesis gas to prevent their conversion to undesirable coke at the higher temperatures present in the autothermal reformer. The pre-reformer is a low-temperature, catalytic[f] steam reformer and the process is net endothermic.[16]

The effluent of the pre-reformer is combined with oxygen in the autothermal reformer. The ATR represents a process intensification of synthesis gas production, commercially proven by Haldor-Topsoe at Oryx,[17] where partial oxidation of the feedstock provides the energy to drive the endothermic reforming of the feedstock to synthesis gas. The reforming is accomplished through contacting the reaction mixture with a nickel supported on alumina catalyst. The ATR is fed 95

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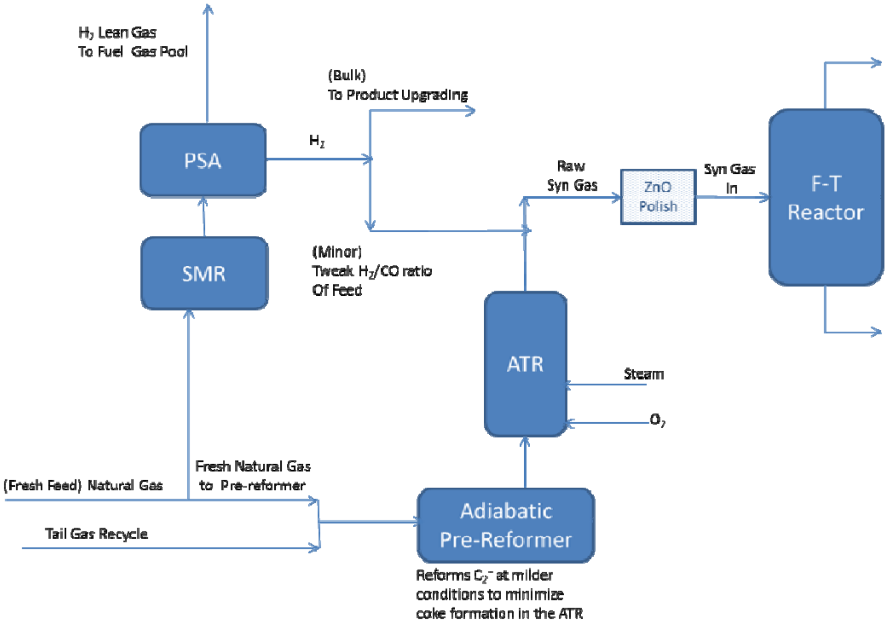
[f] The catalyst is a high activity form of the typical nickel supported on alumina reforming catalyst.

percent pure oxygen feed from the air separation unit[g]; a purified oxygen feed was chosen to minimize the amount of inert gases introduced into the system because of their potential to build up in the recycle loop. The ATR operates at 355.3 psia and 1935 °F (24.5 bar and 1057°C) in order to match the Haldor-Topsoe design.

Exhibit 3-1: Simplified Diagram of the Natural Gas Reforming Section



The operating parameters of the reforming section are provided in **Exhibit 3-2**. The H_2:CO ratio coming out of the reforming section should target the ~2.04 H_2/CO stoichiometricly-derived usage ratio for achieving the desired higher alkane product mix. The generally accepted range of H_2:CO

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[g] The ASU is designed to produce a nominal output of 1.15 million pounds per hour of 95 mole percent O_2 for use in the autothermal reformer. The air compressors are powered by an electric motor. There is no direct integration between the ASU and the rest of the facility.

The air feed to the air separation unit is supplied from stand-alone air compressors. The filtered air is compressed in centrifugal compressors, with intercooling between each stage. The airstream is cooled, and then fed to an adsorbent-based pre-purifier system. The air from the pre-purifier is then split into three streams. About 70 percent of the air is fed directly to the cold box. About 25 to 30 percent of the air is compressed in an air-booster compressor. This boosted air is then cooled in an after-cooler against cooling water before it is fed to the cold box. About 5 percent of the air is fed to a turbine-driven, single-stage, centrifugal booster compressor. This stream is cooled in a shell and tube after-cooler against cooling water before it is fed to the cold box.

All three air feeds are cooled in the cold box to cryogenic temperatures against returning product oxygen and nitrogen streams in plate-and-fin heat exchangers. The first, large-air stream is fed directly to the first distillation column to begin the separation process. The second air stream is liquefied against boiling liquid oxygen before it is fed to the distillation columns. The third, small-air stream is fed to the cryogenic expander to produce refrigeration to sustain the cryogenic separation process. The work produced from the expansion is used to power the turbine booster compressor.

Inside the cold box, the air is separated into oxygen and nitrogen products. The oxygen product is withdrawn from the distillation columns as a liquid and is pressurized in a cryogenic pump. The pressurized liquid oxygen is then vaporized against the high-pressure air feed before being warmed to ambient temperature. The gaseous oxygen exits the cold box and is split into two streams. Essentially, all of the gaseous oxygen is fed to the centrifugal compressor with intercooling between each stage of compression. The compressed oxygen is then fed to the autothermal reformer.

Nitrogen is produced from the cold box at two pressure levels. No process credits were taken for the co-production of nitrogen, since a local process load is not assigned.

ratios from the reforming section that feeds the Fischer-Tropsch synthesis range is 2.0-2.2.[18] The current model produces synthesis gas on the higher end of this range favoring a more saturated product mix.

The stoichiometric number metric (S#), which is calculated by the following equation

$$S\# = (H_2 - CO_2) / (CO + CO_2)$$

is 1.6. In a Fischer-Tropsch system without significant water gas shift activity (e.g., cobalt-based catalyst rather than iron-based), this metric does not have a significant relationship to reactor performance that is dependent on the H_2:CO ratio. In systems with significant water gas shift activity, the S# is targeted to values near 2.1,[19] so the water gas shift functionality works in concert with the reductive oligomerization of carbon monoxide.

The O_2:C[h] and H_2O:C ratios were based on common practice and assure an atmosphere sufficiently oxidative to prevent coking in the ATR. Based solely on the amount of heat required, the O_2:C ratio could be lowered significantly. This is in line with common industrial practice.

A small amount of natural gas is fed to a steam methane reformer (SMR) to produce hydrogen for use in downstream product upgrading. The hydrogen is recovered from this hydrogen-rich synthesis gas via a pressure swing adsorber (PSA). This unit operation also provides a means to augment and regulate the hydrogen content of synthesis gas in the event of process upsets in the ATR itself or in the process in general (affecting the composition of the recycle loop). The hydrogen deficient raffinate from the pressure swing adsorber is combined with the recycled tail gas, which is predominately recycled to the pre-reformer with a small amount used as fuel gas (to create an inert gas purge).

The raw synthesis gas is fed to a zinc oxide bed to remove any residual heteroatom content (i.e., sulfur) that makes it through the natural gas processing and subsequent reforming process. The zinc oxide is a final guard bed to protect the Fischer-Tropsch synthesis catalyst.

Exhibit 3-2: Reforming Section Metrics

Reforming Metrics	GTL
O_2:C[1]	0.73
H_2O:C[1]	0.68
H_2:CO[2]	2.19
Stoichiometric Number (S#)[2]	1.59

1 – Ratio of oxygen and steam injection to hydrocarbon carbon at the input to the natural gas reforming section

2 – Synthesis gas metrics (output of reforming section)

Industrial practice uses the same basic building blocks in a variety of ways to accomplish the production of synthesis gas from natural gas. Choices are driven by project-specific criteria and the

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[h] The carbon basis for this ratio is the amount of hydrocarbon (i.e., non-CO and non-CO_2) in the inlet to the system.

operating experience of the organization developing the project. The Haldor-Topsoe technology employed at Oryx and Escravos[8] matches the chosen technology for this study. South Africa's Mossel Bay facility employs a similar approach by sending a slip stream of natural gas to a steam methane reformer and combines the products of the steam methane reformed with more natural gas, recycled tail gas, and oxygen in an ATR.[8] Excess hydrogen and carbon dioxide are removed prior to introduction of the synthesis gas into the Fischer-Tropsch reactor.[8]

For its Bintulu and Pearl facilities, Shell adapted the Shell Gasification Process (SGP) that it developed in the 1950s for gasifying heavy oil.[8, 15] The SGP thermally reforms natural gas at temperatures above 2,300 ^{o}F (1,260 ^{o}C) via partial oxidation with purified oxygen.[8, 15] The synthesis gas from the SGP is adjusted to meet the desired Fischer-Tropsch usage ratio (~2.15) by using catalytic steam methane reforming and pressure swing absorption to produce additional hydrogen.[8]

Previous NETL-sponsored studies[4, 20] have employed both an adapted SGP and an enriched air-ATR approach. This study chose the Haldor-Topsoe system based on its success in intensifying the production of synthesis gas, its higher reliability, and the amount of open literature available on this system.[21, 16]

3.2 Fischer-Tropsch Synthesis

The intent of this GTL system study is to produce liquid transportation fuels. As such, a low-temperature Fischer-Tropsch process is more appropriate than a high-temperature Fischer-Tropsch process. Specifically, the product distribution from low-temperature Fischer-Tropsch has significantly more weight on fuel-range hydrocarbons and higher hydrocarbons that can be cracked to fuel-range hydrocarbons, a distribution similar to a light crude oil. Since a hydrogen-rich, natural gas feedstock is the source of the synthesis gas in this system analysis, the water gas shift functionality of iron is not required in our Fischer-Tropsch system. Consistent with the intent of this study to make transportation fuels, the higher selectivity to middle distillates of cobalt-based catalysts was preferable to iron systems. As such, this systems study is based on a cobalt-catalyzed, low-temperature Fischer-Tropsch process.

Cobalt-based low-temperature Fischer-Tropsch processes has been established as the basis of commercial GTL processes. While the facility in Mossel Bay, South Africa, employed a more technically ready high-temperature, iron-based system for its initial capacity, once cobalt-based low-temperature Fischer-Tropsch matured technically, it was used for the Mossel Bay capacity expansion. Similarly, cobalt-based low-temperature Fischer-Tropsch was the basis of Shell's Bintulu and Pearl facilities, Oryx, Escravos, and the now-abandoned ExxonMobil GTL project, in Qatar. The previous NETL study also was based on cobalt-based low-temperature Fischer-Tropsch.[4]

Proper reactor technology is crucial to applying the cobalt-based catalytic technology. Shell has chosen a fixed bed reactor technology for its GTL facilities. This approach leads to high paraffin, low oxygenate, lower olefin-product yields consistent with the Shell's 'heavy paraffin synthesis' moniker for their process.[8, 9, 15] Shell's process requires catalyst regeneration every 9-12 months, has a catalyst lifetime of approximately five years,[8] has reduced scale-up risk when implementing new catalyst, and does not risk the potential of high catalyst replacement costs due to attrition.

Despite the risk of catalyst-attrition issues, slurry bed operations have been widely adopted and considered. Advantages of a slurry bed GTL reactor relative to a fixed bed reactor include reduced transport resistances (heat and mass transfer) and the ability to replace catalyst while operating. Unfortunately, these reactor have higher scale-up risks and generally require significant scale to be economical (i.e., this reactor choice encourages mega-scale projects with their higher inherent financial risks). Slurry bed technology was chosen for Oryx, Escravos, the expansion at Mosell Bay,[8] and ExxonMobil's GTL process.[10] Catalyst attrition issues have been reported and have created delays in achieving name-plate capacity at Oryx,[22] but appear to be manageable at this current time. Consequently, a slurry bed reactor system was chosen as the basis of this analysis.

The current study leverages the foundational work of previous NETL-sponsored studies[i, 4, 20, 23] to apply a cobalt-catalyzed, low-temperature Fischer-Tropsch slurry bed process to convert synthesis gas to Fischer-Tropsch liquids. This previous work includes incorporation of DOE-funded kinetic studies of cobalt Fischer-Tropsch catalysts. Details such as catalyst support and binder formulations cannot be captured in the model. The influences of these variables are reflected more in the assumptions regarding catalyst attrition and loss rates used in developing an economic model of the process. A brief description of the process represented by the NETL slurry bed model is provided below.

To scale from previous processes, this study scaled up the Fischer-Tropsch synthesis train of previous NETL studies[4] by 15 percent, and then employed four parallel trains to obtain the desired synthesis-gas conversions capacity. Each train consists of two parallel first-stage slurry bed reactors that feed a common second-stage slurry reactor. The conceptual design of the Fischer-Tropsch synthesis section is represented in **Exhibit 3-3** below. Unconverted synthesis gas is cooled between the first and second stages to 150°F (65.6 °C) to remove water and a hot hydrocarbon condensate (which is sent to product recovery), and then re-heated prior to entering the second stage. Each stage accomplishes approximately 61 percent conversion of carbon monoxide (hydrogen is in excess) leading to a per-pass conversion of 85 percent of carbon monoxide. There is no direct recycle of unconverted synthesis gas back into the second-stage reactor (the implementation of the recycle loop is discussed in Sections 3.1. and 3.4).

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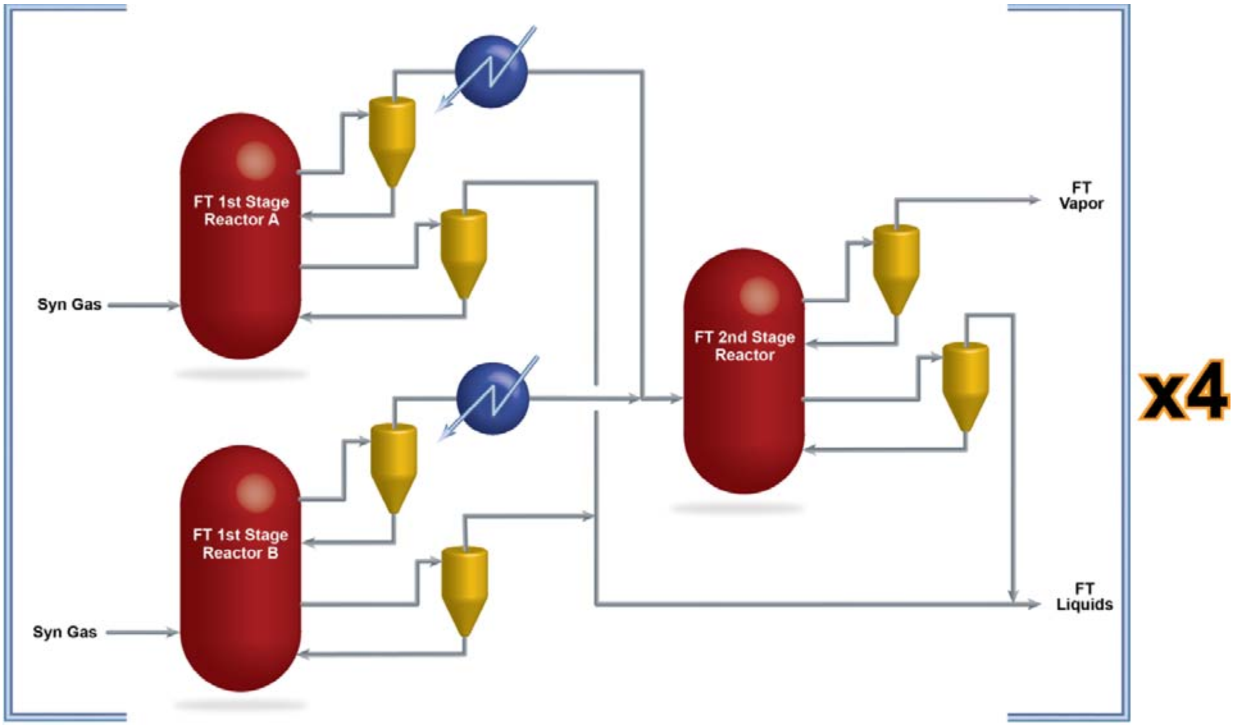
[i] These studies had components of detail engineering design to support their flowsheet model and cost basis that were outside of the scope of the current study. Consequently, the current study leveraged this more detailed work to a large extent. Updating the costs to 2011 dollars and verifying the approaches used in the previous studies are still representative of current technology practice. Areas for potential improvement, such as catalyst activity, are considered in the sensitivity analysis, in Section 5 of this study.

Exhibit 3-3: Simplified Representation of the Fischer-Tropsch Synthesis Section



Each of the three Fischer-Tropsch slurry bed reactors in the train has equivalent external dimensions. A representative geometry for these reactors would be 20.3 feet (6.2 meters) in diameter and approximately 65 feet (19.8 meters) tall.[j] The target superficial gas phase velocities in the slurry bed reactor were 0.39 and 0.48 ft/s (0.12 – 0.13 m/s) in the first- and second-stage reactors, respectively.

Reactor temperature is modulated by removing tremendous heat released by the Fischer-Tropsch process through generation of 150 psig (11.4 bar) steam through bayonet tubes suspended from an internal double tubesheet.[k] The reactors operate at 305 – 319 psia (21 – 22 bar) and nominally 480 °F (249 °C).[l]

Synthesis gas is introduced into the reactors through a cylindrical gas distributor and bubbles up through the slurry phase. Fischer-Tropsch wax and catalyst make up the slurry phase, into which the synthesis gas dissolves, contacts the catalyst, and is oligmerized producing more Fisher-Tropsch

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[j] This represents a reasonable approximate scale-up of dimensions from the previous studies with the diameter of the reactor fixed so as to maintain similar fluid velocities and hydrodynamics. It is fully recognized that scaling up of multiphase reactors is considerably more complex than the approach used to provide approximate geometry. Rigorous scale-up of the reactor design would require significantly more resources than the benefits it would add to the current analysis.

[k] Detailed engineering on the reactor internals was not part of the current study. The number of bayonet tubes is estimated to be on the order of 2000 per reactor.

[l] The outlet temperature is set by the target that was yield in the systems model, with this choice being at the edge of the operating window referred to as low temperature Fischer-Tropsch synthesis.

liquids. Streams of liquid wax are withdrawn from the reactors with a slurry valve, which accomplishes the majority of the catalyst-wax separation.

Vapor is removed from the overhead of the first-stage reactors passing through a cyclone to prevent liquid carryover. The vapor is cooled and sent to a three-phase separator. The hot condensate (light liquid hydrocarbons) is recovered for further use. The water is sent to treatment, and the remaining vapors are reheated and sent to the second-stage reactor.

The wax products are sent to hydrocyclones to accomplish the remaining catalyst-wax separation. Subsequently, the resulting wax stream is degassed with the resulting vapor cooled to remove additional condensate that is sent to product recovery and the residual vapor combined with the tail gas for recycling. The liquid wax is sent to wax clarifying and fed to the hydrocarbon recovery section.

The catalyst that is separated from the bulk of the wax streams is recycled back to the Fischer-Tropsch reactors. To maintain performance, a portion of the catalyst inventory of the reactors is removed for disposal and replaced. Catalyst replacement rates are influenced by loss of catalyst functional lifetime (activity and selectivity) and attrition losses. Cobalt-based Fischer-Tropsch catalysts generally have significantly higher lifetimes than iron-based systems. Physical loss of the catalyst in slurry systems has been an impediment of commercial systems reaching nameplate capacity, and it creates serious operational issues within the Fischer-Tropsch and downstream processes.[8] A composite lifetime (mechanical and functional) of the catalyst in this system analysis is 0.78 years, meaning 0.015 percent of catalyst inventory is replaced per hour.[m]

3.3 Product Recovery

All of the hydrocarbon streams from the Fisher-Tropsch synthesis section (wax and condensate recovered from the overhead vapor streams) are combined in the feed drum of the product fractionator. The product fractionator separates these hydrocarbons into a low-pressure light hydrocarbon vapor steam and three liquid streams: naphtha, middle distillate, and wax. The liquid cuts of the fractionator are defined by the following boiling range cut points:

Naphtha	C_5 saturates to 350 °F (177°C)
Middle distillate	350 – 650 °F (177 – 343 °C)
Wax	Greater than 650 °F (343 °C)

This simple separation section consists of one fractionation column with a steam side-stripper. Required support equipment (heat exchangers, flash drums, and furnace utilities) are included in the capital estimate of this section. The feed drum operates as a flash drum at approximately 135°F (57 °C) and 90 psia (6.2 Bar). The vapor feed rate to the tower is used for pressure control of the flash drum. The vapor serves as stripping media after being fed to the middle of the column. The liquid is heated to approximately 343°F (173 °C) via a fired heater, which burns residual fuel gas.

The overhead vapors are air cooled, then water cooled, and subsequently sent to a three-phase separator. The vapor phase exiting this three-phase separator is combined with the light hydrocarbons

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[m] This is approximately 60 percent higher loss rate than the previous NETL-sponsored GTL study.(Baseline Design Economics for Advanced Fischer-Tropsch Technology, March, 1998) The rate was increased due to the increased attrition observed in slurry bed GTL operations deployed following that study.

recovered from product upgrading process for C_3^+ recovery and eventual combination with the recycle and fuel gas pools (see Section 3.4). The condensed water (introduced to the tower as stripping steam) is sent for waste water treatment. The liquid hydrocarbons are sent back to the tower as a rectification fluid.

The naphtha product is withdrawn from an upper tray and is sent to naphtha hydrotreating section. Steam stripping is used to provide a stable liquid middle distillate stream from the middle of the column, is cooled (via contact with the column feed stream), and then is sent to the distillate hydrotreating section. The tower bottoms are withdrawn and via contact with the tower feed and subsequent water-cooled exchangers prior to being sent to the wax hydrocracking section.

The product recovery scheme used in this analysis is matched to the subsequent fuels-oriented product upgrading (see Section 3.5). Shell[9] and Mossel Bay[n, 8] use simple fractionation to separate their products into fractions aligned with their downstream upgrading operations, which for Shell are more oriented to chemicals and high-grade waxes. Oryx and Escravos do not need fractionation, because they have a very simple, single hydrocracking upgrading operation with some fractionation downstream of their upgrading.

3.4 Recycle Gas Loop

Vapor products of the second-stage reactors (those leaving the Fisher-Tropsch synthesis section) are handled in a similar fashion, with the overhead product being sent to a cyclone to prevent liquid carryover, cooled, and then sent to a three-phase separator. The hot condensate (recovered via a LiBr chiller) and water have the same disposition as following the first stage. The remaining vapors are processed for CO_2 removal, oxygenate removal, dehydration, and additional hydrocarbon recovery. The means of CO_2 removal, capture, and storage is discussed in Section 7.1.4. The resulting tail-gas vapor is pooled with the hydrogen deficient synthesis gas raffinate from the hydrogen production section for recycle to the pre-reformer.

The overhead vapors from the hydrocarbon recovery section, the light gases produced in the hydrocarbon upgrading sections, and the off-gas from the wax stream are collected in one light ends stream. This stream is subjected to moderate pressure and slight cooling (~ 113 ^oF or ~45^oC) to recover residual C_3^+ material in these streams. To the extent possible, this material is alkylated to form a motor gasoline blendstock. The remaining vapor after recovery of the C_3^+ material is combined with the other tail-gas vapors for recycle.

Closing of the recycle loop causes a build-up of non-condensable vapors (e.g., nitrogen and argon). Two steps were taken to contain the build-up of non-condensable build gases. First, 99 percent pure oxygen is fed to the ATR to reduce the amount of nitrogen coming in with oxygen. Second, a 5 percent purge of the tail gas was implemented to remove non-condensable from the process loop. This purge is burned to produce steam used for site utilities and the power cycle.

The recycle gas loop approach used in this study is consistent with industrial practice. Shell recycles a C_1 – C_4 containing tail gas to a steam methane reformer and combines that reformate with the synthesis gas produced by the SGP.[8] Similar to this analysis, the Oryx and Escravos design recycle

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[n] Mossel Bay takes advantage of stepwise cooling of the vapor phase to pre-separate product fractions as well. This is especially true of the lighter product distribution from the original iron-based high temperature Fischer-Tropsch portion of the facility.

their tail gas to the ATR with a slip stream used for fuel gas.[8] The Mossel Bay facility co-feeds their tail gas along with natural gas, oxygen, and the output from a steam methane reformer.[8]

3.5 Product Upgrading

The product upgrading in this systems analysis is more extensive than recent commercial Fischer-Tropsch facilities. This is due to our focus on producing on-specification fuels rather than blendstocks and chemicals. Chemicals are higher margin projects, but an extensive rollout of Fischer-Tropsch technology would flood these markets. Consequently, Fischer-Tropsch projects will increasingly focus on fuels with increasing application of the technology.

The Oryx and Escravos system has a very simple, low-capital expenditure upgrading system that employs Chevron's Isocracking® technology in single reactor.[8] This approach produces liquefied petroleum gas, naphtha,[o] aimed at steam-cracking feedstock (to make prime olefins, i.e., ethylene and propylene), and a middle distillate blendstock, with wax recycled to extinction.[8] Shell also used its naphtha for steam-cracking feedstock for olefin production.[15] The highly paraffinic products from Shell's GTL facilities are sent to hydrocracker[p] to produce fuel blendstocks, and to a hydrotreating[q] operation, which produces chemicals and waxes.[8, 9, 15]

Mossel Bay, with its finished-fuels focus has an array of refining unit operations very similar to those used in this study.[8] The unit operations in this current analysis are consistent with those applied in previous NETL-sponsored GTL studies[4, 20] with adjustments, predominately to the naphtha upgrading, to make on-specification fuel.

The wax fraction is sent to the wax hydrocracker, which breaks down the wax into naphtha and middle distillate, as well as some residual C_4^- vapor[r], which is recovered (see Section 3.4). The single, multibed, downflow wax hydrocracker catalytically cracks the wax under a 1088-1494 psia (75-103 bar) hydrogen atmosphere at a nominal temperature of 700 ^{o}F (371 ^{o}C). The hydrocracking process uses inter-bed cooling to remove the heat of reaction from hydrogen-rich recycle gas. A flash drum is used to recover unutilized hydrogen-rich gas for recycle back to the hydrocracker. A distillation column is used to separate the liquids from the flash into C_5/C_6, naphtha, middle distillate, and wax streams. The residual uncracked wax is recycled to extinction. The middle distillate from the wax hydrocracker is sent for blending with the middle distillate range product from the distillate hydrotreating section. The naphtha stream from the wax hydrocracker is sent to the naphtha reforming unit. The C_5/C_6 stream is sent to the C_5/C_6 isomerization unit.

