Alaska Gasline Development Corporation
Alaska Stand Alone Gas Pipeline/*ASAP*
Gas-to-Liquids Economic Feasibility Study
Final Report

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Executive Summary

Alaska Gasline Development Corporation (AGDC) has commissioned Hatch to conduct an economic feasibility study of a gas-to-liquids (GTL) plant to be situated at the terminus of the proposed Alaska Stand Alone Gas Pipeline/***ASAP*** at either Cook Inlet or Fairbanks. This study, titled ***ASAP*** GTL Study, investigates whether or not this GTL facility can act as an industrial anchor user of the pipeline and whether or not this GTL facility would justify an increase in pipeline capacity.

The GTL facility considered in this study uses proven technology to convert natural gas into liquid fuel products via syngas production and Fischer-Tropsch (FT) synthesis. Pipeline-quality natural gas is delivered to the plant battery limit and terminates with the delivery of the liquid products to the loading facilities for transport to market. The major plant areas included in this GTL facility are:

- Natural Gas Compression
- Air Separation
- Syngas Production
- FT Synthesis & Product Upgrading
- Power Generation
- Utilities

Three GTL facility sizes were considered, forming three process cases, in order to investigate the effect of facility size on the project economics. The number of trains (parallel streams of feed running through the plant) for the GTL facility are based on the number of FT synthesis reactors required. A 17,000 bbl/day FT reactor was considered for this study, based on the capacity of current commercial reactors. The base case (Case B) considers a GTL facility with two FT trains; Cases A and C consider one and four FT trains, respectively.

A GTL facility employing FT synthesis can be configured to produce a combination of diesel, jet fuel and naphtha in varying proportions. A market analysis for each product and various target markets (Alaska, U.S West Coast, Hawaii, Asia) was conducted to determine the optimum product mix for the base case facility. From the netback prices calculated for each product option, it was determined that a product mix of primarily diesel and naphtha should be produced by the base case GTL facility. For the economic analysis, an alternative product mix of jet fuel, diesel and naphtha was also assessed.

The GTL plant performance for all cases were calculated using Aspen Plus® (process simulation software) as summarized below.

	Case A (1 Train)	**Case B – Base Case (2 Trains)**	**Case C (4 Trains)**
Pipeline-quality Natural Gas Required [MMSCFD]	154	309	617
FT Products: Diesel + Naphtha [bbl/day]	16,630	33,260	66,520
Power Export [MWe]	60	119	239
Raw Water Intake [st/h]	228	445	910

The GTL facility design also includes process units to capture carbon dioxide (CO_2) from plant purge gas. In the Cook Inlet region, the recommended end use for the captured CO_2 is for enhanced oil recovery (EOR) in local oilfields, with a secondary option being to sequester the CO_2 in depleted gas reservoirs or saline aquifers. In the Fairbanks region, the captured CO_2 is recommended to be sequestered in deep, unmineable coal seams, which are located near the Healy coal fields.

The base case GTL facility is projected, based on previously completed and current GTL projects under construction, to being start-up and commissioning in Q4-2019, achieving nameplate capacity in the third quarter of 2020, in line with the availability of the ***ASAP***. Permitting and pre-feasibility engineering would begin in 2012.

A capital cost estimate (AACE Level 4 estimate, accuracy range of -30%, +40%) was prepared for the base case (Case B) in 2010 USD assuming a fully integrated gas-to-liquids facility based in Port MacKenzie. The fixed capital investment for the base case GTL facility (33,260 bbl/day of diesel and naphtha) is estimated to be USD 2.93 Billion or 88,000 USD/(bbl/day). Estimates for Cases A and C were scaled from Case B on a per train basis. It was assumed that a savings of 15% is common on subsequent trains. The resulting savings per unit of production from Case A to Case B and to Case C are 7.5% and 11.3%, respectively. The savings are attributable due to duplication of design; however, the majority of the costs are associated with procurement and construction tasks.

For the capital cost estimate of the GTL plant in Fairbanks, 25% was added to the base case CAPEX for Port MacKenzie in order to account for the complexity of construction and transportation of equipment to the Fairbanks facility. This brings the capital cost of the base case GTL facility at Fairbanks up to USD 3.66 Billion.

The operating cost of the Case B facility considers fixed (operations, maintenance, SG&A, insurance) and variable (raw materials, utilities) costs, and is estimated in 2010 USD to be USD 925 Million per year or 83 USD/bbl for the Port MacKenzie GTL facility, and USD 774 Million per year or 70 USD/bbl for the Fairbanks GTL facility. Natural gas accounts for approximately 90% of the GTL facility operating costs, driving the lower costs in Fairbanks since it is closer to the feedstock source. Variable expenses are directly related to plant production, and therefore for Cases A, B and C, the operating cost on a normalized USD/bbl basis are practically equal at this level of accuracy with the exception of natural gas. Since each case represents the anchor tenant for pipelines of different sizes, economies of scale for the pipeline are transferred to the GTL facility by the natural gas transportation tariff, resulting in the largest GTL plant (Case C) having the lowest natural gas price. Economies of scale are also shown by the differences in fixed operating costs. Note that the natural gas transportation cost is levelized in nominal terms, and therefore only the wellhead cost of 1.00 USD/MMBTU is escalated at 3% per annum, while the pipeline tariff is left constant over the valuation horizon.

An economic analysis was performed for the GTL facility considering an 80 USD/bbl crude oil price project and a natural gas price of 7.61 USD/MMBTU delivered to the base case facility at Port MacKenzie and 6.15 USD/MMBTU delivered to the base case facility at Fairbanks. Note that the transportation portion of delivered natural gas price is levelized in nominal terms and therefore not escalated. Other key assumptions include a 3% per year escalation rate; debt to equity ratio of 50:50; a required rate of return on equity of 12%; and the analysis was carried out in 2010 USD.

The economic analysis shows that for both sites, given the base case assumptions, the GTL plant would not meet the equity holders' required rate of return in order to result in a positive value proposition. However, the breakeven analysis reveals that a delivered natural gas price of 4.42 USD/MMBTU to the base case Port MacKenzie facility and 2.19 USD/MMBTU to the base case Fairbanks facility, all else constant, is required to yield the required 12% to its equity holders. Note that the base case Fairbanks facility was assumed in this study to have a 25% higher construction cost than the Port Mackenzie facility, and that the delivered natural gas price at Fairbanks will be inherently lower than Port Mackenzie due to the shorter - roughly 400 miles less - trip from gas source to plant gate. There is an economy-of-scale effect such that, the larger the plant, the higher the allowable delivered natural gas price.

The sensitivity analysis shows that the crude oil price, delivered natural gas price and CAPEX, in that order, are the most important drivers of the GTL plant's economics. It can also be concluded that a significant decrease in the delivered natural gas price or a significant increase in the crude oil price greatly improves the economic attractiveness of the project. However, a significant reduction in CAPEX, all else constant, has a much smaller effect on the project economics.

The sensitivity analysis also shows that under the stated assumptions, a reduction in the natural gas price of USD 1.46/MMBTU delivered to Fairbanks results in the ability of the Fairbanks GTL plant to absorb up to a 14% increase in CAPEX while remaining economically equivalent to the Port MacKenzie GTL plant. Therefore, despite the logistical and locational advantages embodied by the Port MacKenzie site, the advantage of locating a facility closer to the natural gas source, as embodied by Fairbanks, gives significant headroom for increased capital expenditures to overcome the locational and logistical disadvantage. The source of this advantage is that natural gas is a more expensive product to transport than the denser liquid output products, bearing in mind that the volume of the gaseous feedstock is reduced over 1500 times when converted to liquid products. Based on the siting considerations researched and the economic analysis, both the Port MacKenzie and Fairbanks sites are viable locations for the GTL plant.

This study qualifies as a conceptual or FEL1 study in terms of the Hatch guidelines for the deliverables required for this level of work. In a future phase of engineering, both transportation and constructability studies are required to further detail each potential plant location and to associate definitive costs to overcoming the logistical challenges of Fairbanks. Upon completion of these studies, a plant location can be selected. Process optimization and refinement of the cost estimate accuracy would also be performed in the next phase of engineering to determine whether the economics of this GTL facility would improve as an anchor tenant for the proposed Alaska Stand Alone Gas Pipeline/***ASAP***. GTL technology licensor input, for both process and cost information, would be a requirement in the next stage of engineering.

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1. Project Overview

Alaska Gasline Development Corporation (AGDC) has commissioned Hatch to complete an economic feasibility study to address whether a gas-to-liquids (GTL) facility located in Cook Inlet or in Fairbanks could:

- serve as an "anchor tenant" to increase pipeline demand;
- justify an increase in pipeline capacity; and
- support the economic viability of North Slope gas delivered to Cook Inlet and Fairbanks.

The purpose of this work is to determine the economic feasibility of the plant, inclusive of a market analysis for GTL product options, process simulation, plot plan and siting considerations, project execution schedule, capital and operating cost estimates, and an economic model.

The scope of work commences at the pipeline delivery of natural gas to the plant battery limit and terminates at the delivery of FT liquid products to the loading facilities for transport to market.

2. GTL Technology Survey

A gas-to-liquids (GTL) facility fully integrates a variety of processes, and may be summarized in three (3) main units: synthesis gas production, Fischer Tropsch (FT) synthesis and product upgrading. Below are general descriptions of the technologies available for each main process unit, as well as technology options for CO_2 capture.

2.1 Syngas Production

Synthesis gas (syngas) consists primarily of carbon monoxide (CO) and hydrogen (H_2), and is generated from carbonaceous feedstocks such as coal, biomass and natural gas. The conversion of natural gas to syngas is referred to as methane reforming. Methane is reformed over a catalyst at a high temperature (1,470 – 2,010°F) and pressure (290 – 1,450 psi). There are two main processes through which methane can be reformed: steam methane reforming (SMR) and oxidative reforming, the reaction being driven by heat and oxidation, respectively.

The difference in using natural gas as a feedstock for syngas production compared to coal is the hydrogen content of the two feeds. Feedstocks low in hydrogen (relative to the carbon content), such as coal, produce a syngas that is carbon monoxide rich; whereas, a hydrogen-rich feedstock such as natural gas with the main constituent as methane, produces a syngas that is hydrogen rich.

Downstream of syngas generation, FT synthesis requires a particular H_2:CO ratio in the syngas. CO-rich syngas requires a substantial amount of steam to shift a portion of the CO to H_2 in order to achieve the required ratio. However, syngas derived from natural gas can be produced with the desired H_2:CO ratio by modifying the parameters of the methane reforming process and by recycling carbon dioxide (CO_2) rich FT tail gas, eliminating the need for a shift process.

2.1.1 *Steam Reforming*

Steam reforming is an endothermic process carried out in reactors referred to as steam methane reformers (SMR), whereby an external hot gas provides heat to catalyst-filled tubes in which the catalytic reaction takes place, converting steam and methane into H_2 and CO (syngas) (Figure 2-1). This conversion is described by the following reactions:

$$CH_4 + 2H_2O \leftrightarrow CO_2 + 4H_2$$ **Equation 2-1**

$$CH_4 + H_2O \leftrightarrow CO + 3H_2$$ **Equation 2-2**

$$CO_2 + H_2 \leftrightarrow CO + H_2O$$ **Equation 2-3**

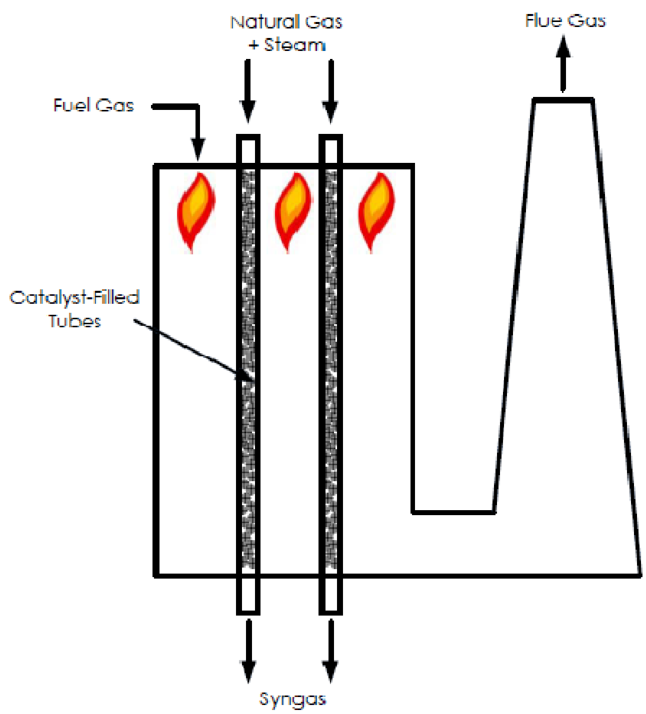


Figure 2-1: Illustration of a Steam Methane Reformer

Steam reforming of hydrocarbons is a mature technology and is the dominating process for manufacture of hydrogen around the world for small and medium size applications. Typical SMR design consists of catalyst-filled tubes suspended in a radiant section of a fired heater, with heat being provided from burners which combust fuel gas. This tubular reactor design converts energy from the hydrocarbon feedstock and fuel gas into hot syngas and flue gas. SMR suppliers offer varying reactor designs to provide sufficient heat to the reactor tubes, but most common include burners mounted on furnace walls (side-fired) or burners mounted on top of the furnace (top-fired).

In order for more efficient reformer operation, the syngas production process typically includes a pre-reforming step, converting higher hydrocarbons (C_2+) in the feedstock into a mixture of methane, steam, carbon oxides and hydrogen as described by the equations. This is typically performed in a fixed-bed adiabatic reactor with a nickel-based catalyst.

Product gas from the pre-reformer is fed directly into the SMR, operating under pressure between 290 – 510 psi. The disadvantage of operating at high pressure is that methane conversion is reduced relative to operating at lower pressures, forcing Equation 2-2 to the left. To counteract this lower conversion rate, the reaction temperature can be increased typically to 1,470 – 1,600°F (some SMR designs suggest operating temperatures above 1,900°F). However, tube materials limit process temperatures. Another way of increasing methane conversion at equilibrium is to raise the steam-to-carbon ratio; however, this increases the SMR heating duty.

Heat is recovered from syngas exiting the SMR in the form of steam, the majority of which is recycled to the SMR as feedstock.

Similar to the pre-reformer, most steam reforming catalysts use nickel as the active component. The reforming catalyst must be designed to have a reasonable lifetime capable of withstanding extreme conditions during start-up and shutdown, and to achieve the desired conversion with low tube wall temperature and pressure drop. Most commercial catalysts have a surplus of catalyst activity, meaning the reforming reactions proceed as fast as the required heat can be supplied through the tube wall.

In steam reforming, the composition of syngas is governed solely by the steam reforming and shift reactions as a function of pressure, temperature and feed composition and catalyst performance. The result is high H_2:CO ratios, hence why SMR is the technology of choice for hydrogen production applications.

2.1.2 *Oxidative Reforming*

Contrast to steam reforming, heat for the oxidative reforming process is provided by internal combustion. Oxidative reforming includes autothermal reforming (ATR) and non-catalytic partial oxidation (POX). ATR has been used to produce H_2 and CO rich syngas for decades. ATR involves a combination of catalytic processes in an adiabatic (without heat loss) reactor. Chemical reactions governing this process are listed below:

$$CH_4 + \frac{3}{2}O_2 \rightarrow CO + 2H_2O$$ **Equation 2-4**

$$CH_4 + H_2O \leftrightarrow CO + 3H_2$$ **Equation 2-5**

$$CO_2 + H_2 \leftrightarrow CO + H_2O$$ **Equation 2-6**

The ATR reactor design consists of a burner, autothermal reforming chamber and fixed bed catalyst section all enclosed in a refractory lined pressure vessel (Figure 2-2). Natural gas from the pre-reformer is mixed with steam and oxygen in the burner where it is combusted in a fuel-rich (substoichiometric) environment. Only a portion of the hydrocarbon feedstock combusts. Following combustion, further conversion to syngas molecules occurs as gases mix at high temperatures in the catalyst bed.

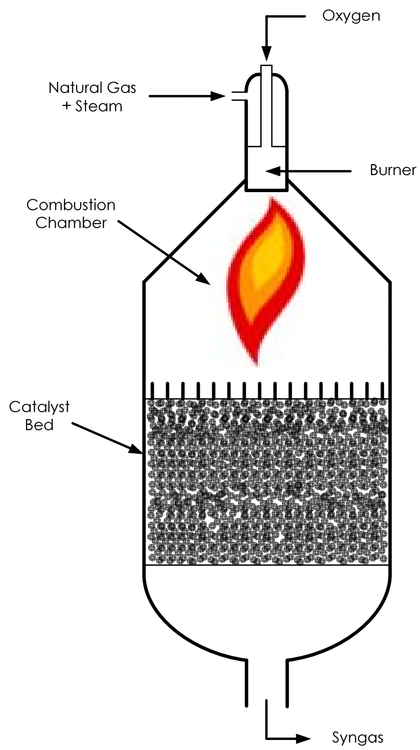


Figure 2-2: Illustration of an Autothermal Reformer

Gas leaving the combustion chamber still contains methane and low concentrations of other hydrocarbons, so final conversion occurs across the fixed catalyst bed. Similar to SMR reactions as described above, these reactions are endothermic, and thus gas temperatures through the bed decrease from approximately 2,200 – 2,400°F down to the exit temperature of about 1,800°F. As mentioned, this catalyst bed operates adiabatically with very little heat loss to the surroundings. The nickel-based catalyst, similar to SMR and pre-reforming, is designed to ensure high activity with low pressure drop in order to maintain efficient, compact reactor design.

Autothermal reforming is exothermic, and steam is generated by recovering heat from the syngas exiting the ATR reactor. This steam is used for mixing with feedstock, as well as distribution and use within the facility.

2.2 Fischer-Tropsch Synthesis

Clean syngas is converted into a mixture of hydrocarbons in Fischer-Tropsch (FT) reactors over metal catalysts. There are two (2) catalyst-based families of FT reactors: iron-based and cobalt-based.

Product selectivity (chain length of hydrocarbon) can be manipulated by the choice of catalyst and operating conditions, and the variety of products typically include distillate (diesel), gasoline, naphtha, Liquefied Petroleum Gas (LPG), oxygenates and olefins.

2.2.1 Fischer-Tropsch Processes

The three main commercially available FT synthesis processes include iron-based Low Temperature Fischer-Tropsch (LTFT), iron-based High Temperature Fischer-Tropsch (HTFT), and cobalt-based Low Temperature Fischer-Tropsch (LTFT).

The iron-based catalyst process requires a H_2:CO ratio between 0.8 and 1.6; whereas cobalt-based is 1.9 to 2.1 (licensor dependent).

2.2.1.1 Iron-Based LTFT

Iron-based catalysts are better suited for synthesis of syngas derived from coal, as it possesses higher concentrations of catalyst poisons. Iron-based catalyst's short lifetime "masks" the effects of syngas poisoning and provides for frequent affordable replacement as it is produced from low cost feedstock (e.g. scrap metal).

Iron-based LTFT processes typically operate at 480°F, with a product selectivity towards distillates (diesel) and naphtha.

Low Temperature Iron-based catalysts process lower H_2:CO syngas ratios – a disadvantage for GTL due to the hydrogen rich syngas produced from methane reforming.

2.2.1.2 Iron-Based HTFT

Iron-based HTFT reactors typically operate at 680°F, roughly 200°F greater than iron-based LTFT reactors. Although iron is the primary catalyst component, the composition and type of promoters differ from LTFT to HTFT. The HTFT process has a higher selectivity towards methane, light hydrocarbons, aromatics and oxygenates relative to iron-based LTFT.

Within the HTFT reaction, a higher degree of shift (increase of H_2:CO) occurs, and with sufficiently high partial pressures of H_2 and CO_2, the reverse shift reaction is possible converting CO_2 into product with sufficient H_2 present. This increases production efficiency. The efficiency gain is driven by higher temperatures, producing higher steam quality for use.

$$H_2 + CO_2 \rightarrow CO + H_2O + \text{heat} \qquad \textbf{Equation 2-7}$$

In terms of H_2:CO ratio, the HTFT process is more flexible than the Iron LTFT process due to the activity of the reverse shift reaction (Equation 2-7).

The HTFT process typically produces gasoline as a final product as opposed to diesel and naphtha.

There is significantly more opportunity to recover high value chemicals such as ethylene, propylene, 1-hexene, 1-octene and oxygenates, although the scope to recover these compounds is limited by market constraints. These high value products increase utility, energy and capital requirements and in turn demand a more complex marketing strategy.

In general, when targeting jet fuel and diesel while limiting capital expenditure, the LTFT process is preferred. HTFT is more suited for locations close to chemical markets, where a market for gasoline and synthetic natural gas drive the selection of HTFT.

2.2.1.3 *Cobalt-Based LTFT*

FT processes using cobalt-based catalysts operate at 440°F. Relative to iron-based LTFT, cobalt-based LTFT has the advantage of a higher resistance against catalyst deactivation due to water oxidation. FT reactions result in the production of H_2O limiting catalyst reactivity. With a higher resistance to deactivation, greater per pass conversion of syngas and efficiency is delivered; additionally, a lesser degree of water is produced as a result of the reduced temperature.

Cobalt-based catalysts have low selectivity towards olefins and oxygenates versus iron-based; therefore, less H_2 is required for upgrading purposes, reducing capital.

Higher conversions per pass can be obtained with Cobalt LTFT rather than Iron LTFT, resulting in higher reactor productivity and lower capital costs in the FT section.

Although cobalt-based catalysts have a higher cost, it is offset by longer life guarantees than iron-based catalysts. However, cobalt-based catalysts require more stringent feed gas constraints against poisoning. FT catalyst poisons include sulfur (S), nitrogen (N) bearing compounds and halides (e.g. florides (F^-), chlorides (Cl^-), bromides (Br^-), etc.). These are more of a concern for coal-to-liquids processes with higher poison concentrations than for gas-to-liquids processes.

Hydrocarbons from the FT process are further upgraded to produce desired products such as diesel, jet fuel, heating oil and naphtha through processes of hydrotreating and wax cracking, with the light-ends from the process being available for use in power generation or as fuel gas. Furthermore, it is easier to obtain higher conversion efficiencies with a higher H_2 content syngas. However, when dealing with stranded natural gas, high conversion to FT products is required, as many co-products are difficult to put to market cost-effectively.

2.2.2 *Fischer-Tropsch Reactors*

It is relatively easy to remove potential FT catalyst poisons from natural gas prior to reforming. Upon reforming, it is possible to produce a syngas with the appropriate H_2:CO ratio for either LTFT or HTFT by recycling the appropriate amount of FT tail gas, or adjusting the steam/carbon ratio to the reforming section. However, it is more expensive to employ an iron-based LTFT process due to the lower per pass conversion, relative to cobalt-based LTFT and HTFT, increasing capital and lowering return.

There are four (4) types of FT reactors in commercial use:

- Circulating fluidized bed reactor (CFB) - HTFT
- Fluidized bed reactor (FB) - HTFT
- Tubular fixed bed reactor (TFB) - LTFT
- Slurry phase reactor (Slurry) - LTFT

Fluidized bed reactors operate between 608°F and 662°F; in this range, there is no liquid phase in the reactor – apart from catalyst particles which is the distinguishing feature between HTFT and LTFT reactors. Any liquid phase in an HTFT reactor results in particle agglomeration and loss of fluidization leading to serious operational problems.

HTFT reactors are utilized when the desired products are shorter chained hydrocarbons (e.g. alkenes). LTFT reactors produce long-chain waxes for upgrading to long-chain hydrocarbons such as gasoline or heavier. Table 2-1 outlines the operating temperatures to maximize specific hydrocarbon chain lengths.

Table 2-1: Production Cut Maximization Temperatures

Cut Maximized by fraction	Minimum Temperature to Avoid Condensation [°F]
$C_2 – C_5$	228
$C_5 – C_{11}$	624
$C_5 – C_{18}$	737
$C_{12} – C_{18}$	874

2.2.2.1 Circulating Fluidized Bed Reactors

A circulating fluidized bed (CFB) reactor synthesizes syngas (CO and H_2) over a catalyst as the syngas-catalyst mixture moves vertically up the FT reactor. Syngas, preheated to 392°F, and catalyst are mixed and flow upwards through the FT reactor. The synthesis of syngas to hydrocarbons is exothermic requiring control of reaction temperature with boiler feedwater (BFW). BFW is raised to saturated steam conditions where it may be utilized throughout the plant to provide heat and – if conditioned – to produce power. Figure 2-3 outlines a CFB reactor.

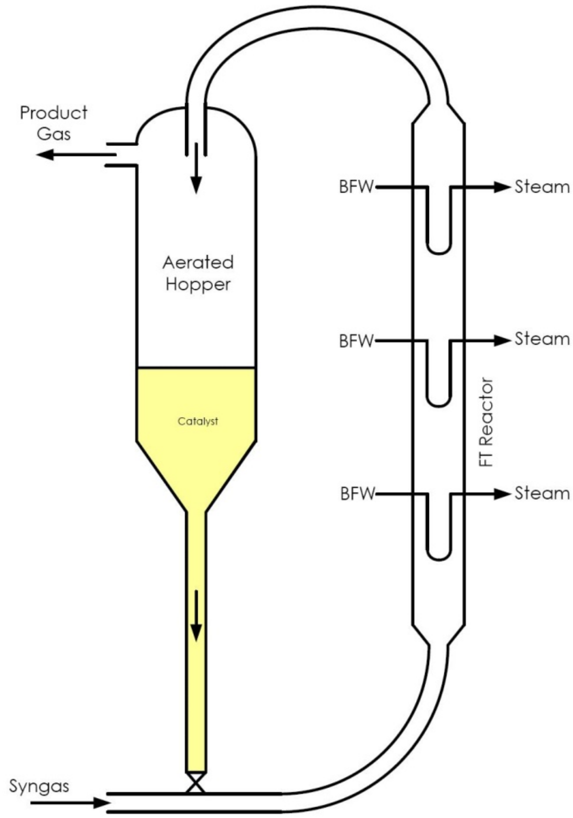


Figure 2-3: Circulating Fluid Bed Reactor

The velocity of the syngas-catalyst mixture up through the reactor dictates the rate of reaction. At temperatures exceeding 640°F, an iron-based catalyst becomes continuously deposited with carbon, causing particle disintegration. Disintegration not only causes catalyst loss, but decreases catalyst density. Reduced catalyst density decreases reaction time per pass due to fixed gas velocity. As a result, on stream catalyst removal and replacement is practiced to maintain consistent production.