The fixed bed distillate hydrotreater catalytically hydrotreats the Fischer-Tropsch middle distillates into a quality acceptable as diesel fuel. This processing section removes oxygenates and any residual heteroatoms (sulfur and nitrogen), and greatly reduces the olefinic and aromatic content. The

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[o] Fischer-Tropsch naphtha is a superior feedstock stream for cracking compared to virgin naphtha cuts from crude oils. Fischer-Tropsch naphtha leads to significantly higher olefin yields and as such is not as vulnerable to the current market conditions that are not favorable to naphtha steam cracking.

[p] This operating section accomplishes hydrogenation, hydrodeoxygenation, hydrocracking, and hydroisomerization.

[q] This operating section accomplishes hydrogenation and hydrodeoxygenation without the cracking and isomerization functionalities present in the hydrocracker.

[r] C_4^- is an industry convention to denote a hydrocarbon chain of four carbons or less.

distillate hydrotreater also produces, and the processing section separates off, a waste-water stream and a residual C_4^- vapor stream, which is recovered (see Section 3.4). The hydrotreated middle distillate from this processing section is blended with the analogous stream from the wax hydrocracker to form a product diesel. The product diesel has an estimated pour point of -52.5 ^oF (-46.9 ^oF), well below the target value of any region in the U.S. The estimated cetane index of this diesel product is 70.9 above the standards set in the World Wide Fuel Charter (minimum requirements between 48 and 55 depending on category).[8] Within the accuracy of the AspenPlus® model, this diesel is a saleable fuel. The properties of the diesel in this system analysis are closer to a No. 1 diesel grade than the more ubiquitous No.2. No. 1 diesel is often used to improve the cold weather performance of diesel.

The diesel fuel produced in this system analysis understandably has the same limitations of GTL-derived diesels. While it is outside of the granularity and capability of an AspenPlus® model to assess the lubricity of the fuel, it is rational to assume that the diesel fuel of this analysis will require additives to improve its lubricity as GTL diesel requires.[24] This can be handled in the blending and distribution phases of the supply chain.

Similarly, the density of the diesel produced in this system analysis is lower than petroleum-derived diesel with subsequently lower energy density, which is endemic with GTL-derived diesel produced at adequate cetane index in high yield due to a lack of cycloalkane compounds[8] with the density being less than even a No.1 diesel. Designers of a GTL system have to select a tradeoff between density and yield.[8] Given the capacity of the refined product system to either use low-density diesel directly or to blend it with a higher density diesel, the choice to operate at higher yield is evident.

The lower-density fuel has a positive impact on nitrogen oxide and particulate matter emissions when combusting the diesel in an engine.[24, 25] Diesel fuel does not have a worldwide specification on density. Additionally, low density diesel can be sold as finished fuel into the distribution chain, which can adjust the density of diesel by blending as part of their normal operations. In the economic analysis of this system, the diesel fuel is sold at a 7 percent discount, based on the reduced energy content.

Production of on-specification gasoline in this analysis requires the coordinated operation of multiple unit operations. Specifically, more than hydrotreating, the naphtha is required to meet on specification gasoline. The Fischer-Tropsch naphtha is run through a fixed-bed hydrotreater to create a saturated naphtha suitable for reforming. Besides the saturated naphtha stream, the naphtha hydrotreater also produces a wastewater stream, a residual C_4^- vapor stream, and a C_5/C_6 stream, which are separated off. Recovery from the C_4^- stream is discussed in Section 3.4. The wastewater is sent to treatment. The C_5/C_6 stream from the naphtha hydrotreater is combined with the analogous stream from the wax hydrocracker and is then subjected to a fixed-bed catalytic isomerization process to raise the octane number of this blending component by converting straight chain paraffins to isoparaffins.

The saturated naphtha streams from the naphtha hydrotreater and the wax hydrocracker are combined and sent to a fixed-bed catalytic reformer. This processing section reforms the naphtha into a high-octane blending component. Besides the reformate product stream, this processing section also produces a residual C_4^- vapor stream whose disposition is covered in Section 3.4.

To supply sufficient iso-butane to the alkylation plant, this analysis includes a processing section isomerize normal butane to iso-butane. A C_3^- lights stream also comes out of this processing section; its disposition is covered in Section 3.4. There is not sufficient normal butane produced in the process to feed this processing section in sufficient quantities to produce the amount of iso-butane required to eventually make on-specification motor gasoline. This requires the import of significant quantities of normal butane (499 tons/day). For purposes of economic evaluation, the normal butane was treated as being purchased from the market. In a practical sense, the normal butane will probably be recovered from the natural gas liquids coming from the facility processing the raw natural gas prior to its introduction to the GTL system. In practice, the gas processing is co-mingled with the GTL facility, so the normal butane purchase is more likely an internal transfer price rather than a true cash expense.

A sulfuric acid alkylation plant converts the iso-butane along with C_3-C_5 stream recovered from the process to a high-octane blendstock with low-Reid vapor pressure. In the AspenPlus® model, this blendstock is represented as single pseudo-compound (ALKYLATE). This processing section also has a rejected normal butane stream (which is recycled back to the butane isomerization unit) and a C_3^- vapor stream whose disposition is covered in Section 3.4 below. A small amount of hydrocarbon is lost with the spent sulfuric acid from the process.

The alkylate, isomerate, and reformate streams are blended to form an on-specification motor gasoline pool. The density of this gasoline is on the low end of the desired density, a research octane number of 90.3, a motor octane number of 85.7, and an (R+M)/2 octane number of 88.0. The motor gasoline from this system study has an approximate Reid vapor pressure of 5.0 psia (0.34 bar), estimated benzene content of 0.26 weight percent, aromatic content of 24.8 weight percent, and olefin content of 0.03 weight percent, all within norms for motor gasoline.

3.6 Power Production

A block flow diagram of the power production used in this system analysis is provided in Section 7. Heat is recovered from the process, as well as steam from the outlet of the reforming sections and Fischer-Tropsch reactor. This steam is combined with the small amount of steam generated from the flue gas boiler. This boiler produces high-pressure steam from steam condensate using the heat from burning the small amount of tail gas, which is burned to prevent buildup of inerts in the gas loop. The high-pressure steam feeds a steam-extraction turbine that produces electricity to satisfy facility loads, as well as excess electric power for sale. Low-pressure steam from the GTL process is fed into later stages of the extraction turbine to recover this energy as well. Steam extracted from intermediate stages of the turbine feeds the ATR and provides the 600 psig (42.4 bar) steam to the Fisher-Tropsch reactor. Energy integration is discussed further in Section 4.4.

4 Technical Performance Summary

Technical details on the performance of the GTL plant are provided below.

4.1 Overall Description

This study targets the production of finished motor gasoline and diesel fuels at a scale of approximately 50,000 bbl/day from processed natural gas. In order to meet fuel specifications, some butane needs to be brought into the system (presumably available from the NGL output of the same gas processing plant that supplies the natural gas). Since the GTL system is equipped with carbon capture and storage equipment, these fuels would be supplied with a GHG impact comparable to the impact of conventional petroleum fuels despite the acquisition and delivery of natural gas having a higher GHG impact than the acquisition and delivery of an energy-equivalent amount of crude oil.

Besides high carbon efficiency, the system has high energy efficiency with power created by the process being used to run system utilities and produce electricity for sale. While the performance parameters of this fuel-centric system cannot be directly translated to other GTL opportunities (such as chemicals and wax production), they may be used to guide discussion on those subjects.

4.2 Scale and Production

Exhibit 4-1 provides a summary comparison of the feedstock consumed and products produced by this GTL system analysis. The scale of the facility investigated by this analysis was chosen to align with other recent NETL studies regarding synthetic fuels (e.g., coal-to-liquids and coal-and-biomass-to-liquids). The optimal scale of a GTL facility is project-specific; the scale used in this analysis should serve as a reasonable basis for understanding the economics associated with any GTL facility. Consistent with shifts in product demand, the analyzed system produces predominately diesel fuel with some gasoline. Electrical power is a significant co-product. This analysis aimed to maximize the production of liquid transport fuels. Production of electricity was subordinate to this goal and was aimed solely at maximized value recovered within this design basis.

Exhibit 4-1: Summary of GTL Performance

Parameter	Value
Feedstock Input	
Natural Gas Feed Flow, kg/hr (lb/hr)	354,365 (781,241)
Butane Feed Flow, kg/hr (lb/hr)	18,843 (41,542)
Power Production and Use	
Total Power Production (Steam Turbine), kWe	303,700
Internal Power Consumption, kWe	262,900
Net Plant (Export) Power, kWe	40,800
Liquid Products	
Diesel Fuel, bbl/day	34,543
Gasoline, bbl/day	15,460
Total Liquids, bbl/day	50,003

4.3 Carbon and Thermal Efficiencies

Exhibit 4-2 provides and compares selected performance data, thermal efficiencies, and carbon efficiencies from this representation of a commercial-scale GTL facility. Approximately 62 percent of the energy content in the feedstocks is recovered as liquid fuels, which is consistent with the open literature.[8] Recovery through electricity production increases the efficiency another 8.5 percent.

Exhibit 4-2: Thermal and Carbon Efficiencies

Parameter	Value
Feedstock Input	
Natural Gas Feed Flow, kg/hr (lb/hr)	354,365 (781,241)
Butane Feed Flow, kg/hr (lb/hr)	18,843 (41,542)
Power Production and Use	
Total Power Production (Steam Turbine), kWe	303,700
Internal Power Consumption, kWe	262,900
Net Plant (Export) Power, kWe	40,800
Liquid Products	
Diesel Fuel, bbl/day	34,543
Gasoline, bbl/day	15,460
Total Liquids, bbl/day	50,003
LHV Thermal Efficiency	
To Liquid Fuels*	61.1%
To All Products**	66.9%
HHV Thermal Efficiency	
To Liquid Fuels*	62.4%
To All Products**	68.0%
Carbon Efficiency	
Natural Gas to Liquid Products	81.8%

** Liquid fuels include naphtha, diesel, and gasoline*

*** Includes liquid fuels and power*

The carbon efficiency reported here is at the system level, which is a combination of the efficiency of butane conversion to making gasoline additives and of the efficiency of the inherent GTL process. **Exhibit 4-3** provides the carbon balance of this analysis. The majority of the carbon comes out in the liquid fuels. Compressed CO_2 contains the majority of the non-product carbon exiting the facility; this carbon is captured and is not an emission. Stack gases and the small amount of CO_2 that is vented accounts for less than 1.7 percent of the carbon input into the system. These later carbon emissions from the plant do not represent the overall GHG profile, which is discussed in Section 4.6.

Exhibit 4-3: GTL Facility Carbon Balance

Carbon In kg/hr (lb/hr)	
Natural Gas	255,956 (564,286)
Air to O_2 (CO_2)	316 (697)
Butane	15,576 (34,339)
Total	**271,848 (599,322)**

Carbon Out kg/hr (lb/hr)	
Diesel	147,524 (325,234)
Gasoline	61,908 (136,485)
Air Vent (CO_2)	306 (675)
CO_2	57,910 (127,669)
Stack gas	4,200 (9,260)
Total	**271,848 (599,322)**

4.4 Energy Integration

A Fischer-Tropsch facility produces considerable high-grade heat captured for use as high-pressure steam as well as produces a significant quantity of tail gas. Production of high-pressure steam is the common mechanism to recover the considerable heat released in both the synthesis gas production and Fischer-Tropsch synthesis process. The amount of heat content in the tail gas is minimized by recycling back into the process and, when the market warrants, recovery of the LPG components. The production of these high-grade heat sources is inherent in the process.

The high-grade heat must be integrated into the thermal and electrical loads of the Fischer-Tropsch facility. Those loads are set by the process scheme and are understandably minimized; however, considerable residual energy remains. When a Fischer-Tropsch facility is integrated into a broader petrochemical process, the residual heat is used to drive other processes. Specifically, the heat can generate steam for other processes, drive broader wastewater treatment, or drive broader thermal operations such as vis-breaking, isomerizations, or cracking. For example, the naphtha output from a Fischer-Tropsch process is an excellent feedstock for thermal cracking to olefins and other chemicals building blocks within a co-located petro-chemical refinery operation. A GTL facility isolated from a broader petrochemical complex has limited process loads (thermal and electrical) to consume the energy released and recovered in the process and therefore has the opportunity to produce excess electricity for sale.

The process for generating excess electricity uses non-condensable gases that must be purged from the fuel gas recycle stream anyways (the purged gases represent 5 percent of the flow rate of the recycle stream). The purged fuel gas is burned by a boiler to produce high-pressure steam. Steam is also produced from excess heat from the reforming and Fischer-Tropsch reactors. The steam is then fed to a turbine to produce power. The energy balance associated with the GTL process is provided in **Exhibit 4-4** below. The GTL facility steam balance (production and consumption) is provided in

Exhibit 4-5 below representing the portion of the energy recovered as thermal energy (e.g., steam).

The energy content of the feed into the current GTL model is approximately 5.0 million KW. Only ~6 percent of that energy content is converted to electrical power (303,700 KW). However, since there is limited load in the GTL facility, 40,800 KW is exported, or ~0.8 percent of the energy of the natural gas input into the system. This reasonably aligned with literature reports. Specifically, Iandoli and Kjelstrup[26] performed an exergy analysis (tool to evaluate the intelligent use of energy for useful work) of the GTL process. In their analysis, the process exports 72 MW of electric power or ~4.2 percent of the 1,720 MW coming into the system. The greater level of energy export in this analysis may indicate that this systems analysis is overly conservative in its estimates of the recovered energy and/or the parasitic-system loads. However, practical energy recovery is typically well below that which is thermodynamically possible.

Exhibit 4-4: GTL Energy Balance

Parameter	Value	Units
Steam Turbine Power	303,700	kWe
Total Gross Power	**303,700**	**kWe**
Air Separation Unit Auxiliaries	2,000	kWe
Air Separation Unit and Compressor	195,500	kWe
Amine System Auxiliaries	11,200	kWe
CO_2 Compressor	6,700	kWe
Boiler Feedwater Pumps	1,000	kWe
Condensate Pump	400	kWe
Circulating Water Pump	6,100	kWe
Ground Water Pumps	500	kWe
Cooling Tower Fans	3,200	kWe
Steam Turbine Auxiliaries	200	kWe
F-T Power	32,300	kWe
Miscellaneous Balance of Plant[s]	1000	kWe
Transformer Losses	2,800	kWe
Total	**262,900**	**kWe**
Net Plant Power	40,800	kWe
Natural Gas Feed Flowrate	354,365 (781,241)	kg/hr (lb/hr)
Condenser Duty	2,808 (2,660)	GJ/hr (MMBtu/hr)
Raw Water Withdrawal	31.8 (8,403)	m3/min (gpm)
Carbon in Feed	255,291 (564,291)	kg/hr (lb/hr)
Raw Water Consumption	24.4 (6,437)	m3/min (gpm)

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[s] Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

Exhibit 4-5: GTL Steam Balance

Stream Description	Steam Production and Consumption (1,000-lb/hr)					
	360 psig/438°F		600 psig/489°F		50 psig/298°F	
	Produced	Consumed	Produced	Consumed	Produced	Consumed
F-T Synthesis	3,563.48	-	-	35.93	8.71	-
Amine CO_2 Removal	-	87.02	-	-	-	537.87
Compression	-	373.50	-	-	376.42	-
Hydrocarbon Recovery	-	605.02	-	102.79	519.85	-
Wax Hydrocracker	-	220.39	-	71.30	205.26	-
Distillate Hydrotreater	-	-	-	-	3.94	-

4.5 Water Balance

The water balance for the GTL system analysis is provided in **Exhibit 4-6** and includes water present in both the gasifier and Fischer-Tropsch plant models. The water produced in the Fischer-Tropsch section is reused following processing.

Exhibit 4-6: GTL Water Balance

Water Use (m^3/min (gpm))	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
Amine System	0.06 (15)	4.5 (1,198)	0.06 (15)	-	0.1 (15)
Pre-reformer	4.53 (1,198)	-	4.53 (1,198)	-	4.53 (1,198)
F-T Area	-1.01 (-266)	-6.57 (-1,735)	-1.01 (-266)	-	-1.01 (-266)
SMR	0.7 (184)	-	0.7 (184)	-	0.7 (184)
Condenser Makeup	0.03 (7)	-	0.03 (7)	-	-
Cooling Tower	31.21 (8,244)	3.71 (979)	27.5 (7,265)	7.42 (1,959)	20.09 (5,306)
Total	**35.51 (9,382)**	**-2.86 (-756)**	**31.81 (8,403)**	**7.42 (1,959)**	**24.37 (6,437)**

Exhibit 4-7 provides another viewpoint on the water balance, where water is converted from gpm to bbl/hr. This allows for an understanding of the water demand associated with the production of liquid fuels.

Exhibit 4-7: Water Balances: Water Relative to Fischer-Tropsch Liquid Production

Process	Value (bbl/hr H_2O)/(bbl/hr F-T Liquids)
Water Demand	4.72
Internal Recycle	-0.38
Raw Water Withdrawal	4.23
Process Water Discharge	0.99
Raw Water Consumption	3.24

4.6 Emissions and Greenhouse Gas Impact

The purpose of the environmental control equipment used in this study was to conform to the BACT guidelines, which details allowable limits on various pollutants, such as SO_2, NOx, particulates, and mercury. Appendix 7.1 describes the technical performance of the environmental control equipment. **Exhibit 4-8** provides the air emission values of those pollutants from all flue gas emitted from the plants.

Exhibit 4-8: GTL Air Emissions

Air Emissions	GTL kg/GJ (lb/MMBtu)	GTL tonne/yr (TPY)	GTL kg/bbl (lb/bbl)
SO_2	0 (0)	0 (0)	0 (0)
NOx	0.003 (0.006)	3 (4)	0.001 (0.003)
PM	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO_2	69.0 (160.4)	87,664 (96,634)	39 (85)

The purpose of CO_2 removal at a GTL plant, and other commercial-scale Fischer-Tropsch (F-T) facilities, is to reduce the circulation of non-reactive gases that would otherwise build up in the F-T recycle loop. CO_2 removal also minimizes equipment sizes and costs. The CO_2 capture technology in this analysis is based on a proprietary technology that uses monoethanolamine (MEA) to absorb CO_2 and is followed by a solvent regeneration process that produces pure CO_2. In this analysis, the CO_2 capture system removes 93 percent of the CO_2 from the synthesis gas stream. If this captured CO_2 is sequestered instead of vented, it could reduce the CO_2 emissions of the GTL plant by 93 percent.

4.6.1.1 Greenhouse Gas Emissions Profile

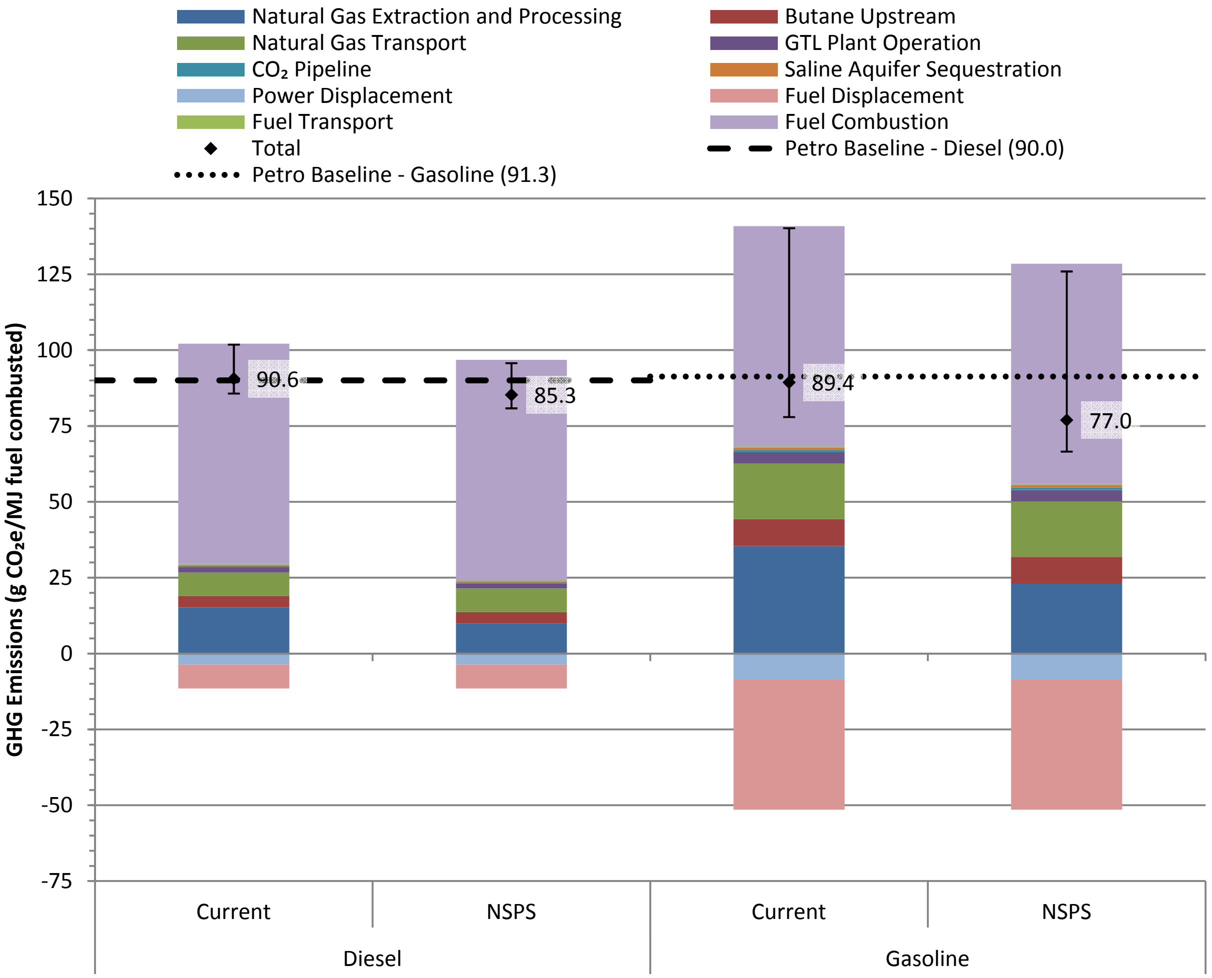
The life cycle analysis (LCA) results of this analysis are restricted to GHG emissions, expressed as carbon dioxide equivalents (CO_2e) using Intergovernmental Panel on Climate Change (IPCC) 2007 100-yr global warming potentials. Scenarios for diesel and gasoline were assessed, using 1 MJ of combusted fuel as the functional unit (the basis of comparison).

The GTL facility produces multiple products, so this analysis uses co-product displacement to express GHG results for one product at a time. The objective of LCA is to assign ownership of environmental burdens to a single product or function. Displacement and allocation are two options for co-product management. Displacement expands system boundaries to envelop alternative pathways that are displaced by co-products, thus isolating one product of interest. Allocation uses a physical or economic relationship as a basis for splitting burdens among co-products. This analysis uses displacement for co-product management. Allocation is not used in this analysis because it requires comparison of two forms of energy (electricity and liquid fuel) that are not used for similar applications and do not have the same degree of usefulness. More details on co-product management are provided in **Section 7.3.6**.

The LCA also accounts for near-term improvement to upstream natural gas emissions. Results were generated for current practices in the natural gas industry and for improved performance once the approved New Source Performance Standards (NSPS) for the Oil and Gas Sector are fully implemented. The combination of the two functional units (diesel and gasoline) and two upstream natural gas practices (current and NSPS) yield four scenarios for the LCA GHG results. The GHG results for these scenarios are shown in **Exhibit 4-9**.

The life cycle GHG emissions for GTL diesel and gasoline when based on current practices in the natural gas industry are 90.6 g CO_2e/MJ and 89.4 g CO_2e/MJ, respectively. If the natural gas extraction and processing sector complies with NSPS, the upstream GHG emissions from natural gas are reduced by 23 percent. With the implementation of NSPS standards, the life cycle values for GTL diesel and gasoline decrease to 85.3 g CO_2e/MJ (5.9 percent reduction) and 77.0 g CO_2e/MJ (13.9 percent reduction), respectively. In **Exhibit 4-9**, these values are compared to the NETL petroleum baseline values for petroleum based fuels, which are 90.0 and 91.3 g CO_2e/MJ for diesel and gasoline, respectively. The expected LCA GHG emissions for GTL diesel are 0.6 percent higher than the baseline without NSPS, and 5.3 percent lower with the implementation of NSPS. For GTL gasoline, the LCA GHG emissions are 2.1 percent lower than the baseline without NSPS and 15.7 percent lower with the implementation of NSPS. The NSPS regulations will reduce the venting rates from well development and maintenance, increase the flaring rates at natural gas wells, and reduce compressor emissions at natural gas processing plants; this will lead to an overall reduction in the GHG emissions from natural gas extraction and processing.

Exhibit 4-9: LCA Greenhouse Gas Results for GTL Diesel and Gasoline



5 Cost Summary

The scoping-level economic assessment of this GTL system study is reported below, providing a basis for discussing the viability of GTL projects and opportunities to improve their viability. Costs are reported on a 2011 dollar basis.

5.1 Capital Expenditures

Capital expenditures include several components that build up to a total-as-spent cost (TASC). Bare erected costs cover the cost of process equipment and required supporting facilities and infrastructure as well as the labor costs of installing such equipment. Total plant costs include the bare erected costs and the engineering, procurement, and construction costs as well as process and project contingency costs. Total overnight capital contains the total plant costs plus overnight costs, including pre-production costs, inventory capital, financing costs, and other owner's costs. The TASC includes the total overnight costs and the escalation of costs and interest on debt during the capital expenditure period, which for this study was taken as five years.