CFB reactors were originally installed at Sasol's Sasolburg plant in the 1950s, operating at 640°F and 290 psi. Today, CFB reactors are only operating at Sasol's Secunda CTL and Statoil, PetroSA and Lurgi's joint venture Mossel Bay GTL at 360 psi.

2.2.2.2 Fluidized Bed Reactors

With the success of the CFB reactor, improvement and optimization were sought based on the principle of linear gas velocity up the FT reactor. This lead to the development of the turbulent fixed fluidized bed (FFB) reactors.

Process design analysis indicated that improving quality and uniformity of fluidization offered potential to enhance the reactor performance. A variety of gas distribution nozzles were investigated by Sasol. A design was recommended and a 3′ 3″ diameter FFB reactor was constructed and commissioned at Secunda in 1984 and renamed the Sasol Advanced Synthol (SAS) reactor. Figure 2-4 outlines a fluidized bed reactor.

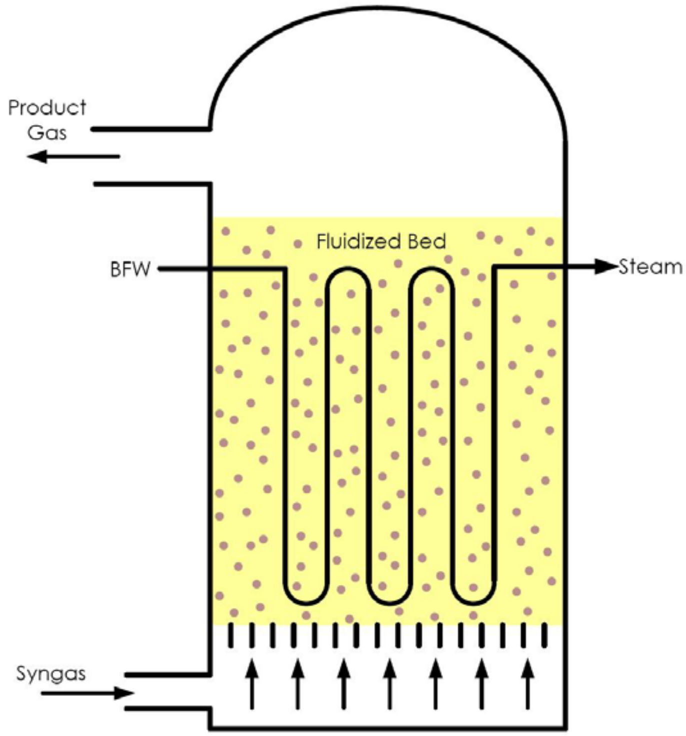


Figure 2-4: Fluidized Bed Reactor

Operating with the same catalyst and product selectivity as CFBs, but with a greater conversion, the design was scaled to a 16′ 4″ diameter unit in 1989 at Sasolburg. In 1995 to 1999, the sixteen (16) existing CFB reactors at Secunda were replaced with eight (8) 26′ 3″ diameter 11,000 bbl/day and eight (8) 32′ 10″ diameter 20,000 bbl/day SAS reactors.

SAS reactors have several advantages over its CFB predecessor, including:

- A 40% lower construction cost, due to smaller size per output, including simpler support structure at 5% of the cost required for a CFB;
- Increased capacity, a result of greater cooling and throughput due to a greater cross-sectional area, permitting increased flow and pressure while maintaining linear gas velocities;
- Complete catalyst charge participation in FT reaction, resulting in higher conversion;
- Allowance for bed expansion due to carbon deposition on the catalyst, resulting in reduced online catalyst replacement and lower catalyst consumption;
- Lower syngas and catalyst linear velocities and pressure drop, resulting in lower compression costs; and

- Larger cross-sections, reducing wear and extending operating time between maintenance shutdowns.

2.2.2.3 *Tubular Fixed Bed Reactors*

Although there is a variety of fixed bed reactors (e.g. vertical spaced, radial flow, multi-tubular), due to the temperature rise effects within individual adiabatic beds, the preferred bed design is the multi-tubular fixed bed (TFB). The catalyst is placed inside the tubes and the cooling medium (boiler feed water) on the shell side. Figure 2-5 outlines a generic TFB reactor.

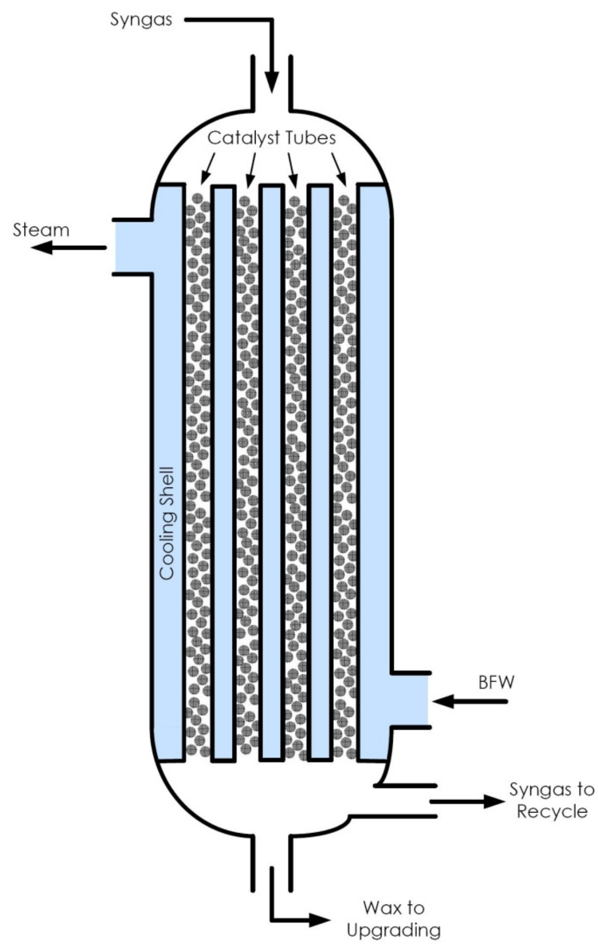


Figure 2-5: Tubular Fixed Bed Reactor

Narrow tube diameters with high linear gas velocities ensure turbulent flow, greatly increasing heat transfer from catalyst particles to the cooling medium. High conversion efficiency is obtained in part by this heat transfer in addition to recycling a portion of the tail gas. The recycle increases linear gas velocities, further improving the rate of heat transfer and efficiency.

Smaller catalyst particles also result in increased conversion efficiency, however, with narrow tube diameters, high gas velocities and small catalyst particles result in unacceptably high differential pressures increasing syngas compression costs and degradation of catalyst particles. Greater catalyst disintegration results in additional cost and downtime necessary to unload and recharge the reactor. TFB reactors must, therefore, balance design and operating requirements in opposition.

Design versus operating constraints ultimately create temperature gradients within the reactor. As a result, there is only a portion of the reactor bed that operates at the optimum temperature, reducing efficiency. TFB reactors are generally not suited for high temperature reactions, as carbon deposition occurs leading to catalyst swelling and tube blockage.

With thousands of tubes per reactor, construction costs are high. TFB reactors are heavy, limiting reactor size and transportation potential.

Advantages of TFB reactors include the elimination of external equipment to separate heavy wax from catalyst. Liquid wax migrates downward through the tubes and collected in a downstream knock-out pot; whereas slurry bed reactors require additional equipment to completely separate fine catalyst particles from the wax external to the reactor.

The single most advantageous aspect of TFB reactors is ease of scale-up. Proof of a single reactor tube is sufficient to predictably scale-up a full size commercial reactor, as each tube operates independently.

As catalyst poisons migrate through the process overtime, TFB reactors have the advantage that the leading face of the catalyst tubes are poisoned and deactivated, serving as a protective barrier allowing the remainder of each tube to continue to carry on.

2.2.2.4 Slurry Phase Reactors

Comparison between fixed and slurry phase FT synthesis was carried out at temperatures common in TFB reactors. See Figure 2-6 for the outline of a slurry phase reactor. The slurry phase reactor performs to a greater hard wax selectivity and conversion efficiency, in spite of a catalyst loading three (3) times less than the fixed bed. The higher conversion is due to reduced catalyst particle size, such that the FT reaction is limited by pore diffusion.

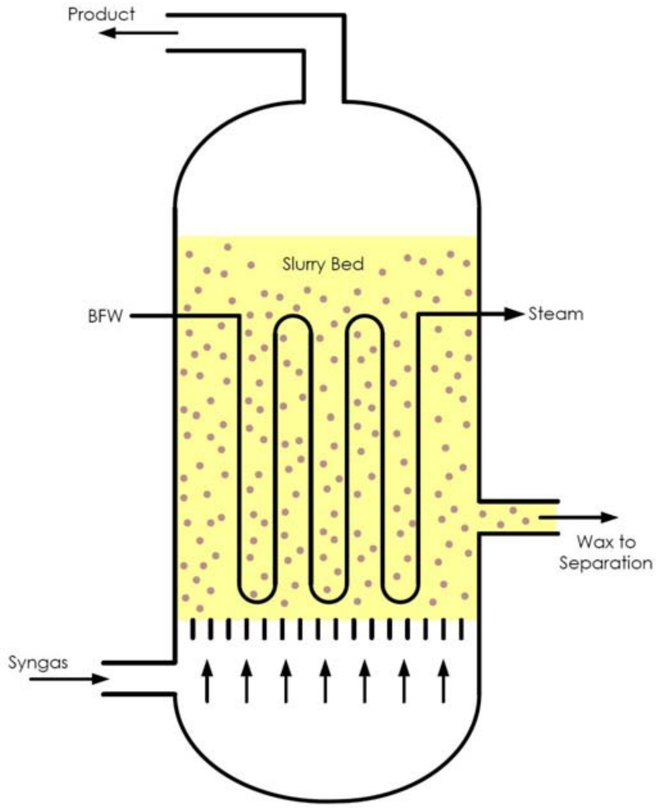


Figure 2-6: Slurry Phase Reactor

Thereafter, comparison between slurry phase and fluidized bed reactors showed that the slurry phase was not well suited for HTFT reactions (615°F). The slurry phase reactor no longer possesses the smaller catalyst particle advantage. At such temperatures, wax is continuously hydrocracked, requiring daily make-up.

Advantages of slurry phase over TFB reactors (for iron-based catalyst only – cobalt comparison data unavailable at time of preparation) include:

- Capital cost of train 75% less than comparative TFB reactor;
- Pressure drop across slurry phase reactor 75% less than comparative TFB reactor, lowers syngas compression costs;
- Higher catalyst activity delivers 75% lower catalyst consumption per barrel of product;
- Isothermal reactor operation versus adiabatic TFB facilitates higher average operating temperature; and
- Online charging and removal of catalyst reduces downtime.

However, it should be noted that many of the slurry phase reactor's advantages are reduced if a cobalt-based catalyst with high activity and long life is utilized.

2.2.3 *Fischer-Tropsch Technology Licensors*

Due to the high cost and complexity required, only a handful of licensors have invested the time and capital necessary to demonstrate FT technologies. The level of development and success within this handful of providers varies widely from multi-billion dollar operating commercial plants to decommissioned pilot plants. An overview of the licensors and their technologies is presented below. It must be noted that as FT technologies are proprietary. The information presented herein is from the public domain and is to serve as a guide and should not be considered complete without licensor engagement.

2.2.3.1 Sasol

South African Synthetic Oil Ltd. (Sasol) is an international energy company headquartered in Johannesburg, South Africa, engaged in the commercial production of chemicals and liquid fuels. Beginning in the 1950s with a licensed FT technology, Sasol has pioneered, developed, commercialized and improved proprietary FT synthesis technologies. These developments include the HTFT Sasol Advanced Synthesis (SAS) process and the LTFT Sasol Slurry Phase Distillate (SPD) process.

The SAS process involves a fluidized bed reactor, and was developed from the 1980s through the 1990s. Today, SAS reactors are capable of 20,000 bbl/day with four (4) currently operating at Secunda in South Africa. SAS reactors employ iron-based catalyst producing light products (e.g. naphtha, paraffins, olefins and aromatics).

The SPD process is a slurry bed reactor; two have been installed at the Oryx facility in Ras Laffan, Qatar, producing 34,000 bbl/day of distillate and naphtha. Discussions to expand the facility to 100,000 bbl/day have been undertaken, but none planned at present.

Currently under construction is the Escravos GTL plant in Nigeria. Located 60 miles southeast of Lagos. Escravos is designed to produce 34,000 bbl/day by 2013. Diesel and naphtha are produced from two (2) 17,000 bbl/day SPD reactors. Escravos is planned to be expanded to 120,000 bbl/day within ten years of completion.

2.2.3.2 Shell

Royal Dutch Shell PLC (Shell), is an international oil and gas company, headquartered in The Hague, the Netherlands. They have been developing FT technology since the early 1980s and since commercially deployed an LTFT technology known as the Shell Middle Distillate Synthesis (SMDS) process – based on an older Sasol tubular reactor.

SMDS' tubular design employs a cobalt-based catalyst. Shell claims the catalyst and reactor design promote higher C_{5+} selectivity, lower CO_2 production and less catalyst consumption than a slurry phase reactor. A demonstration plant in Bintulu, Malaysia was originally commissioned with a 14,700 bbl/day capacity in 1993.

In 2004, Shell began development of a 140,000 bbl/day GTL products plant in Ras Laffan, Qatar. The project, known as Pearl, implements Shell's SMDS technology to produce upstream products (i.e. blendstock) 80 km north of Doha. The complex consists of two (2) 70,000 bbl/day GTL trains, with a total of twenty-four (24) 1,323 short ton reactors and associated facilities. Production ramp-up is expected in 2011, one year behind schedule. Products include: naphtha, distillates, paraffins, kerosene and lubricant-based oils; the range of products is due to the proximity to major downstream off-takers with an appetite for a wide product mix.

2.2.3.3 GTL.F1

GTL.F1 is a joint venture, between Statoil of Norway, PetroSA of South Africa and Lurgi of Germany (subsidiary of Air Liquide), established to commercialize the cobalt-based LTFT technology under development by Statoil over the last twenty years.

The JV developed a cobalt-based LTFT slurry bed reactor and operated a 1,000 bbl/day semi-commercial plant at Mossel Bay in South Africa.

Statoil and PetroSA have indicated intentions to license the technology for the conversion of syngas derived from non-natural gas feedstocks. As Statoil and PetroSA are large natural gas providers, the technology will only be licensed to GTL projects in which Statoil and/or PetroSA take(s) an equity position.

2.2.3.4 BP & Davy Process Technology

BP is an international oil company headquartered in London, UK. Davy Process Technology (DPT) is a licensor of advanced process technologies related to the manufacture of oil and gas, petrochemicals, commodity chemicals, fine chemicals and pharmaceuticals. DPT is a subsidiary of Johnson Matthey PLC and headquartered in London, UK.

Davy developed a LTFT TFB cobalt-based catalyst reactor. A 300 bbl/day demonstration facility located in Nikiski, Alaska was operated from 2003 to 2009. BP announced in 2009 that the project had successfully demonstrated that the process could be scaled up from laboratory to pilot scale to produce diesel and jet fuel.

BP stated its GTL development program will continue in Europe where the company is working with Davy Process Technology on the engineering design of a full scale unit.

2.2.3.5 Syntroleum

Syntroleum Corporation is a synthetic fuels technology company headquartered in Tulsa, Oklahoma. They have developed proprietary technologies for LTFT slurry bed and TFB reactors, both using cobalt-based catalysts.

A 70 bbl/day slurry bed GTL demonstration plant was commissioned in 2004 in Catoosa, Oklahoma. The facility was the company's first green-field demonstration involving syngas production, FT synthesis and upgrading. The facility was shutdown in 2006. In 2007, Syntroleum signed an MOU with Sinopec, selling technology and support services including the demonstration plant which in 2009 was dismantled and shipped to China for re-erection. Sinopec has exclusive rights to the technology within China and will be offering licensing for CTL operations.

2.2.3.6 Rentech

Rentech Inc. is an FT process developer headquartered in Los Angeles, California. Their FT technology utilizes an iron-based catalyst, LTFT, slurry phase bed reactor. Incoming syngas requires a H_2:CO ratio of 0.8–1.0. The iron-based catalyst leads to a mixture of paraffinic, naphthenic, olefinic and aromatic products; although more valuable on the market, these products require more upgrading.

The recently commissioned Rentech demonstration plant is located near Denver, Colorado with a capacity of 10 bbl/day.

Rentech offers worldwide licensing of its proprietary technology for both biomass and fossil fuel feedstock applications. A typical licensing agreement requires licensees to provide up-front payment for services and catalyst in addition to ongoing royalty payments per barrel of product produced. Licensees are responsible for abiding by Rentech's technical recommendations, financing, construction and operations.

2.2.3.7 Other Licensors

There are other FT licensors in addition to those above, listed in Table 2-2; however, they are currently either not marketing their technology or are still developing the technology at the pilot scale.

Table 2-2: FT Licensors in Research and Development

Licensor	Process	Catalyst	Reactor Type	Scale	Location
ConocoPhillips	LTFT	Cobalt	Slurry Phase	400 bbl/day	Ponca City, Oklahoma
ExxonMobil	LTFT	Cobalt	Slurry Phase	200 bbl/day	Baton Rouge, Louisiana
Eni–IFP/Axens	LTFT	Cobalt	Slurry Phase	20 bbl/day	Sannazzaro, Italy
Japan National Oil Corporation	-	-	Slurry Phase	7 bbl/day	Yufutsu, Japan
CompactGTL	LTFT	Cobalt	Mini-channel Fixed Bed	1 bbl/day	Wilton, UK
Velocys	LTFT	Cobalt	Microchannel Fixed Bed	26 gal/day	Plain City, Ohio
World GTL	-	-	Tubular Fixed Bed	2,250 bbl/day	Pointe-Pierre, Trinidad & Tobago
Emerging Fuels Technology	LTFT	Cobalt	-	$<$ 1 lb/day	Tulsa, Oklahoma

2.2.3.8 Licensing

Unless stated above, licensing terms for individual FT technology licensors is not openly disclosed. In recent history, it should be noted, that successful GTL projects have involved equity positions from FT licensors, for example:

- Oryx JV – Qatar Petroleum (51%)
  Sasol (49%)
- Escravos JV – Chevron Nigeria (75%)
  Nigerian National Petroleum Company (25%)
- Pearl JV – Qatar Petroleum (undisclosed)
  Shell (undisclosed)

The ownership requirement is likely due to licensors not wishing to see potential projects strictly in the hands of developers and for them to have more IP control and protection. A technology licensor's stake also provides them access to previously unobtainable reserves.

2.3 Product Upgrading

FT reactors, dependent on their process (LTFT vs. HTFT) and catalyst type (iron-based vs. cobalt-based) produce a variety of intermediate products (e.g. waxes, tail gases, etc.) requiring further processing and upgrading. The following is a description of the upgrading from LTFT, such that FT intermediates are classified by density: 1) light-ends (C_2 – C_5), 2) middle distillates (C_5 – C_{20}) and 3) heavy waxes (C_{20+}). These three densities of intermediates require separation prior to respective upgrading and/or consumption.

2.3.1 Intermediate Upgrading

Hydrocarbon condensate recovered from the FT phase separators contain absorbed gases including H_2, CO and CO_2 requiring removal separation – as they are poisonous to the upgrading catalysts. In addition, these components must be removed as the products de-gas if stored in intermediate storage, posing a health risk to operating personnel. Removal of absorbed gases is achieved by stripping light-ends in a feed stripper column. HP saturated steam is used to vaporize a portion of the contents at the bottom of the column in a reboiler. The resulting overhead vapour, containing dissolved gases and hydrocarbons in the range C_2 to C_5, is routed to fuel gas distribution. Liquid product is sent to intermediate storage or fed to the next stage of upgrading.

Liquid product from the feed stripper contains hydrocarbons in the desired C_5 to C_{20} range. Hydrocracking of these components is not required as they are of the desired chain length – although oxygenates and olefins must be hydrogenated.

The remaining wax from the FT reactors contains hydrocarbons in the C_{20+} range – requiring hydrocracking. This process scheme is illustrated in Figure 2-7.

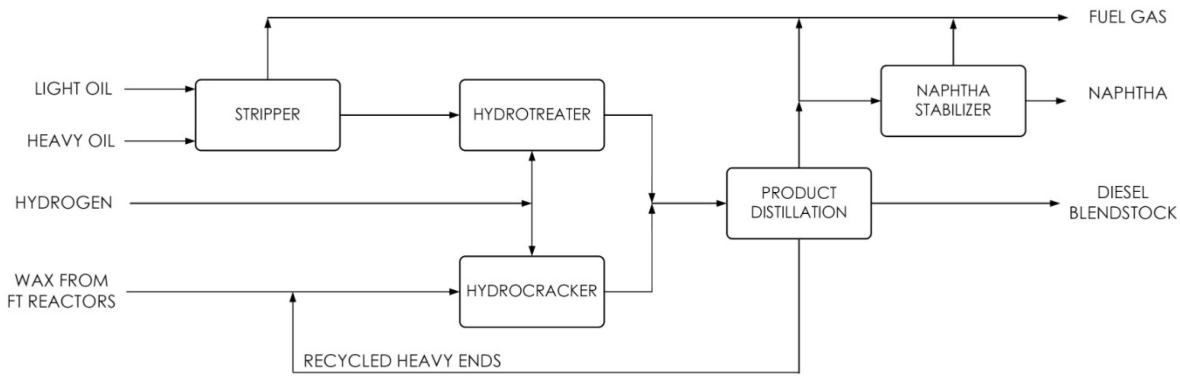


Figure 2-7: Basic LTFT Finished Product Upgrading Flow Scheme

Hydrocracking involves feeding H_2 to the hydrotreater and hydrocracker in stoichiometric excess to maintain high H_2 partial pressure improving hydrogenation and preventing coking of the catalyst. For this reason, there is substantial H_2 circulating in the reactors.

Effluent leaving the reactors is separated via flashing; the liquid stream is sent to fractionation, and the H_2-rich gas stream is combined with fresh H_2 and recycled to the hydrotreating and hydrocracking reactors. The combined liquid product from hydroprocessing is sent to a product distillation column to be fractionated into light-ends, heavy-ends and diesel product. The light-ends are sent to a naphtha stabilizer column where naphtha is recovered and light components stripped out and sent to fuel gas distribution. Heavy-ends are recycled back to the hydrocracking reactor.

2.3.2 Finished Product Quality

With a cobalt catalyst, the FT product is a predominantly paraffinic product with chain lengths ranging from C_5 to as high as C_{80}. These products are unlike those from conventional oil refineries in that olefinic and naphthenic contents are very low and contain no aromatics. Without aromatics, products contain very low octane ratings and are not suited for the manufacture of gasoline.

Cobalt derived FT fuels are considered highly desirable blendstocks for crude derived fuels due to their unique properties of high cetane number and negligible sulfur and aromatic content. Compared to conventional crude derived fuel during combustion, FT fuels show reduced NO_x, CO and particulate emissions. These properties and are extremely well suited for the manufacturing of ultra low sulfur (ULS) jet fuel, ULS diesel and ULS heating oil.

2.4 CO_2 Capture System

A CO_2 capture system can also be included with the overall GTL process design. This CO_2 capture process removes CO_2 from the purge gas stream of the large FT synthesis recycle loop. The resulting lean CO_2 purge gas stream can be used as a fuel gas for power generation and/or process heating. In order to decrease capital and operating costs, it is most effective to capture CO_2 prior to combustion, as the purge gas stream has high CO_2 concentration and partial pressure which results in a lower volume of gas to process.

Different CO_2 capture processes are categorized into three main categories:

- Physical absorption;
- Chemical absorption; and
- Processes utilizing a combination of physical and chemical absorption.

There are many established and developing CO_2 capture processes available in the market; a brief description of some commercially available processes are described herein.

2.4.1 Chemical Solvents

Chemical solvents absorb acid gas (primarily CO_2 and H_2S) by creating a chemical bond between solvent and acid gas molecules. Energy, typically in the form of steam, required to reverse the reaction and regenerate the solvent, producing a pure acid gas stream.

2.4.1.1 Amine Systems

Different amines, or combination thereof, have been used for acid gas removal for many years, and many are commercially available from chemical suppliers. Most applicable for CO_2 capture include: monoethanolamine (MEA), diethanolamine (DEA), methyldiethanolamine (MDEA), as well as inhibited amines or formulated solvents such as aMDEA offered by BASF. Each solvent has specific advantages and disadvantages, but the principle of operation remains similar. The processes typically operate at temperatures of 40-120°F, and are most competitive (compared to physical solvent counterparts) at low pressures. Some of the main criteria in choosing the most suitable amine system for acid gas removal is regeneration energy and the CO_2 removal efficiency at given pressures, temperatures and compositions.

2.4.2 Physical Solvents

In physical absorption systems a physical bond forms between acid gas components and solvent molecules. Physical absorption solvents are regenerated by decreasing the pressure and increasing temperature. The loading capacity (amount of CO_2 captured in the unit mass or volume of the solvent) of physical solvents is governed by Henry's law, stating that absorption is proportional to gas partial pressure (CO_2 in this case). By increasing operating pressure, performance of physical absorption systems is improved.

2.4.2.1 Rectisol®

Rectisol® utilizes cold methanol as a solvent and was originally designed to purify syngas for chemical synthesis applications that require strict impurity limitations. The process operates between -22°F to -77°F to increase absorption rates and achieve greater gas purities than other processes. To reach such temperatures, refrigeration is required. This technology is licensed by Lurgi GmbH and The Linde Group, both headquartered in Germany.

2.4.2.2 Selexol

Selexol utilizes a proprietary solvent, consisting of dimethyl ethers of polyethylene glycol (DMPEG), that is chemically inert and not subject to degradation. The process typically operates between 32°F and 104°F, reducing refrigeration duty required by Rectisol®. Purity levels cannot match those provided by Rectisol®, however, Selexol has a number of reference plants for power applications and other plants that do not critically require tight impurity specifications. The process is licensed by UOP LLC, headquartered in the USA.

2.4.3 Processes Utilizing a Combination of Physical and Chemical Absorption

2.4.3.1 Benfield

The Benfield process uses a potassium carbonate solution to absorb CO_2 at temperatures between 160°F and 260°F. The higher temperature is maintained at the bottom of the absorption column to increase the rate of absorption, while a lower temperature is utilized at the top section of the absorber to maintain a more favorable equilibrium at the exit of the absorber. The absorption section typically operates at pressures of around 435 psi (30 bar). Rich solvent is flashed at close to atmospheric pressure in the regeneration column, while heat is added via steam injection to break chemical bonds.