Estimated capital costs are listed in **Exhibit 5-1**. Account numbers for capital equipment reference EPRI account numbers to allow for comparison with other NETL studies[t]. However, the capital expenditure table is organized based on functional processing area (e.g., synthesis gas production, Fischer-Tropsch reactor). These estimates are consistent with NETL Quality Guidelines for Energy System Studies[5] and have an expected accuracy of -15 percent to +30 percent. Cost escalation and uncertainty between project execution and project completion have plagued GTL projects (e.g., Escravos and Pearl), lending caution on how declarative GTL cost estimates can be.

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[t] The EPRI account numbers are pulled from historical NETL reports to serve as a basis for comparison between current and past NETL studies. These account numbers represent over 20 years of cost analyses. The reconciliation of the EPRI account numbers used in this report to an existing EPRI report would require NETL to purchase an EPRI report.

Exhibit 5-1: GTL Capital Costs

Acct		Equipment	Material	Labor	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST			
No.	Item/Description	Cost	Cost	Cost	Cost $	H.O.& Fee	Process	Project	$	$/bpd	$\$/\text{bpd}_{\text{ECO}}$	$\$/\text{bpd}_{\text{EPD}}$
	Synthesis Gas Production											
4.1	ASU/Oxidant Compression	$248,399	w/equip.	w/equip.	$248,399	$24,077	$0	$27,248	$299,725	$5,994	$6,463	$9,008
5A	Pre-reformer	$188,992	w/equip.	w/equip.	$188,992	$18,143	$51,028	$64,541	$322,704	$6,454	$6,958	$9,698
5A	Autothermal Reformer	$238,116	w/equip.	w/equip.	$238,116	$22,859	$64,291	$81,316	$406,582	$8,131	$8,767	$12,219
	SUBTOTAL Synthesis Gas Production	**$675,507**	**$0**	**$0**	**$675,507**	**$65,080**	**$115,319**	**$173,105**	**$1,029,011**	**$20,579**	**$22,188**	**$30,925**
	Fischer Tropsch Synthesis											
5A	FT Synthesis	$144,254	w/equip.	w/equip.	$144,254	$13,848	$38,949	$49,263	$246,314	$4,926	$5,311	$7,403
	SUBTOTAL Fischer Tropsch Synthesis	**$144,254**	**$0**	**$0**	**$144,254**	**$13,848**	**$38,949**	**$49,263**	**$246,314**	**$4,926**	**$5,311**	**$7,403**
	Product Recovery											
5A	Overhead recovery	$24,115	w/equip.	w/equip.	$24,115	$2,315	$6,511	$8,235	$41,176	$823	$888	$1,237
5A	Liquids recovery and fractionation	$33,838	w/equip.	w/equip.	$33,838	$3,248	$9,136	$11,556	$57,779	$1,156	$1,246	$1,736
5B.1	CO$_2$ Removal System	$141,953	$0	$42,776	$184,729	$15,413	$36,946	$47,418	$284,506	$5,690	$6,135	$8,550
5B.2	CO$_2$ Compression & Drying	$12,275	$0	$4,698	$16,974	$1,420	$0	$3,679	$22,072	$441	$476	$663
	SUBTOTAL Product Recovery	**$212,181**	**$0**	**$47,475**	**$259,656**	**$22,397**	**$52,593**	**$70,887**	**$405,533**	**$8,110**	**$8,744**	**$12,188**
	Product Upgrading											
5C	Hydrogen Production: Steam Methane Reformer	$34,439	w/equip.	w/equip.	$34,439	$3,306	$9,299	$11,761	$58,805	$1,176	$1,268	$1,767
5C	Hydrogen Production: Pressure Swing Absorber	$7,303	w/equip.	w/equip.	$7,303	$701	$1,972	$2,494	$12,469	$249	$269	$375
5C	Hydrogen Production: Miscellaneous	$19,814	w/equip.	w/equip.	$19,814	$1,902	$5,350	$6,767	$33,833	$677	$730	$1,017
5D	Naphtha Hydrotreater	$10,454	w/equip.	w/equip.	$10,454	$1,004	$2,822	$3,570	$17,850	$357	$385	$536
5D	Naphtha Reformer	$40,895	w/equip.	w/equip.	$40,895	$3,926	$11,042	$13,966	$69,828	$1,396	$1,506	$2,099
5D	C5/C6 Isomerization	$10,686	w/equip.	w/equip.	$10,686	$1,026	$2,885	$3,649	$18,246	$365	$393	$548
5D	C4 Isomerization	$13,591	w/equip.	w/equip.	$13,591	$1,305	$3,670	$4,641	$23,207	$464	$500	$697
5D	C3/C4/C5 Alkylation	$52,118	w/equip.	w/equip.	$52,118	$5,003	$14,072	$17,798	$88,991	$1,780	$1,919	$2,674
5E	Diesel Hydrotreater	$29,555	w/equip.	w/equip.	$29,555	$2,837	$7,980	$10,093	$50,465	$1,009	$1,088	$1,517
5F	Wax Hydrocracker	$82,299	w/equip.	w/equip.	$82,299	$7,901	$22,221	$28,105	$140,526	$2,810	$3,030	$4,223
	SUBTOTAL Product Upgrading	**$301,154**	**$0**	**$0**	**$301,154**	**$28,911**	**$81,312**	**$102,844**	**$514,221**	**$10,284**	**$11,088**	**$15,454**
	Gas Loop											
5G	Raw Fuel Gas Compressor, HP	$35,229	w/equip.	w/equip.	$35,229	$2,947	$0	$7,635	$45,811	$916	$988	$1,377
5G	Fuel Gas to Boiler, HP	$6,613	w/equip.	w/equip.	$6,613	$553	$0	$1,433	$8,599	$172	$185	$258
	SUBTOTAL Gas Loop	**$41,842**	**$0**	**$0**	**$41,842**	**$3,500**	**$0**	**$9,068**	**$54,410**	**$1,088**	**$1,173**	**$1,635**
	Product Storage											
5G	Gasoline Storage Tank	$823	w/equip.	w/equip.	$823	$79	$0	$226	$1,128	$23	$24	$34
5G	Diesel Storage Tank	$1,311	w/equip.	w/equip.	$1,311	$126	$0	$359	$1,796	$36	$39	$54
	SUBTOTAL Product Storage	**$2,134**	**$0**	**$0**	**$2,134**	**$205**	**$0**	**$585**	**$2,924**	**$58**	**$63**	**$88**

Exhibit 5-1: GTL Capital Costs (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Cost	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST			
							Process	Project	$	$/bpd	$/bpd$_{ECO}$	$/bpd$_{EPD}$
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	Feedwater System	$1,162	$1,203	$983	$3,348	$270	$0	$543	$4,160	$83	$90	$125
3.2	Water Makeup & Pretreating	$4,374	$452	$2,245	$7,071	$584	$0	$1,531	$9,186	$184	$198	$276
3.3	Other Feedwater Subsystems	$544	$180	$150	$875	$68	$0	$141	$1,083	$22	$23	$33
3.4	Service Water Systems	$528	$1,053	$3,394	$4,975	$423	$0	$1,080	$6,478	$130	$140	$195
3.5	Other Boiler Plant Systems	$3,559	$1,330	$3,059	$7,947	$647	$0	$1,289	$9,884	$198	$213	$297
3.6	Natural Gas, incl. pipeline	$25,176	$818	$708	$26,701	$2,201	$0	$4,335	$33,238	$665	$717	$999
3.7	Waste Treatment Equipment	$1,528	$0	$884	$2,412	$208	$0	$524	$3,144	$63	$68	$94
3.8	Misc. Equip. (cranes, Air Comp., Comm.)	$1,563	$209	$759	$2,531	$216	$0	$549	$3,296	$66	$71	$99
	SUBTOTAL 3.	**$38,434**	**$5,245**	**$12,182**	**$55,860**	**$4,617**	**$0**	**$9,993**	**$70,469**	**$1,409**	**$1,519**	**$2,118**
7	BOILER HRSG, DUCTING & STACK											
7.1	Heat Recovery / Steam Generation	$13,176	$0	$2,440	$15,616	$1,297	$0	$1,691	$18,605	$372	$401	$559
7.2	Waste Heat Boiler	$3,173	$0	$588	$3,760	$315	$0	$611	$4,686	$94	$101	$141
7.3	Ductwork	$0	$58	$42	$100	$10	$0	$27	$137	$3	$3	$4
7.4	Stack	$113	$0	$43	$156	$15	$0	$43	$214	$4	$5	$6
7.9	HRSG,Boiler,Duct & Stack Foundations	$0	$44	$41	$85	$7	$0	$19	$111	$2	$2	$3
	SUBTOTAL 7.	**$16,462**	**$102**	**$3,153**	**$19,717**	**$1,644**	**$0**	**$2,391**	**$23,752**	**$475**	**$512**	**$714**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$38,812	$0	$6,543	$45,355	$3,582	$0	$4,894	$53,831	$1,077	$1,161	$1,618
8.2	Turbine Plant Auxiliaries	$239	$0	$531	$770	$66	$0	$84	$920	$18	$20	$28
8.3	Condenser & Auxiliaries	$5,060	$0	$3,427	$8,488	$715	$0	$920	$10,123	$202	$218	$304
8.4	Steam Piping	$4,541	$0	$1,840	$6,381	$435	$0	$1,022	$7,838	$157	$169	$236
8.9	TG Foundations	$0	$1,229	$2,030	$3,259	$275	$0	$707	$4,241	$85	$91	$127
	SUBTOTAL 8.	**$48,652**	**$1,229**	**$14,371**	**$64,252**	**$5,073**	**$0**	**$7,627**	**$76,952**	**$1,539**	**$1,659**	**$2,313**
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$13,378	$0	$4,054	$17,432	$1,455	$0	$1,889	$20,775	$415	$448	$624
9.2	Circulating Water Pumps	$2,622	$0	$162	$2,784	$212	$0	$300	$3,296	$66	$71	$99
9.3	Circulating Water System Auxiliaries	$201	$0	$27	$227	$19	$0	$25	$271	$5	$6	$8
9.4	Circulating Water Piping	$0	$6,232	$1,411	$7,643	$573	$0	$1,232	$9,449	$189	$204	$284
9.5	Make-up Water System	$607	$0	$780	$1,386	$115	$0	$225	$1,726	$35	$37	$52
9.6	Component Cooling Water Sys	$401	$479	$308	$1,188	$95	$0	$192	$1,475	$30	$32	$44
9.9	Circulating Water System Foundations	$0	$2,832	$4,703	$7,536	$636	$0	$1,634	$9,806	$196	$211	$295
	SUBTOTAL 9.	**$17,209**	**$9,544**	**$11,444**	**$38,197**	**$3,104**	**$0**	**$5,497**	**$46,798**	**$936**	**$1,009**	**$1,406**

Exhibit 5-1: GTL Capital Costs (Continued)

Acct		Equipment	Material	Labor	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST			
No.	Item/Description	Cost	Cost	Cost	Cost $	H.O.& Fee	Process	Project	$	$/bpd	$\$/\text{bpd}_{ECO}$	$\$/\text{bpd}_{EPD}$
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$4,807	$0	$2,843	$7,650	$628	$0	$621	$8,899	$178	$192	$267
11.2	Station Service Equipment	$4,719	$0	$406	$5,125	$422	$0	$416	$5,963	$119	$129	$179
11.3	Switchgear & Motor Control	$5,807	$0	$1,009	$6,816	$565	$0	$738	$8,119	$162	$175	$244
11.4	Conduit & Cable Tray	$0	$3,035	$8,743	$11,778	$981	$0	$1,914	$14,673	$293	$316	$441
11.5	Wire & Cable	$0	$9,752	$5,545	$15,298	$920	$0	$2,433	$18,650	$373	$402	$560
11.6	Protective Equipment	$0	$661	$2,294	$2,956	$254	$0	$321	$3,531	$71	$76	$106
11.7	Standby Equipment	$90	$0	$84	$174	$15	$0	$19	$208	$4	$4	$6
11.8	Main Power Transformers	$8,774	$0	$112	$8,886	$603	$0	$949	$10,439	$209	$225	$314
11.9	Electrical Foundations	$0	$97	$248	$345	$29	$0	$75	$449	$9	$10	$13
	SUBTOTAL 11.	**$24,197**	**$13,546**	**$21,284**	**$59,027**	**$4,417**	**$0**	**$7,485**	**$70,930**	**$1,419**	**$1,529**	**$2,132**
12	INSTRUMENTATION & CONTROL											
12.3	Steam Turbine Control	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$9,151	$0	$5,826	$14,977	$1,258	$749	$2,548	$19,531	$391	$421	$587
12.5	Signal Processing Equipment	$7,294	$0	$668	$7,962	$0	$0	$0	$7,962	$159	$172	$239
12.6	Control Boards,Panels & Racks	$2,734	$0	$1,672	$4,406	$370	$220	$749	$5,745	$115	$124	$173
12.7	Computer & Accessories	$14,588	$0	$1,336	$15,924	$1,307	$796	$1,803	$19,830	$397	$428	$596
12.8	Instrument Wiring & Tubing	$0	$8,142	$14,405	$22,547	$1,620	$1,127	$3,794	$29,089	$582	$627	$874
12.9	Other I & C Equipment	$10,876	$0	$25,182	$36,058	$3,086	$1,803	$4,095	$45,041	$901	$971	$1,354
	SUBTOTAL 12.	**$44,643**	**$8,142**	**$49,088**	**$101,873**	**$7,641**	**$4,696**	**$12,988**	**$127,198**	**$2,544**	**$2,743**	**$3,823**
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$338	$7,177	$7,515	$656	$0	$1,634	$9,806	$196	$211	$295
13.2	Site Improvements	$0	$3,097	$4,092	$7,190	$639	$0	$1,566	$9,395	$188	$203	$282
13.3	Site Facilities	$6,324	$0	$6,634	$12,958	$1,155	$0	$2,823	$16,936	$339	$365	$509
	SUBTOTAL 13.	**$6,324**	**$3,435**	**$17,904**	**$27,663**	**$2,451**	**$0**	**$6,023**	**$36,137**	**$723**	**$779**	**$1,086**
14	BUILDINGS & STRUCTURES											
14.2	Steam Turbine Building	$0	$6,402	$8,518	$14,919	$1,147	$0	$2,410	$18,477	$370	$398	$555
14.3	Administration Building	$0	$3,607	$2,444	$6,050	$471	$0	$978	$7,499	$150	$162	$225
14.4	Circulating Water Pumphouse	$0	$574	$284	$858	$66	$0	$138	$1,062	$21	$23	$32
14.5	Water Treatment Buildings	$0	$1,789	$1,630	$3,419	$270	$0	$553	$4,243	$85	$91	$128
14.6	Machine Shop	$0	$3,129	$2,000	$5,129	$398	$0	$829	$6,356	$127	$137	$191
14.7	Warehouse	$0	$1,010	$609	$1,620	$125	$0	$262	$2,007	$40	$43	$60
14.8	Other Buildings & Structures	$0	$605	$440	$1,045	$82	$0	$169	$1,296	$26	$28	$39
14.9	Waste Treating Building & Structures	$0	$1,184	$2,113	$3,297	$269	$0	$535	$4,100	$82	$88	$123
	SUBTOTAL 14.	**$0**	**$18,300**	**$18,038**	**$36,338**	**$2,828**	**$0**	**$5,875**	**$45,041**	**$901**	**$971**	**$1,354**

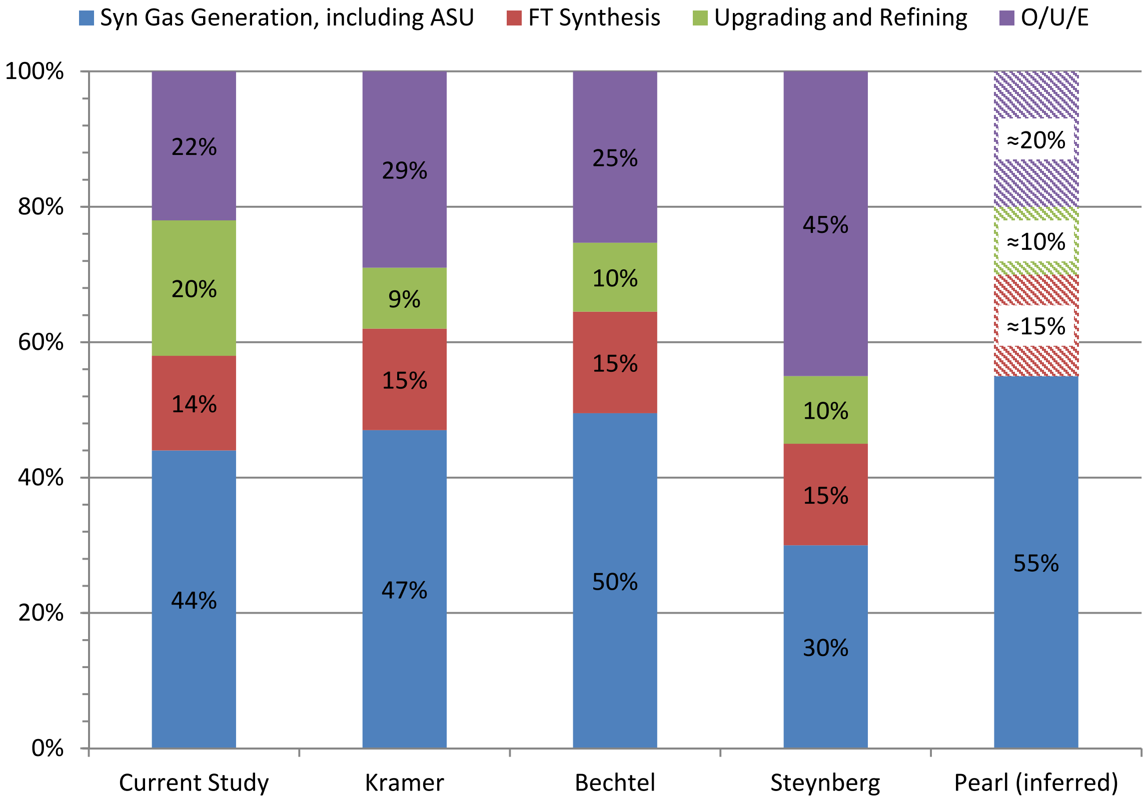
Exhibit 5-1: GTL Capital Costs (Continued)

	Equipment	Material	Labor	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST			
Item/Description	**Cost**	**Cost**	**Cost**	**Cost $**	**H.O.& Fee**	**Process**	**Project**	**$**	**$/bpd**	**/bpd_{ECO}**	**/bpd_{EPD}**
TOTAL COST	**$1,572,992**	**$59,544**	**$194,939**	**$1,827,475**	**$165,715**	**$292,868**	**$463,631**	**$2,749,689**	**$54,990**	**$59,290**	**$82,637**
Owner's Costs											
Preproduction Costs											
6 Months All Labor								$109,557	$2,191	$2,362	$3,293
1 Month Maintenance Materials								$6,004	$120	$129	$180
1 Month Non-feedstock Consumables								$1,946	$39	$42	$58
25% of 1 Months Feedstock Cost at 100% CF								$22,221	$444	$479	$668
2% of TPC								$54,994	$1,100	$1,186	$1,653
Total								**$194,722**	**$3,894**	**$4,199**	**$5,852**
Inventory Capital											
60 Day Supply of Consumables at 100% CF								$178,566	$3,571	$3,850	$5,366
0.5% of TPC (spare parts)								$13,748	$275	$296	$413
Total								**$192,314**	**$3,846**	**$4,147**	**$5,780**
Initial Cost for Catalyst and Chemicals								$24,247	$485	$523	$729
Land								$900	$18	$19	$27
Other Owner's Costs								$412,453	$8,249	$8,893	$12,396
Financing Costs								$74,242	$1,485	$1,601	$2,231
Total Overnight Costs (TOC)								**$3,648,567**	**$72,967**	**$78,672**	**$109,651**
TASC Multiplier				Commercial Fuels (5 years of construction) + 30 years life				1.181			
Total As-Spent Cost (TASC)								**$4,309,687**	**$86,188**	**$92,927**	**$129,520**

Exhibit 5-2 breaks down GTL capital costs between synthesis gas production, Fischer-Tropsch synthesis; product upgrading and refining; and offsites, utilities, and electrical (O/U/E) sections of this analysis, previous analyses, and the Shell Pearl GTL project. The studies labeled Kramer[20] and Bechtel[11] were also sponsored by NETL.[u] The study labeled Steynberg refers to an open literature study.3 The entry labeled Pearl provides an estimate of capital expenditure breakdown of Shell's GTL project in Qatar inferred from open literature sources.[8, 9, 15] A high-level summary of these reference projects are provided in **Exhibit 2-2** above.

Reconciling allocations between these broad cost accounts requires some judgment and interpretation. As such, these numbers are provided solely to give general guidance on which components drive the capital requirements of GTL projects. The breakdown of the capital outlay in this study is consistent with previous work and industrial projects. This system analysis has a higher allocation of capital to product upgrading and refining, consistent with its production of finished fuels products rather than chemicals and blendstocks.

Exhibit 5-2: Breakdown of Capital Costs by Processing Section



The total capital expenditure estimate of this GTL system analysis, summarized in the metric of $86,188 per daily barrel of Fischer-Tropsch liquids, is in line with expectations based on

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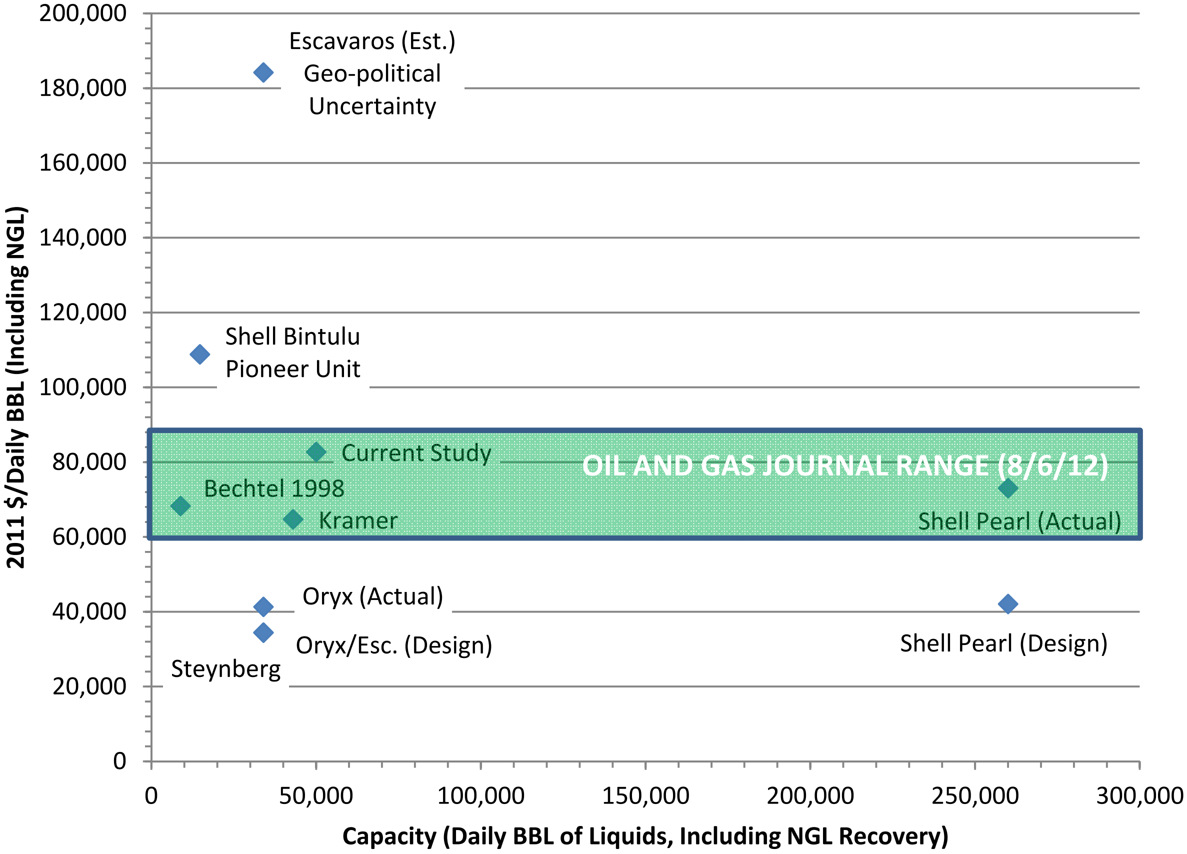
[u] These studies were sponsored by the Pittsburgh Energy Technology Center and the Federal Energy Technology Center, which became the National Energy Technology Laboratory.

previous studies and actual GTL projects. A comparison of the capital costs per daily barrel of Fischer-Tropsch liquids are summarized in **Exhibit 5-3** below. To allow for consistent comparison, overall system cost estimates were crudely[v] adjusted to 2011 dollars via cost index. The green bar indicates the range of capital costs provided in a recent Oil and Gas Journal article.[27]

Exhibit 5-3: Capital Costs Per Daily Barrel of Fischer-Tropsch Liquids from GTL Projects and Studies



The studies used in the capital cost breakdown in **Exhibit 5-2**, Kramer,[20] Bechtel,[11] and Steynberg,[3] are also present in **Exhibit 5-3**. The entries for implemented GTL projects in Qatar (Shell Pearl and Oryx) and Nigeria (Escravos) are based on open literature sources.[8, 9, 15] Best efforts were made to compare projects on a consistent basis, such as not including contributions from natural gas processing and natural gas liquids recovery in the capital estimates since these were outside the boundary of analysis.

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[v] The bulk capital cost per daily barrel number was moved based on the ratio of the 2011 cost index to the relevant year’s cost index. To be done rigorously, each sub-account in the capital costs estimate should be indexed separately using a more sub-account specific cost index; such detail was not readily available for all projects and such fidelity was not required for these general context comparisons.

This current systems analysis is higher than, but fairly consistent with, the previous NETL-sponsored systems analysis, which is expected based on the use of institutional knowledge in conducting this analysis. A 2011 capital cost estimate from Eni[28] of ~ $120,000/bbl for a 17,000 bpd facility has been previewed in the synthetic fuels community. Assuming a 0.7 exponent for economies of scale, this Eni estimate would predict a capital cost for the scale of this system study (50,003 bbl/day) of $86,819. However, without the details of the Eni estimate, it is not possible to definitively assess how well it does or does not align to the cost estimates of this study or the $60,000-$85,000 range reported in the Oil and Gas Journal.[27] The Eni estimate is less than 0.75 percent above the estimate from this study.

The Steynberg study estimated capital outlays very close to the design estimate of the Oryx and Escravos projects. Operating issues at the Oryx facility led to additional capital outlays, which increased the required capital (reflect as a separated data point in **Exhibit 5-3**). The Escravos project, using the same technology as Oryx, has had severe cost escalation related to uncertainty in the region. The estimated Escravos capital cost is reported in the figure but should be considered an anomaly. The design number and realized capital costs per daily barrel for Oryx are significantly less than those of this study. This is to be expected since these projects have fairly simplified product recovery and upgrading.

The Shell Bintulu facility was a pioneer plant leading to higher capital costs per daily barrel. This facility was a first-of-a-kind and could not benefit from cost reductions of the experience curve. Also, the facility had the objective of finishing the development of Shell’s GTL technology that was later used at Pearl and therefore had additional costs. Furthermore, the Bintulu facility produces some high-grade specialty chemical products that also skew the capital costs estimates.

Shell’s Pearl facility capital costs per daily barrel are expected to be significantly less than this study due to economies of scale. The initial capital estimates of the Pearl project are consistent with this expectation. However, the realized capital costs were significantly higher for the Pearl project. Several spurious factors during the period of construction contributed to the escalation at Pearl, such as the 2008 world banking crisis and the impact of rapid economic growth in India and China on the costs of capital goods and construction. There is not a fundamental reason to believe the realized capital costs at Pearl reflect a true increase in capital required for a GTL facility.

5.2 Operating Expenditures

The estimated operating and maintenance costs are listed in **Exhibit 5-4** and were developed using NETL’s standard method.[5, 12] A base-case natural gas price of $6.13/MMBTU was chosen to be consistent with NETL Quality Guidelines for Energy System Studies for a 2011 base year. Market conditions leading to considerably lower natural gas prices are reflected in the sensitivity analysis discussed in Section 5.3. Liquid product prices and butane input prices were based on market conditions of October 5, 2012.

Exhibit 5-4: Operating and Maintenance Costs

					Cost Base (Jun):	2011
Gas-to-Liquids Facility					bbl/day:	50,003
					Effective Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate (base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
Operating Labor Requirements (O.J.) per Shift:			Total Plant			
Skilled Operator	21.0		84.0			
Operator	42.0		168.0			
Foreman	15.0		60.0			
Lab Tech's, etc.	15.0		60.0			
TOTAL-O.J.'s	93.0		372.0			
					Annual Cost $	Annual Unit Cost $/(bbl/day)
Annual Operating Labor Cost					$146,789,042	$2.936
Maintenance Labor Cost					$28,501,783	$0.570
Administrative & Support Labor					$43,822,706	$0.876
Property Taxes and Insurance					$36,549,496	$0.731
TOTAL FIXED OPERATING COSTS					**$255,663,028**	**$5.113**
VARIABLE OPERATING COSTS						$/bbl
Maintenance Material Cost					**$64,845,675**	**$3.94773**
Consumables	Consumption Initial Fill	Consumption /Day	Unit Cost	Initial Fill Cost		
Water (/1000 gallons)	0	6,050	1.67	**$0**	**$3,327,101**	**$0.20255**
Chemicals		6				
MU & WT Chem. (lbs)	0.0	36,046	0.27	$0	$3,171,504	$0.19308
MEA Solvent (ton)	401.4	0.56	3,481.91	$1,397,776	$644,827	$0.03926
Activated Carbon (lb)	0.0	629.07	1.63	$0	$335,853	$0.02045
Corrosion Inhibitor	0.0	0.00	0.00	$54,997	$2,619	$0.00016
Natural Gas Reforming Catalyst (ft^3)	21283.5	19.43	650.00	$13,834,277	$4,148,659	$0.25257
FT Catalyst (lbs)	1103614.9	3,858.22	7.15	$7,890,846	$9,062,083	$0.55169
Hydrotreating Catalyst (ft^3)	843.9	0.77	700.00	$590,751	$177,156	$0.01079
Naphtha reforming catalyst (ft^3)	304.6	0.28	900.00	$274,174	$82,220	$0.00501
Isomerization Catalysts (ft^3)	407.6	0.37	500.00	$203,778	$61,110	$0.00372
Subtotal Chemicals				**$24,246,599**	**$17,686,030**	**$1.07671**
Other						
Butane (tons)		499	$651.34	$0	$106,663,845	$6.49358
Subtotal Other				**$0**	**$106,663,845**	**$6.49358**
By-products						
Supplemental Electricity (for sale) (MWh)		980	58.59	$0	($18,858,803)	-$1.14810
Subtotal By-products				**$0**	**($18,858,803)**	**-$1.14810**
TOTAL VARIABLE OPERATING COSTS				**$24,246,599**	**$173,663,847**	**$10.57246**
Natural Gas Feedstock (MMBtu)		423,745	6.13	**$0**	**$853,297,428**	**$51.94780**

NETL's Power System Financial Model was used to determine appropriate capital recovery factors in setting a cost of production of Fischer-Tropsch liquids. The cost model accounts for significant capital and operation and maintenance (O&M) expenses incurred by the GTL system during construction and operation. It is a discounted cash flow analysis over the lifetime of a natural gas power plant, which includes a construction and operating period. Capital costs are expressed in terms of total-as-spent capital (TASC), which, as discussed above, is the sum of total overnight costs (TOC) as well as interest during construction, escalation, and other financial assumptions. Key financial parameters are listed in **Exhibit 5-5**. The construction period is 5 years, making 2015 the first year of operation. The operating period is 30 years, making 2045 the last year of operation. Therefore, the total time frame of the GTL model is 35 years (5 years of construction and 30 years of operation). As a discounted cash flow model, it includes the nominal dollar expenditures during each year of construction and operation; all costs are escalated with respect to annual inflation rates, and the interest accumulated on the debt portion of capital is accounted for during the construction period. All cost results are expressed in 2011 dollars, because capital expenditures start in 2011, the first year of construction.

Exhibit 5-5 Economic and Financial Parameter Assumptions

Parameter	Value
Taxes	
Income Tax Rate	38% (Effective 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 years
Contracting and Financing Terms	
Contracting Strategy	Engineering Procurement Construction Management (owner assumes project risks for performance, schedule and cost)
Type of Debt Financing	Non-Recourse (collateral that secures debt is limited to the real assets of the project)
Repayment Term of Debt	15 years
Grace Period on Debt Repayment	0 years
Debt Reserve Fund	None
Analysis Time Periods	
Capital Expenditure Period	5 Years
Operational Period	30 years
Economic Analysis Period (Used for IRROE)	35 Years (capital expenditure period plus operational period)
Treatment of Capital Costs	
Capital Cost Escalation During Capital Expenditure Period (Nominal Annual Rate)	3.6%[w]
Distribution of Total Overnight Capital Over the Capital Expenditure Period (Before Escalation)	5-Year Period: 10%, 30%, 25%, 20%, 15%

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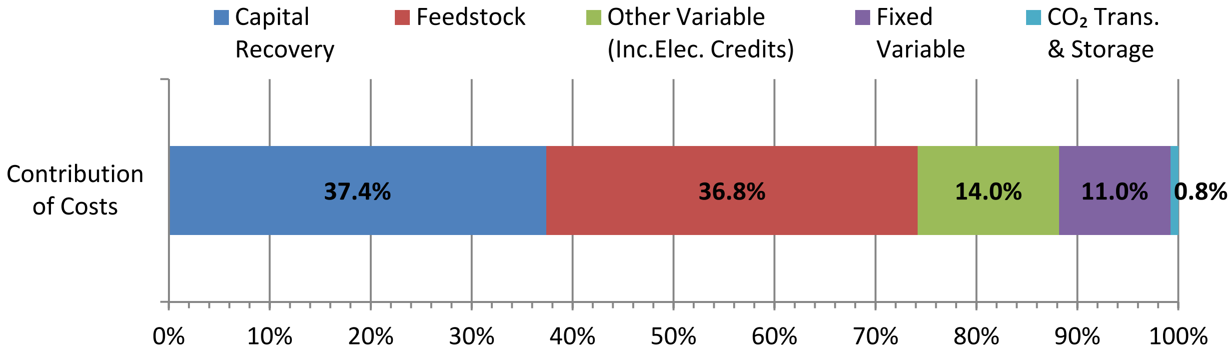
[w] A nominal average annual rate of 3.6 percent is assumed for escalation of capital costs during construction. This rate is equivalent to the nominal average annual escalation rate for process plant construction costs between 1947 and 2008 according to the Chemical Engineering Plant Cost Index. (http://www.che.com/pci/)

Parameter	Value
Working Capital	zero for all parameters
% of Total Overnight Capital Depreciated	100% (this assumption introduces a very small error even if a substantial amount of TOC is actually non-depreciable)
Escalation of Operating Revenues and Costs	
RSP, O&M, COE, and Fuel Costs (Nominal Annual Rate)	3.0%[x]
Financing Structure for Commercial Fuels Projects[y]	
Debt % of Total	50.00%
Debt Interest Rate	8.0% (LIBOR + 4.5%)
Equity Percent of Total	50.00%
Required Return on Equity	20.00%
Capital Charge Factor	0.236
TASC/TOC	1.181

By combining the capital recovery with the operating costs (**Exhibit 5-4**), an estimate of the relative contributions to the cost of production of Fischer-Tropsch liquids can be made. These estimates are reported in **Exhibit 5-6** below. Carbon capture, transportation, and storage would constitute less than one percent of the cost of production. Close to three quarters of the cost of producing Fischer-Tropsch liquids is the cost of feedstock and capital recovery (roughly split evenly between the two items). This indicates that favorable feedstock pricing and reducing capital costs are the levers to target for improving the viability of Fischer-Tropsch based GTL projects.

Exhibit 5-6 Approximate Contribution to Cost of Production of Fischer-Tropsch Liquids



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[x] An average annual inflation rate of 3 percent is assumed. This rate is equivalent to the average annual escalation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods, the so-called "headline" index of the various Producer Price Indices (http://www.bls.gov/ppi/). The Producer Price Index for the Electric Power Generation Industry may be more applicable, but that data does not provide a long-term historical perspective since it only dates back to December 2003.

[y] Source: "Recommended Project Finance Structures for the Economic Analysis of Fossil-Based Energy Projects," Rev 1,2011. http://www.netl.doe.gov/energy-analyses/refshelf/PubDetails.aspx?Action=View&PubId=401

5.3 Sensitivity Analysis

Sensitivity analyses were performed to provide insight on the impact of shifting market conditions and the potential impact of technology improvements. These analyses also expanded the results of the study beyond the base point and mitigate the effect of uncertainty in the system analysis.

5.3.1 Sensitivity to Natural Gas and Liquid Fuels Prices

Exhibit 5-7 provides a viability envelope for a GTL project. Specifically, the cost estimates reported above were used to develop a simplified after-tax cash flow model, which was used to determine the set of natural gas feedstock and product prices that would generate a given internal rate of return (IRR). To make the analysis tractable, the credit for co-produced electricity was held constant, and motor gasoline prices were assumed to have perfect covariance with diesel prices.[z] Due to its low density, the diesel fuel produced in this study was assumed to sell at a 7 percent discount to petroleum diesel prices.

A 20 percent internal rate of return was considered a reasonable required return for a GTL project. Conditions to the upper left of the 20 percent internal rate of return curve in **Exhibit 5-7** are considered to have prohibitive economics. Conditions to the lower right of the 20 percent internal rate of return curve in **Exhibit 5-7** are considered to have favorable economics. For reference, the price of Brent crude oil implied by a given diesel price (based on their relative New York Harbor prices on May 17, 2013, as reported in the Oil and Gas Journal) is provided as a second reference for the viability window.

Establishing the required rate of return for a project does have a subjective component. Consequently, a curve was established for 14 percent and 26 percent internal rates of return as well. A 14 percent rate of return establishes a lower bound on acceptable returns given the market risk (uncertainty of natural gas and liquid product prices) associated with GTL projects. Though 26 percent may appear to be a high hurdle rate, it is often employed due to erosion of internal rate of return between project conception and implementation resulting from risks such as capital cost escalation, increased feedstock prices, and reduced product pricing.

Market conditions at the time this GTL system analysis was completed are favorable to undertaking a GTL project. However, the value at risk in a GTL project still remains considerably high. Specifically, there is a high probability of market conditions that are not favorable to GTL existing for extended periods of time throughout the plant life. The risk is heightened by the decoupling of natural gas prices and distillate fuels prices, being a recent phenomenon leaving analysts with limited historical basis to leverage. Ideally, reasonable probability distributions for these prices and their co-variance would be developed and translated into a Monte Carlo-style financial model to quantify the probability that a GTL project would provide sufficient returns to be viable. With a limited historical basis, production of such distributions is highly debatable and is left for future analysis.

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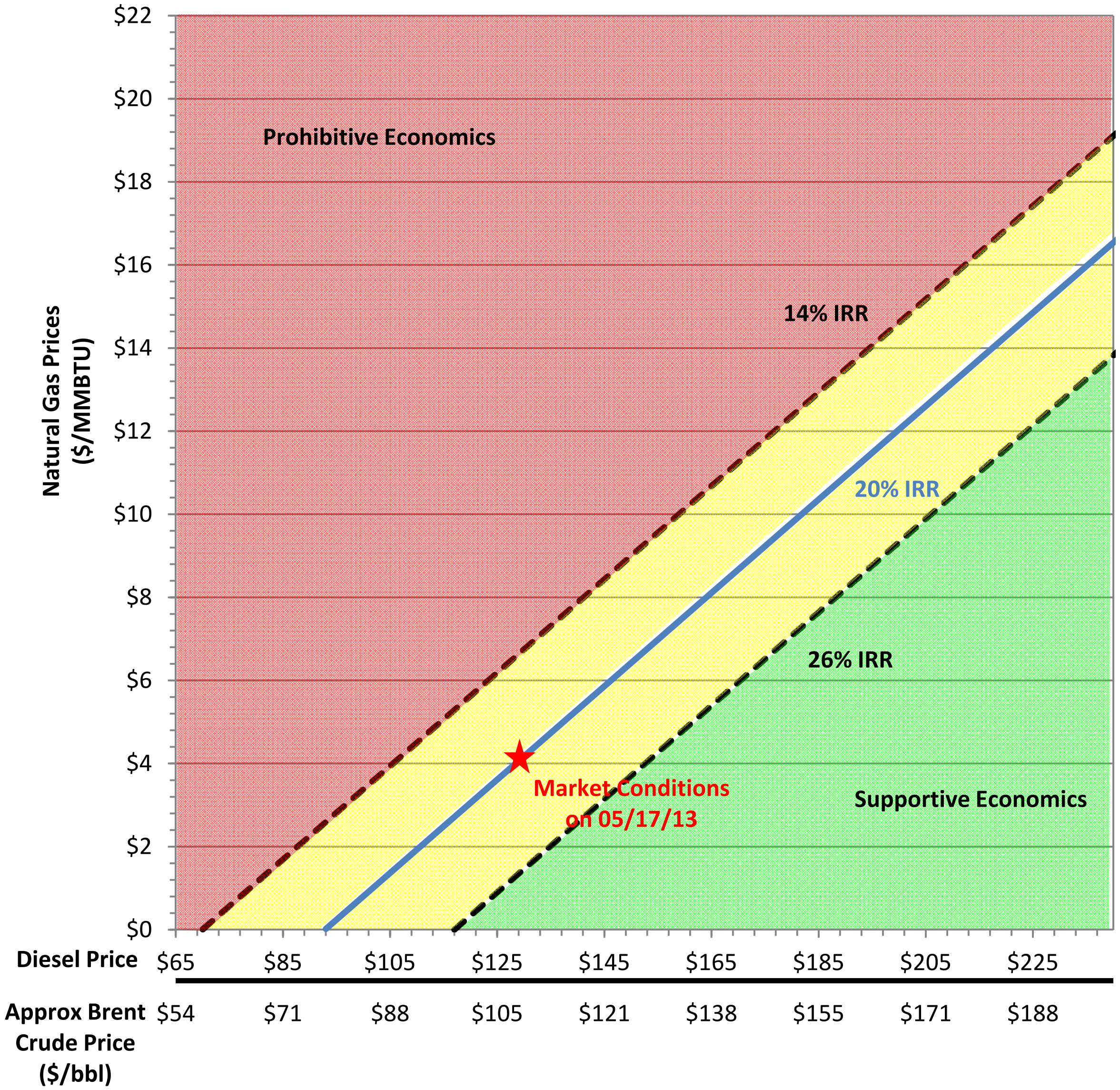
[z] The ratio of motor gasoline prices to diesel prices was set to their ratio in New York harbor on October 5, 2012, as reported in the Oil and Gas Journal. The prices used for comparing gasoline and diesel are $137.634 and $137.466 per barrel, respectively.

Exhibit 5-7: Viability Envelope for GTL Projects



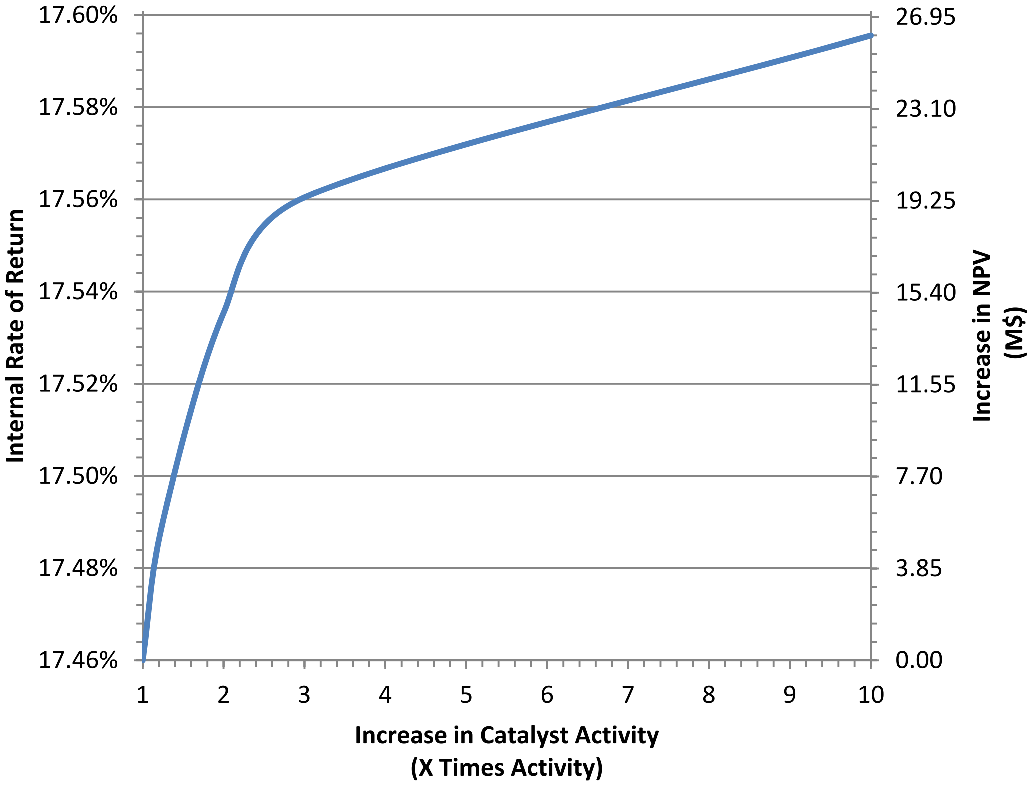
5.3.2 Sensitivity to Catalyst Performance

The Fischer-Tropsch reactor section of this analysis rested on a foundation of kinetic studies sponsored by the Department of Energy in the late 1980s and early 1990s. Catalyst improvements have been reported since that time and are believed to be in initial deployment. However, sufficient details on these performance gains were not publicly available to the extent that a reasonable update of the kinetic model could be developed.

To cover the impact of claimed catalyst improvements, sensitivity analyses were performed. Specifically, it was assumed that activity gains and liquid selectivity improvements occurred while catalyst costs remained constant. The assumption of constant catalyst costs is unlikely to hold true given the options (such as additions of expensive metals) to improve catalysts. The window of catalyst improvements are bound by public claims of catalyst improvements.[9, 10]

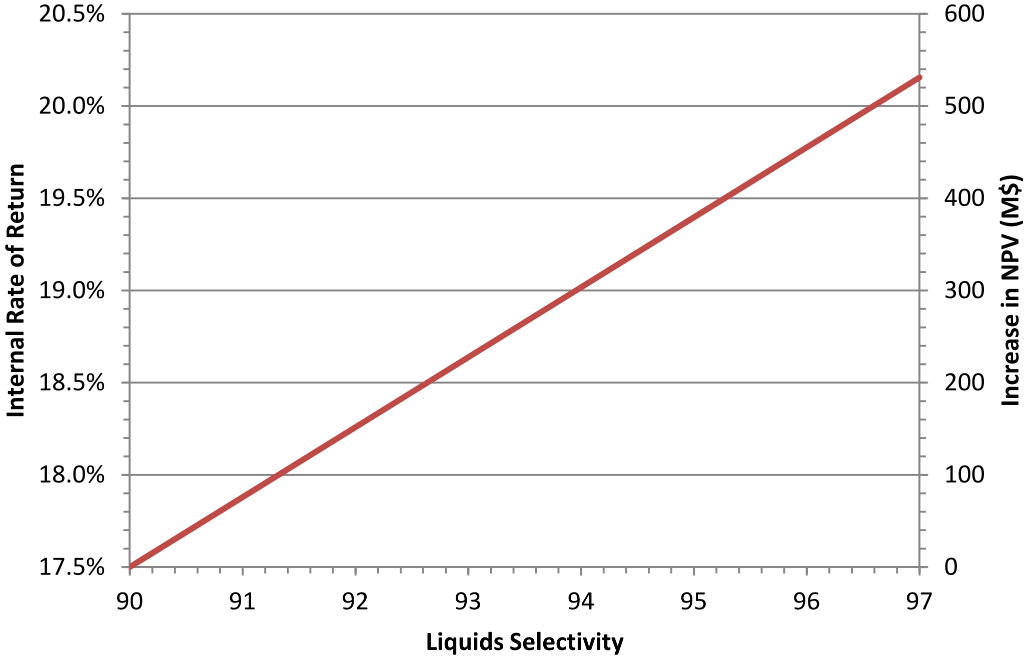
The impact of activity gains is reported in **Exhibit 5-8** below. A three-to-four-fold increase in activity is the highest plausible claim of potential catalyst activity improvement.[9, 10] The analysis was conducted at 10-fold increase in order to gauge how much a step change in catalyst activity could impact project viability. A gain in activity was also assumed to cut down on the rate of catalyst losses due to reduced catalyst requirements in the reactor. A 10-fold increase in activity would only increase the internal rate of return from 17.46 percent to 17.59 percent; adding only 26.1 MM$ to the net present value to a GTL facility of this scope. Consequently, given the high cost of catalyst development, the benefits of a catalyst program would need to be spread over a large number of projects in order to be a judicious investment.

Exhibit 5-8: Impact of Improved Catalyst Activity on Project Viability



Similarly, the impact of improvement in liquids selectivity is reported in **Exhibit 5-9** below. An increase to 94-95 percent in liquids selectivity is the highest plausible claim of potential catalyst activity improvement.9 The analysis was conducted up to 97 percent selectivity in order to gauge how much a step change in catalyst activity could impact project viability. Liquids selectivities at that high of a range are unlikely, given the high implied chain length propagation probability ("alpha" greater than 90 percent). An increase in selectivity to 97 percent would increase the internal rate of return from 17.46 percent to 20.18 percent. Such an improvement would add 531.0 MM$ to the net present value to a GTL facility of this scope. Similar to activity gains, given the high cost of catalyst development, the benefits of a catalyst program would need to be spread over a large number of projects in order to be a judicious investment.

Exhibit 5-9: Impact of Improved Liquid Selectivity on Project Viability



5.3.3 Sensitivity to Catalyst Consumption

The impact of catalyst losses on project economics are reported in **Exhibit 5-10** below. Catalyst losses are a critical issue that significantly separated the success of Pearl from the difficulties faced at Oryx. Pearl utilizes a multi-tubular fixed-bed reactor system[15] with “minimal catalyst attrition”[8] and associated losses. On the other hand, Oryx uses slurry bed technology and their high catalyst losses were one of the operational issues they had to struggle through upon start-up.[22]

Losses for the slurry phase system used for this study were assumed to be 3,858 lbs/day (~0.015 percent of the catalyst inventory per hour)[aa,bb] based on previous NETL studies and experience with multiphase catalytic reactor systems.[11, 20] Catalyst losses result from a complex interplay of reactor hydrodynamics and catalyst formulation.[cc] It is not possible to assess these factors in a scoping economics studies, since to do so requires project specific detailed design, physical testing, and firm detailed definition and large-scale production of the catalyst.