An amine promoter (i.e. DEA, LRS-10 or ACT-1) is used to enhance the performance of the potassium carbonate solution. The Benfield process uses highly toxic Vanadium as a corrosion inhibitor, which must be carefully managed during solution discharges from the plant.

2.4.3.2 Sulfinol

Sulfinol is customized to each individual application. A mixture consisting of water, a chemical-reacting alkanolamine and the physical solvent sulfolane to the appropriate proportions is more effective than aqueous amine processes at removing COS, mercaptans and other organic sulfur compounds. Applications also include complete or partial removal of CO_2 from natural, synthetic and refinery gases. Loaded solvent is regenerated while impurities are flashed post-absorption and consumed as a fuel gas. Over 200 units have been licensed worldwide by Shell Global Solution International B.V., headquartered in The Hague, The Netherlands.

3. GTL Product Market Analysis

As with the variety of technologies that may be utilized in the GTL facility process design, there is also a variety of FT liquid fuels that may be produced. The GTL facility may be configured to produce a combination of liquid fuels including diesel, jet fuel, naphtha, liquefied petroleum gas (LPG, e.g. propane and butane), oxygenates and olefins.

A market analysis for various GTL product options (FT products) was performed in order to prioritize and determine the products to be included in the process design.

3.1 Methodology and Assumptions

The potential markets for FT products were analyzed focusing primarily on intrastate Alaska, U.S. West Coast, Hawaii and Asia (China and Japan). Methodology includes:

- analysis of historical consumption patterns of relevant fuel in order to derive a projected trend;
- analysis of historical relationship between price of a specific fuel on a particular market with price of Western Texas Intermediate (WTI) at Cushing, OK; additionally, historical relationship between specific fuels and Alaska North Slope (ANS) crude oil derived for products sold in Alaska;
- linear regression from analysis to quantify price relationship, confidence levels and analysis of statistical significance of price relationship; additionally, price models utilized in the economic model to forecast revenue generated by GTL plant; and
- derivation of preliminary conclusions in terms of attractiveness of particular markets and of recommended prioritization of certain products given market prices.

This market analysis contains various assumptions. Although FT jet fuel and diesel are considered superior to crude derived, they cannot be directly consumed without the use of certain additives and/or blending to improve lubricity, increase density and inhibit corrosion. For this reason, FT products are assumed to be sold in the wholesale or resale market.

Secondly, as previously mentioned, FT products exhibit advantageous characteristics; in particular, FT diesel is extremely low in sulfur and aromatics with a high cetane rating. It also exhibits lower particulate emissions during combustion relative to crude derived diesel. FT naphtha is an ideal feedstock for steam crackers and literature suggests it may help increase yield [1]. FT jet fuel exhibits a desired cetane rating, no sulfur, no aromatics and excellent smoke and flash points. These differences may suggest a potential price premium. However, factual evidence for such a premium is currently anecdotal and cannot be verified. For this reason, no price premiums in the wholesale price for FT products is assumed.

In order to quantify markets and model price relationships it is assumed that FT jet fuel's best proxy product corresponds to kerosene-type jet fuel and FT diesel's to No. 2 Diesel Fuel. The latter is a simplified assumption as a better comparison would be against No. 2 Diesel, Ultra Low-Sulfur. However, historical data for this product is not sufficiently available to conduct a meaningful analysis. Furthermore, the consumption of No. 2 Diesel Fuel is assumed equal to distillate fuel oil for transportation consumption.

For the purpose of this analysis, it is assumed that the U.S. West Coast market is defined as California, Washington State, Oregon, Nevada and Arizona. This definition is equivalent to the Petroleum Administration for Defense District (PADD) V excluding Hawaii and Alaska, which are analyzed separately. In terms of Asian markets, countries selected for analysis are chosen in part based on market size. The Chinese market is larger than the Japanese for diesel, the former representing 36% of the Asia and Oceania's market for diesel (Japanese market represents 14% of said market). Jet fuel in both countries is approximately equal. The predominant market for naphtha is Japan. Therefore, China is focused upon for FT diesel and jet fuel, and Japan for FT naphtha.

Current hydrocarbon shipping costs including time in ports, fees and additional costs due to north of 54ºN are also used in Table 3-1. Quoted rates are based on 330,000 bbl ship size ("handymax") departing from either Anchorage or Nikiski, Alaska. Quotes were also benchmarked to Hatch's procurement group data.

Table 3-1: Shipping Costs from Alaska

Information	Unit	Anchorage AK	Los Angeles CA	Honolulu HI	Tokyo Bay JPN	Shanghai CHN	Houston TX
Ship Capacity	Bbl	330,000	330,000	330,000	330,000	330,000	330,000
One-Way Trip from AK	nautical miles	-	2,181	2,480	3,278	4,118	6,639
	day(s)	-	10	10	14	18	28
Time in Ports	day(s)	2	3	3	3	3	3
Time Cost	USD/day	-	65,000	65,000	-	-	65,000
Trip Cost	USD/trip	-	-	-	735,000	785,000	-
Port Fees	USD/port	66,000	40,000	50,000	45,000	45,000	32,000
Canal Fees	USD/trip	-	-	-	-	-	120,000
Canal Time	day(s)	-	-	-	-	-	1
Insurance	USD	22,000	-	-	-	-	-
Additional Costs	%	5%	5%	5%	5%	5%	5%
Total	**USD**	**-**	**1,465,050**	**1,475,550**	**851,550**	**901,550**	**3,207,150**
Total	**USD/bbl**		**4.44**	**4.47**	**2.58**	**2.73**	**9.72**
Total	**US_{cents}/US_{gal}**	**-**	**10.57**	**10.65**	**6.14**	**6.50**	**23.14**

Source: OSG (Overseas Shipholding Group) quotes provided November 30, 2010.

For a Fairbanks plant location, additional logistic means are required to transport the FT products to the Anchorage or Nikiski ports. A potential method for transporting FT products to the coast would include using the existing TAPS pipeline to send products down to the Valdez Oil terminal. A number of companies in Alaska have performed studies regarding the feasibility of transporting synthetic fuels through the TAPS pipeline, and the general consensus is that fuel contamination is a major issue that would cause this method to be uneconomic. For this reason, rail transport is the preferred transport method at this time, but advances in technology may make the pipeline option more favorable. For this study, rail transport of products from Fairbanks to Anchorage has been assumed.

The Flint Hills refinery in North Pole, Alaska currently transports approximately 15 million barrels of refined products annually to a storage facility in Anchorage by rail. Approximately 80 cars are shipped daily along the 425 mile railroad from North Pole to Anchorage. The Alaska Railroad Corporation (AKRR) provides a tariff rate for petroleum products, but industry experts have stated this rate is negotiable for large capacity, long term supply contracts. Based on current fuel transport contracts and recent rail extension studies, it has been recommended to use a cost of 0.045 USD/ton/mile for FT products being shipped from Fairbanks to Anchorage. This cost includes the time for drivers/operators during transport and loading/unloading activities, but requires the addition of a fuel surcharge, equal to 19.5% as of February 2011.

The transport cost also does not include provision of rail cars, which the Flint Hills refinery currently leases from GATX in Houston, Texas. The AKRR typically ships fuel in 30,000 gallon capacity rail cars, which are filled to 28,500 gallons. From GATX, the cost for leasing these cars is approximately 900 USD/car/month, in addition to the initial investment of 12,000 USD/car to have the rail cars shipped to Alaska from continental U.S. The lease costs cover regularly scheduled maintenance as well as normal wear and tear, and the life of these cars is typically 35 years before requiring replacement. Representatives from both AKRR and GATX recommended that the required operational cars (based on the Case B GTL fuel production of 33,260 bbl/day) be increase threefold (3x) to account for cars down for maintenance and for the shifting and realignment of cars. Table 3-2 presents the assumptions made for this calculation for both fuel transport and rail car leasing costs. The resulting transport cost for shipping FT products from Fairbanks to Anchorage by rail is 3.20 USD/bbl for FT Diesel, 2.83 USD/bbl for FT Naphtha and 3.11 USD/bbl for FT Jet Fuel.

Table 3-2: Transport Cost from Fairbanks to Anchorage

Information	Unit	FT Diesel [1]	FT Naphtha [1]	FT Jet Fuel [2]
Fuel production	bbl/day	24,680	8,580	12,555
	gal/day	1,036,560	360,360	527,310
Density	bbl/metric tonne	8.25	9.37	8.51
	bbl/short ton	7.49	8.50	7.72
Transport Cost	USD/short ton/mile	0.045	0.045	0.045
Transport Distance	miles	425	425	425
Fuel Surcharge		19.5%	19.5%	19.5%
Annual Fuel Transport Cost [3]	*USD/a*	*25,114,000*	*7,687,000*	*12,386,000*
Rail Car Capacity	gal	28,500	28,500	28,500
No. of Operational Cars (daily)		37	13	19
Total No. of Cars		111	39	57
Lease Cost	USD/car/month	900	900	900
Annual Rail Car Leasing [3]	*USD/a*	*1,199,000*	*421,000*	*616,000*
Total Annual Transport Cost	**USD/a**	**26,313,000**	**8,108,000**	**13,002,000**
Total Transport Cost	**USD/bbl**	**3.20**	**2.83**	**3.11**
	US_{cents}/US_{gal}	**7.62**	**6.74**	**7.40**

Note: 1) Fuel production for Case B GTL facility
2) Fuel production for Case B Alternative Mix GTL facility, as described in section 12.1.1
3) Annual transport costs assume 8,000 operating hours per year

With three (3) different products (jet fuel, diesel and naphtha) and four (4) markets (Alaska, U.S. West Coast, Hawaii and Asia) a significant amount of research was conducted to determine the target markets for each product as indicated in Table 3-3, based on price, consumption and transportation costs. The subsequent charts and equations are presented in USD/bbl. Appendix F contains charts on a U.S. Cents per U.S. Gallon basis.

Table 3-3: Summary of Products and Markets Analyzed

Market	Jet Fuel	Diesel	Naphtha(a)
Alaska	☑	☑	☑
U.S. West Coast	☑	☑	☑
Hawaii	☑	☑	☑
Asia(b)	☑	☑	☑

Notes: (a) naphtha market for the United States as a whole was examined; (b) jet fuel and diesel markets for China only; and naphtha market for China, Japan and Korea were analyzed.

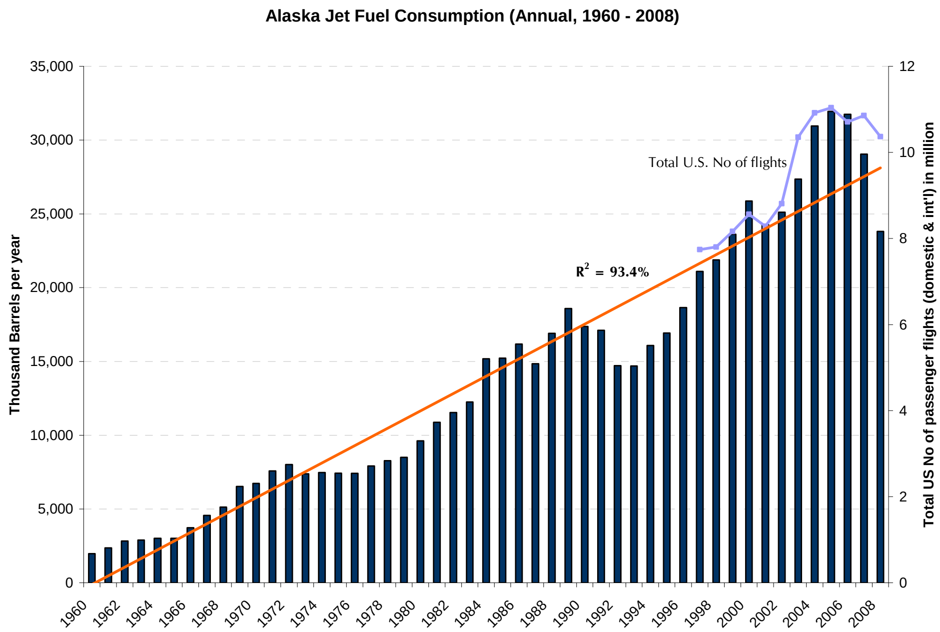
3.2 FT Jet Fuel Analysis

The Alaska, Hawaii, U.S. West coast, and China markets for FT jet fuel were analyzed. It is important to point out that, subsequent to successful testing, a 50:50 blend was approved for use in civil aviation in September 2009 [2]. This effectively reduces markets by half when analyzing consumption patterns.

3.2.1 Alaska

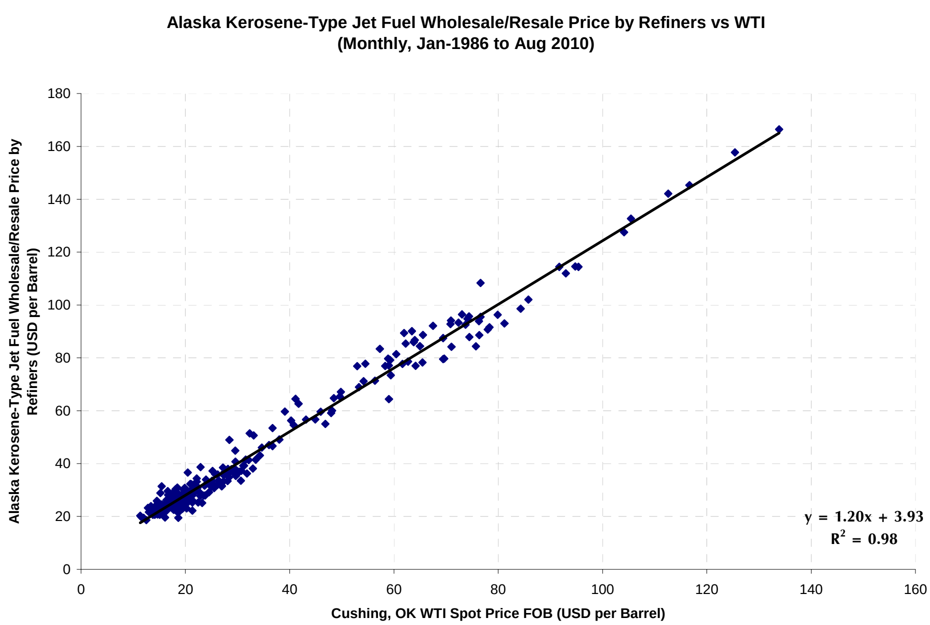
The Alaskan market for kerosene-type jet fuel represented 23,817,000 bbl/a in 2008. Its evolution is shown in Figure 3-1 and exhibits an increasing trend and a cyclicality mostly associated with the overall airline industry (trend lines are shown on all charts). Indeed, Figure 3-1 depicts total number of U.S. passenger flights (as a proxy for the overall industry activity), correlated to total consumption. Research indicates Alaska's refineries supply approximately 88% of in-state jet fuel consumed [3]. From historical price analysis, the jet fuel wholesale price in Alaska as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-2:

$$\text{Jet Fuel}_{AK}\,[\text{USD/bbl}] = 1.20 \times \text{WTI} + 3.93\,[\text{USD/bbl}]$$



Source: EIA, State Energy Data System (SEDS); Bureau of Transportation Statistics

Figure 3-1: Alaska Jet Fuel Consumption

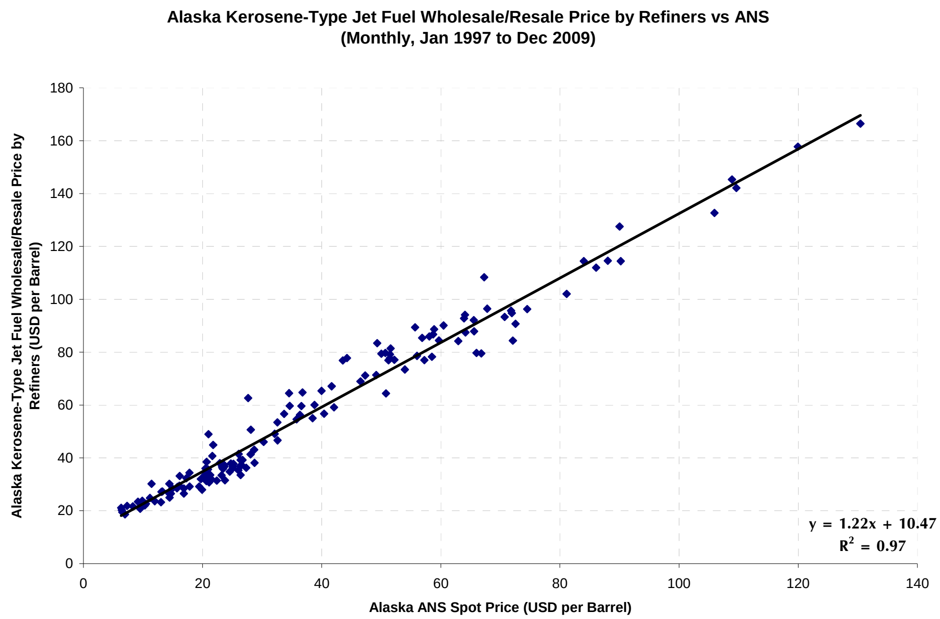


Source: EIA, Petroleum Product Prices

Figure 3-2: Alaska Jet Fuel vs. WTI

Additionally, historical price relationship between jet fuel wholesale price in Alaska and Alaska North Slope (ANS) crude oil was analyzed and is given by the following equation and depicted in Figure 3-3:

$$\text{Jet Fuel}_{AK}\ [\text{USD/bbl}] = 1.22 \times \text{ANS} + 10.47\ [\text{USD/bbl}]$$



Source: EIA, Petroleum Product Prices; Alaska Department of Revenue

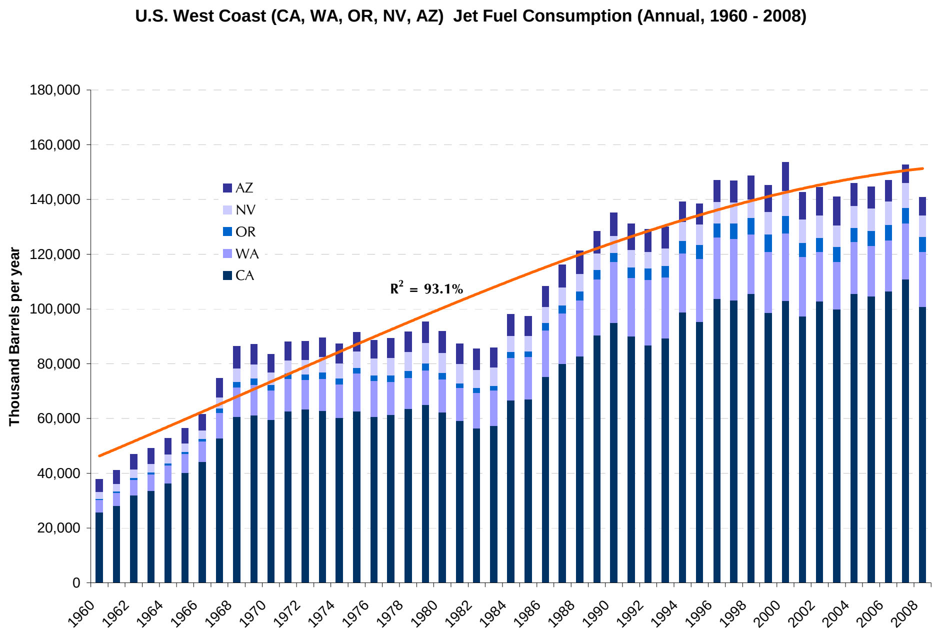
Figure 3-3: Alaska Jet Fuel vs. ANS

3.2.2 U.S. West Coast

The U.S. West Coast market for kerosene-type jet fuel represented 140,890,000 bbl/a in 2008; evolution over time is shown in Figure 3-4 exhibiting an increasing trend at a decreasing rate, with flatter consumption over recent 18 years. From the analysis of historical prices, the jet fuel wholesale price in the U.S. West Coast as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-5:

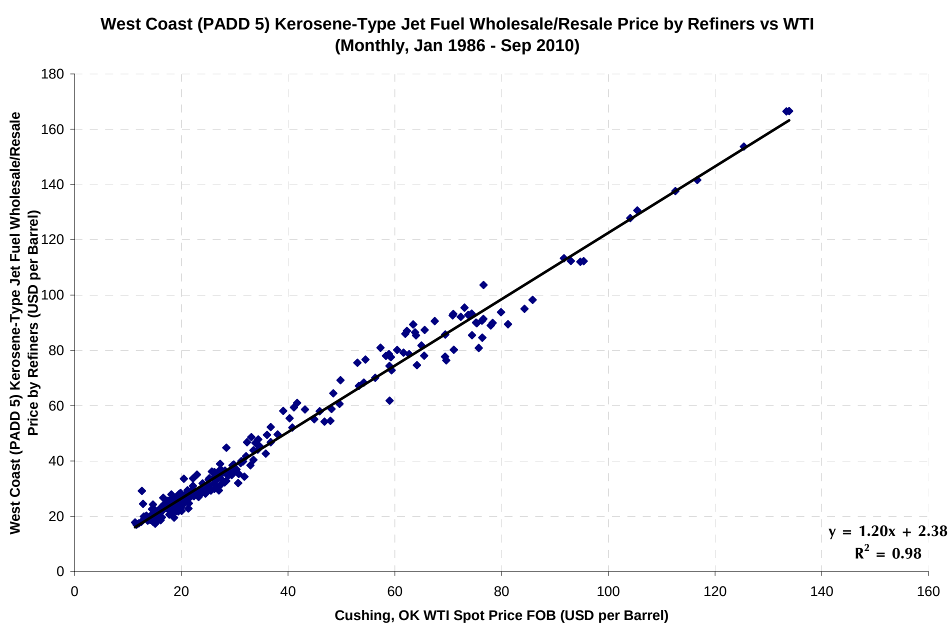
$$\text{Jet Fuel}_{\text{U.S. West cost}}\ [\text{USD/bbl}] = 1.20 \times \text{WTI} + 2.38\ [\text{USD/bbl}]$$

In the price analysis, the reported average for PADD V as a proxy for U.S. West Coast was used.



Source: EIA, State Energy Data System (SEDS)

Figure 3-4: U.S. West Coast Jet Fuel Consumption



Source: EIA, Petroleum Product Prices

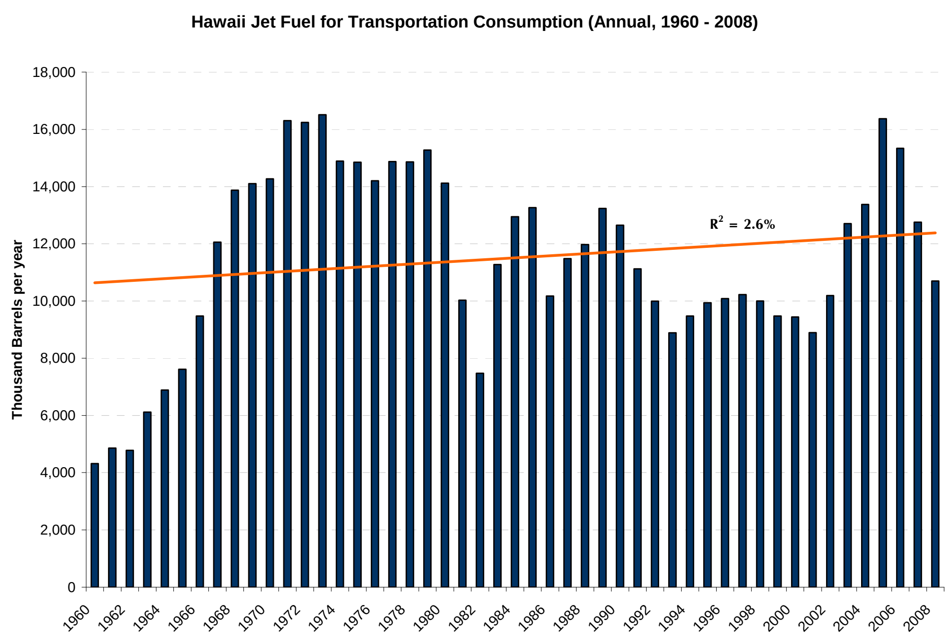
Figure 3-5: U.S. West Coast Jet Fuel vs. WTI

3.2.3 Hawaii

The Hawaiian market for kerosene-type jet fuel represented 10,702,000 bbl/a in 2008; evolution over time is shown in Figure 3-6 and exhibits no clear trend. From the analysis of historical prices, jet fuel wholesale price in Hawaii as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-7:

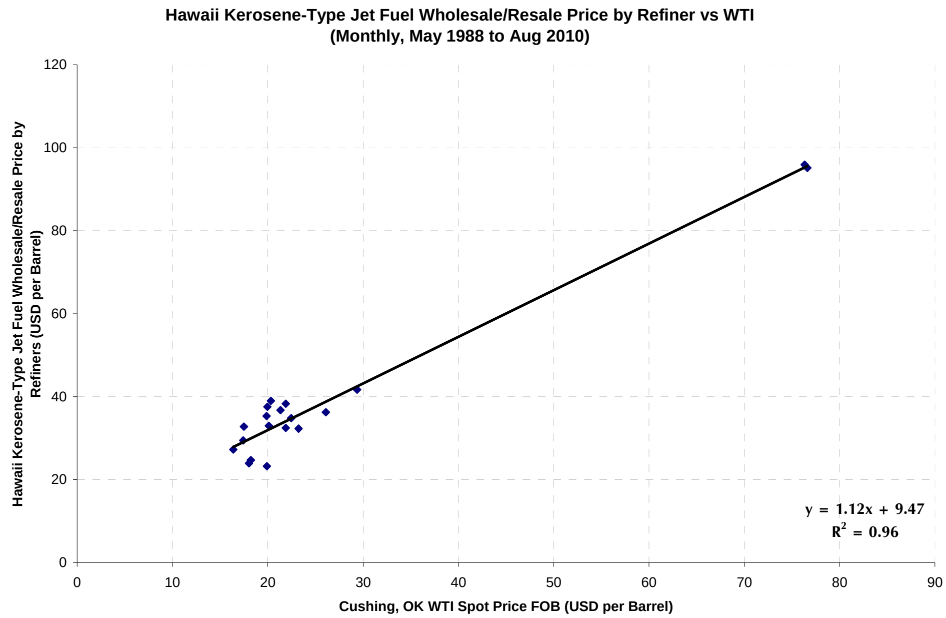
$$\text{Jet Fuel}_{\text{HI}}\ [\text{USD/bbl}] = 1.12 \times \text{WTI} + 9.47\ [\text{USD/bbl}]$$

Given the lack of available price data, a caveat needs to be made that the price model, although statistically significant, is based on only 19 observations.



Source: EIA, State Energy Data System (SEDS)

Figure 3-6: Hawaii Jet Fuel Consumption



Source: EIA, Petroleum Product Prices

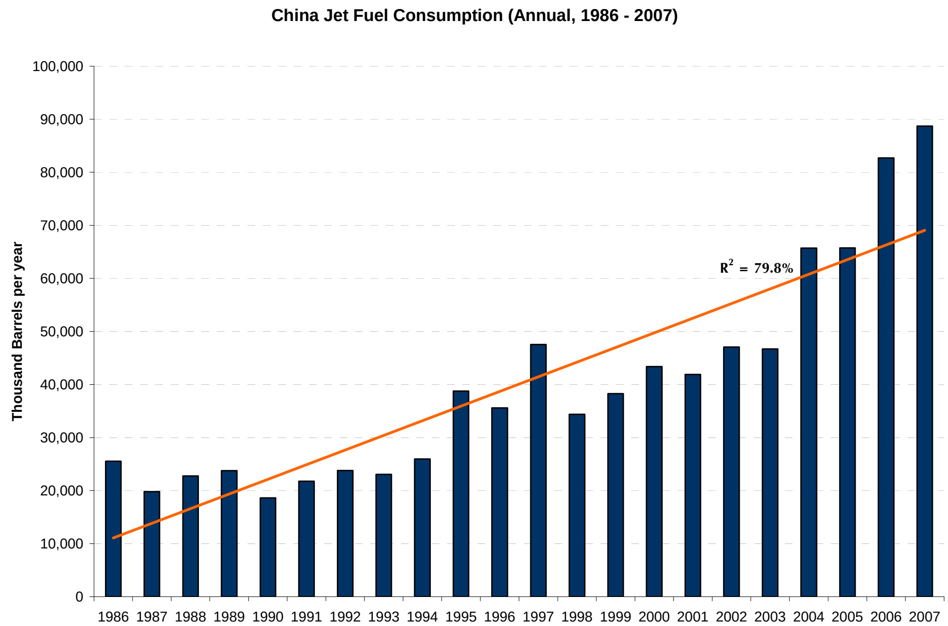
Figure 3-7: Hawaii Jet Fuel vs. WTI

3.2.4 China

The Chinese market for kerosene-type jet fuel represented 88,681,000 bbl/a in 2007; evolution over time is shown in Figure 3-8 and exhibits an increasing trend. The Chinese jet fuel market is one of the fastest growing oil product markets in China. In 2007, China imported approximately 40 million barrels of jet fuel; however, it is reported that in 2009, China imported nearly all (99.4%) of its jet fuel from Asia. Furthermore, there are only a handful of trading companies that have the authority to import jet fuel in China [4]. From the analysis of historical prices, the jet fuel imported price in China as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-9:

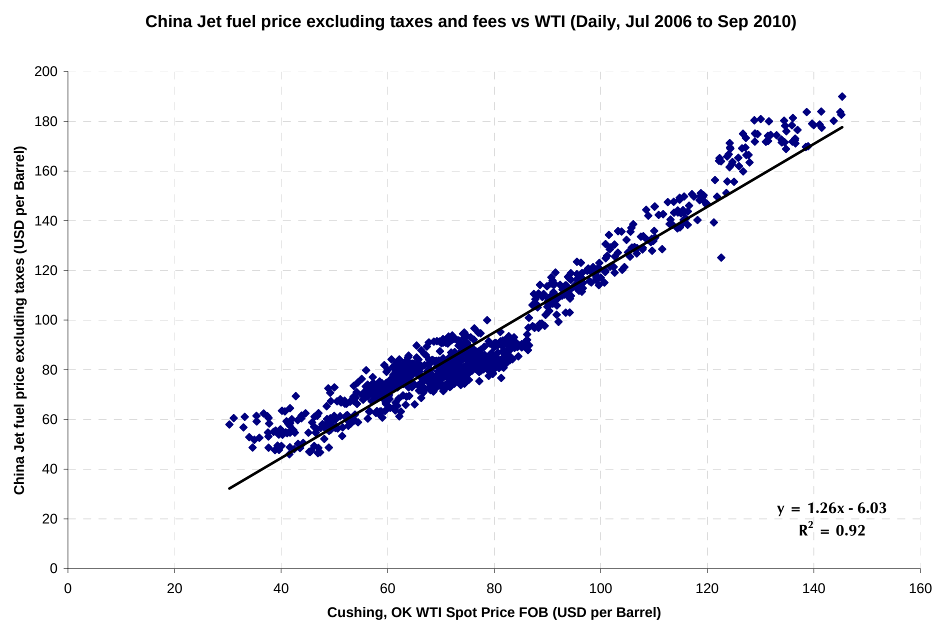
$$\text{Jet Fuel}_{\text{CHINA}}\ [\text{USD/bbl}] = 1.26 \times \text{WTI} - 6.03\ [\text{USD/bbl}]$$

For valid comparisons, imported jet fuel prices in China exclude import duties, value added tax and consumer tax. The data used in this analysis has a daily frequency from July 2006 to September 2010. We consider this to be a representative period to derive statistically significant conclusions.



Source: EIA, International Energy Statistics

Figure 3-8: China Jet Fuel Consumption



Source: Chemsin (http://www.chemsin.com/)

Figure 3-9: China Jet Fuel vs. WTI

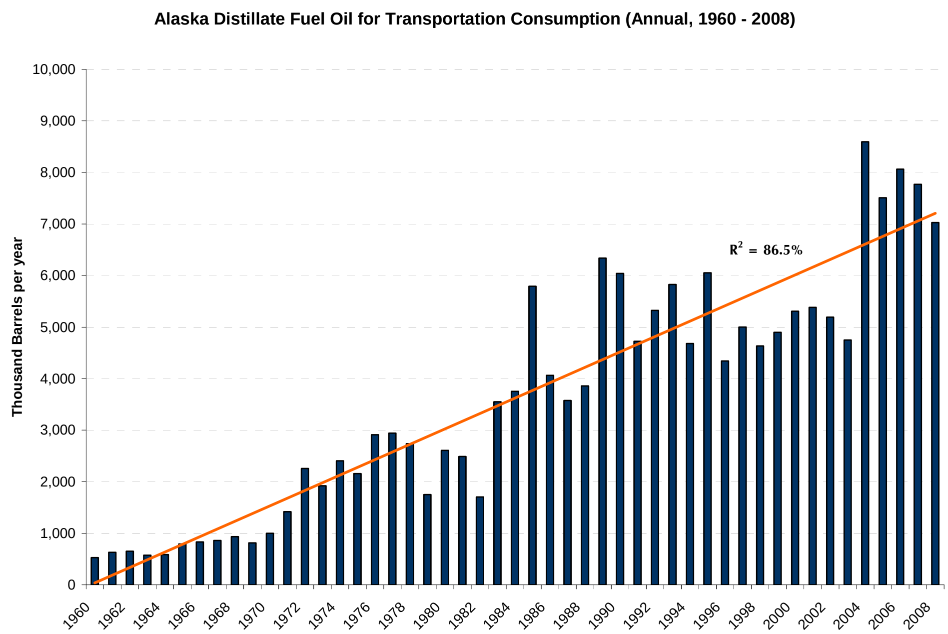
3.3 FT Diesel Analysis

The Alaska, Hawaii, U.S. West Coast, and China markets for FT diesel fuel were analyzed. As pointed out in the assumptions, for consumption analyses distillate fuel oil for transportation as a proxy for total usage of No. 2 diesel was employed. Additionally, No. 2 diesel, as a proxy for FT diesel, was selected due to lack of historical data available for ultra low-sulfur No. 2 diesel, although the latter is more similar to the FT diesel produced.

3.3.1 Alaska

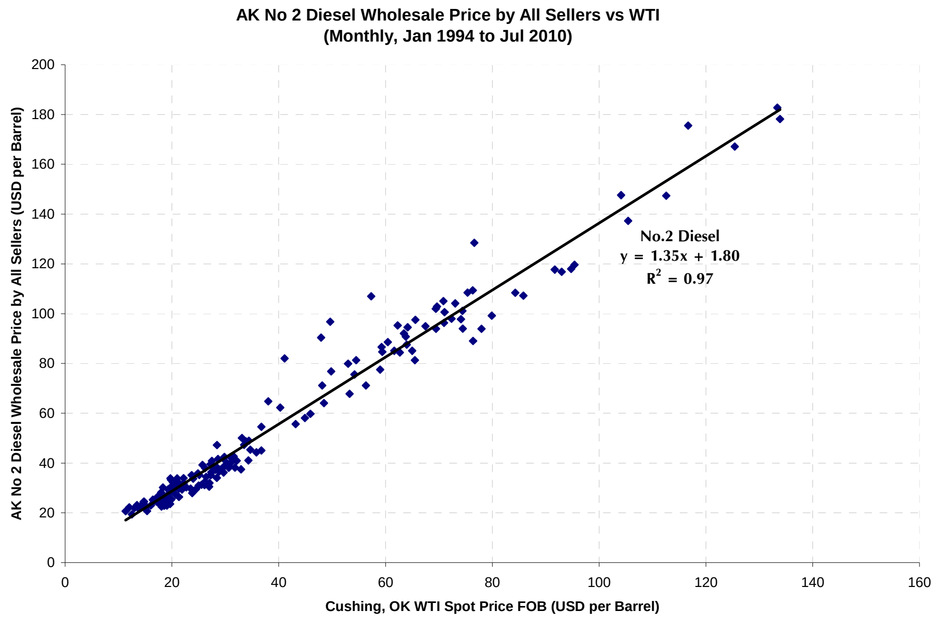
The Alaskan market for distillate fuel oil for transportation represented 7,026,000 bbl/a in 2008. Its evolution is shown in Figure 3-10 and exhibits an increasing trend. Research also indicates that Alaska's refineries currently supply the in-state needs and export excess [3]. From the analysis of historical prices, No. 2 diesel wholesale price in Alaska as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-11:

$$\text{No. 2 Diesel}_{AK}\ [\text{USD/bbl}] = 1.35 \times \text{WTI} + 1.80\ [\text{USD/bbl}]$$



Source: EIA, State Energy Data System (SEDS)

Figure 3-10: Alaska Fuel Oil for Transportation Consumption

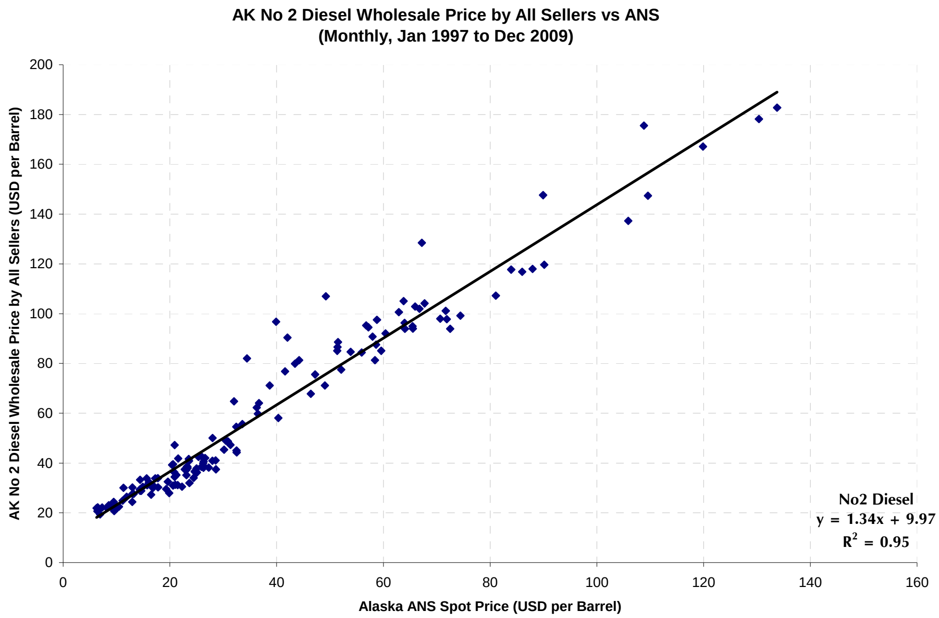


Source: EIA, Petroleum Product Prices

Figure 3-11: Alaska No. 2 Diesel vs. WTI

Additionally, the historical price relationship between No. 2 diesel wholesale price in Alaska and the Alaska North Slope (ANS) crude oil was analyzed and is given by the following equation and depicted in Figure 3-12:

$$\text{No. 2 Diesel}_{AK}\ [\text{USD/bbl}] = 1.34 \times \text{ANS} + 9.97\ [\text{USD/bbl}]$$



Source: EIA, Petroleum Product Prices ; Alaska Department of Revenue

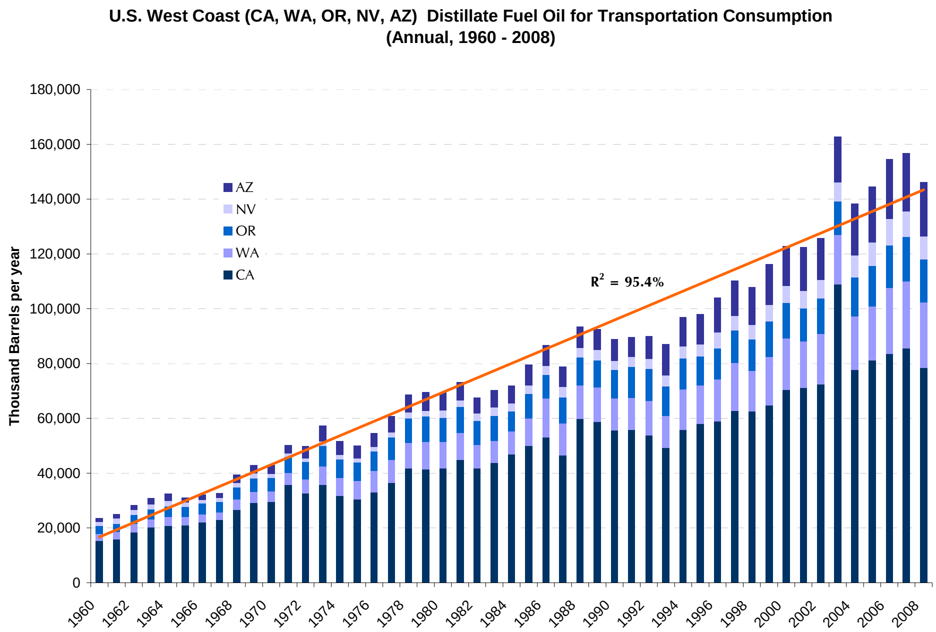
Figure 3-12: Alaska No. 2 Diesel vs. ANS

3.3.2 U.S. West Coast

The U.S. West Coast market for distillate fuel oil for transportation represented 146,061,000 bbl/a in 2008. Its evolution over time is shown in Figure 3-13 and exhibits a clear increasing trend. From the analysis of historical prices, No. 2 Diesel wholesale price in the U.S. West Coast as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-14:

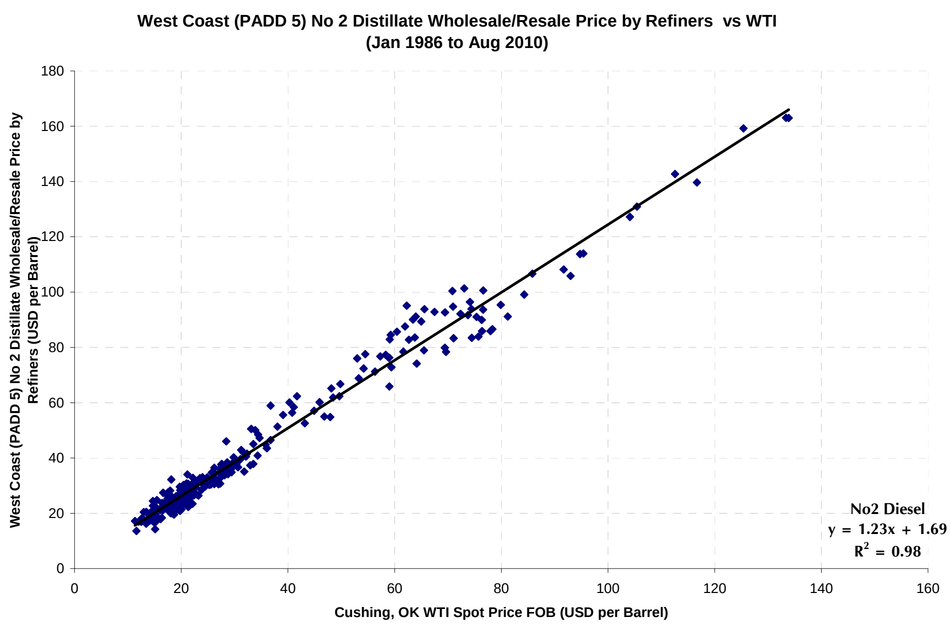
$$\text{No. 2 Diesel}_{\text{U.S. west cost}} \text{ [USD/bbl]} = 1.23 \times \text{WTI} + 1.69 \text{ [USD/bbl]}$$

In the price analysis, the reported average for PADD V as a proxy for U.S. West Coast was used.



Source: EIA, State Energy Data System (SEDS)

Figure 3-13: U.S. West Coast Fuel Oil for Transportation Consumption



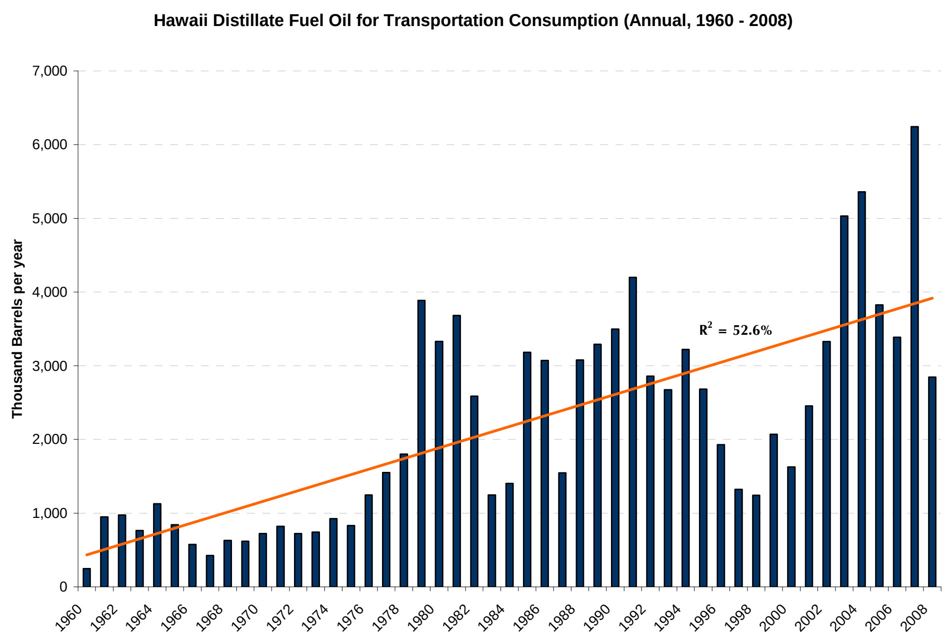
Source: EIA, Petroleum Product Prices

Figure 3-14: U.S. West Coast No. 2 Diesel vs. WTI

3.3.3 Hawaii

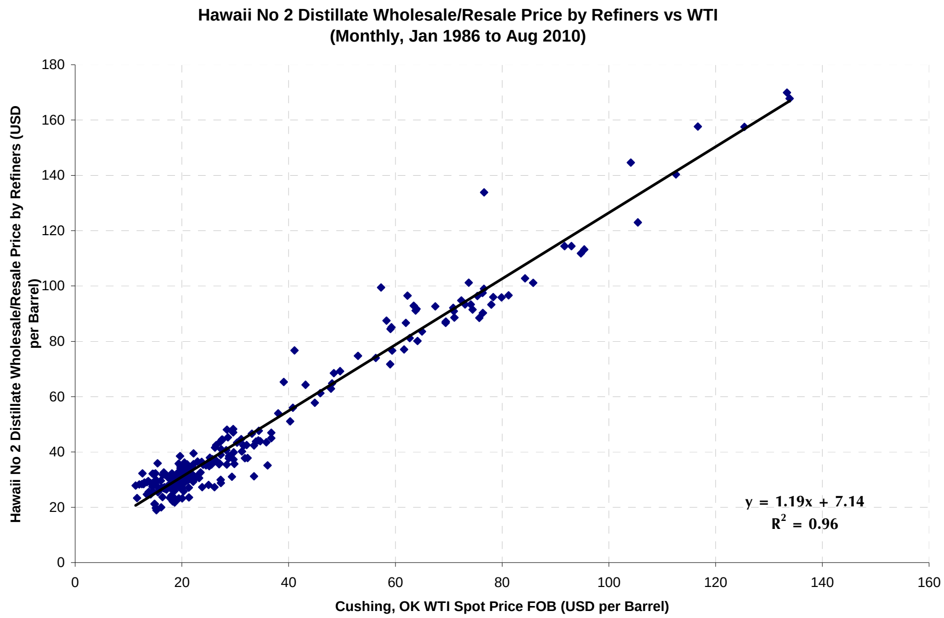
The Hawaiian market for distillate fuel oil for transportation represented 2,845,000 bbl/a in 2008; evolution is shown in Figure 3-15 exhibiting an increasing yet volatile trend. From the analysis of historical prices, No. 2 diesel wholesale price in Hawaii as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-16:

$$\text{No. 2 Diesel}_{\text{HI}} \text{ [USD/bbl]} = 1.19 \times \text{WTI} + 7.14 \text{ [USD/bbl]}$$



Source: EIA, State Energy Data System (SEDS)

Figure 3-15: Hawaii Fuel Oil for Transportation Consumption



Source: EIA, Petroleum Product Prices

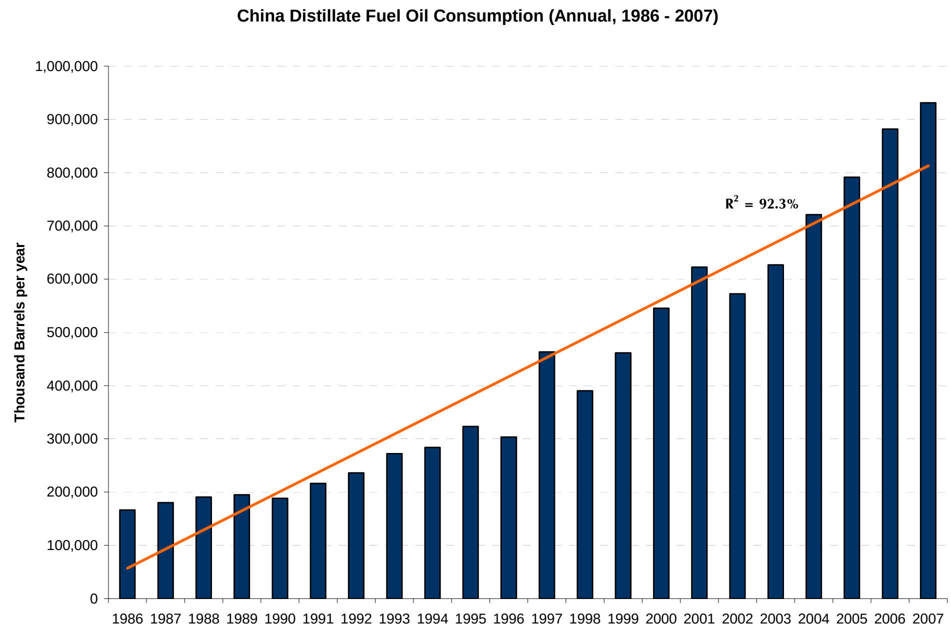
Figure 3-16: Hawaii No. 2 Diesel vs. WTI

3.3.4 China

The Chinese market for distillate fuel oil represented 931,127,000 bbl/a in 2007. This figure is more than nine times larger than the U.S. West Coast, Alaskan and Hawaiian markets combined for 2007. Its evolution is shown in Figure 3-17 and exhibits an increasing trend. From the analysis of historical prices, diesel import price in China as a function of WTI was obtained and is given by the following relationship, shown in Figure 3-18:

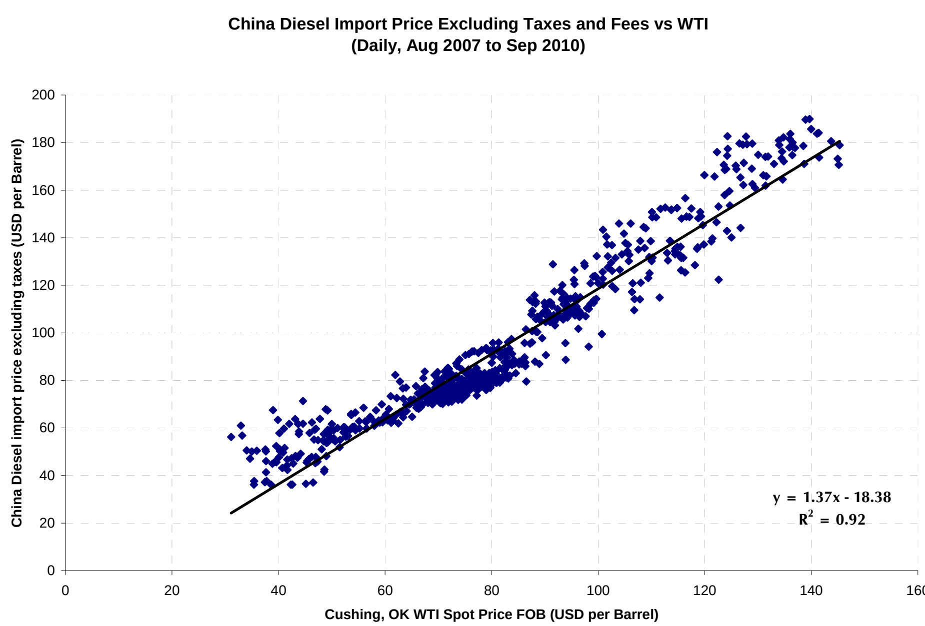
$$Diesel_{CHINA}\ [USD/bbl] = 1.37 \times WTI - 18.38\ [USD/bbl]$$

For valid comparisons, imported diesel prices in China exclude import duties, value added tax and consumer tax. The data used in this analysis has a daily frequency from August 2007 to September 2010, considered this to be a representative period to derive statistically significant conclusions.