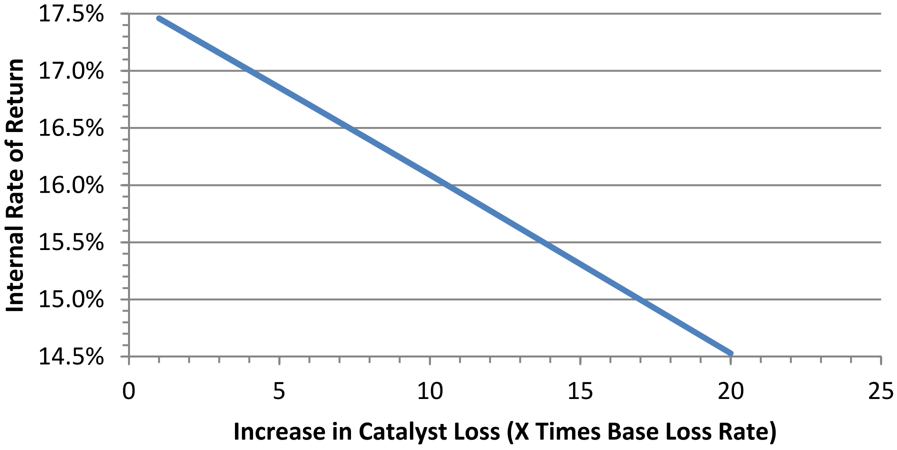
[aa] The 1lbs of catalyst loses per bbl of Fischer-Tropsch liquids would be 12.96 times higher than the loss rate used for this study. This convenient mnemonic (1lbs/bbl) has often been utilized to gain a rough order of magnitude of catalyst losses from a slurry bed Fischer-Tropsch reactor, but this loss rate implies that the catalyst bed turns over approximately every 22 days. If losses were this high, a different reactor choice (i.e., fixed bed) would be recommended.

[bb] As best can be discerned from the open literature reports regarding closely-guarded, proprietary catalyst attrition performance this attrition rate parameter is in line with post-learning curve, established, commercial performance in slurry reactors such as Oryx.

[cc] Catalyst formulations (adjustments to the support and binder portions of the catalyst) have been adapted to minimize the contribution of catalyst attrition to catalyst losses.

The impact of catalyst losses are reported in **Exhibit 5-10**. The internal rate of return of a GTL system falls appreciably as catalyst losses increase. Specifically, if losses are greater than 20 times the predicted level, the GTL system becomes economically marginal (an internal rate of return ~14 percent or less). Variance in catalyst losses of this order of magnitude between predicted and realized are common in multi-phase reactors.

Exhibit 5-10: Impact of Catalyst Losses on Project Viability



5.3.4 Sensitivity to Capital Project Costs

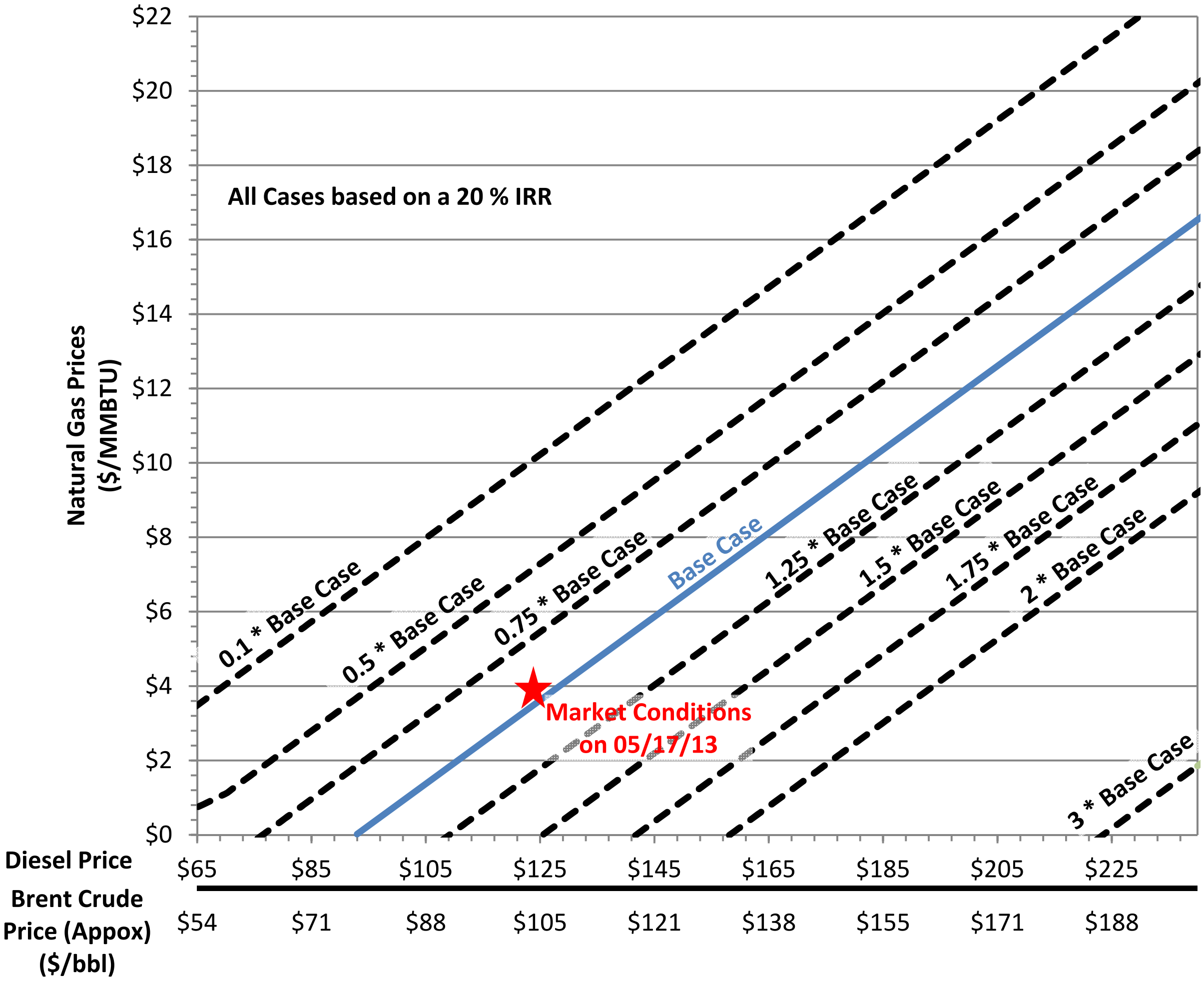
As mentioned in Section 5.2 above, capital costs are a significant portion of the cost of production. Furthermore, there is inherent uncertainty in the capital estimates of this study. Additionally, there has been a history of exogenous cost escalation during the long period of project development. Consequently, it is prudent to assess the sensitivity of GTL viability to capital cost estimates.

Exhibit 5-11 demonstrates the impact of changes in the capital expenditures on the viability window for GTL projects. The analysis is the same construct discussed in Section 5.3.1 above with a target internal rate of return of 20 percent. The labels refer to adjusting the capitalized expenses in the analysis by the given multiple (0.1, 0.5, 0.75, 1.25, 1.5, 2, and 3 respectively). The horizontal axis of **Exhibit 5-7** is petroleum diesel prices. Due to its low density, the diesel fuel produced in this study was assumed to sell at 7 percent discount to petroleum diesel prices.

Capital expenses that are two-to-three times the base estimate of this analysis have a viability window of prices that have not been realized. Projects where the capital is twice the amount of the estimate in this study still may be viable at high, but observed, diesel prices and low gas prices such as those in stranded gas niche situations or where the risk of the narrow window of viability can be mitigated. Current market conditions appear viable for capital expenditures roughly 24 percent higher than the estimate from this study, near the higher end of (but within) the error bounds associated with our cost estimate.

Reduced capital estimates were investigated in order to assess what level of capital reduction would be required to widen the viability window to increase the application of GTL technology. Essentially, a major step change in capital requirements, such as those targeted by process intensification efforts, (e.g. microchannel reactors combined with true scalability) is required to achieve this goal.

Exhibit 5-11: Impact of Variation of Capital Costs on Project Viability



6 Conclusions and Recommendations

This updated evaluation of the cost and performance of a commercial-scale GTL facility is consistent with the significant amount of previous work in this area.[3, 8, 11, 20, 27, 28] Furthermore, this and other GTL studies have been congruent with commercial applications[8, 9, 10, 15, 22, 26] within the significant variances in approach, scope, and intent. This consistency of capital costs reflects the limited experience curve for large petrochemical projects due to the number applications being limited by a tight constraint on the number of market-justified facilities. The 50,003 bbl per day of Fischer-Tropsch liquids GTL facility of this system analysis has an estimated total as-spent capital cost of 4.3 B$ (3.7 – 5.6 B$) or $86,188 ($73,260 - $112,045) per bbl of daily production of Fischer-Tropsch liquids. The addition of carbon capture to the facility contributes less than 1 percent to the cost of production of the Fischer-Tropsch liquids. Due to the incorporation of carbon capture, the liquid transportation fuels produced by the Fischer-Tropsch process have life cycle GHG emissions of GTL fuels comparable to those for conventional petroleum fuels despite the acquisition and delivery of natural gas having a higher GHG impact than the acquisition and delivery of an energy-equivalent amount of crude oil.

A domestic commercial GTL facility at the scale used for this system analysis would be viable at the market conditions (liquid fuel and natural gas prices) of the beginning of October 2012, generating approximately a 24 percent internal rate of return if these conditions were to hold over the life of the plant. Based on the latest market conditions from May 2013, the internal rate of return has fallen to 18 percent. A key concern that inhibits commitment of investment in domestic GTL facilities is the certainty that market conditions will change. Specifically, a high probability exists that there will be significant periods when market conditions will be insufficiently profitable to justify the significant capital investment required for GTL facilities. Additionally, the window of viability closes quickly as capital costs escalate. Doubling of the capital estimates in this analysis would require market prices, which have never been realized. This is not an extreme concern considering that cost escalation at Shell's Pearl facility was nearly double its design basis.[22] Significant reduction in the capital requirements, possibly beyond those enabled by process intensification, is needed to fully remove this market risk.

Improvement of key process parameters has a limited potential to widen the window of viability for GTL projects. Further catalyst activity gains as high as 10-fold would improve the internal rate of return less than 0.2 percentage points, given the small portion of the total project capital expenditures that such improvements could impact. This guides further work toward liquids selectivity, which may be nearing its practical limit, and improvement in effective catalyst lifetime (through either attrition resistance and/or improved ability to forestall deactivation and loss of selectivity). Liquids selectivity could increase the internal rate of return 2.7 percentage points if 97 percent liquid selectivity could be achieved. Catalyst losses have significant potential to erode the profitability of GTL projects, dropping the internal rate of return by 0.5, 1.4, and 3 percentage points as losses increase 4-, 10-, and 20-fold. This lends credence to the reports that higher-than-expected attrition had a significant deleterious impact on the profitability of the early phases of the Oryx project.[22]

With the technical risks fairly well-mitigated and the market risks understood, interest in domestic GTL has grown. Current market projections are within the viability window quantified by this study for a moderately sized GTL facility. Application of GTL domestically will continue the progression of the technology down the experience curve to further reduce technical and market risk and create ample opportunity to incorporate additional advances in catalyst and processing technology. Most importantly, GTL provides a disposition for expanding supplies of domestic natural gas, which creates significant economic value and improves the country's energy security.

A robust research and development program will enable GTL to provide an environmentally and economically advantaged disposition for domestic natural gas. First, such a program will contribute to the management of extensive knowledge regarding GTL as the work force transitions to another generation. Second, such a program can drive down capital and environmental costs of GTL. Third, research and development can enable GTL to be applied to other energy issues, such as defense installation energy, and reducing the amount of flared natural gas.

Reductions in capital costs can be realized through research and development of non-cryogenic gas separation (which can lower the need for air separation units by enabling an easier purge of inert, fixed gases) and application of process intensifications such as small-scale modular processing. Barriers to entry into the GTL space will be lowered by reducing the scale of the system required to be economically viable (and therefore the minimum capital required to enter the GTL space), allowing more institutions to participate and drive innovation. Further research into the recovery and upgrading of oxygenated co-products (which will also help biofuels development as well) will improve the carbon efficiency of GTL and widen the window of economic viability.

Improving the environmental impact of applying GTL technology requires research and development. The operation of GTL plants or the combustion of a fixed quantity of GTL fuel by mobile sources (i.e., vehicles) do not have significant opportunities for GHG emission reductions. However, significant improvements to the life cycle GHG emissions of GTL-derived fuel can be obtained through reduction of the emissions from natural gas extraction, processing, and transport. Reducing gas losses in the upstream natural gas sector is a research and development goal that could reduce the life cycle GHG emissions from GTL to values below the petroleum baseline.

Additionally, research and development can allow GTL technology to mitigate challenges presented by developing shale gas and shale oil. Specifically, natural gas stranded at the well-head, such as currently flared natural gas associated with the development of shale oil in the Bakken formation, could be converted to more easily and more economically transportable liquid fuels. Translating GTL technologies to a more intense process that could function at smaller scales, more amenable to well-head processing, could support further development of domestic shale oil resources by reducing the amount of flared associated natural gas.

Research and development can also allow GTL to solve critical military problems. The military has emerging needs to shrink the energy supply lines in forward areas (due to causalities and costs related to defending supply lines) and to assure security of supply to installations independent of standard energy infrastructure. Using the GTL technology basis to synergistically pair natural gas and biomass feedstocks could allow the military to meet these goals. This would allow production of liquid fuels from a combination of a limited number of wells, landfill gas, and locally sourced biomass, thus reducing the amount of energy supplies needing to be delivered to a forward area and/or for a military installation to be self-sufficient. Consequently, another research and development goal for GTL would be the combination of thermochemical conversion processes for gas and biomass with carbon capture, preferably at small scale.

GTL has an established technology platform with market risks that require mitigation. This platform can provide the foundation of solving defense, resource development, and energy security needs through a well-targeted wave of research and development. Additionally, further research and development can help reduce the market risks associated with these projects and maintain the knowledge base of GTL technology.

7 Appendix

This appendix provides details on the operating parameters of the GTL plant and a description of the LCA method.

7.1 Modeling Design Basis

The system modeled for this analysis is described in Sections 2.2 and 3 above. The gas-to-liquid (GTL) system modeled here converts natural gas following gas processing into synthesis gas, which is subsequently converted to Fischer-Tropsch liquids. The Fischer-Tropsch liquids are upgraded to salable-grade diesel and motor gasoline fuels. Energy is recovered from the process to satisfy system utilities loads with excess energy converted to exported electricity. This GTL system was equipped for the capture of carbon dioxide.

7.1.1 Site Description

The plant in this study was assumed to be located at a generic plant site in the Midwestern U.S., with ambient conditions and site characteristics as presented in **Exhibit 7-1** and **Exhibit 7-2**.

Exhibit 7-1: Site Ambient Conditions

Condition	Value
Elevation, ft	0
Barometric Pressure, psia	14.696
Design Ambient Temperature, Dry Bulb, °F	59
Design Ambient Temperature, Wet Bulb, °F	51.5
Design Ambient Relative Humidity, %	60

Exhibit 7-2: Site Characteristics

Characteristic	Value
Location	Greenfield, Midwestern U.S.
Topography	Level
Size, acres	300[dd]
Transportation	Rail, Road, Pipeline
Ash Disposal	Off Site
Water	Municipal (50%) / Groundwater (50%)
Access	Landlocked, with access by rail and highway
Captured CO_2	Exported from plant boundary at 15.2 MPa (152 bar, 2,215 psia)

[dd] Based on the acreage used for the Shell facility in Qatar[9] and the reduction of scale, our site would be approximately 198 acres. An additional 102 acres was added to account for require rights of way and less efficient footprint for a smaller scale facility.

The following design parameters were considered site-specific and are not quantified for this study. Allowances for normal conditions and construction were included in the cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise regulations and their impact on site and surrounding area

7.1.2 Facility Feedstock

Natural gas – specifically, natural gas after it has gone through midstream natural gas processing between the well-head and the battery limits of the GTL plant – was used as the feedstock for this study. As such, the feedstock will be relatively free of containments, natural gas liquids, water, and condensate. The composition of the natural gas feedstock used in this study is presented in **Exhibit 7-3**.[12] The integration of natural gas liquids and condensate recovery into GTL facilities is common in practice but is outside of the scope of this study and can only be intelligently done for a specific project opportunity versus a general study.

Exhibit 7-3: Natural Gas Composition

Component		Volume Percentage
Methane	CH_4	93.1
Ethane	C_2H_6	3.2
Propane	C_3H_8	0.7
n-Butane	C_4H_{10}	0.4
Carbon Dioxide	CO_2	1.0
Nitrogen	N_2	1.6
	LHV	HHV
kJ/kg MJ/scm	47,454 34.71	52,581 38.46
Btu/lb Btu/scf	20,410 932	22,600 1,032

Note: Feedstock composition is normalized and heating values are calculated.

7.1.3 Environmental Requirements

The environmental control equipment to be used in the conceptual design conforms to best available control technology (BACT) guidelines, which are summarized in **Exhibit 7-4**.

Exhibit 7-4: BACT Environmental Design Basis

Pollutant	Control Technology	Limit
Sulfur Oxides (SO_2)	Rectisol®/Selexol/Econamine Plus/Sulfinol-M + Claus Plant	99+% or ≤ 0.050 lb/106Btu
Nitrogen Oxides (NOx)	LNB and N2 Dilution primarily with humidification as needed	15 ppmvd (@ 15% O2)
Particulate Matter (PM)	Cyclone/Barrier Filter/Wet Scrubber/AGR Absorber	0.006 lb/106Btu
Mercury (Hg)	Activated Carbon Bed	95% removal

The current regulation governing new fossil-fuel-fired power plants is a New Source Performance Standard (NSPS) for Electric Utility sector,[29] published in February 2006 and shown in **Exhibit 7-5**. These standards supersede the previous performance standards, established in 1978.[29] The new standards apply to units with the capacity to generate greater than 73 MW of power by burning fossil fuels, as well as cogeneration units that sell more than 25 MW of power and more than one third of the potential output capacity to any utility power distribution system.

Exhibit 7-5: Standards of Performance for Electric Utility Steam-generating Units Built, Reconstructed, or Modified after February 28, 2005

Pollutant	New Units		Reconstructed Units		Modified Units	
	Emission Limit	% Reduction	Emission Limit ($lb/10^6Btu$)	% Reduction	Emission Limit ($lb/10^6Btu$)	% Reduction
PM	0.015 $lb/10^6Btu$	99.9	0.015	99.9	0.015	99.8
SO_2	1.4 lb/MWh	95	0.15	95	0.15	90
NOx	1.0 lb/MWh	N/A	0.11	N/A	0.15	N/A

The BACT emission limits assumed for this study are more stringent than the emission requirements of the 2006 NSPS. It is possible that state and local requirements could supersede NSPS or BACT and impose even more stringent requirements.

7.1.4 Carbon Dioxide

In the commercial-scale Fischer-Tropsch (F-T) facilities examined in this analysis, CO_2 is separated ("captured") from other process gases as a function of normal plant operations. This is done to reduce the circulation of non-reactive gases, which would otherwise build up in the F-T recycle loop, and to minimize equipment sizes and cost. The CO_2 capture technology in this analysis is based on a proprietary technology that uses monoethanolamine (MEA) to absorb CO_2 and is followed by a solvent regeneration process that produces a pure CO_2 stream.

Unlike current commercial F-T systems, which vent captured CO_2 to the atmosphere, the scenarios of this analysis send captured CO_2 to a geologic sequestration site. The CO_2 is dehydrated and compressed to 2,215 psia for pipeline transport at the plant boundary. CO_2 is transported 80 km

(50 miles) via pipeline to a geologic sequestration field for injection into a saline formation. The CO_2 is transported and injected as a supercritical fluid in order to avoid two-phase flow while achieving maximum efficiency. Currently, the CO_2 transport, sequestration, and monitoring (TS&M) costs are calculated at $7/tonne on a first-year basis.

7.1.5 Raw Water Usage

Raw water makeup was assumed to be provided 50 percent by publicly owned treatment works (POTW) and 50 percent from groundwater. Raw water usage is defined as the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, condenser makeup, ash handling makeup, synthesis gas humidification, and quench system makeup. The plant is equipped with an evaporative cooling tower, and all process blowdown streams are treated and recycled to the cooling tower.

Since the F-T plant differs from a natural gas combined cycle (NGCC) power plant, the technique for estimating water usage is different and based on the following:

1. The power plant was modeled in the AspenPlus® process simulation program, which predicts the required cooling water load and stack water loss
2. The AspenPlus® model for the F-T synthesis area and the product upgrading area was used to predict the required F-T cooling water load
3. The total of the cooling tower blowdown and vaporization losses were predicted based on the estimated cooling water circulation rate

Process water and cooling water came from two different treatment facilities. The composition and physical properties of each are shown in **Exhibit 7-6**.

Exhibit 7-6: Process and Cooling Water Properties

Property	Process Water	Cooling Water
Total Dissolved Solids (TDS)	200 µS/cm	1250 µS/cm
Hardness	100 mg/l as $CaCO_3$	75 mg/l as $CaCO_3$
Alkalinity	100 ppm	350 ppm
Sulfate	4 ppm	50 ppm
Chloride	10 ppm	200 ppm
Silica	30 mg/l	30 mg/l
Aluminum	Not Available	Not Available
Iron	0.25 mg/l	0.25 mg/l
Calcium	70 mg/l	25 mg/l
Magnesium	25 mg/l	45 mg/l
Phosphate	0.4 mg/l	6.0 mg/l (ortho)
Ammonia	<1 mg/l	19 mg/l
Chlorine	<0.1 mg/l	<0.1 mg/l
pH	8.0	8.0

7.1.6 Balance of Plant

The assumed balance of plant requirements is listed in **Exhibit 7-7**.

Exhibit 7-7: Balance of Plant Design Requirements

Item	Value
Cooling System	Recirculating Evaporative Cooling Tower
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors 250 hp and below	480 volt
Motors above 250 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam & Gas Turbine Generators	24,000 volt
Grid Interconnection Voltage	345 kV
Water and Waste	
Makeup Water	Process water is available at a flow rate of 1,500 gpm
Feed Water	Process water treatment is included and produces boiler feed quality water.
Process Wastewater	Process wastewater and storm water that contacts equipment surfaces is collected and treated for recycle as slurry or quench makeup. Selected blowdown is discharged through a permitted discharge permit.
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site.
Water Discharge	Most of the wastewater is recycled for plant needs. Blowdown is treated for chloride and metals, and then discharged.

7.2 Process Block Flow Diagrams and Stream Tables

Block flow diagram of the process and associated stream tables for the GTL process are provided below. **Exhibit 7-8** was used in Section 2 (**Exhibit 2-1**) above and is repeated here for ease of use of this Appendix as a standalone section.