Source: EIA, International Energy Statistics

Figure 3-17: China Distillate Fuel Oil Consumption



Source : Baichuan (http://www.baiinfo.com/default.html)

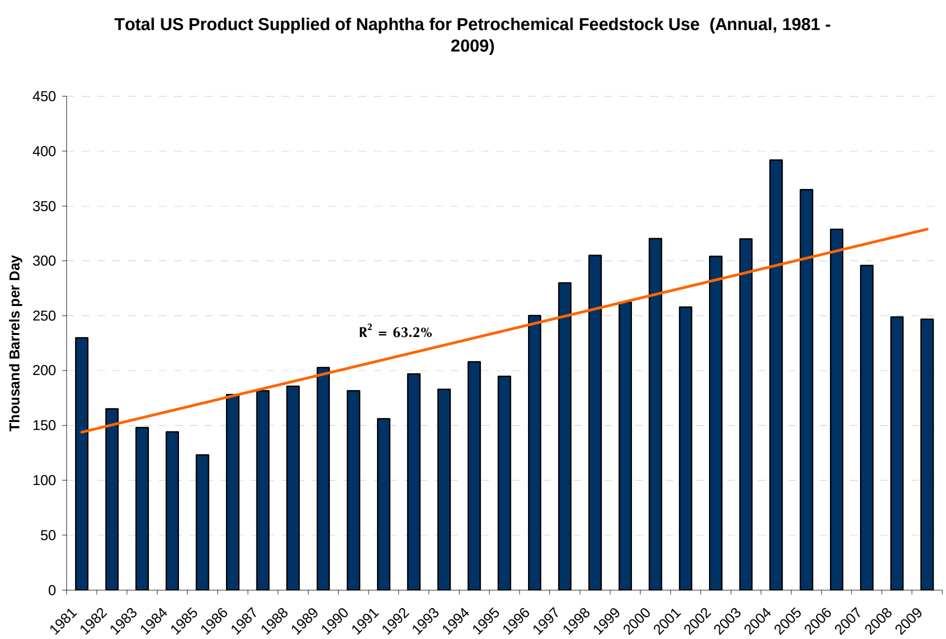
Figure 3-18: China Diesel vs. WTI

3.4 FT Naphtha Analysis

The U.S. and Asian markets for naphtha consumption as a petrochemical feedstock were analyzed. Specifically, naphtha consumption by U.S. PADD as well as in Alaska were analyzed and compared with current and projected consumption in China, Japan and Korea.

U.S. naphtha market is primarily driven by PADD III – U.S. Gulf Coast, which, over the past six (6) years, has accounted, on average, for 81% of the total U.S. consumption. Figure 3-19 shows the historical consumption of naphtha as a petrochemical feedstock in the U.S. However, current shipping costs, including insurance, time in ports and fees for Alaska to Houston are 356% of Alaska to China/Japan. The current shipping costs for Alaska to Los Angeles is 163% of Alaska to China/Japan. Moreover, based on our analysis, the total U.S. market currently represents approximately 7% of the combined naphtha consumption in China, Japan and Korea making the latter a more attractive market. Finally, there is a projected combined deficit in these three Asian markets going forward in the order of 23,000,000 bbl/month, and therefore import from foreign sources is necessary [5].

Local naphtha consumption in Alaska appears to be driven by the Golden Valley Electric Association (GVEA) in the Fairbanks region, which has a demand of 2,000 bbl/day of turbine fuel. Contracted pricing requires investigation to quantify this market. However, both the Flint Hills North Pole and Tesoro Nikiski refineries report naphtha production. Therefore, there may be insufficient demand in Alaska for additional FT naphtha consumption.

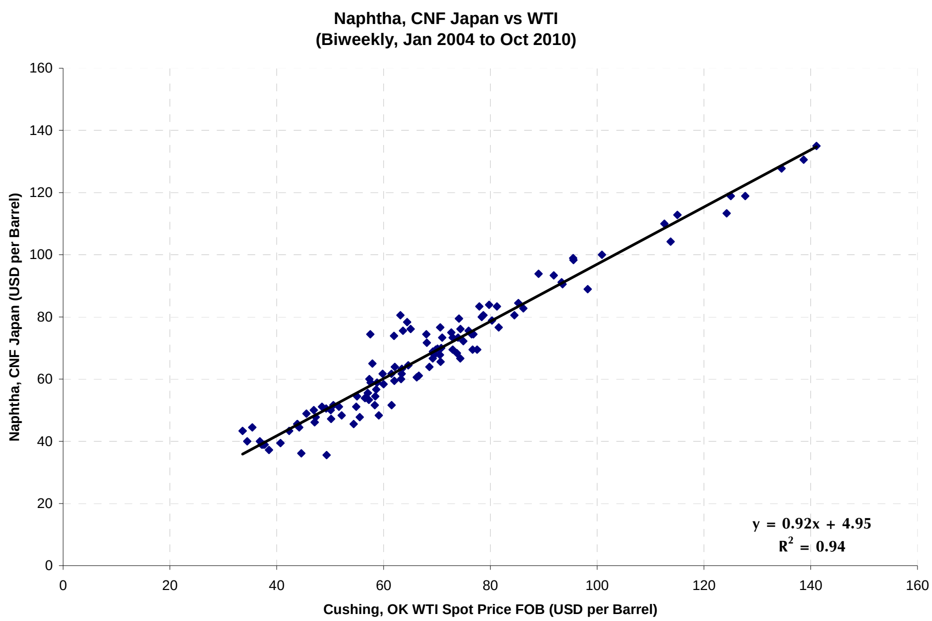


Source: EIA, Petroleum Product Supplies

Figure 3-19: Total U.S. naphtha as a petrochemical stock consumption

Based on the foregoing discussion, FT naphtha analysis in the Asian markets focused upon and the price of naphtha in Japan (quoted as CNF – cost and freight or, equivalently, exporter pays ocean freight and importer pays insurance) was analyzed as a function of WTI. The price relationship is given by the following equation and is shown in Figure 3-20:

$$\text{Naphtha}_{\text{CNF, Japan}}\ [\text{USD/bbl}] = 0.92 \times \text{WTI} + 4.95\ [\text{USD/bbl}]$$



Source: EIA, Petroleum Product Prices; http://www.plastemart.com/polymer-feedstock-prices.asp
[Accessed on November 8, 2010]

Figure 3-20: Naphtha, CNF Japan vs. WTI

3.5 Recommendations from Market Analysis

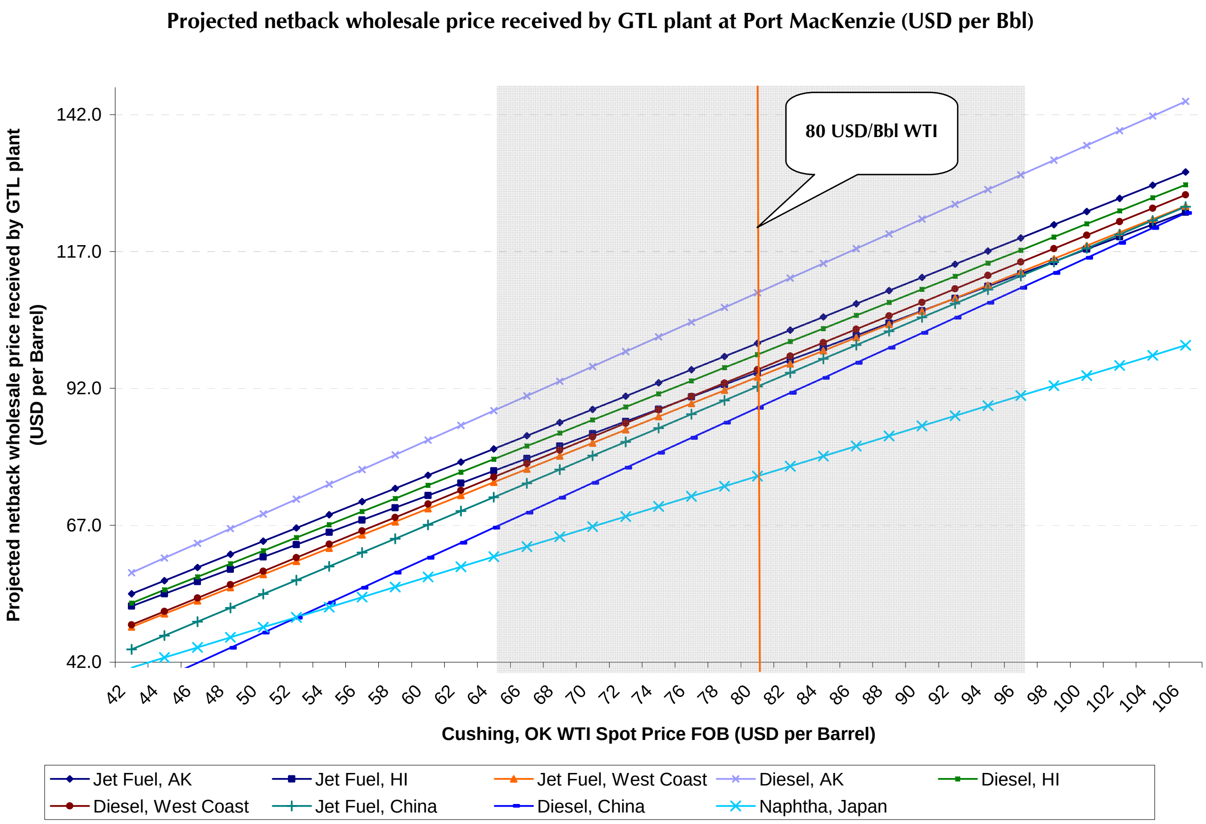
The foregoing analysis was prepared to determine the base case GTL plant product mix. From the market analysis conducted, it is recommended the GTL facility produce primarily diesel, followed by naphtha as a by-product.

A summary chart (Figure 3-21) is provided to show the netback (price less shipping costs) wholesale price as a function of WTI to be received by the GTL plant located at Port MacKenzie for jet fuel, diesel and naphtha for each market. A similar chart (Figure 3-22) is presented for the GTL plant located at Fairbanks. The latter includes the transportation of the FT liquids from Fairbanks to the export port at Anchorage. A WTI base projection of 80 USD/bbl ±20% has been overlaid on this chart. Given the projection, current shipping rates and identified fuel price relationships to WTI, it may be concluded that FT diesel should be maximized. Although FT jet fuel sold in Alaska ranks second, assessment shows the market to be small and currently supplied by local refineries (88%). Only half of the Alaskan jet fuel market requirements remains (given the required 50:50 blend), leading to an opportunity of approximately 4,000 bbl/day using 2008 consumption data. Pursuing this option may put the GTL plant at the risk of having to export jet fuel to other markets and decreasing value. As mentioned earlier, further research into the Chinese market for jet fuel reveals imports from Asia only resulting in a non-viable market.

In terms of export market prioritization for FT diesel, after trying to pursue the local market, focus should be the Hawaiian market followed by the U.S. West Coast, from a purely netback price perspective. However, given the significant historical volatility of Hawaiian diesel for transportation consumption relative to steady growth shown by the U.S. West Coast consumption, the latter market is recommended as a better alternative. Selling FT diesel in China, from a purely netback price perspective, appears to lower revenue.

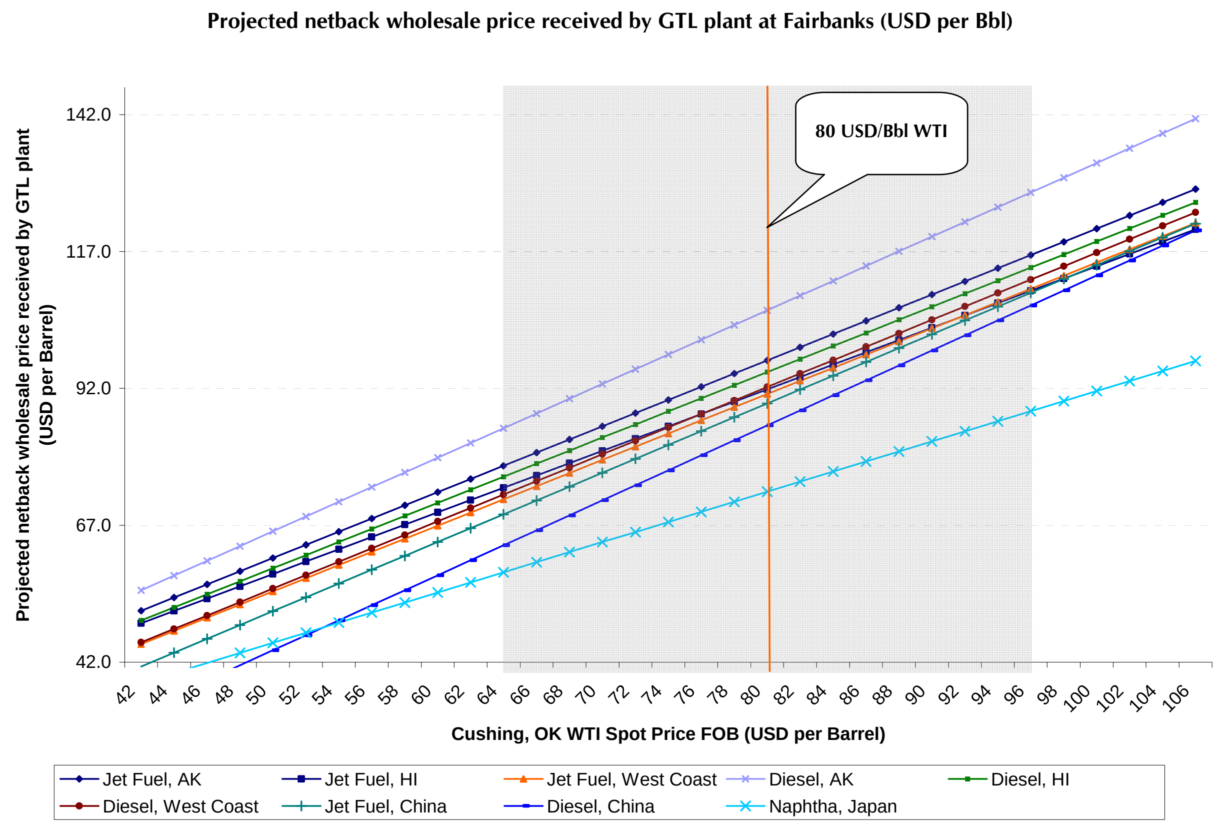
It is important to point out that the relative historical premium netback of the Alaskan market is based on the existing supply and demand equilibrium. If the GTL plant owner were to focus on the local market, the excess supply, by the amounts considered in the project, may erode this premium, bringing prices closer to export parity.

Finally, market trends show a significant demand for FT naphtha from Asian steam crackers, combined with the product's advantageous characteristics, the Asian market for naphtha is attractive. Alternatively selling naphtha to the U.S. Gulf Coast would result in higher shipping costs, reducing revenue. Currently, naphtha is used in Alaska as a refinery feedstock; however, given the low octane number of FT naphtha, it is not preferential to use it as a refinery feedstock and also not preferred for use as a gasoline blendstock. Therefore, petrochemical applications are preferred for this product and there is currently no market for petrochemical use of FT naphtha in Alaska.



Source: Hatch Analysis

Figure 3-21: Netback wholesale price by fuel type at Port MacKenzie



Source: Hatch Analysis

Figure 3-22: Netback wholesale price by fuel type at Fairbanks

3.6 Relationship between GTL Facility Costs, Products Thermal Efficiency and Economic Impact

3.6.1 Jet Fuel Production

A portion of the distillate fraction of the FT product slate can be separated to form a blendstock for kerosene-type jet fuel. During a high level assessment of the potential jet fuel recovery that may be attained, Hatch estimated that 38% jet fuel yield can be achieved through a relatively simple configuration to separate jet fuel blendstock from the rest of the distillates and by operating the hydrocracker at a marginally higher level of severity to keep the diesel blendstock within specifications. More complex refinery configurations can result in higher jet fuel yields and new designs for Cobalt LTFT GTL is reported to achieve up to a 60% jet fuel blendstock yield. However these typically involve extra refining steps to combine the lighter molecules to produce additional molecules in the jet fuel range.

The production of approximately 38% jet fuel would require a minimal increase in capital spending compared to a naphtha and diesel only case. The costs for the additional distillation step and storage facilities would not be a significant contributor to the overall capital costs of a product workup unit. Furthermore, the upgrading section typically has a 20% impact on the overall capital costs of a GTL plant.

The most common type of jet fuel is Jet A, which mostly competes with diesel, and is a lighter cut suitable in arctic climates. While, there is potential to route some of the FT naphtha towards jet fuel when producing wide-cut aviation fuel (i.e. JP-4 (U.S. military) or Jet B (Canadian specification), the market for wide-cut aviation fuel is however limited and confined to the local consumption in Alaska/Northern Canada.

3.6.2 Diesel No. 1

Similar to the jet fuel production cases, Diesel No. 1 fuel can be produced in addition to No. 2 Diesel to reduce the naphtha yield, therefore increasing the yield of the higher value products. Diesel No. 1 is also sold at a higher price than No.2 Diesel. However, Hatch concluded a high level market analysis for No. 1 Diesel and it was found that the market size is 2% of the No. 2 Diesel consumption in the PADD V district. Furthermore, the demand is seasonal (with No. 1 Diesel mostly being consumed in winter months) and appears to be declining.

3.6.3 LPG, Thermal Efficiency and other GTL products

The thermal efficiency of the GTL process can be increased through recovering lighter components, suitable for use as liquefied petroleum gas (LPG, e.g. propane and butane) and through more efficient utility system designs. The addition of LPG to the product slate typically requires a refrigeration system to recover these light components from the FT tail gas, which in turn requires additional energy consumption. LPG production typically reduces the amount of energy available for power production.

Since LPG is a volatile product which is not easily stored like diesel and naphtha, specialized storage systems are required, adding further additional capital costs for this option. Furthermore, LPG is a relatively low value product compared to the premium FT products (diesel and naphtha). For these reasons, it was decided not to include LPG in the base case due to the added complexity it would bring to the process and the relatively low impact on overall production.

Although the preferred GTL product distribution for the main FT products: LPG, naphtha, jet fuel and diesel, may change subject to more rigorous optimization, the marketing study shows that such changes would not markedly improve economic viability.

More exotic valuable products such as lube oil feedstock, paraffins and specialized FT waxes may also be extracted from the FT product slate. However, the worldwide market for these products are very limited and overproduction from GTL plants would depress values.

4. Base Case GTL Facility Definition

Using the GTL technology survey and the FT product market analysis, the process units and product mix were defined to form the base case gas-to-liquids facility for this study. Rationale for the technology chosen for this conceptual study, along with process design of the base case GTL plant are detailed herein. A summary of the base case GTL facility for this study is included in Table 4-1.

Table 4-1: Process Summary of the Base Case GTL Facility

Parameter	Value
Feedstock	Pipeline-quality natural gas delivered to plant battery limit
Products	Base Product Mix: Diesel + Naphtha Alternative Product Mix: Diesel + Naphtha + Jet Fuel
Plant Capacity	Case A – 1 FT train; 154 MMSCFD natural gas; 16,630 bbl/day FT product Case B – 2 FT trains; 309 MMSCFD natural gas; 33,260 bbl/day FT product Case C – 4 FT trains; 617 MMSCFD natural gas; 66,520 bbl/day FT product
Power Generation	Yes
CO_2 Capture	Yes
Process Selection	Syngas Production – Autothermal reforming (ATR) FT Synthesis – Cobalt-based Low Temperature Fischer-Tropsch CO_2 Capture - Selexol, Benfield or other amine-based system

4.1 Process Selection

With a variety of process technologies available for the main components of the GTL facility, careful consideration must be made in selecting technology which is appropriate for commercial scale production.

4.1.1 Syngas Production

The production of syngas from natural gas can be accomplished by steam methane reforming (SMR) or oxidative reforming (which includes autothermal reforming (ATR) and non-catalytic partial oxidation (POX)). Oxidative reformers are exothermic, providing steam for plant use. ATRs require lower operating temperatures than POX units and are therefore thermally more efficient. Furthermore, ATRs are also more thermally efficient due to their energy being generated directly in the vessel as opposed to SMR where heat is provided through heat exchange. This benefit is offset by the additional utility requirements from the air separation unit (ASU); however, the air separation unit (ASU) may be driven directly or indirectly from steam produced by the ATR.

For commercial scale use in an integrated facility the preferred method is ATR. This is especially the case for GTL capacities above 10,000 bbl/day as the ATR and oxygen units have considerably greater economies of scale at higher capacities, relative to SMRs.

Furthermore, the high H_2:CO ratio of SMR product gas (in the order of 3:1) is not ideal for cobalt-based Fischer Tropsch synthesis, which requires an H_2:CO ratio in the order of 2:1.

Additionally, the use of a single chamber for ATR eliminates the potential of tube failure associated with steam methane reformers (SMRs). Such failure results in considerable maintenance costs.

4.1.2 Fischer-Tropsch

Securing product off-takers is important in demonstrating economic feasibility; a small number of long-term, high volume contracts are preferred. Due to Alaska's remoteness, it is more likely that a GTL operator is to find off-takers outside of the local region. Therefore, easily transportable products are preferred. This preference indicates that the heavier, longer-chained products (e.g. distillates (diesel), jet-fuel and naphtha) should be targeted. Therefore, a LTFT process is most desirable.

Relative to coal-derived syngas, natural gas derived syngas is low in catalyst poisons. Therefore, the higher reactivity, longer lifetime guarantee and improved efficiency of cobalt-based catalyst processes over iron-based is preferred.

Based on demonstrated production, product selectivity, process family, catalyst type, and reactor type, the following FT technologies have been ranked in order of preference for this project only, Table 4-2.

Table 4-2: FT Technologies Ranked

Rank	Licensor	Process	Catalyst	Reactor
1	Sasol	LTFT	Cobalt	Slurry Phase
2	Shell	LTFT	Cobalt	Tubular Fixed Bed
3	GTL.F1	LTFT	Cobalt	Slurry Phase
4	BP & DPT	LTFT	Cobalt	Tubular Fixed Bed
5	Syntroleum	LTFT	Cobalt	Slurry Phase
6	Rentech	LTFT	Iron	Slurry Phase

4.1.3 Upgrading

Upgrading is widely practiced by many operators with a variety of crudes and synthetic feedstocks. FT licensors have developed proprietary technology in conjunction with upgrading providers; therefore, limited selection is available as to which upgrading technology may be employed. Upgrading technology is therefore dictated by the FT technology selected as outlined in Table 4-3.

Table 4-3: Upgrading Providers for FT Technologies

FT Licensor	Upgrading Partner
Sasol	Chevron
Shell	Shell
GTL.F1	Axens/UOP
BP & DPT	BP
Syntroleum	Syntroleum/ExxonMobil
Rentech	UOP

4.1.4 CO_2 Capture

The purge stream produced by the GTL plant will be used as fuel gas. It, therefore, does not require high purity levels to be reached as is the case with feed streams to the process section. Since in this case the CO_2 capture process has no effect on the liquid fuel synthesis process, choosing the best CO_2 capture technology for the GTL facility will be based on minimizing the capital and operating cost required for such a system.

Selexol, Benfield or one of the other amine-based systems can satisfy the CO_2 capture requirement of the GTL plant, assumed to be 95% capture efficiency. The fact that the purge gas stream is at high pressure, favors the technologies that are based on physical absorption, as they will typically have lower energy requirements than processes relying on chemical absorption (which is more favored to low pressure applications). For the purposes of this study we have selected Selexol as the CO_2 capture system.

4.2 Base Case Plant Areas

The GTL facility is designed to process the pipeline-quality natural gas supplied from Alaska's North Slope via the Alaska Stand Alone Gas Pipeline (***ASAP***). The major plant areas in the GTL facility included in this design are listed in Table 4-4, and an overall block flow diagram of the facility is provided in Figure 4-1.

Table 4-4: GTL Plant Areas

Plant Area Number	Plant Area
0100	Natural Gas Compression, Purification and Conditioning
0200	Air Separation Unit (ASU)
0300	Syngas Production
0400	Fischer-Tropsch (FT) Synthesis
0500	Hydrogen Plant
0600	Product Upgrading
0700	CO_2 Capture, Dehydration & Compression
0800	Boiler Feed Water and Steam System
0900	Power Generation System
1000	Water Treatment
1100	Fuel Gas System
1200	Cooling Water System
1300	Infrastructure and Services

Figure 4-1 illustrates the main utilities of process importance, and further detail of the GTL process design is provided in the block flow diagrams in Appendix A. Additional utility requirements are identified in the block flow diagrams.

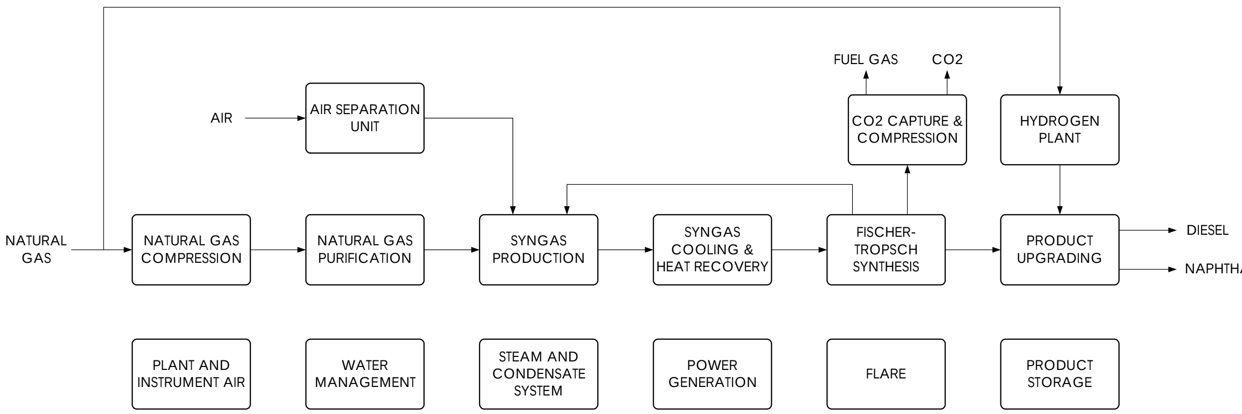


Figure 4-1: GTL Overall Block Flow Diagram

4.3 Case Definition

Three GTL facility sizes were considered in order to investigate the effect of facility size on project economics.