Exhibit 7-8: GTL System Block Flow Diagram

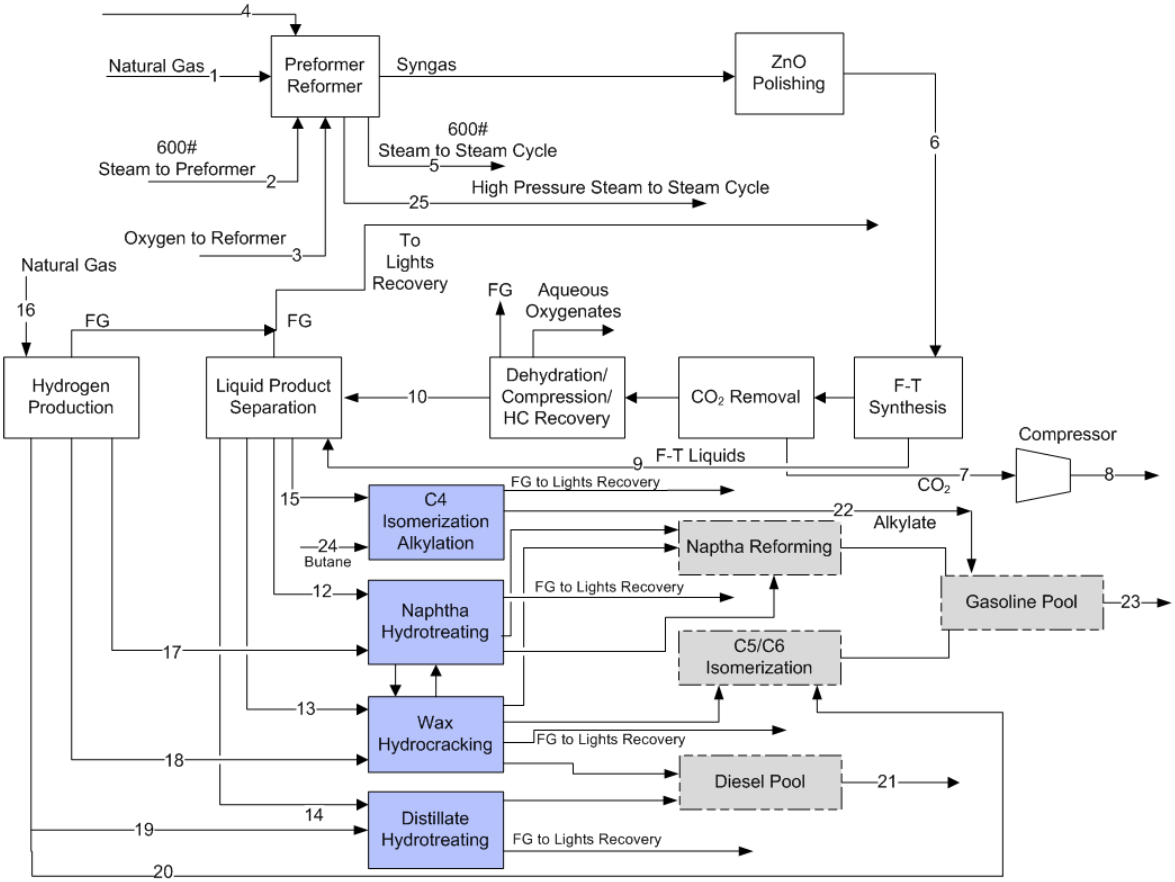
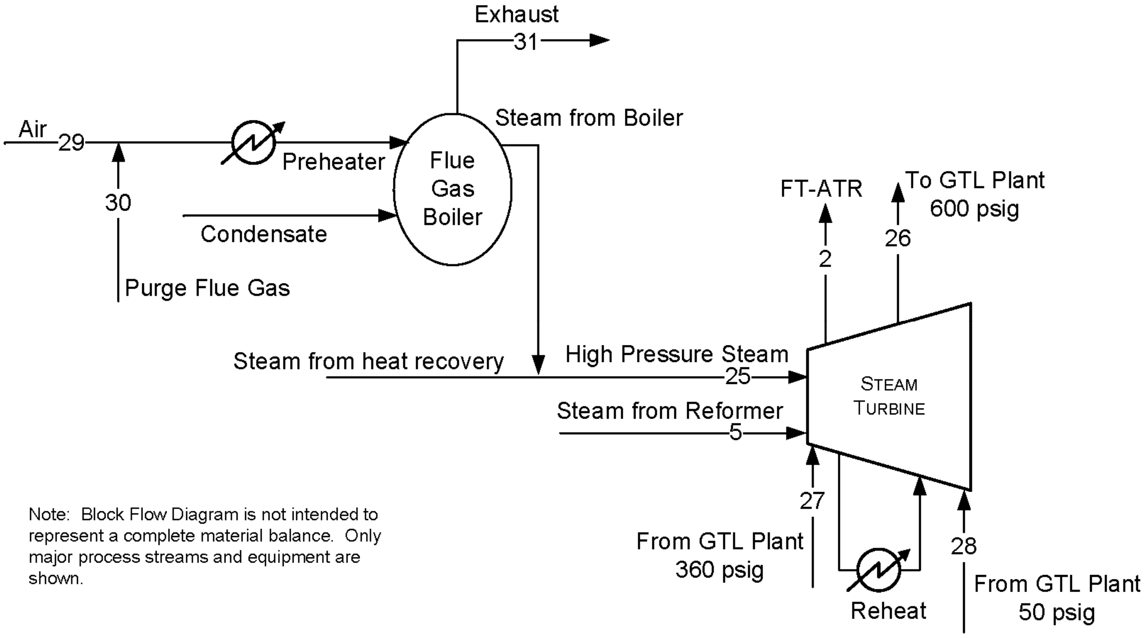


Exhibit 7-9: Power Production Block Flow Diagram



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 7-10: System Stream Tables

	1	2	3	4	5	6	7	8	9	10	12	13	14	15	16
V-L Mole Fraction															
H2	0.0000	0.0000	0.0000	0.4051	0.0000	0.5418	0.0000	0.0000	0.0007	0.0000	0.0000	0.0000	0.0000	0.0010	0.0000
N2	0.0160	0.0000	0.0100	0.4007	0.0000	0.1173	0.0000	0.0000	0.0015	0.0000	0.0000	0.0000	0.0000	0.0023	0.0160
O2	0.0000	0.0000	0.9900	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.1380	0.0000	0.2471	0.0000	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0007	0.0000
CO2	0.0100	0.0000	0.0000	0.0099	0.0000	0.0580	1.0000	1.0000	0.0028	0.0003	0.0000	0.0000	0.0000	0.0043	0.0100
H2O	0.0000	1.0000	0.0000	0.0013	1.0000	0.0328	0.0000	0.0000	0.0688	0.0145	0.0000	0.0000	0.0000	0.0243	0.0000
NH3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SULFUR	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH4	0.9310	0.0000	0.0000	0.0319	0.0000	0.0030	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0003	0.9310
C2H4	0.0000	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0320	0.0000	0.0000	0.0044	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0000	0.0000	0.0000	0.0031	0.0320
C3H6	0.0000	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000	0.0000	0.0001	0.0575	0.0000	0.0000	0.0000	0.2244	0.0000
C3H8	0.0070	0.0000	0.0000	0.0046	0.0000	0.0000	0.0000	0.0000	0.0000	0.0250	0.0000	0.0000	0.0000	0.0974	0.0070
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0032	0.0000	0.0000	0.0000	0.0124	0.0000
NC4H8	0.0000	0.0000	0.0000	0.0020	0.0000	0.0000	0.0000	0.0000	0.0002	0.0606	0.0000	0.0000	0.0000	0.2364	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0016	0.0000	0.0000	0.0000	0.0063	0.0000
NC4H10	0.0040	0.0000	0.0000	0.0007	0.0000	0.0000	0.0000	0.0000	0.0001	0.0314	0.0000	0.0000	0.0000	0.1225	0.0040
C5H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0514	0.0000	0.0000	0.0003	0.2003	0.0000
NC5H12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0349	0.0000	0.0000	0.0588	0.0491	0.0000
IC5H12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0039	0.0000	0.0000	0.0000	0.0151	0.0000
C6H12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0395	0.0000	0.0000	0.1040	0.0000	0.0000
NC6H14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0385	0.0000	0.0000	0.1014	0.0000	0.0000
IC6H14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0043	0.0000	0.0000	0.0113	0.0000	0.0000
C7H14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0262	0.0000	0.0000	0.0692	0.0000	0.0000
C7H16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0487	0.0000	0.0000	0.1284	0.0000	0.0000
C8H16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0157	0.0000	0.0000	0.0415	0.0000	0.0000
C8H18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0013	0.0524	0.0000	0.0000	0.1388	0.0000	0.0000
C9H18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0105	0.0000	0.0000	0.0280	0.0000	0.0000
C9H20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0020	0.0513	0.0000	0.0000	0.1367	0.0000	0.0000
C10H20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0073	0.0000	0.0000	0.0196	0.0000	0.0000
C10H22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0030	0.0487	0.0000	0.0000	0.1309	0.0000	0.0000
C11H22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0046	0.0000	0.0135	0.0000	0.0000	0.0000
C11H24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0041	0.0461	0.0000	0.1361	0.0000	0.0000	0.0000
C12H24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0027	0.0000	0.0082	0.0000	0.0000	0.0000
C12H26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0056	0.0428	0.0000	0.1284	0.0000	0.0000	0.0000
C13H26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0020	0.0000	0.0062	0.0000	0.0000	0.0000
C13H28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0074	0.0386	0.0000	0.1186	0.0000	0.0000	0.0000

Exhibit 7-10: System Stream Tables (Continued)

	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
V-L Mole Fraction															
H2	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4051	0.4051
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7804	0.4007	0.4007
O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2093	0.0000	0.0000
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1380	0.1380
CO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0099	0.0099
H2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0100	0.0013	0.0013
NH3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SULFUR	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0319	0.0319
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0044	0.0044
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0006
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0046	0.0046
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0020	0.0020
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0004
NC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007	0.0007
C5H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NC5H12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC5H12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NC6H14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC6H14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C11H22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C11H24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C12H24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C12H26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C13H26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C13H28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Exhibit 7-10: System Stream Tables (Continued)

	1	2	3	4	5	6	7	8	9	10	12	13	14	15	16
V-L Mole Fraction															
C14H28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0014	0.0000	0.0046	0.0000	0.0000	0.0000
C14H30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0099	0.0344	0.0000	0.1095	0.0000	0.0000	0.0000
C15H30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0010	0.0000	0.0031	0.0000	0.0000	0.0000
C15H32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0121	0.0306	0.0000	0.1010	0.0000	0.0000	0.0000
C16H32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0006	0.0000	0.0019	0.0000	0.0000	0.0000
C16H34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0142	0.0271	0.0000	0.0932	0.0000	0.0000	0.0000
C17H34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0002	0.0000	0.0009	0.0000	0.0000	0.0000
C17H36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0179	0.0231	0.0000	0.0861	0.0000	0.0000	0.0000
C18H36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0002	0.0000	0.0008	0.0000	0.0000	0.0000
C18H38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0209	0.0193	0.0000	0.0786	0.0000	0.0000	0.0000
C19H38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0002	0.0000	0.0007	0.0000	0.0000	0.0000
C19H40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0239	0.0158	0.0000	0.0718	0.0000	0.0000	0.0000
C20H40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0001	0.0006	0.0000	0.0000	0.0000	0.0000
C20H42	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0271	0.0123	0.0627	0.0000	0.0000	0.0000	0.0000
ALKYLATE	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
ISOMERAT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
REFORMAT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7-300HC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
3-350HC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
350-5HC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
500+HC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7-300HT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
3-350HT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
350-5HT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
500+HT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
OXVAP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0030	0.0000	0.0000	0.0000	0.0000	0.0000
OXHC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0246	0.0000	0.0368	0.0312	0.0000	0.0000
OXH2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10AP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Exhibit 7-10: System Stream Tables (Continued)

	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
V-L Mole Fraction															
C14H28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C14H30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C15H30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C15H32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C16H32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C16H34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C17H34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C17H36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C18H36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C18H38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C19H38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C19H40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H42	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
ALKYLATE	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.3301	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
ISOMERAT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2929	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
REFORMAT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3770	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7-300HC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
3-350HC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
350-5HC	0.0000	0.0000	0.0000	0.0000	0.6556	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
500+HC	0.0000	0.0000	0.0000	0.0000	0.0500	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7-300HT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
3-350HT	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
350-5HT	0.0000	0.0000	0.0000	0.0000	0.2025	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
500+HT	0.0000	0.0000	0.0000	0.0000	0.0919	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
OXVAP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
OXHC	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
OXH2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10N	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9A	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10AP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Exhibit 7-10: System Stream Tables (Continued)

	1	2	3	4	5	6	7	8	9	10	12	13	14	15	16
V-L Mole Fraction															
C21OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0291	0.0103	0.0595	0.0000	0.0000	0.0000	0.0000
C22OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0311	0.0082	0.0556	0.0000	0.0000	0.0000	0.0000
C23OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0323	0.0063	0.0519	0.0000	0.0000	0.0000	0.0000
C24OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0329	0.0049	0.0485	0.0000	0.0000	0.0000	0.0000
C25OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0329	0.0037	0.0454	0.0000	0.0000	0.0000	0.0000
C26OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0325	0.0028	0.0424	0.0000	0.0000	0.0000	0.0000
C27OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0316	0.0021	0.0396	0.0000	0.0000	0.0000	0.0000
C28OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0305	0.0016	0.0370	0.0000	0.0000	0.0000	0.0000
C29OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0293	0.0012	0.0346	0.0000	0.0000	0.0000	0.0000
C30WAX	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4862	0.0001	0.5222	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg$_{mol}$/hr)	19,849	15,081	16,334	23,275	18,433	83,584	4,821	4,821	311	789	289	276	301	202	602
V-L Flowrate (kg/hr)	343,930	271,689	522,011	405,651	332,081	1,211,199	212,188	212,188	135,800	103,919	139,925	55,395	32,696	11,068	10,435
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	38	343	152	44	343	93	38	47	253	166	38	38	38	38	15
Pressure (MPa, abs)	3.10	4.24	2.76	3.10	4.38	2.24	1.83	15.27	0.59	2.00	0.34	0.34	0.34	0.34	3.10
Density (kg/m^3)	22.2	16.4	25.0	20.4	17.0	10.6	34.2	686.3	669.0	581.0	816.9	744.9	687.1	8.0	24.4
V-L Molecular Weight	17.327	18.015	31.959	17.429	18.015	14.491	44.010	44.010	437.305	131.698	483.591	200.406	108.670	54.677	17.327
V-L Flowrate (lb$_{mol}$/hr)	43,759	33,248	36,010	51,312	40,639	184,271	10,629	10,629	685	1,740	638	609	663	446	1,328
V-L Flowrate (lb/hr)	758,235	598,971	1,150,836	894,308	732,113	2,670,238	467,794	467,794	299,388	229,103	308,482	122,124	72,083	24,400	23,006
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	100	650	305	111	650	199	100	117	488	330	100	100	100	100	59
Pressure (psia)	450.0	615.0	400.0	450.0	635.0	325.0	265.0	2,214.7	85.0	290.0	50.0	50.0	50.0	50.0	450.0
Density (lb/ft^3)	1.384	1.022	1.560	1.273	1.059	0.662	2.138	42.846	41.762	36.270	50.998	46.506	42.891	0.497	1.525
	A - Reference conditions are 32.02 F & 0.089 PSIA														

Exhibit 7-10: System Stream Tables (Continued)

	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
V-L Mole Fraction															
C21OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C22OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C23OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C24OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C25OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C26OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C27OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C28OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C29OP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C30WAX	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	277	216	1,090	12	889	238	722	324	6,093	5,619	57,345	14,511	2,583	1,225	23,275
V-L Flowrate (kg/hr)	559	436	2,197	25	174,497	27,216	73,331	18,843	109,775	101,219	1,033,079	261,410	74,280	21,350	405,651
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	38	38	38	38	115	38	49	38	521	345	226	140	15	31	44
Pressure (MPa, abs)	4.14	4.14	0.83	4.14	0.21	0.34	0.21	0.10	12.76	4.24	2.58	0.34	0.10	2.72	3.10
Density (kg/m^3)	3.2	3.2	0.6	3.2	687.7	676.8	679.0	2.4	38.9	16.3	12.6	1.9	1.2	18.7	20.4
V-L Molecular Weight	2.016	2.016	2.016	2.016	196.304	114.230	101.590	58.123	18.015	18.015	18.015	18.015	28.753	17.429	17.429
V-L Flowrate (lb_{mol}/hr)	611	477	2,403	27	1,960	525	1,591	715	13,434	12,387	126,424	31,990	5,695	2,701	51,312
V-L Flowrate (lb/hr)	1,231	962	4,843	55	384,700	60,000	161,668	41,542	242,012	223,150	2,277,550	576,310	163,759	47,069	894,308
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	100	100	100	100	239	100	120	100	970	653	438	283	59	89	111
Pressure (psia)	600.0	600.0	120.0	600.0	30.0	50.0	30.0	15.0	1,850.0	614.7	374.7	50.0	14.7	395.0	450.0
Density (lb/ft^3)	0.197	0.197	0.040	0.197	42.932	42.254	42.391	0.149	2.430	1.018	0.785	0.116	0.076	1.166	1.273

Exhibit 7-11: Summary and Design Basis Capacity of GTL Processing Sections

Processing Section	Capacity (1,000 lb/hr)	Capacity (HP)
Natural Gas Pre-Reformer	1,653	
NG Reformer/ATR	3,402	
F-T Synthesis	2,670	
Amine CO_2 Adsorption	1,380	
Amine Regeneration Section	468	
Gas Compression and Dehydration	911	
Hydrocarbon Recovery	528	
H_2 Production (SMR and PSA)	115	
Naphtha Hydrotreater	82	
Naphtha Isomerization	39	
Naphtha Reformer	74	
C_4 Isomerization	60	
Alkylation	61	
Diesel Hydrotreater	123	
Wax Hydrocracker	313	
Fuel/Recycle Gas Compressor, HP		2,828 HP
Recycle Gas Compressor, HP		3,254 HP
CO_2 Compression & Drying	468	9,036 HP

7.3 LCA GHG Method

The GHG accounting method used in this study is based on an life cycle analysis (LCA) approach,[6] and is consistent with the guidelines for performing LCAs developed by International Standards Organization (ISO).[30, 31]

7.3.1 Purpose

The purpose of this LCA is to understand the life cycle greenhouse gas footprint of liquid transportation fuels produced from natural gas. In this case, there are two products of interest: diesel and gasoline produced by a GTL plant.

7.3.2 Basis for Comparison

All LCA results are expressed in terms of a functional unit, which serves as a basis of comparison among scenarios. The functional unit of this analysis is 1 MJ of fuel combusted in a vehicle. Converting this energy combustion to the basis of vehicle-miles traveled would improve the interpretation of results, but would require assumptions about vehicle type and fuel efficiency, which would introduce further uncertainty to this analysis.

The 2005 petroleum baseline was selected as the comparative baseline year in conformance with the Energy Independence and Security Act of 2007.[32] The GHG results from the petroleum baseline are 90.0 g CO_2e/MJ diesel and 91.3 g CO_2e/MJ gasoline.[33]

Comparative analysis of the life cycle GHG to petroleum-derived diesel fuel is based on an industry size representing one million bbl/d (or more) production of F-T diesel. No future energy-efficiency or learning-curve estimates are accounted for in the comparative assessment modeling approach.

7.3.3 Study Boundary

The time period represented by the energy-conversion facilities is considered a near-term technology based on the integration of commercially-available technology.

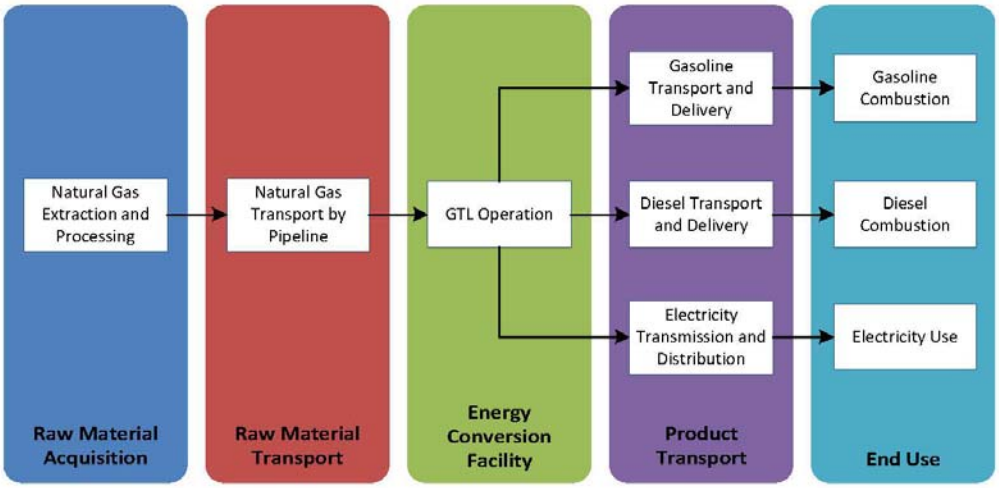
The study boundary for the analysis is from the extraction of raw materials from the earth to the consumption of the diesel fuel to move a passenger vehicle. The boundary applied is commonly referred to as a “cradle-to-grave” LCA. The “cradle” refers to extraction of raw materials from the earth, and the “grave” is the combustion of fuel in the vehicle.

Contributions of GHG emissions are accounted for within five life cycle stages:

- **Raw Material Acquisition (RMA)** includes the extraction and processing of natural gas.
- **Raw Material Transport (RMT)** is the movement of raw material, in this case natural gas via pipeline, from the site of acquisition to the liquid fuels production facility.
- The **Energy Conversion Facility (ECF)** converts raw materials to liquid fuels. Carbon capture and sequestration (CCS) operations are included here.
- **Product Transport (PT)** moves fuel from the energy conversion facility to the refueling station, on-site storage, and dispensing of the fuel into a vehicle.
- **Use** of fuel in a passenger vehicle, using a functional unit of 1 MJ of combusted diesel.

Exhibit 7-12 is a high-level illustration of the life cycle boundaries of the GTL system and includes the transport and use of co-products. The life cycle begins with the extraction and delivery of natural gas; includes the operation of the GTL facility, which produces liquid fuels and electricity; accounts for the transport and delivery of liquid fuels and electricity to the end user; and ends with combustion of fuel and consumption of electricity.

Exhibit 7-12: LCA Boundaries for GTL System, Including Co-product Transport and Use



The boundaries of **Exhibit 7-12** include the transport and use of all co-products. A co-product management method is necessary to determine the GHG emissions attributable to only one product. Co-product management is discussed in **Section 7.3.6**.

7.3.4 Environmental Metrics

The scope of the LCA is limited to GHG emissions, including CO_2, CH_4, SF_6, and N_2O. The effects of each GHG emission are normalized and reported in terms of their GWP. Normalized values are expressed in terms of CO_2e. Intergovernmental Panel on Climate Change (IPCC) publishes the international standard for calculating GWP based on the weighted contribution of various emissions.[34, 35] The IPCC publishes values for three timeframes: 20, 100, and 500 years. The U.S. standard is based on the 100-year timeframe. Within this study, the 2007 IPCC values are used. **Exhibit 7-13** lists the primary GHGs and their corresponding GWP reported in mass of CO_2 equivalents.

NO_X and its impact on global warming are currently being reviewed by climatologists around the world. As a result, there is a lack of agreement about the impact of NO_X in relation to global warming. The 2007 IPCC report “Climate Change 2007: The Physical Science Basis” notes that “nitrogen oxides have short lifetimes and complex nonlinear chemistry with opposing indirect effects through ozone enhancements and methane reduction.”[36] Most current research suggests that the GWP for surface/industrial NO_X emission may be negative. Wild, et al. reported a GWP for industrial NO_X emissions of -12.[37] Since there is a lack of agreement on NO_X effects, the IPCC has opted to omit them from consideration.

Exhibit 7-13: GHG Emissions Include in Study Boundary and Their 100-year GWP

Emissions to Air	Abbreviation	This Study (GWP, CO_2e)
Carbon Dioxide	CO_2	1
Methane	CH_4	25
Sulfur Hexafluoride	SF_6	22,800
Nitrous Oxide	N_2O	298

7.3.5 Cut-off Criteria

Cut-off criteria define the selection of materials and processes to be included in the system boundary. Following the requirements of ISO14040, the criteria of mass, cost, and environmental relevance was used for material and energy inputs.[30]

A significant, or relevant, material input is defined as a material that has a mass greater than one percent of the principal product that is produced by a unit process. Similarly, a significant energy input is defined as one that contributes more than 1 percent of the total energy used by the unit process.

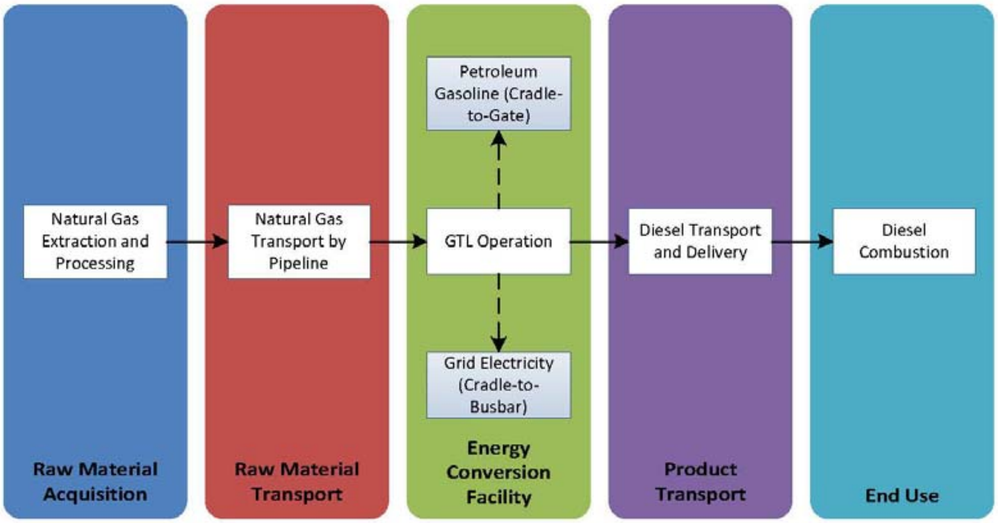
In some systems, there are inputs that represent a relatively low fraction of total system inputs, but represent a disproportionately high share within a particular category in the LCA results. For example, the relatively small mass fraction of silver in lead-free solder represents a disproportionately high share of the water quality impacts from soldered electronic systems.[38] The scope of this analysis is restricted to GHG emissions. GHG emissions are produced by any activity that involves the handling of fossil feedstocks or the combustion of a fuel. GHG emissions are common to most unit processes in an LCA, especially an LCA of an energy system with fossil feedstocks. Based on the scope of this analysis and the type of system being studied, it is unlikely that there are any system inputs that represent both a small fractional input and a large contribution to total results.

7.3.6 Co-product Management

The objective of LCA is to assign ownership of environmental burdens to a single product or function. The GTL facility produces three products (diesel, gasoline, and electricity), making it necessary to use a co-product management method to assign burdens to the functional unit. System expansion and allocation are two different methods for co-product management.

System expansion alters system boundaries so co-products are enveloped by other processes in the system. The goal of system expansion is to draw boundaries such that only one product, the functional unit, exits the system. This can be accomplished by including processes that consume co-products or by assuming that co-products displace alternative production routes. This analysis uses system expansion with displacement to manage GTL co-products. When diesel is the functional unit, co-produced gasoline displaces conventional petroleum gasoline and co-produced electricity displaces average power produced by the U.S. electricity grid. When gasoline is the functional unit, co-produced diesel displaces conventional petroleum diesel and co-produced electricity displaces average power produced by the U.S. electricity grid. No scenarios of this analysis have electricity as a functional unit, so it is not necessary to model the displacement of gasoline and diesel in the same scenario. **Exhibit 7-14** illustrates the system expansion with displacement scheme with diesel combustion as the functional unit.

Exhibit 7-14: GTL System Expansion for Life Cycle Modeling of GTL Diesel



Co-product allocation is an alternative to system expansion and uses a physical basis for apportioning burdens. For example, if energy-based co-product allocation is applied to the GTL plant in this analysis, the heating values of diesel and gasoline and electrical energy are ratioed to calculate the energy share of each co-product. Diesel is 69 percent of total energy produced by the GTL plant, so it is assigned 69 percent of the cradle-to-gate burdens of GTL operation, with the exception of the butane input, which is 100 percent allocated to gasoline. The remaining 31 percent of the cradle-to-gate burdens (29.6 and 1.4 percent for gasoline and electricity, respectively) are excluded from the boundaries of the system. The problem with this method is that it requires a comparison of two forms of energy – electricity and heat of combusted fuel. Further, a megajoule of electricity accounts for the efficiency losses of thermoelectric power generation, while a megajoule of combusted fuel does not account for the efficiency of converting heat to useful work. Since a megajoule of electricity and a megajoule of heat from combusted fuel are not providing equivalent services, it is hard to defend the use of energy allocation as a co-product management method for GTL co-products.

7.4 Life Cycle GHG Data

An LCA model is a network of interdependent unit processes scaled to the basis of a functional unit. The data and assumptions for key unit processes in the LCA model of GTL systems are summarized below.

7.4.1 Natural Gas Acquisition

The source of natural gas used in this study is considered to be a 2009 average mixture. It was assumed that the distance of transmission would be 605 miles, based on average well-to-gate transmission distances.