- **Case A:** A GTL facility utilizing one (1) FT synthesis train with pipeline-quality natural gas consumption of 154 MMSCFD which produces 16,630 bbl/day of liquid products
- **Case B (Base Case):** A GTL facility utilizing two (2) FT synthesis trains with pipeline-quality natural gas consumption of 309 MMSCFD which produces 33,260 bbl/day of liquid products
- **Case C:** A GTL facility utilizing four (4) FT synthesis trains with pipeline-quality natural gas consumption of 617 MMSCFD which produces 66,520 bbl/day of liquid products

Case B is selected as the base case for this study as the required natural gas flow rate is within the range for typical delivery capacity of the natural gas pipeline. Conceptual engineering and design of the facility, including process simulation, is performed only for this base case. A mass and heat balance of the base case facility is included with the block flow diagrams in Appendix A.

4.4 Process Description

4.4.1 Natural Gas Conditioning & Syngas Production

Natural gas enters the plant battery limit from the neighboring receiving terminal, and is measured for consumption and quality purposes. Natural gas is then compressed from the pipeline pressure of 435 psi to 725 psi suitable for ATR operating pressure, and purified (Area 0100) to remove sulfur (S) and chlorine (Cl) compounds, as required. The purified natural gas then reports to syngas production (Area 0300), with a portion reporting to the hydrogen plant (Area 0500) to be converted to hydrogen for product upgrading downstream.

In the syngas production unit, natural gas is pre-reformed after it is mixed with steam and recycle gas from FT synthesis unit. A portion of the FT synthesis purge gas is routed to syngas production where it is compressed and recycled to the syngas production unit. This recycle is necessary to produce the required H_2:CO ratio in the syngas for FT synthesis. The pre-reformed natural gas then is reformed over a catalyst with oxygen (O_2) in an autothermal reformer (ATR). The O_2 is supplied from the Air Separation Unit (ASU, Area 0200). The syngas produced primarily consists of H_2 and CO.

Syngas is produced at a high temperature (approximately 1,050°C/1922°F). Much of this heat is recovered through convective boilers. Syngas exits the heat recovery boiler and heat exchangers at approximately 100°C/212°F. The heat recovered is used to generate steam, as well as pre-heat wash water and boiler feed water.

4.4.2 FT Synthesis & Product Upgrading

The cooled syngas then enters the FT synthesis unit (Area 0400) where the H_2 and CO react in the presence of a metal-based catalyst to produce straight chain hydrocarbons. The current design assumes a cobalt-based catalyst. The FT synthesis reaction produces both oils and wax, which are separated and stabilized prior to the upgrading process. During the separation process a tail gas is produced which is divided into three portions:

1. The first portion is recycled and mixed with syngas from the syngas production unit prior to being fed to the FT synthesis reactor.
2. The second portion is mixed with purified natural gas and is recycled to the pre-reformer in the syngas production area.
3. The third portion of the tail gas is purged from the FT synthesis loop, sent to a CO_2 capture process and then used in the plant as fuel gas. This purge is required in order to control the accumulation of inert components, such as CO_2, in the synthesis loop. The captured CO_2 is compressed to 2,000 psig and can be sold as a by-product or mitigated by sequestration.

Wax and oils are combined with compressed H_2, from the hydrogen plant (Area 0500), and undergo hydrotreating and hydrocracking processes producing the required products (e.g. diesel, naphtha) in the upgrading process (Area 0600). The hydrogen plant (Area 0500) consists of a Steam Methane Reformer (SMR) and a Pressure Swing Adsorption (PSA) unit to produce H_2 for product upgrading.

4.4.3 Power Generation

Power generation (Area 0900) equipment will be installed within the GTL facility, using steam produced throughout the process to generate power. The plant produces both high pressure (HP) and low pressure (LP) steam from the ATR and FT Synthesis processes respectively. HP steam generated from the process is used within the plant to directly drive large rotating machinery (e.g. ASU compressors), and excess steam is sent to an HP steam turbine for electricity generation. LP steam will be distributed to various processes throughout the plant, but the majority will be sent to the LP steam turbine generator for power production. Heat tracing of the facility and equipment will likely also require LP steam.

Exhaust from both steam turbine generators is cooled by an air cooling condenser, and the resulting condensate is sent to the water management facility for re-use in the plant.

4.4.3.1 Start-Up Power

The GTL facility also includes gas-fired boilers to produce steam during start-up conditions. In order to ramp up production of the plant, both steam and electrical power must be provided to machinery at the front end of the plant. The boilers have been designed to produce HP steam in order to drive the required steam-driven machinery, and excess steam will be sent to the HP steam turbine generator to produce power.

Major consumers during start-up include the natural gas compressor and the air separation unit, which are both required before operation of the ATR can commence. Once the ATR is in operation, HP steam will be produced and the start-up boilers can be ramped down as necessary. For the Case B facility, start-up boilers have been sized to provide approximately 100 MWe of power.

4.4.4 Water Management

The GTL plant requires water as an input for various processes with the major users by volume being steam services (heat exchange and natural gas processing) and open circuit cooling. The GTL process itself generates recoverable water as process condensate from the FT and ATR reactions. Overall, however, the GTL plant is a net consumer of water. For the purposes of this study, it was assumed that the plant would be supplied with aquifer groundwater that requires clarification and minimal softening before use as process water, cooling and firefighting water. The plant water management configuration for this study incorporates a water reclamation system in order to reduce net fresh water demand by 60% compared to a “once-through” system.

The plant heat load of almost 900 MMBTU/hr is initially proposed to be removed by re-circulating cooling water through an open circuit cooling tower loop. During future phases of study, further consideration is required for selection of the most appropriate heat rejection technology for the Alaskan climate. Definition of the cooling technology and feed water source allows for further optimization of the cooling water recycle rate to minimize make-up requirements.

Demineralized water for the plant’s boilers is produced by treating media filtered process water in a two-pass reverse osmosis (RO) system to remove dissolved solids. A common demineralized water make-up stream is produced for the plant’s low, medium, and high pressure boilers, but requires optimization during the next phase of study. Steam condensate from the plant’s turbines (high and low pressure) and upgrading column reboilers are collected and recycled as boiler feed water after passing through cartridge filtration units.

Process wastewaters contaminated with organic compounds are treated in a two stage biological process. In the first stage, an anaerobic process converts the majority of organic compounds to an energy rich biogas, and in the second stage an aerobic polishing step converts the residual organic compounds to carbon dioxide and landfillable biosolids.

Various partially treated wastewater streams are reclaimed to reduce raw demands and recover water that is generated through the GTL process reactions. Reclaim water streams are pre-treated for suspended solids before being fed to a single pass reverse osmosis membrane unit to produce a product suitable for use as process water.

Operation of the water reclamation processes generate a brine stream of concentrated dissolved solids with concentration of 5,000 – 10,000 mg/L and approximate flow of 1,200 gpm for Case B. The local water utility in the Anchorage area is likely capable of accepting this stream at the wastewater facility for a surcharge. Further investigation is required to determine an appropriate brine discharge point in the Fairbanks area and if marine disposal is feasible for Port MacKenzie.

Nearly 100% water recovery may be achieved by treating the brine stream on site through the use of evaporator and crystallizer equipment at the expense of greater capital and energy (steam) requirements. A trade-off study would be performed in the next phase of study to determine if the marginal costs associated with zero liquid discharge would balance the costs of water import and brine discharge fees. Environmental regulatory restrictions on maximum water intake and brine discharge also require consideration.

4.4.5 Balance of Plant

Additional plant areas are required to complete the balance-of-plant (BOP), including: quality control, tank farm, flare system, power distribution, plant and instrument air, security, rail, roads, maintenance shop, administration buildings, firefighting equipment, and medical emergency building. These BOP are classified as infrastructure and services (Area 1300).

5. GTL Plant Performance

5.1 Performance Summary

Performance of the GTL facility is summarized in Table 5-1. These values are considered preliminary estimates and require confirmation from technology suppliers/licensors during the next phase of study.

Table 5-1: GTL Plant Performance Summary

Parameter	Unit	Quantity		
		Case A	Case B	Case C
INPUTS				
Natural Gas	MMSCFD	154	309	617
	MMBTU/hr	6,675	13,350	26,700
Oxygen (99 vol%)	st/hr	168	336	671
Total Raw Water Intake	st/hr	228	455	910
OUTPUTS				
Total Liquid Fuels Produced	bbl/day	16,630	33,260	66,520
Diesel (74 vol%)	bbl/day	12,340	24,680	49,360
Naphtha (26 vol%)	bbl/day	4,290	8,580	17,160
Power Export	MW_e	60	119	239
CO_2 for Export	st/hr	38	77	153
Treated Water Discharged	st/hr	141	281	562
Thermal Efficiency†	%	57	57	57
Carbon Efficiency*	%	70	70	70

† Thermal efficiency refers to higher heating value of the FT liquid products (diesel + naphtha) over the feedstock (natural gas)

* Carbon efficiency refers to amount of carbon in feedstock converted to FT liquid products (diesel + naphtha)

Major inputs and outputs have also been calculated on a normalized unit/bbl of plant production (diesel + naphtha) value and are presented in Table 5-2.

Table 5-2: Normalized Performance of the GTL Facility

Parameter	Unit	Quantity
Natural Gas	SCF/bbl	9,280
	MMBTU/bbl	9.6
Power Export	kWh/bbl	86
CO_2 for Export	lbs/bbl	110
Total Raw Water Intake	lbs/bbl	657
Treated Water Discharged	lbs/bbl	406

5.2 Feedstock

The primary feedstock for the GTL facility is natural gas, and approximately 9.6 MMBTU/bbl, equivalent to 9,280 SCF/bbl, is consumed by the plant. Table 5-3 presents the natural gas composition, as provided by AGDC, used for simulation purposes.

Table 5-3: Natural Gas Composition

Gas Component	Composition
CH_4 (vol%)	91.0
C_{2+} (vol%)	6.4
CO_2 (vol%)	1.5
N_2 (vol%)	0.7
O_2 (vol%)	0.6
Total S (max ppmv)	16.0

5.3 Synthetic Fuel Products

The primary products from the GTL facility are diesel and naphtha. Each Fischer-Tropsch train produced approximately 16,630 bbl/day, resulting in 16,630 bbl/day for Case A (1 train), 33,260 bbl/day for Case B (2 trains) and 66,520 bbl/day for Case C (4 trains). The product split described in Table 5-1 is estimated based on Hatch's in-house data and requires licensor verification during future phases of study.

5.3.1 FT Diesel

Although FT Diesel conforms to most of the ASTM D975 specifications, lubricity enhancing additives are required to conform to the lubricity requirements. In addition, blending with higher density diesel is required to obtain an acceptable energy density. A comparison of typical FT Diesel parameters with ASTM D975 specifications is presented in Table 5-4.

Table 5-4: Typical FT Diesel Properties

Parameter	Unit	Method	No.2-D S15		Typical FT Diesel
			Min.	Max.	
ASTM D975 Specification					
Flash Point	°C		52	-	65
Distillation Temperature 90%	°C	ASTM D86	282	338	338 *
Kinematic Viscosity @ 40°C	mm^2/s		1.9	4.1	3
Sulfur Content	ppmw		-	15	<10
Cetane Number			40	-	>70
Typical Properties					
Density	bbl/tonne		7.33		8.25
API Gravity	°API		22	41	54
Energy Density	MMBTU/bbl		5.825		5.44

* Distillation temperature can be controlled by adjusting recycle rate to hydrocracker

5.3.2 FT Naphtha

The FT naphtha is highly paraffinic, making it a premium feedstock for steam crackers producing ethylene and propylene. A higher yield of ethylene and propylene is obtained through cracking FT-derived naphtha compared to its crude-derived counterpart. Cracker run times are also said to be longer with FT-derived naphtha. FT naphtha may also be used as gasoline blendstock or directly for power production, but the same premium properties are not utilized in these applications.

Similar to diesel fuel, there is small variance between typical density of crude-derived naphtha and FT naphtha, 8.31 bbl/tonne and 9.37 bbl/tonne respectively.

5.3.3 FT Jet Fuel

Alternative product mixes can also be produced by the GTL facility, one example including diesel (42 vol%), jet fuel (38 vol%) and naphtha (20 vol%). Similar to FT Diesel, FT Jet Fuel also has enhanced properties compared to crude-derived Jet A, but does not meet density requirements without blending. A comparison of FT Jet Fuel properties with Jet A is presented in Table 5-5. Further discussion regarding the impacts on facility costs and efficiency are summarized in section 3.6.

Table 5-5: FT Jet Fuel Properties

Parameter	Unit	Method	Jet A, A-1		Typical FT Jet Fuel
			Min.	Max.	
Specifications					
Aromatics	vol%	ASTM D1319	-	25	<0.1
Sulfur	wt%	ASTM D4294	-	0.3	0
Sulfur (mercaptans)	wt%	ASTM D3227	-	0.003	0
Distillation 10% recovered	°C	ASTM D86	-	205	158
Distillation end point	°C	ASTM D86	-	300	300
Flash Point	°C	ASTM D56	38	-	41
Density @ 15°C	kg/m3	ASTM D1298	775	840	739
Freezing Point					
Jet A	°C	ASTM D5972	-	-40	< -70
Jet A-1	°C	ASTM D5972	-	-47	< -70
Net heat of combustion, LHV	MJ/kg	ASTM 4809	42.8	-	44.2
Typical Properties					
API Gravity	°API		37	50	60
Energy Density	MMBTU/bbl		5.670		5.30

5.4 Utilities

Utility balances for the Case B facility are summarized in Appendix A, and described in subsequent sections. Plant distribution of major utility streams is also presented in the block flow diagrams (Appendix A).

5.4.1 Power Balance

Major power consumers and producers have been estimated and tabulated in Appendix A. Power consumption has been divided up into both electrically-driven and steam-driven machinery in order to represent how much steam will be used for electrical generation, see section 5.4.2. A 20% contingency has been applied to account for smaller miscellaneous consumers not investigated at this stage of study.

Case B generates approximately 183 MWe of power, of which 119 MWe are available for export to the local grid.

5.4.2 Steam Balance

The GTL facility is designed based on two levels of steam pressure: low-pressure (LP) steam produced by FT synthesis, and high-pressure (HP) steam produced by the ATR. The hydrogen plant also produces medium-pressure (MP) steam but this is consumed by the hydrogen production process.

Steam is first distributed to process users and steam-driven machinery, the excess steam is used to generate power through steam turbines. Approximately 46 wt% of steam produced by the plant is used for power generation, while 27 wt% is sent to steam-driven machines and 27 wt% to the process.

5.4.3 Cooling Balance

Major cooling loads throughout the GTL facility are maintained by a cooling water system. In order to reduce the water circulation rate, large cooling loads are designed to be handled by air cooling systems. These loads include condensers in the power generation and FT synthesis areas of the plant. A 20% contingency has been applied to account for miscellaneous cooling loads not included at this stage of study.

5.4.4 Water Balance

Case B water balance is presented in Appendix A. Fresh water intake from a local source and water generated by the process account for 61% and 39% respectively. The major source of water losses is through the evaporative cooling towers. At future phases of study the use of a glycol cooling system will be investigated which could potentially remove this large source of water loss. Approximately 38% of total water is discharged from the water reclamation plant.

5.4.5 Fuel Gas Balance

The majority of fuel gas produced in the GTL facility originates from the FT synthesis loop. A purge is required to maintain a consistent level of inert gases in the recycle stream, as shown in the block flow diagrams in Appendix A. This purge stream is sent to the CO_2 capture plant to remove CO_2 and provide a cleaned tail gas for plant distribution. Small amounts of purge gas from various plant areas are also combined prior to distribution.

Fuel gas is required as a heat source for various fired heaters throughout the plant, with the largest consumers being the ATR heater and steam superheater. The Case B GTL facility produces excess fuel gas of about 65 MMBTU/hr, equivalent to 3.5% of total fuel gas production. One potential solution would be to utilize the start-up gas boilers and produce more steam, potentially increasing power generation capabilities.

5.5 Effluents & Emissions

5.5.1 Solid & Liquid Effluents

Solid and liquid effluents from the GTL facility are only produced by the on-site water treatment plant, and include waste activated sludge and a brine water stream for discharge. Waste activated sludge can be sent to the local landfill for disposal, while the brine water stream can likely be treated by the local water utility, or possibly discharged back to the local water source. Further investigation of these options will occur in the next phase of study.

5.5.2 Gaseous Emissions

Gaseous emissions from the GTL facility primarily originate from the fired heaters located throughout the plant. These units combust fuel gas to provide heat for various streams in the natural gas purification, ATR, hydrogen production and product upgrading areas of the plant. Standard practice for these emission sources involves venting directly to the atmosphere, and compositions are similar to those from gas-fired power generation facilities.

5.5.2.1 Carbon Footprint

A CO_2 emission balance is presented in Appendix A. Fired heaters throughout the plant produce CO_2 during the combustion of fuel gas, and emit this CO_2 into the atmosphere with a dilute concentration in low pressure flue gas. These streams are considered “difficult to capture” emission sources.

The major portion of fuel gas produced in the plant is purged from the FT synthesis loop. This stream is purged at a pressure of approximately 390 psi(g) and contains high concentrations of CO_2. For these reasons this stream is considered an “easy to capture” source, and is sent to the CO_2 capture facility in order to remove CO_2 prior to distribution throughout the plant. Although this gas is used as the heat source for combustion in fired heaters, removing CO_2 prior to combustion greatly reduces the CO_2 emissions from the plant. Approximately 38 st/h, 77 st/h, and 153 st/h of CO_2 is captured for Cases A, B and C, respectively.

5.5.2.2 Air Quality Control

The Department of Environmental Regulation in Alaska requires that stationary emission sources meet specific air quality control regulations, based on information published by the U.S. Environmental Protection Agency (EPA); Table 5-6 below establishes the significant impact level for various pollutants. As long as ambient impacts from emissions of a stationary source are below the concentrations listed, the emissions are not considered to cause a violation of ambient air quality standards or maximum allowable increases for a Class II area. Proposed sites in both Cook Inlet and Fairbanks fall under the Class II distinction, although Fairbanks is currently classified as an EPA PM2.5 Non-Attainment region and subject to strict particulate matter regulations, as described in section 7.1.1.

Table 5-6: Significant Impact Levels for Ambient Impacts of Emissions from Stationary Sources [6]

Pollutant	Significant Impact Level (micrograms per cubic meter)				
	Annual	Averaging Time (hours)			
		24	8	3	1
Sulfur Dioxide	1.0	5	N/A	25	N/A
PM10	1.0	5	N/A	N/A	N/A
Nitrogen Dioxide	1.0	N/A	N/A	N/A	N/A
Carbon Monoxide	N/A	N/A	500	N/A	2 000

The ambient impact of emissions from the GTL facility will be determined during the next phase of study using dispersion modeling techniques combined with more detailed information from technology suppliers, but the following qualitative arguments can be made:

- Sulfur Dioxide (SO_2): Natural gas is assumed to be delivered to the plant with total sulfur content less than 16 ppm. This sulfur is then captured by the on-site desulfurization unit, down to levels below 10 ppb, prior to any fuel gas production. For this reason SO_2 emissions during normal operation are likely not to be a concern. During start-up, when raw natural gas is burned in plant start-up boilers, there is potential for SO_2 emissions to increase but further analysis of this emission source will need to be investigated during the next phase of study.
- Particulate Matter (PM): Should not be a concern as the GTL facility uses only gas-fired processes.
- Nitrogen Dioxide (NO_2): NO_2 emissions from the combustion of fuel gas is assumed to be similar to that of gas-fired power plants, which without the use of de-NOx systems will not meet EPA standards. Supply of fired heaters is typically a package unit offered by technology vendors, and therefore in the next phase of study further details regarding low-NOx systems for these packages require investigation. Typical options used in power plants to reduce NOx emissions include use of low-NOx burners, selective catalyst reduction (SCR) and selective non-catalyst reduction (SNCR), all of which could potentially be applied to the fired heaters.
- Carbon Monoxide (CO): CO is one of the key components of syngas produced in the GTL facility, and is also contained in fuel gas fed to the fired heaters. Heaters will be designed for complete combustion and therefore minimal CO emissions are expected, meeting the necessary regulations.

6. Plant Layout

A facility plot plan was developed for the Case B facility (33,260 bbl/day), based on a two (2) FT reactor design each with a capacity of 17,000 bbl/day. The facility is broken down into the plant areas as defined in Table 4-4.

Additionally, there are miscellaneous services which may be classified as general utilities, infrastructure and offsites, including:

- Process flare
- Maintenance shop & stores
- Emergency response
- Storage yard
- Product export
- Security
- Administration

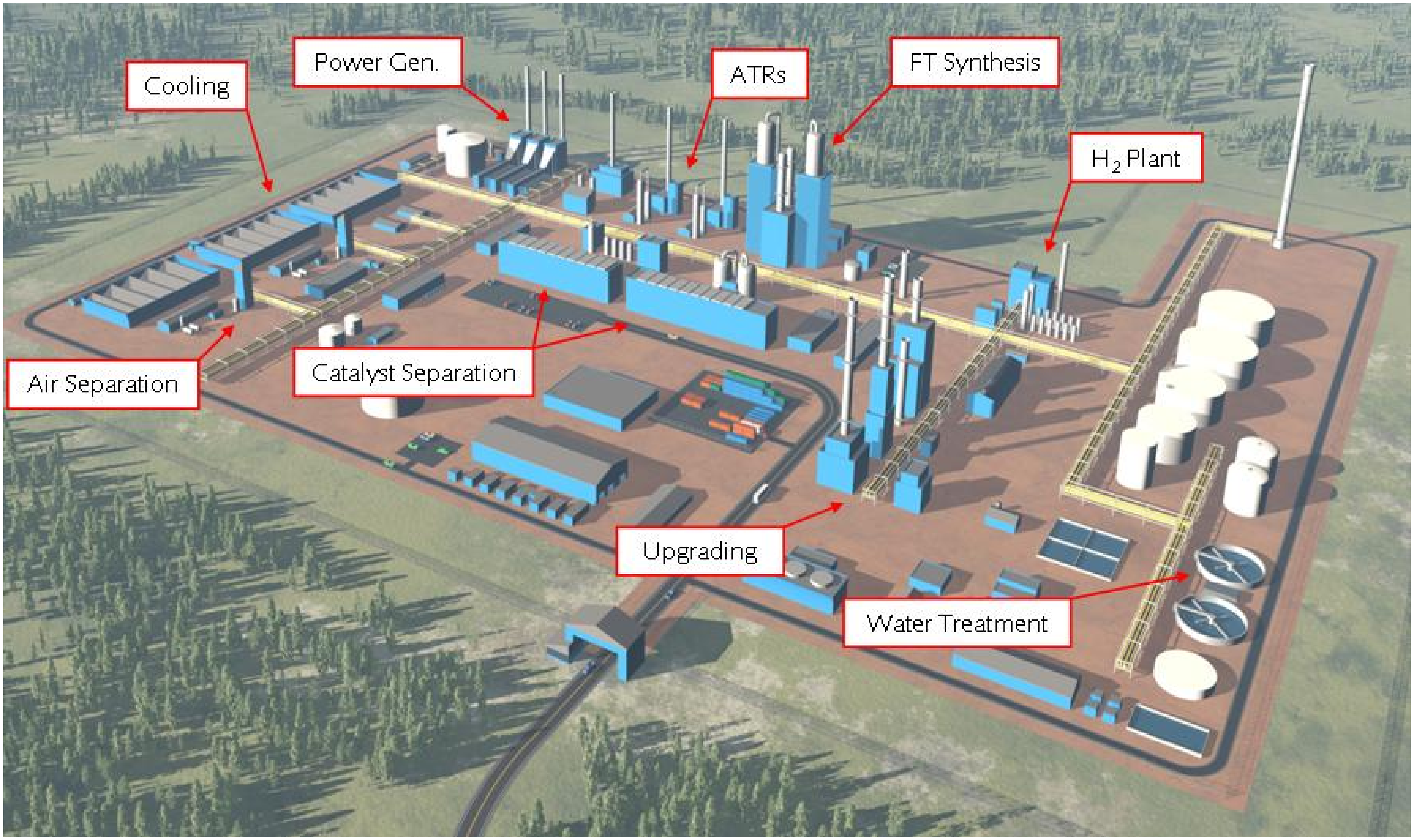


Figure 6-1: Rendition of Conceptual 33,260 bbl/day GTL Facility in Alaska with Process Units Identified

Due to the in-series design of the GTL facility and integration of various process units, the plant has been laid out along a central pipe rack. A scaled plot plan may be found in Appendix B, outlining the major process units and plant areas. Renditions of the conceptual facility are also included in Appendix B.

Considerations for the layout of the plant include:

- Prevailing wind direction, such that the intake for the ASUs is up-wind of any gas processing units (e.g. ATR, FT and upgrader).
- The water import and conditioning, boiler feed water, steam recovery, cooling, ASUs, power generation and primary power distribution units have been grouped in order to minimize large pipe runs, notably steam lines.
- The ATRs and FT synthesis units are located adjacent to one another, minimizing syngas pipe runs.
- Integral to syngas production and FT synthesis are the catalyst separation units, located flanking their respective processes.
- The neighboring miscellaneous utilities to FT synthesis include fuel gas recovery and quality control systems for the FT units feeding the upgrading units.
- Upgrading and hydrogen production are separated from upstream units in order to provide access to the various plant areas – for maintenance and operational purposes.
- As the process water treatment facility has open units, it has been removed from high traffic areas to avoid contamination and limit access lying next to the tank farm.
- The tank farm is removed from the facility such that access to the area is strictly controlled and only as necessary.
- The flare stack – operational only during start-up, shutdown and emergency procedures – is sufficient to flare the complete flow of gas while providing for personnel to evacuate the area safely in an adequate amount of time without the need for protective equipment.
- The maintenance shop, stores, storage yard, staff house and emergency response areas have been grouped in the vicinity of the main gate in order to reduce the number of personnel deep within the plant areas. This area of high personnel concentration is also removed from the hydrocarbon processing units and located up-wind to mitigate potential exposure should an incident occur.
- Access to the facility is thereby heavily controlled, surrounded by a double fence with a 150 ft recess; access is restricted to the main gate and controlled by security personnel. Administration and other non-essential staff to plant operations reside in the office block outside of the plant battery limits.