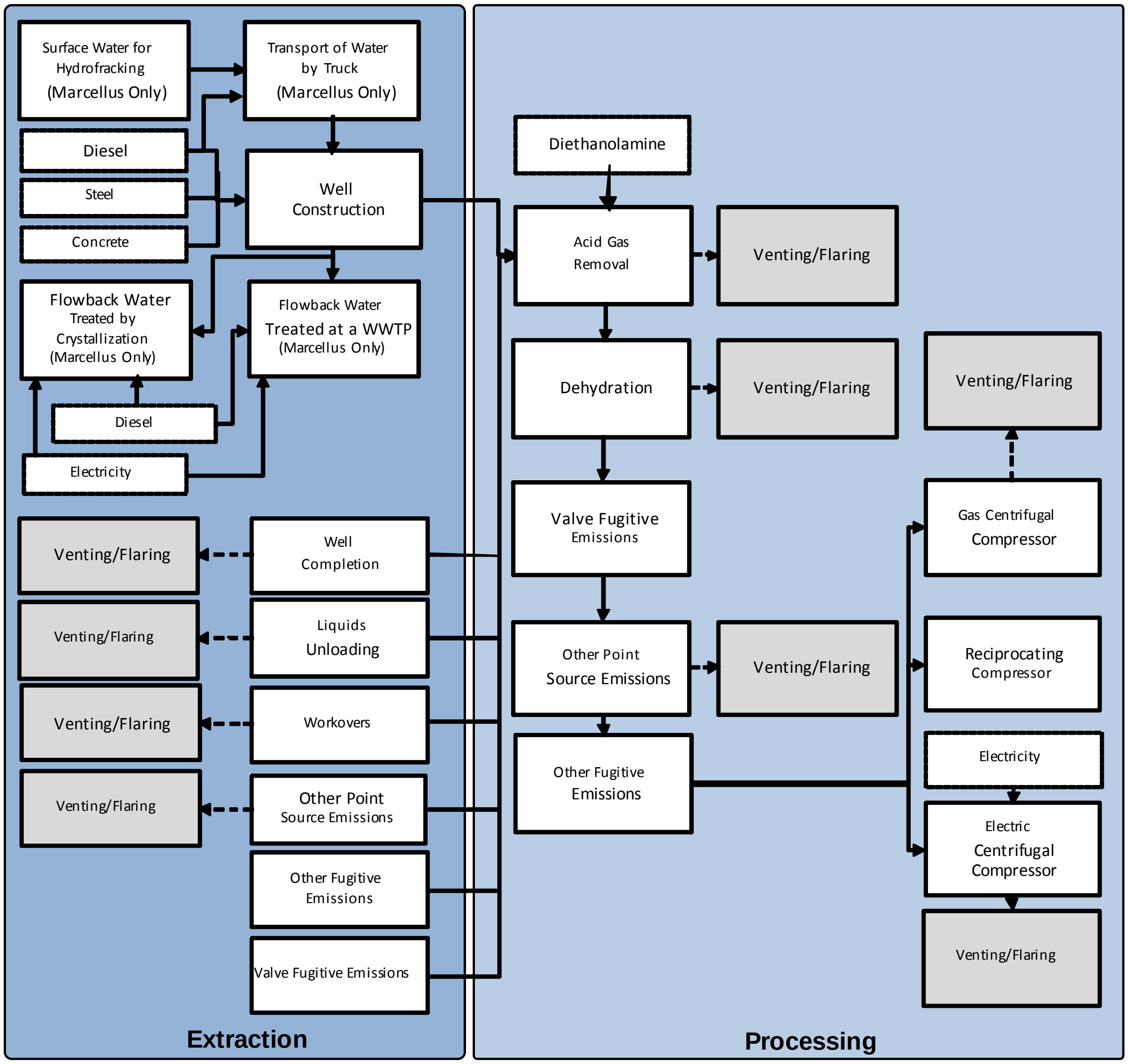
The energy and material flows tracked by NETL's LCA method in support of this study are used to quantify emissions of GHGs (CO_2, CH_4, N_2O, and SF_6) that would result from natural gas extraction and transport. The methods for calculating these flows for the raw material acquisition (RMA) and raw material transport (RMT) of natural gas and coal are provided below.

In this analysis, the boundary of the RMA for natural gas begins with the extraction of natural gas from nature and ends with processed natural gas ready for pipeline delivery. Key activities in the RMA of natural gas are as follows:

- Well construction and installation
- Natural gas sweetening (acid gas removal)
- Natural gas dehydration
- Natural gas venting and flaring
- Other point source and fugitive emissions
- Natural gas compression
- Well decommissioning

The relationships among natural gas extraction and processing activities, as modeled by NETL's LCA model, are illustrated in **Exhibit 7-15**.

Exhibit 7-15: Unit Process Network for Natural Gas Extraction and Processing



The data sources and assumptions for calculating the GHG emissions from each RMA activity are provided below. In most cases, the methane emissions are calculated by using standard engineering calculations around key gas-field equipment, followed by the application of the Environmental Protection Agency's (EPA) AP-42 emission factors as necessary.

7.4.1.1 Well Construction and Installation

NETL's LCA model of natural gas extraction includes the construction and installation activities for natural gas wells. Construction is defined as the cradle-to-gate burdens of key materials that embody key equipment and structures. Installation is defined as the activity of preparing a site, erecting buildings or other structures, and putting equipment in place.

The construction of natural gas wells requires a well casing that provides strength to the well bore and prevents contamination of the geological formations that surround the gas reservoir. A well is lined with a carbon steel casing that is held in place with concrete. A typical casing has an inner diameter of 8.6 inches, is 0.75 inches thick, and weighs 24 pounds per foot.[39] The weight of concrete used by the well walls is assumed to be equal to the weight of the steel casing. The total length of a natural gas well is variable, based on the natural gas extraction profile under consideration. The total weight of materials for the construction of a well bore is estimated by factoring the total well length by the linear weight of carbon steel and concrete.

The installation of natural gas wells includes the drilling of the well, followed by the installation of the well casing. An advanced drilling rig has a drilling speed of 17.8 meters per hour, which translates to the drilling of a 7,000-foot well in approximately 10 days.[39] A typical diesel engine used for oil and gas exploration has a power of 700 horsepower and a heat rate of 7,000 Btu/hp-hr.[40] The methane emissions from well installation is the product of the following three variables: heat rate of drilling engine (7,000 Btu/hp-hr), methane emission factor[40] for diesel combustion in stationary industrial engines (6.35E-05 lb/hp-hr), and the total drilling time (in hours).

The daily production rate of a natural gas well is an important factor in apportioning one-time construction activities or intermittent operations to a unit of natural gas production. Typical production rates vary considerably based on well type. Production rates also vary based on well specific factors, such as the age of the natural gas well.

The construction and material requirements are apportioned to one kilogram of natural gas product by dividing them by the lifetime production of the well. The natural gas wells considered in this study are presumed to produce natural gas at the rates discussed above, with a lifetime of 30 years. Thus, construction and material requirements, and associated GHG emissions, are apportioned over the lifetime production rate specific to each type of natural gas well, based on average well production rates.

7.4.1.2 Natural Gas Sweetening (Acid Gas Removal)

Raw natural gas contains varying levels of hydrogen sulfide (H_2S), a toxic gas that reduces the heat content of natural gas and causes fouling when combusted in equipment. The removal of H_2S from natural gas is known as sweetening. Amine-based processes are the predominant technologies for the sweetening of natural gas.

The H_2S content of raw natural gas is highly variable, with concentrations ranging from one part per million on a mass basis to 16 percent by mass in extreme cases. An H_2S concentration of 0.5 percent by mass is modeled in this analysis. This H_2S concentration is based on raw gas composition data compiled by the Gas Processors Association.[41]

The amine reboiler combusts natural gas to generate heat for amine regeneration. This analysis applies EPA emission factors for industrial boilers[40] to the energy consumption rate in order to estimate the combustion emissions from amine reboilers.

The sweetening of natural gas is also a source of vented methane emissions. In addition to absorbing H_2S, the amine solution also absorbs a portion of methane from the natural gas. This methane is released to the atmosphere during the regeneration of the amine solvent. The venting of methane from natural-gas sweetening is based on emission factors developed by the Gas Research Institute.[42]

Raw natural gas contains naturally-occurring CO_2 that contributes to the acidity of natural gas. Most of this CO_2 is absorbed by the amine solution during the sweetening of natural gas and is ultimately released to the atmosphere when the amine is regenerated. This analysis calculates the mass of naturally-occurring CO_2 emissions from the acid gas recovery (AGR) unit by balancing the composition of production gas (natural gas that has been extracted but has not undergone significant processing) and pipeline-quality gas.

The majority (84 percent by mass) of the AGR vent stream is non-methane volatile organic compounds (NMVOC). At this concentration, NMVOCs are a high-value energy product. Thus, from an LCA perspective, NMVOCs are a valuable co-product of the AGR process. Co-product allocation is used to apportion life cycle emissions and other burdens between the natural gas and NMVOC products. In this analysis, the relative energy contents of the natural gas and NMVOC outputs from the AGR process are used as the basis for co-product allocation. When these mass flows are converted to an energy basis using their respective heating values, 88.1 percent of the product leaving the AGR process is natural gas and 11.9 percent is NMVOCs. Thus, the natural gas model allocates 88.1 percent of the energy requirements and environmental emissions of acid gas removal to the natural gas product.

7.4.1.3 Natural Gas Dehydration

Dehydration is necessary to remove water from raw natural gas, which makes it suitable for pipeline transport and increases its heating value. The configuration of a typical dehydration process includes an absorber vessel in which glycol-based solution comes into contact with a raw natural gas stream, followed by a stripping column in which the rich glycol solution is heated in order to drive off the water and regenerate the glycol solution. The regenerated glycol solution (the lean solvent) is recirculated to the absorber vessel. The methane emissions from dehydration operations include combustion and venting emissions. This analysis estimates the fuel requirements and venting losses of dehydration in order to determine total methane emissions from dehydration.

In addition to absorbing water, the glycol solution also absorbs methane from the natural gas stream. This methane is lost to evaporation during the regeneration of glycol in the stripper column. Flash separators are used to capture most of methane emissions from glycol strippers; nonetheless, small amounts of methane are vented from dehydrators. The emission of methane from glycol dehydration is based on emission factors developed by the Gas Research Institute,[42] Based on this emission factor, 8.06E-06 lb of methane is released for every pound of natural gas that is dehydrated.

7.4.1.3.1 Natural Gas Venting and Flaring

Venting and flaring are necessary in situations where a natural gas (or other hydrocarbons) stream cannot be safely or economically recovered. Venting and flaring may occur when a well is being prepared for operations and the wellhead has not yet been fitted with a valve manifold, when it is not financially preferable to recover the associated natural gas from an oil well, or during emergency operations when the usual systems for gas recovery are not available.

The combustion products of flaring include CO_2, CH_4, and NO_X. The flaring emission factors published by the American Petroleum Institute[42] are based on the following recommendations by the IPCC:

- If measured data are not available, assume flaring has a destruction efficiency of 98 percent. Destruction efficiency is a measure of how much carbon in the flared gas is converted to CO_2[41]
- The CO_2 emissions from flaring are the product of the destruction efficiency, carbon content of the flared gas, and the molar ratio of CO_2 to carbon (44/12). Methane is 75 percent carbon by mass, and the other hydrocarbons in natural gas are approximately 81 percent carbon by mass[27]; the composite carbon content of natural gas is calculated by factoring these carbon compositions with the natural gas composition
- Methane emissions from flaring are equal to the two-percent portion of gas that is not converted to CO_2[42]
- N_2O emissions from flaring are based on EPA AP-42 emission factors for stationary combustion sources[42]

Recent data indicate that only 51 percent of vented natural gas from conventional natural gas extraction operations is flared and the remaining 49 percent is released to the atmosphere.[42] The flaring rate at natural gas processing plants is assumed to be 100 percent.

Sources of venting include:

- Venting from well completion
- Venting from well workovers
- Venting from liquid unloading (not necessary for unconventional wells or associated gas wells)
- Venting from wet seal degassing
- Fugitive emissions from pneumatic devices

Unlike well completions, well workovers occur more than one time during the life of a well. The frequency of well workovers was calculated using EPA's accounting of the total number of natural gas wells in the U.S. and the total number of workovers performed per year (all data representative of 2007).

Liquid unloading is a routine operation for conventional gas wells. The frequency of liquid unloading was calculated using EPA's assessment of two producers and the unloading activities for their wells.[43]

7.4.1.4 Other Point Source and Fugitive Emissions

The emissions described above account for natural gas emissions from specific processes, including the episodic releases of natural gas during well completion, workovers, and liquid unloading, as well as routine releases from wet-seal degassing, AGR, and dehydration. Natural gas is also released by other extraction and processing equipment. To account for these other emissions, NETL’s model includes two additional emission categories: other point source emissions and other fugitive emissions. Other point source emissions account for natural gas emissions that are not accounted for elsewhere in the model and can be recovered for flaring. Other fugitive emissions include emissions that are not accounted for elsewhere in the model and cannot be recovered for flaring.

EPA's Background Technical Support Document "Greenhouse Gas Emissions Reporting from the Petroleum and Natural Gas Industry Background"[43] was used for quantifying the other point source and fugitive emissions from natural gas extraction and processing. A three-step process was used to filter EPA's venting and flaring data to maintain consistency with the boundary assumptions of this analysis.

Emissions that are accounted for by NETL's existing natural gas unit processes were not included in the categories for other point source and fugitive emissions. For example, EPA provides emission rates for well construction, well completion, dehydration, and pneumatic devices. The emissions from these activities are accounted for elsewhere in NETL's model and, to avoid double counting, are not included in the emission factors for other point and fugitive emissions.

Emissions that fall within NETL's boundary definitions for natural gas processing were moved from the natural-gas-extraction category to the natural-gas-processing category.

The EPA data[43] does not discern between point source and fugitive emissions, so emissions were assigned to the point source or fugitive emission categories based on another EPA reference that provides more details on point source and fugitive emissions.[44]

7.4.1.5 Natural Gas Compression

Compressors are used to increase the gas pressure for pipeline distribution. This analysis assumes that the inlet pressure to compressors at the natural gas extraction and processing site is 50 psig, and the outlet pressure is 800 psig. The inlet pressure depends on the pressure of the natural gas reservoir and pressure drop during gas processing and introduces uncertainty to the model. The outlet pressure of 800 psig is a standard pressure for pipeline transport of natural gas.

7.4.1.6 Well Decommissioning

This analysis assumes that the de-installation of a natural gas well incurs ten percent of the energy requirements and emissions as the original installation of the well.

7.4.2 Natural Gas Transport

The boundary for natural gas transport begins with receipt of processed natural gas at the extraction site and ends with the delivery of natural gas to an energy conversion facility. The data used to represent the pipeline transport of natural gas are described below.

7.4.2.1 Pipeline Construction and Decommissioning

Carbon steel is the primary material used in the construction of natural gas pipelines. The mass of pipeline per unit length was determined using an online calculator.[45] The weight of valves and fittings were estimated at an additional 10 percent of the total pipeline weight. The pipeline was assumed to have a life of 30 years. The mass of pipeline construction per kilogram of natural gas was determined by dividing the total pipeline weight by the total natural gas flow through the pipeline for a 30-year period.

The decommissioning of a natural gas pipeline involves cleaning and capping activities. This analysis assumes that the decommissioning of a natural gas pipeline incurs 10 percent of the energy requirements and emissions as the original installation of the pipeline.

7.4.2.2 Pipeline Operation (Energy and Combustion Emissions)

The U.S. has an extensive natural gas pipeline network that connects natural gas supplies and markets. Compressor stations are necessary every 50 to 100 miles along the natural gas transmission pipelines in order to boost the pressure of the natural gas. Compressor stations consist of centrifugal and reciprocating compressors. Most natural gas compressors are powered by natural gas, but, when electricity is available, electrically-powered compressors are used.

A 2008 paper published by the Interstate Natural Gas Association of America provides data from its 2004 database, which shows that the U.S. pipeline transmission network has 5,400 reciprocating compressors and over 1,000 gas turbine compressors.[46] Further, based on written communication from El Paso Pipeline Group, approximately three percent of transmission compressors are electrically driven.[47] El Paso Pipeline Group has the highest transmission capacity of all natural gas pipeline companies in the U.S., and it is thus assumed that the share of electrically-powered compressors in their fleet is representative of the entire natural gas transmission network. Based on written communication with El Paso Pipeline Group,[47] the share of compressors on the U.S. natural gas pipeline transmission network is approximately 78 percent reciprocating compressors, 19 percent turbine-powered centrifugal compressors, and 3 percent electrically-powered compressors.

The use rate of natural gas for fuel in transmission compressors was calculated from the Federal Energy Regulatory Commission (FERC) Form 2 database, which is based on an annual survey of gas producers and pipeline companies.[48] The 28 largest pipeline companies were pulled from the FERC Form 2 database. These 28 companies represent 81 percent of NG transmission in 2008. The FERC data for 81 percent of U.S. natural gas transmission is assumed to be a representative sample of the fuel use rate of the entire transmission network. This data shows that 0.96 percent of natural gas product is consumed as compressor fuel. This fuel use rate was converted to a basis of kg of natural gas consumed per kg of natural gas transported by multiplying it by the total natural gas delivered by the transmission network in 2008[49] and dividing it by the annual tonne-km of pipeline transmission in the U.S.[50] The total delivery of natural gas in 2008 was 21 Tcf, which is approximately 400 billion kg of natural gas. The annual transport rate for natural gas transmission was steady from 1995 through 2003, at approximately 380 billion tonne-km per year. More recent transportation data are not available; thus, this analysis assumes the same tonne-km rate for 2008 as shown from 1995 through 2003.

The air emissions from the combustion of natural gas by compressors are estimated by applying EPA emission factors to the natural gas consumption rate of the compressors.[51] Specifically, the emission profile of gas-powered, centrifugal compressors is based on emission factors for gas turbines; the emission profile of gas-powered, reciprocating compressors is based on emission factors for 4-stroke, lean-burn engines. For electrically-powered compressors, this analysis assumes that the indirect emissions are representative of the U.S. average fuel mix for electricity generation.

The average power of electrically-driven compressors for U.S. natural gas (NG) transmission is assumed to be the same as the average power of all compressors on the transmission network. An average compressor on the U.S. natural gas transmission network has a power rating of 14,055 horsepower (10.5 MW) and a throughput of 734 million cubic feet of natural gas per day (583,000 kg NG/hour).[52] Electrically-driven compressors have efficiencies of 95 percent.[46, 53] This efficiency is the ratio of mechanical power output to electrical power input. Thus, approximately 1.05 MWh of electricity is required per MWh of compressor energy output.

In addition to air emissions from combustion processes, fugitive venting from pipeline equipment results in the methane emissions to air. The fugitive emission rate for natural gas pipeline operations is based on data published by the Bureau of Transportation Statistics (BTS) and EPA. The transport data for natural gas transmission is based on ton-mileage estimates by BTS, which calculates 253 billion ton-miles of natural gas transmission in 2003.[54] The 2003 data are the most recent data point in the BTS reference, and thus EPA's inventory data for the years 2000 and 2005 were interpolated to arrive at a year 2003 value of 1,985 million kg of fugitive methane emissions per year.[55] Dividing the EPA emission by the transport requirements and converting to metric units gives 5.37E-06 kg/kg-km.

7.4.2.3 Pipeline Operation (Fugitive Methane)

Methane emissions from pipeline operations are a function of pipeline distance. This analysis uses a pipeline transport distance of 605 miles, which is the average distance for natural gas pipeline transmission in the U.S. and was calculated by balancing national emission inventory[55] and natural gas consumption data[49] with NETL's unit process emission factor for fugitive methane emissions from pipeline operations. The data sources and assumptions for calculating the greenhouse gas emissions from construction and operation of natural gas transmission pipelines are discussed below.

7.4.3 GTL Operation

The design of a GTL facility that uses natural gas feedstocks to produce diesel and gasoline is discussed in detail earlier in this study (see Section 3). These design specifications were modeled using AspenPlus®, and the resulting stream flows for feedstocks, products, and emissions were adapted to NETL's unit process format for LCA modeling. With respect to the LCA model, the GTL facility is a black box. The LCA model does not have the capability to change parameters that affect the performance of the GTL facility.

Exhibit 7-16 shows input and output flows of the GTL facility, as accounted for by the LCA model. All flows are expressed on the basis of 1 kg of Fischer-Tropsch diesel production.

Exhibit 7-16: Unit Process Flows for GTL Operation

Flow Name	Diesel Reference Flow	Gasoline Reference Flow	Units
Inputs			
Natural Gas	2.03	4.8	kg
Butane (n-butane)	0.11	0.26	kg
Water (Ground)	5.47	13.0	L
Water (Surface)	5.47	13.0	L
Outputs			
Diesel (Co-product)	1	2.4	kg
Gasoline (Co-product)	0.42	1	kg
Carbon Dioxide (Air Emission)	0.07	0.17	kg
Nitrogen Oxides (Air Emission)	1.96E-06	4.67E-06	kg
Carbon Dioxide (Captured)	1.22	2.90	kg
Electricity (Co-product)	2.34E-04	5.57E-04	MWh
Water (Waste)	2.55	6.07	L

7.4.4 Carbon Capture and Sequestration

Saline aquifers are geological formations that are saturated with brine water and can be used for long-term CO_2 storage. The development and use of a saline aquifer site for CO_2 sequestration includes site preparation, construction, operation, monitoring, brine management, and well closure.

7.4.4.1 Site Preparation

The preparation of a saline aquifer site requires a seismic survey conducted by vibroseis trucks, or specialized explosives, which vibrate the ground and use seismic equipment to measure the geological characteristics of a site. When vibroseis trucks cannot be used, due to complications in terrain or access, specialized explosives are used. This involves the drilling of shot holes, loading the charges and detonation. These trucks consume diesel for transport and equipment operation. The survey of a typical site takes seven 12-hour days.[56]

7.4.4.2 Well Construction

The construction and installation of wells includes the drilling of the well bore followed by the installation of a well casing. The well casing provides strength to the well bore and prevents contamination of groundwater that surrounds the well.

Eight different well types of varying depths are required for CO_2 sequestration in a saline aquifer: stratigraphic test, injection, reservoir monitoring, above-seal monitoring, groundwater monitoring, vadose zone monitoring, water production, and water disposal. NETL's saline aquifer storage cost model contains a representative list of possible storage formations in the U.S. For each formation, the model provides the depth of each well type based on the special characteristics of the formation. This analysis uses the average well depths from the selected list of storage formations for the calculation of drilling and casing requirements.[57]

Vertical drilling is used for the wells required for saline aquifer sequestration. Most drilling rigs use diesel fuel. A diesel-powered drilling rig has a power output of 600 horsepower[51] and a top drilling speed of 17.8 meters per hour.[58]

Wells are lined with threaded chromium and molybdenum-alloy casings that are held in place with concrete. For the purposes of the model, each well at a saline aquifer site, with the exception of the groundwater and vadose zone monitoring wells, is assumed to have three concentric casing sections of varying diameters and depths.[59] The model uses average casing diameters and casing string lengths for calculating the environmental impacts. The top string of casing, known as the conductor casing, is assumed to have a 16-inch diameter casing set in a 26-inch diameter hole to a depth of 40 feet (12.2 m). The next casing section, known as the surface casing, consists of an 8 5/8-inch diameter pipe set in a 12 1/4-inch hole extending from the surface to a depth of 2,167 feet (660.5 m). The final casing section in the series, known as the production casing, consists of a 5 1/2-inch pipe set in a 7 7/8-inch hole extending from the surface down to the depth determined from the NETL saline storage cost model for each well type.[57]

7.4.4.3 Sequestration Operations

The operation of CO_2 injection site uses electricity to pressurize and inject incoming CO_2 into an underground formation. The electricity requirements of a given injection site are a function of injection pressure and the number of injection wells. The required injection pressure is calculated based on the representative list of possible storage formations provided in the NETL saline aquifer storage cost model.[57] For each formation, the model provides the pressure at the midpoint based on the corresponding geological parameters.

CO_2 arrives by pipeline as a pressurized fluid at a pressure of at least 7.38 MPa (1,070 psia), a required pressure that ensures all CO_2 is in a supercritical state at standard ambient temperatures. To achieve the correct injection pressure, a CO_2 injection pump must overcome the pressure drop that occurs in the injection well, between the surface and the aquifer.[60] The boosting of supercritical CO_2 from its critical point to the required injection pressure is provided by electrically-powered injection pumps. Fugitive emissions of CO_2 escape through pump seals during the operation of CO_2 injection pumps.

In addition to the fugitive CO_2 emissions from the injection pump, this analysis also assumes leakage of CO_2 from the underground storage formation. It is assumed that a maximum of one percent of the stored CO_2 eventually migrates to the surface and is released to the atmosphere over a 100-year monitoring period. (This conservative assumption is consistent with other NETL reports on CCS and is used to bracket the current range of potential loss until measurement data from operating storage sites can validate this loss factor.) The expected parameter value for the model (0.5 percent) was selected as the midpoint between the maximum leakage rate of 1 percent and no leakage from the formation.

Brine water production from the saline aquifer is one method to control the pressure in the underground formation, but it is not always required.[61] Extraction of water from the aquifer storage formation occurs at a safe distance from the injection wells to prevent migration of CO_2 to the surface with the produced brine.

7.4.4.4 Site Monitoring

With respect to site monitoring, this LCA model accounts for the construction of monitoring wells and seismic testing during site operations. Other types of monitoring activities are a negligible contribution to the environmental burdens of a saline aquifer storage site. This conclusion is supported by a detailed list of capital and O&M costs developed by NETL in support of its CO_2 transport and storage cost model, which shows that constructing wells and conducting seismic surveys accounts for over 90 percent of lifetime capital and operating costs.[57] This is further supported by EPA’s analysis of the costs of geologic CO_2 sequestration, which concludes that the drilling of monitoring wells accounts for the majority of site monitoring costs.[62] Seismic monitoring during the operation phase of the aquifer is modeled in the same way as for site preparation**.**

7.4.4.5 Brine Management

The management of brine at a saline aquifer site consumes electricity, which is used by water treatment processes and/or injection pumps. Two water treatment technologies were used in this analysis: reverse osmosis and vapor compression distillation. The choice between these technologies depends on the concentration of total dissolved solids (TDS) in the brine. Brine water quality is highly variable among saline aquifer injection sites, with TDS levels ranging from 10,000 to over 300,000 mg/L.[61] Reverse osmosis is effective at handling water with a TDS of less than 50,000 mg/L, while distillation is effective at higher TDS concentrations.[63] The concentrated waste stream produced from water treatment can either be re-injected into a suitable formation or transported offsite for additional processing.