7. Plant Siting

7.1 Siting Considerations

The Alaska Stand Alone Gas pipeline is proposed to deliver natural gas from the north slope to the Cook Inlet and Fairbanks regions in Alaska. Therefore, Fairbanks North Star (FNS) and Matanuska-Susitna (MS) boroughs were identified as the two regions to investigate with regards to potential sites for the GTL facility; both the geographic regions and potential site locations were evaluated. Screening criteria included:

- For each geographic region consider:
  - Access – road, rail, marine
  - Labor force
  - Environmental conditions
  - Seismic conditions
- For each individual site consider:
  - Availability and access to land (ownership, expansion, zoning)
  - Facility access – road, rail, marine
  - Suitable separation from residential, commercial and military zones
  - Sufficient divide from airports and flights paths
  - Prevailing wind direction
  - Local utility capacity (water, power, gas, etc.)
  - Site conditions (topography, rock mechanics, soil conditions, drainage, etc.)
  - Access to water
  - Air quality restrictions

7.1.1 Air Quality - Particulate Matter

Fairbanks North Star Borough is subject to strict particulate matter regulations, EPA PM2.5 Non-Attainment. Surrounded by hills on the north, east and west sides, Fairbanks is susceptible to temperature inversions - trapping a layer of cold air close to the ground. Small amounts of particulate matter and other air pollutants can remain suspended for days and become more pronounced in periods of cold weather (below -15°F). EPA PM2.5 limits the products of combustion arising from increased fuel consumption during cold periods. Although clean burning fuels (natural gas and syngas) are consumed to provide heat, the GTL facility would be required to show sufficient modeling that it is not violating PM2.5 restrictions. The EPA's PM2.5 non-attainment boundary includes the majority of the FNS borough, Appendix B, therefore it is unlikely that a suitable site within the FNS borough would be exempt from PM2.5 restrictions.

7.1.2 Ice Fog

Alaska suffers from the effects of ice fog, a result of the cold air's inability to hold any additional moisture. When warm, moist exhaust interacts with a freezing atmosphere, water quickly freezes and produces fine particles. Air temperature inversions aggravate the effects of ice fog, suspending it over the city, Figure 7-1.

Fairbanks is more susceptible to ice fog than Cook Inlet due to the surrounding hills, exacerbating and prolonging the effects. Ice fog results in poor visibility, to which aviation is most sensitive.

Figure 7-1: Ice Fog over Anchorage [1]

The current design of the GTL facility utilizes both water (17%) and air (83%) cooling. However, Case B's design (33,260 bbl/day) requires 896 MMBTU/h of water cooling. For perspective, a 100 MWe natural gas-fired combined cycle power plant requires approximately 1,000 MMBTU/h of cooling, making the GTL facility one of the largest point sources of ice fog within the state. To minimize ice fog it is recommended to relocate the facility well outside of city limits and away from airfields, install a closed-loop cooling system or maximize the use of air cooling.

A closed loop cooling system is comprised of a glycol-based coolant – not unlike that within an automobile. The coolant removes heat from the various processes within the plant and then passed through large radiators across which air is drawn. A closed loop system increases capital and operating costs, the latter a result of the power required for cooling fans.

7.1.3 Aviation

In addition to the consideration of ice fog for aviation, the proximity of the facility to airports and flight paths also requires consideration. The tallest structure within the facility is a 510 ft emergency flare stack, sized to accommodate the total gas flow of the facility at full load. The flare stack must be outside of imaginary spaces surrounding the air strips which limit structure heights. Of the proposed sites (FNSB-2 and Port MacKenzie) minimum safe distances from the following air strips have been confirmed:

- Fairbanks International Airport
- Fort Wainwright Airfield
- Eielson AFB Airfield
- Ted Stevens Anchorage International Airport
    - Runway 14/32
    - Runway 7R/25L
    - Runway 7R/25R
- Elmendorf AFB Airfield
    - Runway 6/24
    - Runway 16/34

With appropriate spacing and lighting of all facility structures, consideration of flare operation is also necessary. A minimum safe distance of 510 ft provides 2-3 minutes of safe exposure without protective equipment. Classifying the air space above the GTL facility as restricted shall ensure that flare radiation does not affect air craft operations. Concern of pilot distraction or passenger discomfort upon view of flaring is low such that syngas and fuel gas have low radiative properties, producing a dim flame.

7.1.4 Climate

Climate impacts supply and distribution lines; Alaskan ports require vessels capable of navigating subarctic waters or contingency for icebreakers to maintain port access. In addition, relative to a facility located in the lower forty-eight states, storage capacity should be increased to accommodate of more frequent service interruptions. It has been indicated that storage capacity of up to 45 days is required for locations dependent on rail access.

7.1.5 Inlet Currents

Currents at Port MacKenzie should not be a concern at the Knik Arm narrowing; the maximum current speeds during spring time ebb were documented by Knik Arm Bridge And Toll Authority (KABATA) at less than 4 ft/s, (Figure 7-2) and are considered navigable.

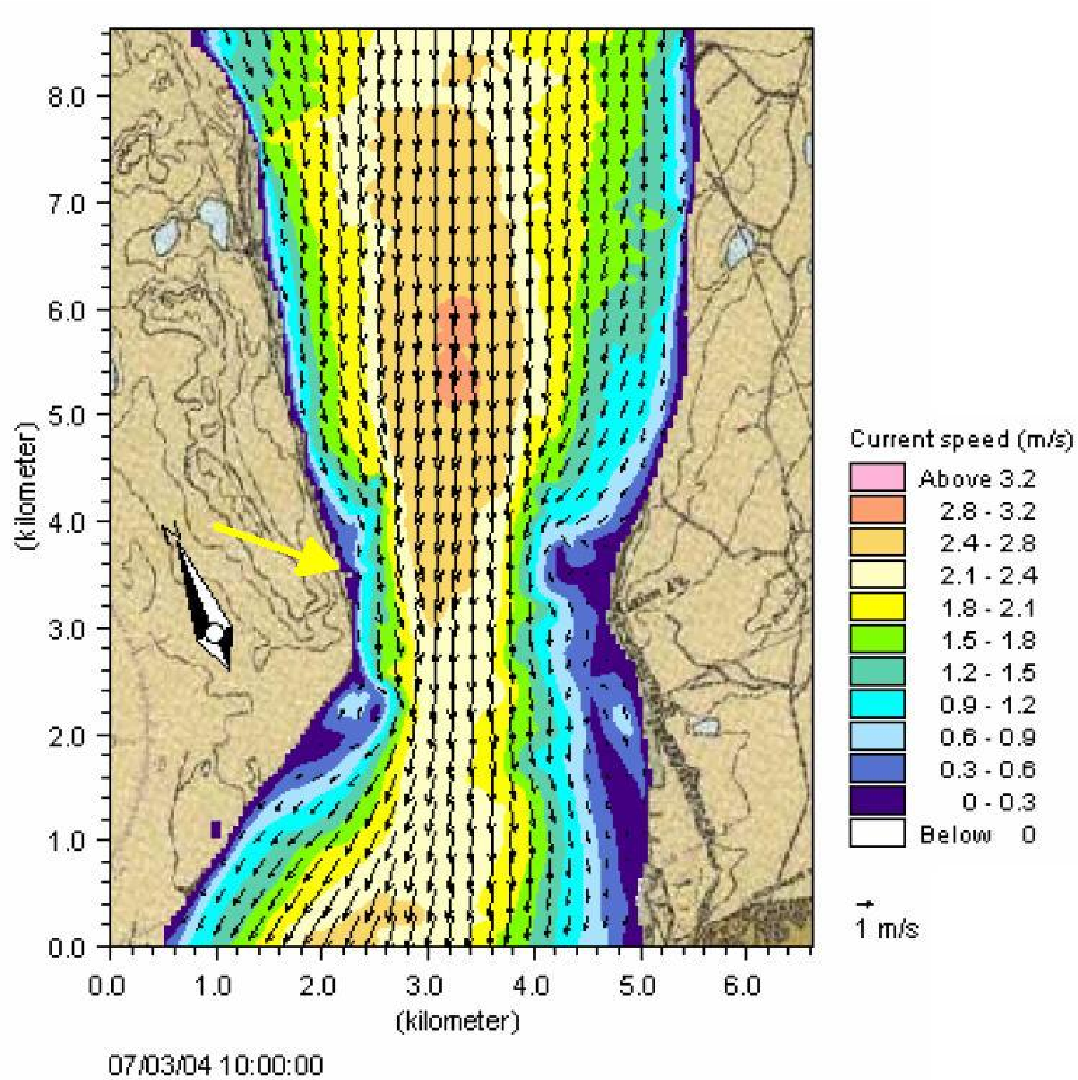


Figure 7-2: Maximum Spring Time Current Ebb at Port MacKenzie [9]

7.1.6 Seismic

The location of a heavy industrial plant, such as the GTL facility, will affect the design aspects that directly relate to the facility's superstructure. Alaska is a seismically active region and different soil conditions alter seismic design aspects. The selected site requires extensive sub-surface soil exploration to ensure the foundations and superstructure are properly designed for expected environmental design conditions.

Typical soil conditions in the Port MacKenzie region include sands and gravels at depth from the Elmendorf moraine or deposited by the Knik and Matanuska rivers. The soils are suitable for direct load bearing of foundations and superstructure; however, these soil conditions are susceptible to liquefaction during seismic events. Liquefaction is a momentary loss of load bearing capacity under seismic activity and require additional design considerations and cost to accommodate.

The Fairbanks region also has high seismic activity, however from a design perspective is less stringent than Port MacKenzie. Seismic activity in the Fairbanks region is estimated to be approximately 20% less violent than Port MacKenzie. Soils in the Fairbanks region consist of sands and gravels to a depth deposited by the Tanana River, with smaller gravels and a greater percentage of sands relative to Port MacKenzie. From a seismic perspective, it is considered easier and cheaper to construct a superstructure in Fairbanks than Port MacKenzie.

Seismic design requirements and related cost impacts are not considered in the concept or cost estimates included herein. In the future phase of study, foundation and superstructure designs will be based on the latest issue of the International Building Code.

7.1.7 Transportation & Construction

GTL facilities constructed to date have been carried out at sites with excellent transportation access. Ease of access is important in both the construction and operations period of the facility.

The transportation of large quantities of equipment and bulks is important during construction. With a limited outdoor construction season in Alaska, it is vital to ensure that materials and equipment are delivered on time in order to avoid schedule delays. Specific to GTL facilities, the single largest equipment units to be transported are the FT reactors. With internals removed to reduce shipment weight, a 17,000 bbl/day, slurry-phase FT reactor ships at roughly 2,000 tons, 200 ft long and 33 ft diameter (Figure 7-3). Transport of FT reactors by rail is therefore not possible.

Figure 7-3: Transport of one 17,000 bbl/day Slurry-Phase FT Reactor [8]

After shipping to port, reactors are driven to site under heavy haul provisions. A site with easy port access is ideal in minimizing costs and risks associated with transportation. Further discussion related to the benefits of modular construction may be found in section 7.1.8.

Fairbanks supplies the greatest transportation challenge during construction. Due to the limited marine access and distance from the coast, it has been determined – upon consultation with Crowley Marine Transport, Lynden Transport and the Alaska Department of Transportation – that it is very difficult to transport the 17,000 bbl/day reactors from the coast. Whether by land or river barge transport, low river water levels, bridges, underpasses and other obstructions pose constraints that would require modification in order to allow for the passage of these large items. However, the reactors may be engineered for transportation and assembled at site as accomplished at Sasol's Secunda facility. In Alaska, reactor segments may be transported by road, barge, river transport or over ice roads. Taking advantage of the winter conditions, large, heavy loads may be transported during the winter months, which is a common practice in Alaska. Furthermore, a greater number of smaller reactors may be specified and transported to the interior, minimizing heavy haul transport and infrastructure improvements. With the existing operation of industrial facilities in the Fairbanks region, transportation should not be considered a limiting factor for construction in Fairbanks; however, further study is required to evaluate transportation options, alternatives and costs in the next phase of study.

7.1.8 Modularization

Alaska is a well serviced region, however, geographic and climatic conditions add complexity to the execution of major capital projects. The subarctic conditions of both the Anchorage and Fairbanks areas require careful consideration regarding construction techniques.

The construction season is relatively short while labor productivity is adversely affected by temperature drop. In order to maintain schedule and minimize cost overruns, it is recommended that construction of the facility be modularized.

Modularization involves designing and engineering the facility as blocks. These blocks are fabricated in offsite yards and shipped to site for assembly.

Although transportation logistics are complicated, the time and costs of modularization are less than the alternative – stick-built – in areas with labor and material constraints. Module scale and weight require the project to have a deep draft port with heavy haul road and stable site conditions.

In Alaska, modularization would minimize productivity loss by: 1) extending the construction season as mechanical construction occurs offsite at a location not affected by winter conditions; 2) cladding of modules at site allows completion to occur locally, unencumbered by the elements year round; 3) threat of skilled labor shortage is reduced; and 4) temporary construction and camp costs are minimized.

Hatch's experience with modularized construction is proving successful. At the time of this report, Xstrata's Koniambo FeNi (ferro-nickel) smelter modules, on the island of New Caledonia in the south Pacific, are being assembled. The modules were constructed in China, transported to site and are currently being assembled. The approach was undertaken to minimize costs associated with stick-built construction. Although the scale of the modules may be intimidating, up to 90 ft x 115 ft x 140 ft and 3,500 tons, assembly is progressing well (Figure 7-4).

7.1.9 *Operations*

During operations, the ability to easily distribute product is essential, and it is not acceptable that the GTL facility be shutdown due to a lack of product storage. GTL products may be transported throughout the state by road or rail and exported via tanker. In order to export, a facility located in FNS must transport product to terminal via rail; trucks are impractical (more than 150 tankers per day). Due to the volume it is preferred that for export purposes the facility be located near a port.

Port MacKenzie, in southern Alaska, is geographically adjacent to the largest population centre within the state. Although not currently serviced by rail, there are plans to tie into the Alaskan network as part of the Port MacKenzie development plan. Additionally, Port MacKenzie is equipped with a deep draft port (60 ft). Conversely, FNSB-1 may be easily tied into the Alaskan rail network to supply the interior or transport its product to port. However, FNSB-1's marine access is the Tanana River with a 3-4 ft, resulting in limited barging capability.

Figure 7-4: Xstrata's Koniambo Module Unloading at Site [10]

7.2 Geographic Regions

The Fairbanks North Star (FNS) borough in the interior of Alaska offers potential such that it is:

- Third largest population centre in Alaska
- Well serviced by road and rail
- Marine accessible
- Sited along the Trans-Alaskan Pipeline
- Closest major city to North Slope gas fields

The Matanuska-Susitna (MS) borough was also put forth as a potential region for a GTL facility, more specifically, the area of Port MacKenzie was proposed, which offers similar and additional benefits relative to FNS, including:

- Adjacent to largest Alaskan metropolis (Anchorage)
- Easy accessible to well developed road and rail networks
- Possesses a dedicated deep draft port

The technical, logistical and environmental considerations for these two regions are compared in Table 7-1.

Table 7-1: Geographic Region Comparison

	Port MacKenzie	Fairbanks North Star
Port Access	Year-round deep draft port. Facilitates modularization of facility (see Section 7.1.8). Minimizes construction and operating expenses related to transportation.	Seasonal barging only (May-October). Only directly accessible by shallow water barge (3-4 ft draft). Potential to barge (4-5 ft draft) up Yukon River closer to Fairbanks (Tanana; Haul Road; or possibly Nenana) and truck materials/equipment to Fairbanks. Heavy haul overland transport to Fairbanks during winter only.
Road Access	Developed highway network. Road access minimally required, imports and exports primarily conducted by marine.	Developed highway network. Distance of 360 miles to nearest deep draft port. Roads and bridges not suitable for heavy haul transportation.

	Port MacKenzie	Fairbanks North Star
Rail Access	Rail spur tying into Alaskan railroad planned as part of Port MacKenzie development plan.	Well serviced with rail. Rail critical to distribution and export of liquid products throughout Alaska. Current Fairbanks Refinery (Flint Hills Resources), distributes and exports product by rail.
Modularization	Possible due to port proximity.	Impossible due to scale and weight of modules. Stick built construction required.
Labor	Imported labor required. Most populous region in Alaska. Modularization takes advantage of low-cost labor regions (e.g. China).	Imported labor required. Half the local labor force of Port MacKenzie. Increased indirect costs associated with increase of imported labor – camp required. Lower productivity due to environmental conditions (winter) Increase in labor for onsite construction, relative to Port MacKenzie.
Schedule	Modularization facilitates dual workforces – Alaska and low-cost region (e.g. China) simultaneously. Minimize impact of winter construction due to modularization.	Extended schedule due to stick built construction. Winter conditions reduce productivity.
Capital Cost	Considered base case.	Increases due to: • Stick built • Increased indirects • Prolonged schedule • Lower productivity • Lack of low-cost labor region See section 9.

	Port MacKenzie	Fairbanks North Star
Environmental	Potential for CO_2 sequestration for enhanced oil recovery or saline acquires (see section 7.4).	Temperature inversions cause ice fog. EPA PM2.5 limits stack dispersions (see section 7.1.1). CO_2 sequestration and enhanced oil recovery research undeveloped at present (see section 7.4).
Geotechnical / Seismic	Port MacKenzie impact of low-value wetlands to be determined. Soil liquefaction potential during seismic event; increased superstructure and foundation design requirements.	Challenging terrain. Soil liquefaction potential during seismic event, increased superstructure and foundation design requirements; seismic potential (magnitude) estimated to be 20% less violent than Port MacKenzie.
Set-back	Sufficiently set-back from obstructing flight paths and restricted military zones.	Sufficiently set-back from flight path obstruction and restricted military zones.

7.3 Individual Sites

7.3.1 Fairbanks North Star (FNS)

Identification and evaluation of three potential sites within the FNS borough was conducted. The three identified sites included:

1. FNSB-1 (Old Richardson Highway)
2. FNSB-2 (Badger Road)
3. FNSB-3 (Bethany Street)

This list should not be considered exhaustive and it should be noted that other potential sites may exist, requiring further screening in the next phase of study. However, at this level of study, with the tools available, the three identified sites illustrated in Figure 7-5 met the majority of the screening criteria.

Figure 7-5: Potential Sites in Fairbanks North Star Borough [Google Earth, 2010]

7.3.1.1 FNSB-1: Old Richardson Highway

FNSB-1 consists of five (5) lots, all owned by the Bureau of Land Management and reside outside the city boundary of North Pole. The site is next to Flint Hills Resources' North Pole Refinery. The refinery's processing capacity is 220,000 bbl/day and produces gasoline, jet fuel, heating oil, diesel, gasoil and asphalt for the Alaskan market – 60% destined for the aviation market. FNSB-1 offers many advantages as a result of its proximity to the existing Flint Hills Resources' refinery: 1) zoning; 2) supply and distribution access; 3) access to utilities.

Unfortunately, the lot available is too near residential homes to the northeast of Old Richardson Highway (300 ft). Due to the size of the facilities, it is unlikely that FNSB-1 be approved for development.

7.3.1.2 FNSB-2: Badger Road

North of Interstate A2 at Rentals Street, a 619 acre lot, owned by the FNS Borough offers a potential site. The facility is zoned as General Use (GU-1), accommodating to petrochemical and petroleum refineries. A conditional use permit is required for the development of a GTL facility. Interstate A2 provides road and a rail access to the southwest boundary of the lot. Adequate separation between the plant and the residential communities are observed (2,000 ft), Figure 7-6.

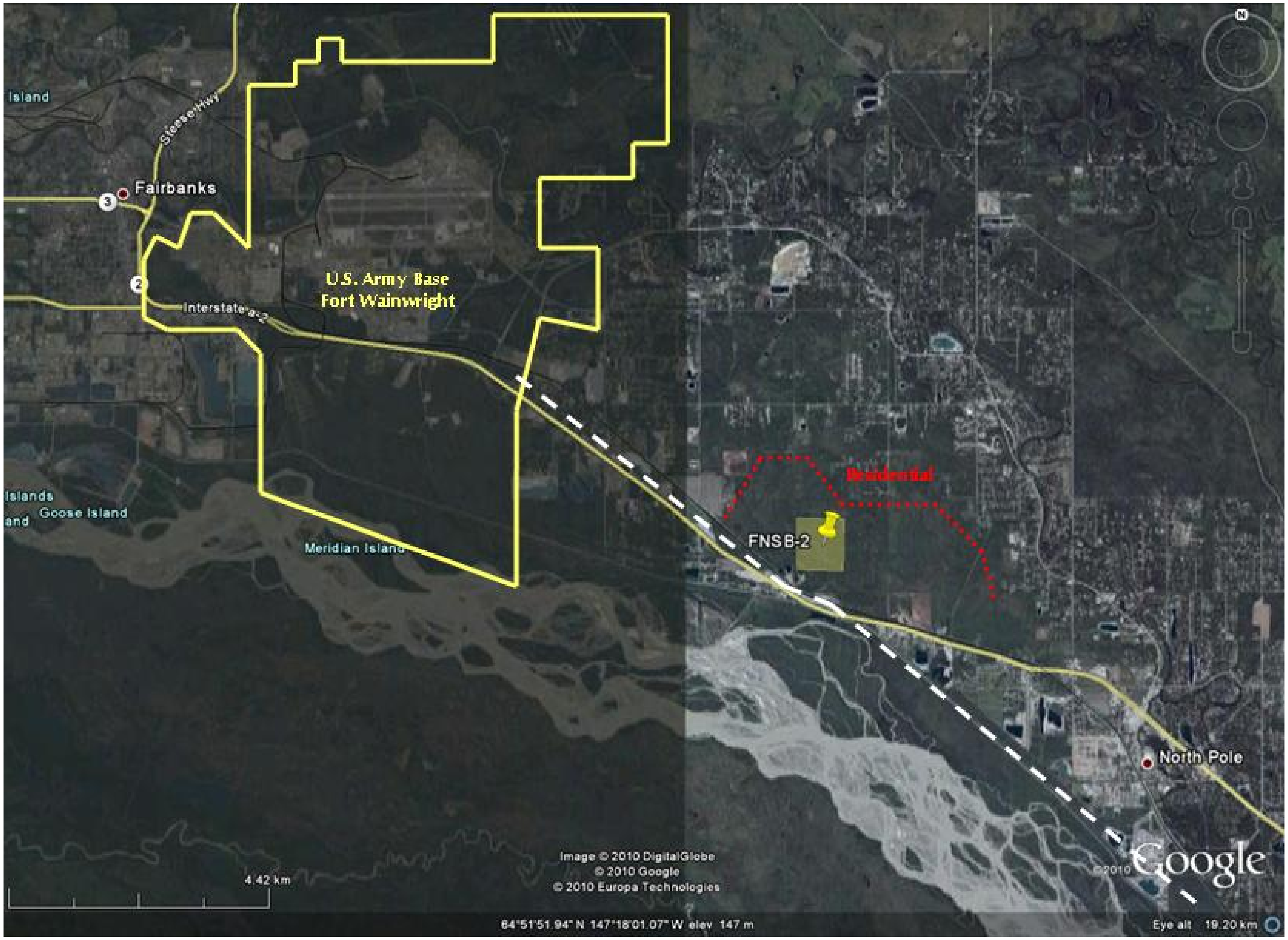


Figure 7-6: FNSB-2 Relative to U.S. Army Base Fort Wainwright and Rail Line [Google Earth, 2010]

7.3.1.3 FNSB-3: Bethany Street

Accessed by Bethany Street, southwest of Interstate A2, a 585 acre lot owned by the University of Alaska is zoned for GU-1. Although sufficiently removed from residential communities and easily accessible by road, FNSB-3 is divided by an east-west levee, which is part of the flood protection system for the Tanana River, limiting the potential for future plant expansion.

7.3.2 Matanuska-Susitna

Along the coast, across from the city of Anchorage, lies the Matanuska-Susitna (MS) borough. The MS borough was identified as a potential area for locating a GTL facility as it is the planned termination point of the natural gas pipeline.

Additionally, MS borough provides retreat from the populated region of Anchorage, across the water to the southeast. Upon review of MS borough's land ownership maps and review of the Port MacKenzie Master Plan Update, a suitable lot was identified at Port MacKenzie (Figure 7-7).

Figure 7-7: Port MacKenzie (Photo courtesy of Port MacKenzie and Alaska Aerial Technologies)

7.3.2.1 Port MacKenzie

A master development plan previously identified an area for a gas derived products facility within PID-I (Port Industrial District I). PID-I seeks to preserve and protect the coastal resources by limiting uses to strictly marine/rail-oriented industrial operations. Regrettably, aerial and topographic maps indicate that the identified site within the Master Plan for the gas derived facility is within a wetland.

At the time of this report, construction costs of the Escravos GTL facility in Nigeria have escalated more than expected from previous estimates. It is believed that siting within a wetland has contributed substantially to the cost and should be mitigated in future development plans. As a result, Hatch proposes another site in the northwest corner of PID-II, shown Figure 7-8, planned to house commercial and industrial operations not requiring close proximity to the ports or railroad. The identified site may provide good drainage, soil conditions, access (rail, road and marine) and sufficient retreat from neighbouring planned and existing commercial and residential facilities. However, further evaluation is required at the next phase of study.

Figure 7-8: Matanuska-Susitna Potential Site Identification for GTL Facility [Google Earth, 2010]

7.4 Considerations for CO_2 Sequestration

The GTL facility captures CO_2 from fuel gas prior to combustion to produce a pure CO_2 stream available for export. Case A (1 FT train) facility captures 38 st/hr, Case B (2 FT trains) captures 77 st/hr and Case C (4 FT trains) captures 153 st/hr, equivalent to approximately 9 million, 18.5 million and 37 million short tons of CO_2 respectively over the 30 year economic life of the facility. The GTL plant also includes a CO_2 compressor in order to compress the gas to a pressure of 2,000 psi, suitable for pipeline transport. The following sections describe potential end-users for this CO_2 in both Cook Inlet and Fairbanks plant locations.

7.4.1 Cook Inlet

Due to the large amount of oil and gas activity and industrial CO_2 sources in the Cook Inlet area, it is one of the few regions in Alaska where subsurface characteristics have been studied for their ability to sequester CO_2. Potential end-use options for CO_2 produced by the GTL facility in Cook Inlet include enhanced oil recovery (EOR), as well as sequestration in either depleted oil and gas reservoirs or saline aquifers.

7.4.1.1 Enhanced Oil Recovery (EOR)

EOR with CO_2 is a tertiary method of recovering oil from partially depleted oil reservoirs, after primary (removing easily recovered oil) and secondary (water flooding) methods have been applied. During tertiary recovery, CO_2 is injected into the reservoir increasing pressure and improving fluid flow, incrementally increasing the recovery of oil.