Instead of treating the produced brine water at the surface, it may be desirable to instead re-inject the stream into a suitable underground formation near the production site. This practice is common in the oil and gas industry for onshore wells.[61] The key operating requirement for re-injection is the electricity used by injection pumps. Injection pumps are sized to overcome the head losses in a water injection well.

7.4.4.6 Well Closure

The purpose for plugging wells prior to abandonment is to ensure that the abandoned wells do not allow the injected fluids (in this case, CO_2) or natural brines to migrate up the well bore and into underground sources of drinking water (USDW).[64] The EPA provides guidance on plugging and abandoning wells of various types but has yet to provide specific guidance for wells that would be used for CO_2 sequestration, which are defined as Class VI wells.[65] This analysis uses the EPA guidance for Class II wells, defined as wells that inject fluids that are brought to the surface in connection with conventional oil or natural gas production, are used for enhanced recovery of oil or natural gas, or are used for storage of hydrocarbons.[64] Class II guidance is a good approximation since the well depth, usage, and pressure would be similar to Class VI wells. The EPA guidance[64] also includes options for how to implement the bottommost plug in Class II wells, which significantly changes the amount of concrete used. Both methods, either cast iron bridge plug or the use of a cement retainer (also made out of cast iron), are accounted for in this analysis. Additionally, the shallow vadose zone and ground water monitoring wells are assumed to be plugged entirely with concrete.

7.4.5 Displaced Products

The GTL facility produces three co-products: diesel, gasoline, and electricity. When using system expansion with displacement to manage these co-products, it is necessary to model the cradle-to-gate life cycle of each displaced co-product. When diesel is the functional unit, co-produced gasoline displaces conventional petroleum gasoline and co-produced electricity displaces average power produced by the U.S. electricity grid. When gasoline is the functional unit, co-produced diesel displaces conventional petroleum diesel and co-produced electricity displaces average power produced by the U.S. electricity grid. No scenarios of this analysis have electricity as a functional unit, so it is not necessary to model the displacement of gasoline and diesel in the same scenario.

Displacement factors for electricity account the cradle-to-gate generation of electricity at the power plant busbar and are based on the average 2010 U.S. grid mix. However, to model uncertainty, extreme life cycle values for electricity are also used. Fleet coal is used to represent a high value for displaced electricity, and the predicted EIA 2035 grid mix is used to represent a low value of displaced electricity. These factors were generated by the electricity mixers in NETL’s LCA library (www.netl.doe.gov/LCA).

The displacement factors for diesel and gasoline account for the cradle-to-gate production of conventional petroleum fuel, beginning with crude oil extraction and ending with refined product exiting a petroleum refinery. The values for these displacement factors were generated using NETL’s baseline petroleum model.[33] The production of diesel and gasoline from imported (non-North American) crude was used to represent a high value for displacement. A displacement value of zero was used to represent the lowest possible value for displacement of diesel or gasoline; in this context, a displacement value of zero is representative of a liquid fuels market that does not reduce production of conventional fuels in response to the introduction of GTL products.

Exhibit 7-17 shows the displacement factors for GTL co-products. The low and high values represent the low and high values from an input perspective, and not necessarily the low and high values from a results perspective. Since the magnitude of a displacement value has an inverse relationship with a life cycle result, the low displacement values in **Exhibit 7-17** correspond to the *high* values in the final LCA results, and the high displacement values in **Exhibit 7-17** correspond to the *low* values in the final LCA results.

Exhibit 7-17: Displacement Factors for GTL Co-products

Co-Product	Low Value	Expected Value	High Value
Electricity	AEO 2035 U.S. Grid Mix (671 kg CO_2e/MWh)	U.S. Grid Mix (707 kg CO_2e/MWh)	Fleet Coal (1,161 kg CO_2e/MWh)
Diesel	No Displacement	100% Displacement of Diesel from Imported Crude Mix	100% Displacement of Diesel from Imported Crude Mix
Gasoline	No Displacement	100% Displacement of Gasoline from Imported Crude Mix	100% Displacement of Gasoline from Imported Crude Mix

7.4.6 Fuel Transport and Delivery

There are five transport modes for liquid fuels in the U.S.: pipeline, ocean tanker, barge, railroad, and truck. The energy consumption and GHG emissions for these transport modes are based on NETL's baseline LCA for petroleum-based fuels,[33] which uses data representative of petroleum transport activities in 2004.

The energy intensity of petroleum pipeline transport is 260 Btu/ton-mile.[66] The energy intensity of water carriers (ocean tankers and barges) is 514 Btu/ton-mile, the energy intensity of rail transport is 337 Btu/ton-mile, and the energy intensity of a tanker truck with a 25-ton capacity is 822 Btu/ton-mile.[67]

In 2004, natural gas and electricity were the primary fuels for pipeline transport. All crude oil and petroleum product transport by pipeline is assumed to be fueled by electricity. Diesel is the only type of fuel used for freight railroad operations. Water carriers transporting freight are powered by diesel fuel and residual fuel oil at a 30.5 and 69.5 percentage split, respectively. Tanker trucks used for road transport of petroleum are powered by diesel fuel.[68]

In 2004, there were 528 billion ton-miles of domestic petroleum transport. Pipelines accounted for 59.8 percent, water carriers accounted for 29.9 percent, trucks accounted for 6.3 percent, and rail accounted for 4.0 percent of total ton-mileage.[68]

The total energy consumption for transport of petroleum products in 2004 and their associated emissions are allocated to the volume of petroleum products produced or imported in 2004. This total volume is 7,781,225 thousand barrels and was calculated as follows: *total refinery products* minus *refinery products used on-site as a fuel* minus *petrochemical feedstocks* plus *total imported petroleum products* plus *natural gas liquids production*.[33]

Emission factors specific to each fuel type and combustion technology were applied to the total fuel consumed by the above transport modes. The electricity consumed by pipeline transport is representative of the average North American electricity grid as modeled by NETL's power mixer. Emission factors for water carriers and rail transport are cited by EIA's technical documentation for voluntary emissions reporting.[69] Emission factors for heavy duty trucks are cited in API's documentation of GHG estimation methods for the oil and gas industry.[70]

7.4.7 Fuel Combustion

The final stage of this LCA is the combustion of fuel in a vehicle. This analysis uses the combustion of 1 MJ of fuel as a functional unit. Converting this energy combustion to the basis of vehicle-miles traveled would improve the interpretation of results, but would require assumptions about vehicle type and fuel efficiency, which would introduce further uncertainty to this analysis.

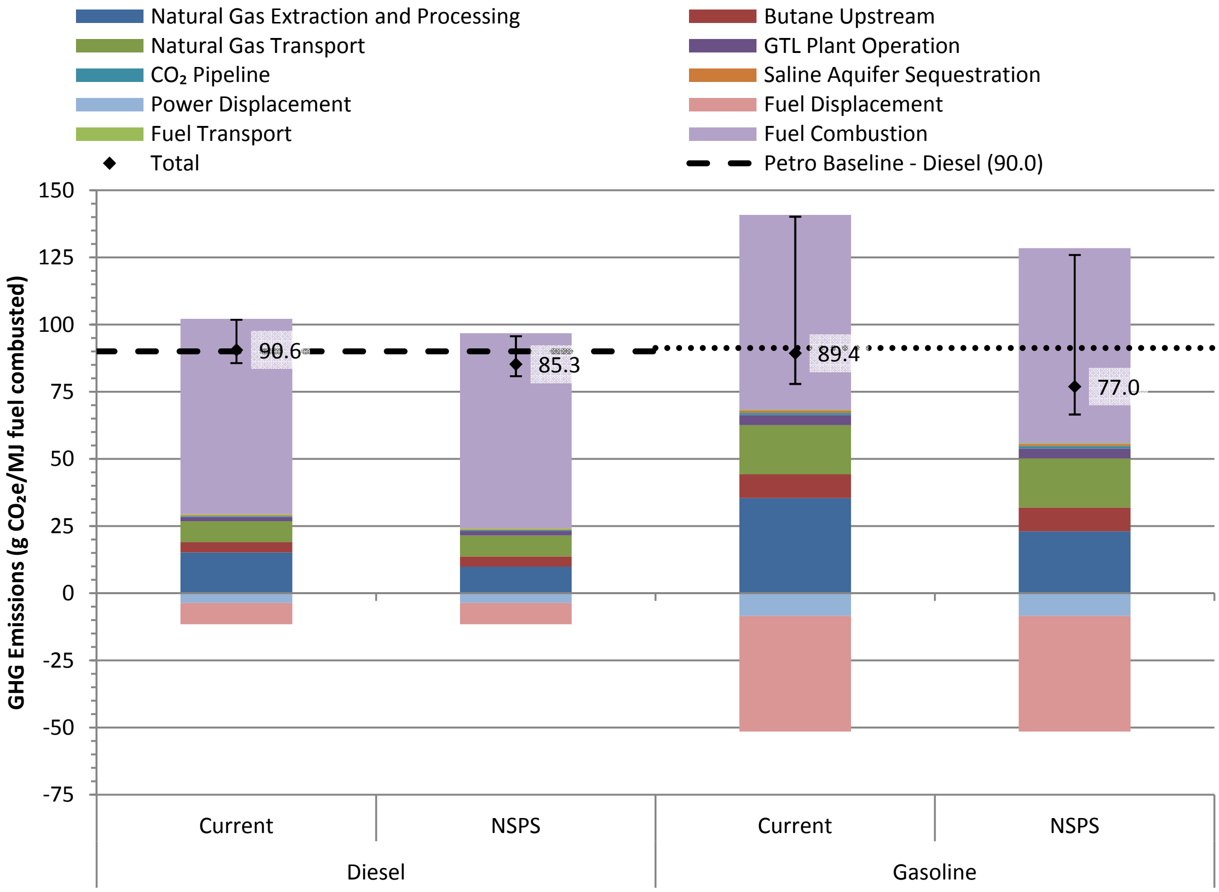
Combustion emissions per unit of fuel are calculated by EPA Motor Vehicle Emission Simulator (MOVES) model.[71] This analysis uses a conventional internal combustion engine in the passenger car vehicle class to represent diesel combustion. The combustion of diesel on an LHV basis produces 76.7 kg CO_2/MMBtu (72.7 g CO_2e/MJ. Small amounts of methane and nitrous oxide are also emitted from the combustion of these fuels. On the basis of 2007 IPCC 100-year global warming potentials, CO_2 emissions account for 99.9 percent of GHG emissions from diesel combustion. The combustion of gasoline on an LHV basis produces 75.0 kg CO_2/MMBtu (71.1 g CO_2/MJ). Small amounts of methane and nitrous oxide are also emitted from the combustion of gasoline. On the basis of 2007 IPCC 100-year global warming potentials, CO_2 emissions account for 97.8 percent of GHG emissions from gasoline combustion. The total emission factor for combusted gasoline is 76.6 kg CO_2e/MMBtu (72.6 g CO_2e/MJ).

7.5 LCA Results

The LCA results of this analysis are restricted to GHG emissions, expressed as carbon dioxide equivalents (CO_2e) using IPCC 2007 100-yr global warming potentials. Scenarios for diesel and gasoline were assessed, using 1 MJ of combusted fuel as the functional unit. The LCA also accounts for near-term improvement to upstream natural gas emissions. Results were generated for current practices in the natural gas industry and for improved performance once the approved New Source Performance Standards (NSPS) for the Oil and Gas Sector are fully implemented. The combination of the two functional units (diesel and gasoline) and two upstream natural gas practices (current and NSPS) yield four scenarios for the LCA GHG results. The GHG results for these scenarios are shown **Exhibit 7-18**.

The life cycle GHG emissions for GTL diesel and gasoline when based on current practices in the natural gas industry are 90.6 g CO_2e/MJ and 89.4 g CO_2e/MJ, respectively. If the natural gas extraction and processing sector complies with NSPS, the upstream GHG emissions from natural gas are reduced by 23 percent. With the implementation of NSPS standards, the life cycle values for GTL diesel and gasoline decrease to 85.3 g CO_2e/MJ (5.9 percent reduction) and 77.0 g CO_2e/MJ (13.9 percent reduction), respectively. In **Exhibit 7-18**, these values are compared to the NETL petroleum baseline values for petroleum based fuels, which are 90.0 and 91.3 g CO_2e/MJ for diesel and gasoline, respectively. The expected LCA GHG emissions for GTL diesel are 0.6 percent higher than the baseline without NSPS, and 5.3 percent lower with the implementation of NSPS. For GTL gasoline, the LCA GHG emissions are 2.1 percent lower than the baseline without NSPS and 15.7 percent lower with the implementation of NSPS. As previously noted, the NSPS regulations will reduce the venting rates from well development and maintenance, increase the flaring rates at natural gas wells, and reduce compressor emissions at natural gas processing plants; this will lead to an overall reduction in the GHG emissions from natural gas extraction and processing.

Exhibit 7-18: LCA Greenhouse Gas Results for GTL Diesel and Gasoline



In addition to calculating expected values, this analysis also calculates uncertainty around expected values. The sources of uncertainty are illustrated by the error bars in **Exhibit 7-19** through **Exhibit 7-22**. The majority of the uncertainty is based on the range of displacement values used for fuel and electricity as detailed in **Exhibit 7-17**. Additional uncertainty is based on the upstream emissions that result from the extraction, production, and transport of natural gas. As shown by the following figures, the majority of methane emissions come from natural gas extraction, processing, and transport, while the majority of carbon dioxide emissions are from the GTL plant operations.

The uncertainty range for GTL gasoline using displacement is wider than the uncertainty range for GTL diesel using displacement. The GTL facility produces gasoline and diesel in a fixed proportion, with gasoline being produced at a lower rate than diesel. When using co-product displacement and fixing the results on the basis of 1 MJ of fuel combusted, the GTL gasoline scenario must displace 55.4 g of diesel to produce 1 MJ of gasoline, but the GTL diesel scenario must displace 9.97 g of gasoline to produce a 1 MJ of diesel.

Exhibit 7-19: Detailed GHG Results for GTL Diesel with Current Natural Gas

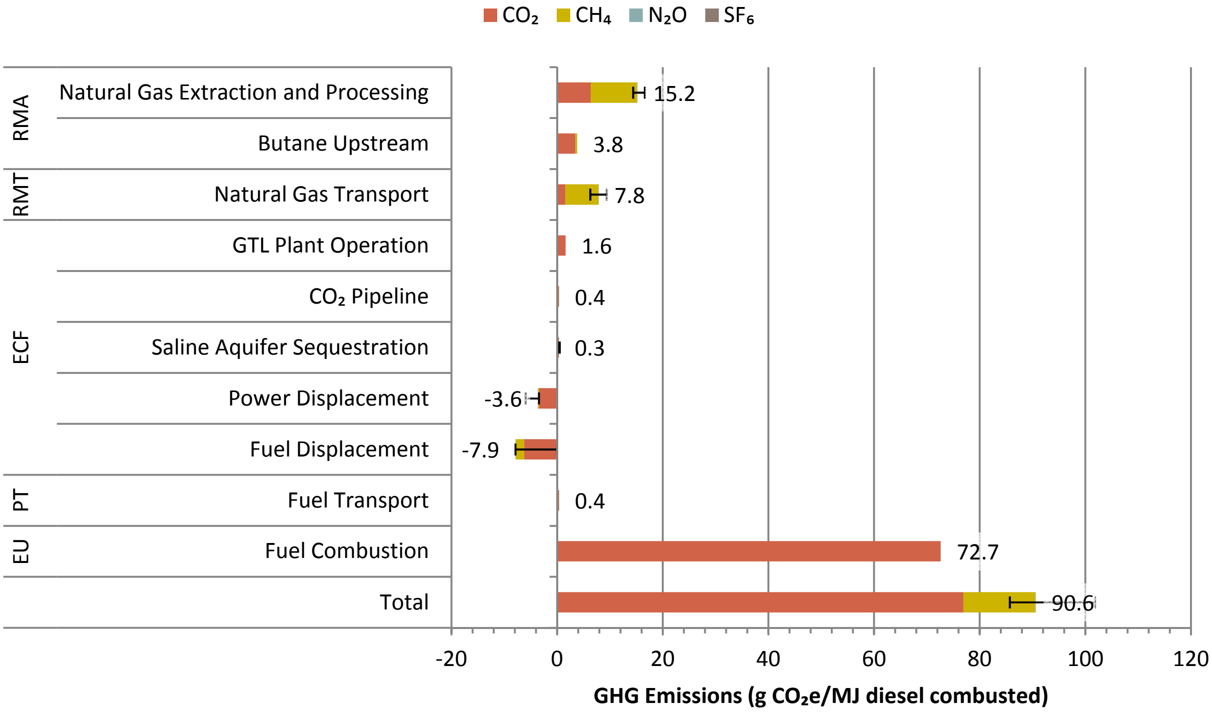


Exhibit 7-20: Detailed GHG Results for GTL Diesel with NSPS Natural Gas

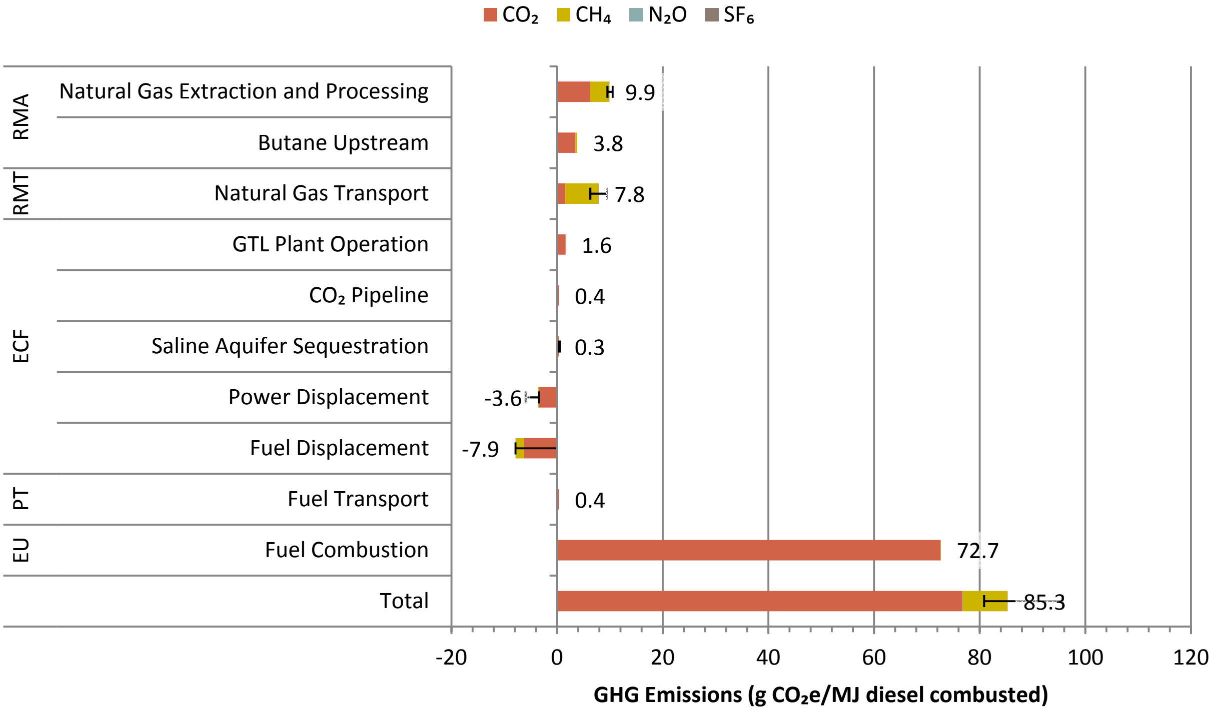


Exhibit 7-21: Detailed GHG Results for GTL Gasoline with Current Natural Gas

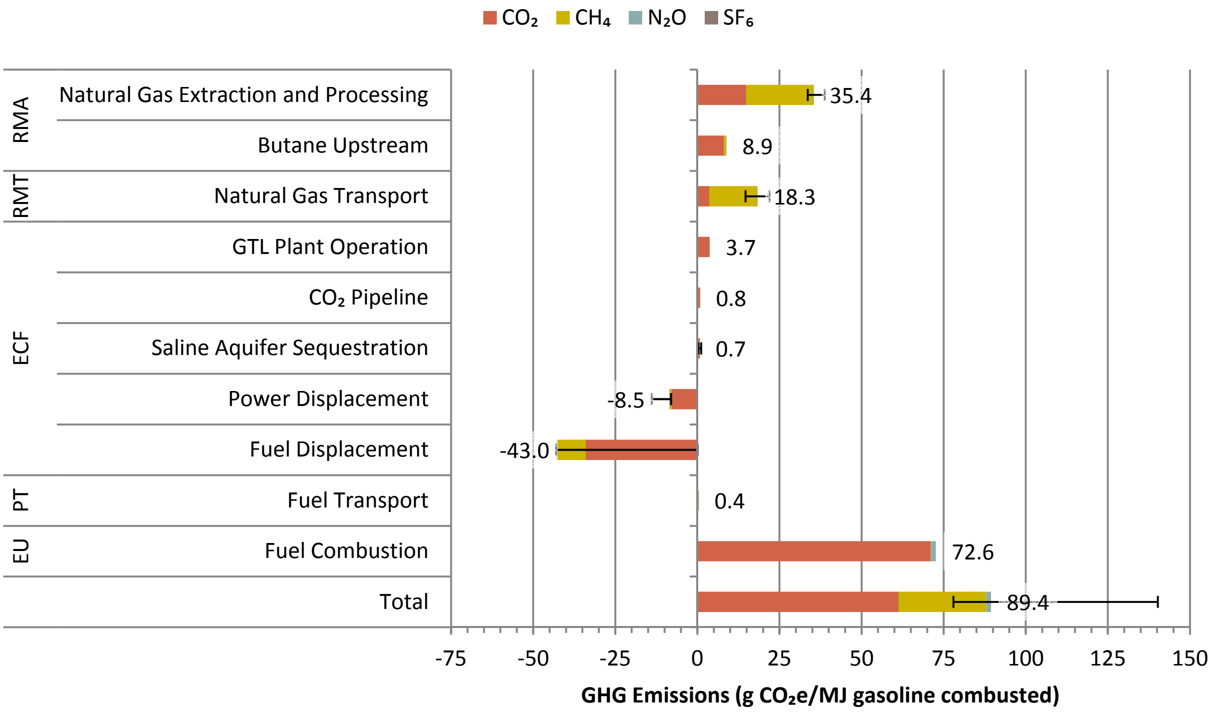
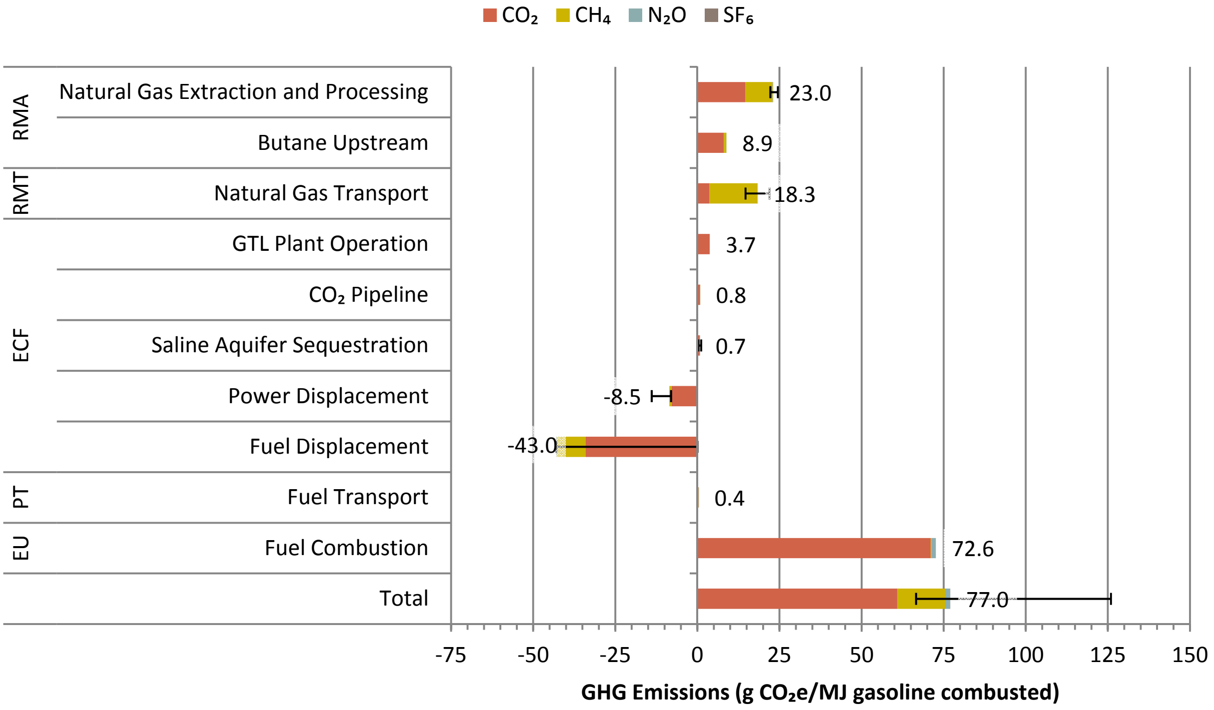


Exhibit 7-22: Detailed GHG Results for GTL Gasoline with NSPS Natural Gas



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