EOR candidate reservoirs must meet a number of screening criteria, including parameters such as appropriate depth, pressure and temperature, as well as oil gravity and oil pressure. If these conditions are met, then it must be determined if the appropriate technique would be miscible CO_2 flooding, where CO_2 dissolves into the oil and reduces oil viscosity, or immiscible flooding, where CO_2 remains a separate state and simply forces the remaining oil out.

EOR using CO_2 is a mature technology and has been applied in the U.S. since the 1970's, extending the life of oil fields by many years. In 2009 over 3,500 miles of CO_2 pipeline [18] were used to transport the gas to fields throughout the U.S., producing over 250,000 bbl/day of incremental oil from CO_2 EOR [19].

The potential of EOR is proven by EnCana's Weyburn Oil Field in Saskatchewan, Canada. Beginning in 2000, CO_2 captured from the Great Plains Synfuels Plant in North Dakota (a coal gasification facility producing synthetic natural gas) has been transported 205 miles by pipeline to the Weyburn oil field. There CO_2 is injected into depleted wells to increase production through tertiary EOR. The use of CO_2 EOR is estimated to have extended the life of the reservoir by 25 years and is expected to recover 155 million barrels of incremental oil by 2035 [20].

The Cook Inlet is the second largest oil producing region in Alaska, and locations of the major oil fields are shown on Figure 10-1. These fields range from 50 to 75 miles from Port MacKenzie.

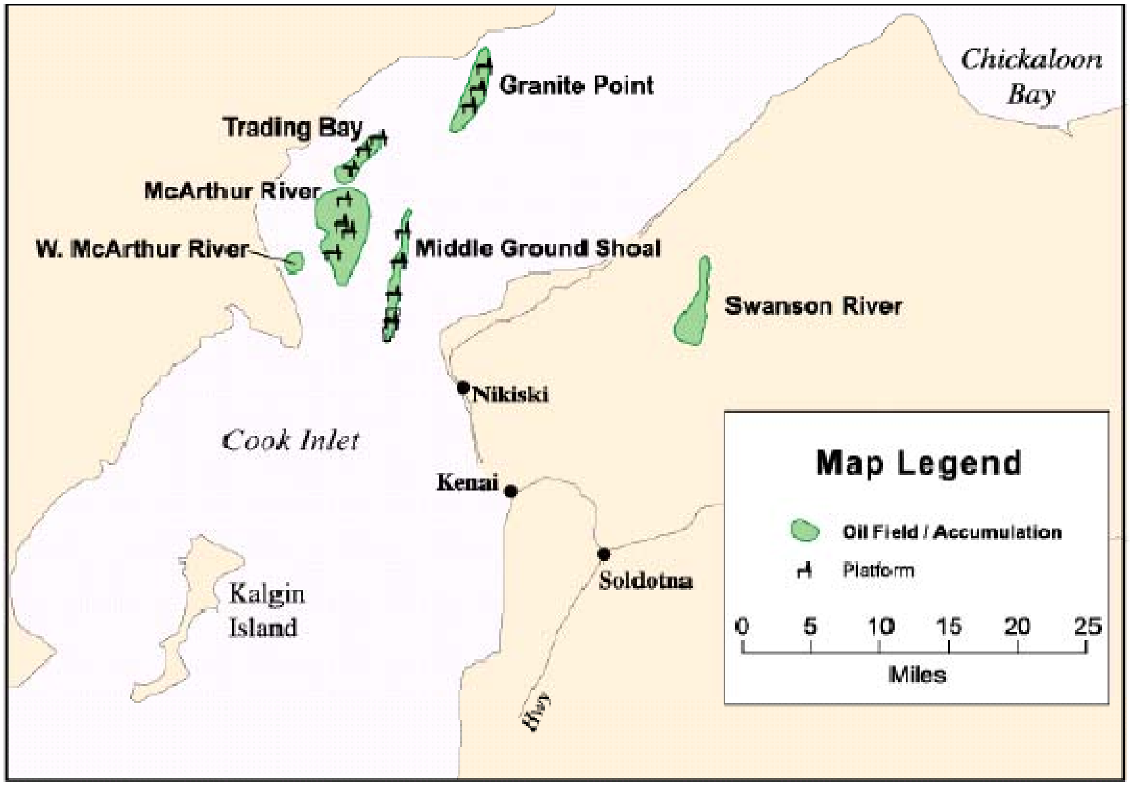


Figure 7-9: Locations of the major oil fields in Alaska [14]

Oil production began in the region in 1959, but has been declining since the 1970's. Historic and projected oil production rates are displayed in Figure 7-10 [15]. The majority of the fields in the region have produced greater than 95% of the estimated ultimately recoverable (EUR) oil, meaning that without any further development they will soon be depleted leaving almost two-thirds of the original-oil-in-place (OOIP) stranded.

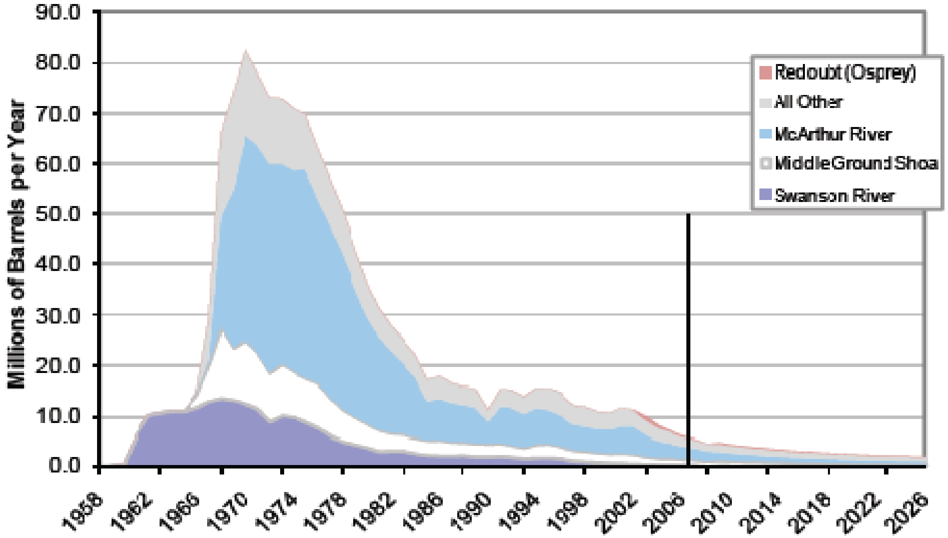


Figure 7-10: Historic and projected oil production for Alaska [15]

In 2005, Advanced Resources International Inc. produced a study for the U.S. Department of Energy (DOE) investigating CO_2 EOR potential in Alaska, which concluded that twelve reservoirs in the Cook Inlet region are technically favourable for miscible floods. Assuming the average range of incremental increase in production of 8% to 11% of original-oil-in-place (OOIP) for miscible floods, the Cook Inlet has potential to produce an incremental 290 to 400 million barrels of oil. In order to reach incremental production of this magnitude, consistent virgin CO_2 supply combined with recycled CO_2 (produced during miscible flood, separated from oil and re-injected) is required. Between 1,200 and 1,700 BCF of CO_2, equivalent to approximately 57 to 81 million short tons, of purchased CO_2 would be required.

As mentioned, Case C produces approximately 37 million short tons of CO_2 throughout the 30-year lifetime of the plant, and therefore EOR operations in the Cook Inlet have the capability of sequestering all captured CO_2 from the GTL facility (one, two or four trains) for the entire lifetime of the plant. This would, in turn, extend the life of these oil fields by up to an additional twenty years.

Although technically feasible based on the screening exercises summarized above, there are many critical factors that require further investigation at the next phase of study to determine if EOR operations of this magnitude are possible in the region. In addition to the required construction of a dedicated CO_2 pipeline from the GTL facility to Cook Inlet reservoirs, current facilities also require updated oil separation equipment in order to recover CO_2 contained in the oil, and compress it to a pressure suitable for re-injection. Due to the fact the majority of production in the Cook Inlet occurs on off-shore platforms, cost of this infrastructure would be greatly increased compared to onshore operations.

There are currently no offshore CO_2 EOR facilities in the world, although secondary recovery techniques are being employed in the North Sea and tertiary recovery operations are being investigated [21]. The major onshore field in the Cook Inlet is Swanson River, as shown in Figure 7-9. If CO_2 EOR was applied to this field alone it could store 8 to 11 million short tons of CO_2, producing 40 to 55 million barrels of incremental oil, providing enough storage for the entire lifetime of the Case A facility, but additional end-users would be required for Cases B and C.

With high oil prices, the incremental oil production from CO_2 EOR could provide a credit for the GTL facility, but further investigations into costs of additional infrastructure and willingness of plant operators to extend the lifetime of aging production facilities is necessary. However, even with uncertainty regarding these issues, CO_2 EOR is still the preferred end-user for CO_2 produced by the GTL facility.

7.4.1.2 Sequestration

Two potential options for CO_2 sequestration in the Cook Inlet region include injection into depleted gas reservoirs and injection into saline aquifers. The burden of costs for either option lies with the GTL facility and therefore taxation on CO_2 emissions would likely be required before pursuing these options.

Requirements for both cases of CO_2 injection for underground storage include the following geological formations:

- Porous and permeable reservoir rocks to provide sufficient storage volume for CO_2 and allow injected CO_2 to move away from injection point;
- Sealing rocks with no porosity or permeability to provide barriers preventing CO_2 from leaking out of the reservoir rock; and
- Trap geometric configuration of the reservoir rock and seal to ensure CO_2 remains isolated in one place.

7.4.1.2.1 Depleted Gas Reservoirs

Injection into depleted gas reservoirs is the most attractive option for two main reasons: 1) the integrity of the formation is more well known, as the same permeable reservoir rock has stored natural gas for thousands of years; by injecting CO_2 into these formations there is greater confidence that it will be permanently sealed into the formation; and 2) the majority of required infrastructure is already in place and may be re-used for CO_2 injection, providing capital investment savings.

In the Cook Inlet region, there are a number of large gas reservoirs within proximity to the proposed site, but all are currently producing gas and are expected to continue producing gas in the short-to-medium term future. For this reason it is unlikely that depleted gas reservoirs can be utilized for CO_2 sequestration. Further investigation into the expected production timeline and co-operation of field owners for this option will be conducted in future phases of study.

7.4.1.2.2 Saline Aquifers

Saline aquifers are sedimentary formations with no economic value that contain high salinity water. These formations are estimated to have the largest CO_2 storage potential throughout the U.S. Injection into saline aquifers would require development of all necessary infrastructure including distribution pipelines, injection wells and supporting facilities.

In the Cook Inlet region, saline aquifer formations underlie all of the Kenai Peninsula at shallow depths between approximately 3,000 to 5,000 feet. Further details regarding the formation characteristics are required prior to making any definite conclusions, but it is likely that aquifers with sufficient storage volume for all of the CO_2 produced by the GTL facility could be located in the Cook Inlet region.

7.4.2 Fairbanks

For interior Alaska, information regarding CO_2 storage resources is limited due to harsh working environments and lack of local industrial CO_2 sources. Most of the research to date has been focused on Cook Inlet and North Slope regions which have large sources of industrial CO_2 and extensive characterization data from oil and gas exploration [17].

Exporting CO_2 to either Cook Inlet or North Slope regions for EOR operations is impractical due to high costs associated with constructing a dedicated CO_2 pipeline to either site from Fairbanks. There are no depleted oil and gas reserves in the Fairbanks region for sequestration of CO_2, and although Fairbanks does have various sedimentary formations nearby, deep saline aquifers with necessary sealing characteristics have not been identified and may not exist.

The primary option for CO_2 sequestration in the Fairbanks region involves injecting CO_2 into deep unmineable coal seams near the Healy coalfields, approximately 115 miles southwest of Fairbanks. Coal has the affinity to adsorb many gases, and gaseous CO_2 injected will displace gases with lower affinity to adsorption, such as methane. Although potential for CO_2 sequestration in coal seams is large, the process for CO_2 trapping is still not well understood and various pilot projects around the world are investigating the technology [22]. In addition to sequestration potential in deep coal seams, by displacing methane, CO_2 injection has the potential of enhanced coal bed methane (ECBM) production, but the Healy coalfields currently do not have any ECBM facilities so this would likely be impractical.

Similar to sequestration methods discussed above, costs for CO_2 pipeline, injection wells and associated facilities would be the responsibility of the GTL plant owner.

8. Project Execution Schedule

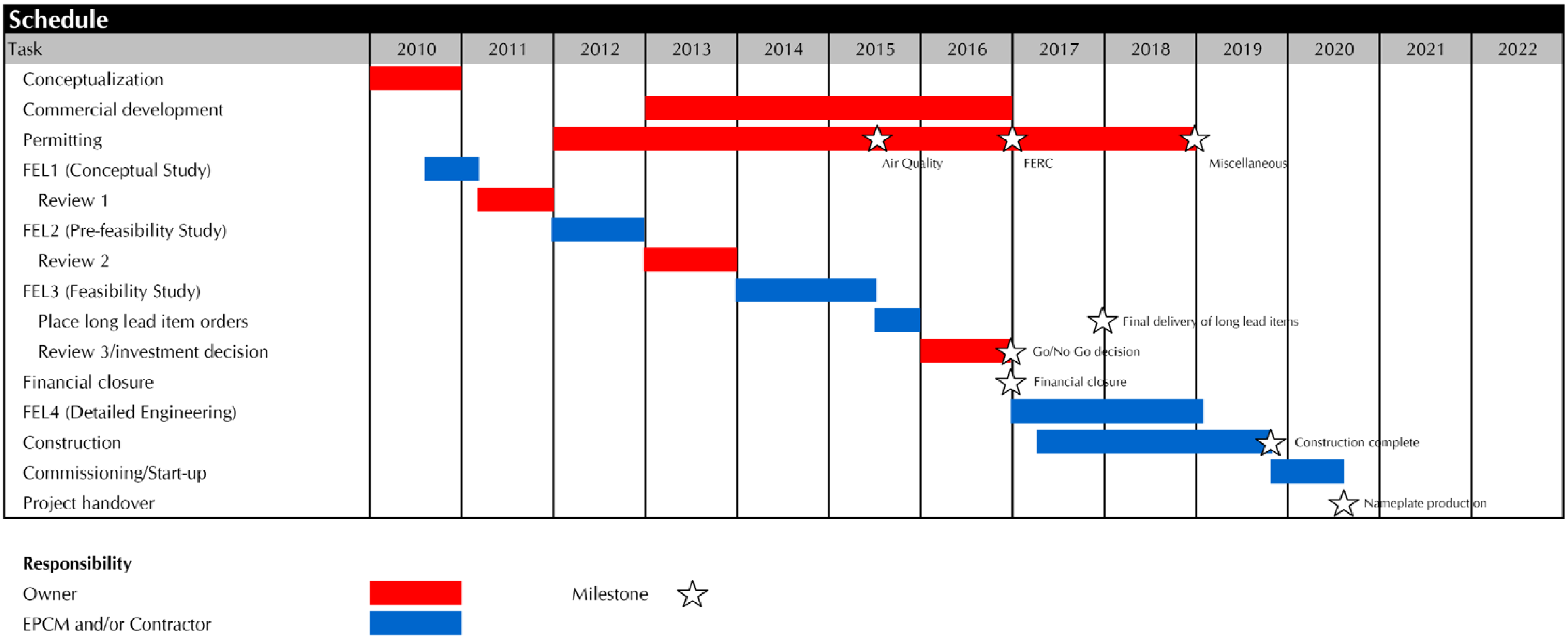
The schedule assumes an increasing level of detail per FEL (Front End Loading) phase. The phased approach ensures no plant areas, systems, procedures or other items/tasks are omitted from the design. This phased approach has proven successful in the development of capital projects across a wide variety of industries, including: mining, metals, infrastructure, energy and others.

A projected schedule, Table 8-1, outlines the major steps for the Case B facility (33,260 bbl/day). The schedule assumes an immediate notice to commence without major delays – attributable, but not limited to:

- Access to technology
- Permitting
- Financing
- Equipment fabrication
- Labor sourcing
- Construction
- Commissioning
- Start-up

The schedule is based on GTL projects previously completed or currently under construction. In order to accommodate for the shortened transportation and construction season, the schedule assumes modular construction, as discussed in section 7.1.8.

Table 8-1: Project Schedule for Southern Alaskan GTL Facility



With FEL1 complete, it is expected that FEL2 (pre-feasibility study) and appropriate owner's review to span Q2-2011 through Q4-2012. Upon owner's satisfaction, order to proceed to FEL3 (feasibility study/basic engineering) is anticipated for Q4-2013. The FEL3 owner's review is scheduled for 2016, with authorization to place long-lead item orders (most critical are the FT reactors). The long-lead item orders are placed on condition of successful board approval from the owner. It is assumed that such approval requires appropriate financial closure for the project.

The period required for financial closure is assumed to be in line with the schedule, however, it must be noted that financing for capital projects of this scale requires considerable time to confirm. Prior to construction, the greatest risk to schedule delay is a lack of financial closure.

With financing confirmed, FEL4 (detailed engineering) immediately follows FEL3 – Q4-2016. FEL4 is expected to take two years with construction – site preparation – starting in Q2-2017. The overlap is required in order to meet the schedule and accommodate the shortened construction season in Alaska and made possible by the modular construction approach.

Mechanical completion of the facility is planned for Q4-2019. Commissioning and start-up are expected to be completed in Q3-2020 and project handover upon attainment of nameplate production capacity.

The projected periods are based on projects of similar type and scale, notably Oryx, Pearl and Escravos GTL facilities. Although each facility had differing timelines, the projections for development and construction were similar. Schedule contraction relative to existing projects was assumed in the commissioning and start-up phase due to the expected knowledge gained by technology licensors in existing and under development projects.

9. GTL Capital Cost Estimate (CAPEX)

A capital cost estimate was prepared for Case B (33,260 bbl/day) assuming a fully integrated gas-to-liquids facility based in southern Alaska. The basis for the estimate includes:

Project Type:	Greenfield gas-to-liquids facility
Plant Capacity:	33,260 bbl/day of diesel and naphtha based fuels
Location:	Southern Alaska
Construction Type:	Modular
Currency:	USD
Estimate Type:	Order-of-Magnitude
Estimate Class:	AACE Level 4
Level of Engineering:	1% complete
Contingency:	30% (minor variances per plant)

Accuracy: -30%, +40%

Location Factor: 1.25

The fully integrated facility includes the plant areas as listed in Table 4-4. Each plant area's disciplines have been scaled from respective delivered direct equipment costs and include:

- Erection
- Piping
- Structural steel
- Electrical
- Instrumentation
- Island services
- Excavation & site preparation
- Auxiliaries
- Field expense
- Engineering
- Contractor's fee, overhead, profit
- Contingency

In addition, auxiliaries for each unit include the necessary infrastructure and offsites required:

- Process flare
- Maintenance shop & stores
- Emergency response
- Storage yard
- Product export
- Security
- Administration

There are also other costs captured, but not explicitly broken out within the estimate which include:

- Freight
- Spares
- Commissioning
- Temporary construction facilities
- Camp

Exclusions from the estimate – although some have been captured within the economic model, include:

- Land
- Infrastructure outside battery limit
- Labor transportation
- License fees
- Owner's management costs
- Working capital
- Insurance
- Permits
- Legal fees
- Financing fees
- Taxes, duties, brokerage fees
- Escalation
- Foreign exchange
- Other statutory charges

The methodology associated with an AACE Class 4 estimate is appropriate for a conceptual study and for identification of requirements necessary for the next phase of study. The estimate includes the use of equipment quotations, factored equipment costs and parametric models.

Estimates for Case A and C were factored from Case B on a per train basis. It was assumed that a savings of 15% (based on Hatch experience) is common on subsequent trains. The resulting savings per unit of production from Case A (single-train) to Case B (double-train) and to Case C (quadruple-train) are 7.5% and 11.3%, respectively. Savings are attributable due to duplication of design; however, the majority of the costs are associated with procurement and construction tasks. The estimates for Cases A, B and C are presented in Table 9-1.

Table 9-1: Capital Cost Estimate for GTL Facility in Port MacKenzie

Capital Estimate Summary [USD]		Case A [16,630 bbl/day]	Case B [33,260 bbl/day]	Case C [66,520 bbl/day]
No.	**Plant Area**			
0100	Natural Gas Compression, Purification & Conditioning	26,182,000	48,436,000	92,946,000
0200	Air Separation Unit	209,321,000	387,244,000	743,090,000
0300	Autothermal Reforming	162,128,000	299,936,000	575,554,000
0400	Fischer-Tropsch Synthesis*	619,290,000	1,145,687,000	2,198,480,000
0500	Hydrogen Plant	37,466,000	69,313,000	133,004,000
0600	Product Upgrading	included in area 0400 - FT	included in area 0400 - FT	included in area 0400 - FT
0700	CO2 Capture, Dehydration & Compression	99,570,000	184,205,000	353,474,000
0800	Boiler Feed Water & Steam System	91,175,000	168,673,000	323,671,000
0900	Power Generation System	180,615,000	334,138,000	641,183,000
1000	Water Management System	112,384,000	207,911,000	398,963,000
1100	Fuel Gas System	3,548,000	6,563,000	12,595,000
1200	Cooling Water System	9,451,000	17,485,000	33,551,000
1300	Tankage	31,034,000	57,412,000	110,171,000
	Fixed Capital Investment‡	1,582,164,000	2,927,003,000	5,616,682,000
	Unit Cost [USD/(bbl/day)]	95,100	88,000	84,400

* FT Synthesis includes: FT Synthesis, Hydrogen Plant, and Upgrading

‡ FCI includes indirect costs: camp, temporary construction facilities, freight, spares, commissioning

The AGDC estimate has been benchmarked against previously constructed (or in construction) GTL facility estimates – escalated to the current southern Alaskan market (using CEPCI index and location factor) – see Appendix D for additional details.

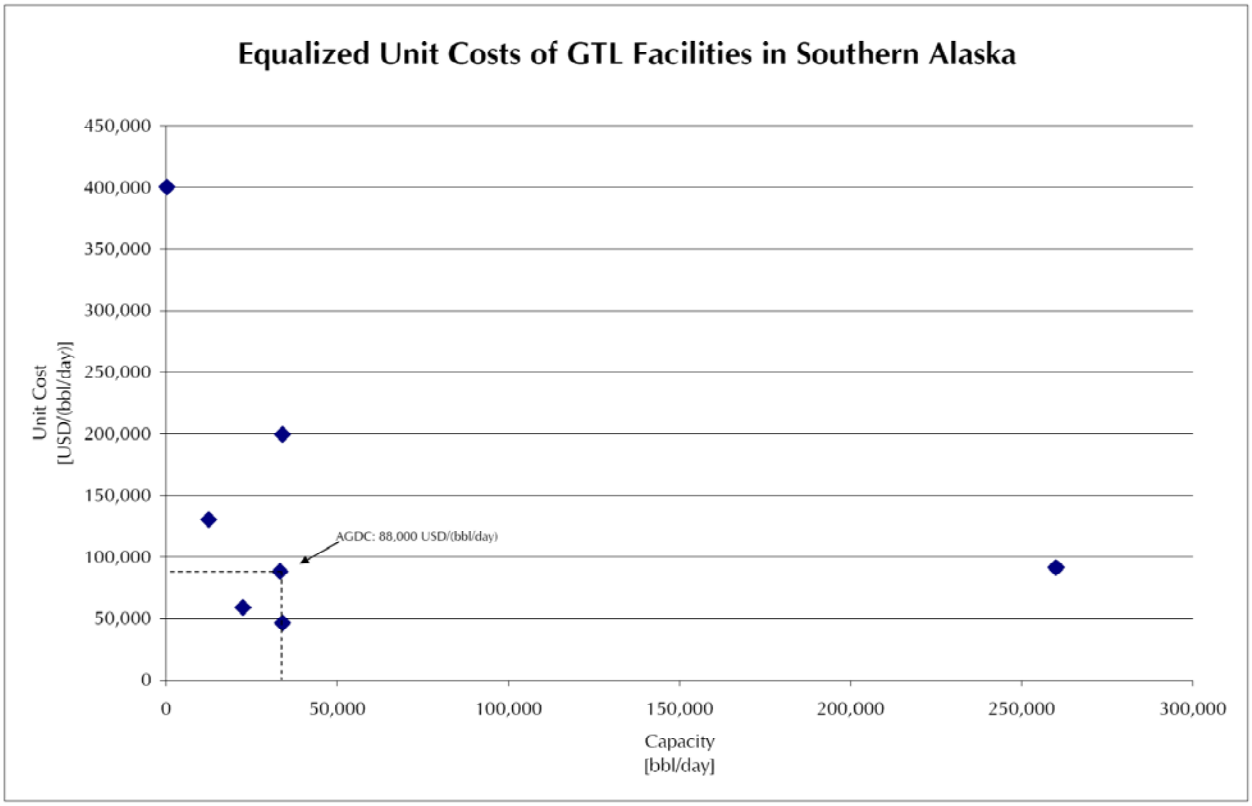


Figure 9-1: Benchmarked Southern Alaska GTL Capital Cost Estimates

10. GTL Operating Cost Estimate (OPEX)

Conceptual level operating cost estimates for the Case A (1 FT train); B (2 FT trains) and C (4 FT trains) GTL facilities are summarized in Appendix E. The analysis is presented in 2010 USD, and for calculation purposes the OPEX considers the Port MacKenzie plant location. Operating cost variation between Port MacKenzie and Fairbanks is discussed in section 11.2.

10.1 Variable Costs

Variable operating costs consist of estimates for raw materials and utilities, as described below.

10.1.1 Raw Materials

The primary raw material for the GTL facility is natural gas. Case B consumes 309 MMSCFD, which is equivalent to 13,350 MMBTU/hr. The delivered price of natural gas to the GTL plant inlet at Port MacKenzie for Cases A, B, and C are given by AGDC to be 8.75, 7.61, and 6.88 USD/MMBTU respectively. The differences are due to economies of scale of the natural gas pipeline for the larger facilities, which is reflected in the transportation tariff portion of these values. This tariff represents a major component of the gas cost from the North Slope to the GTL facility; therefore AGDC assumes a GTL facility in Fairbanks will have delivered natural gas prices of 7.25, 6.15, and 5.74 USD/MMBTU for Cases A, B and C respectively.

Other raw materials include chemicals and catalysts, which are required by process units such as the autothermal reactors (ATR), FT synthesis reactors and water treatment. Annual costs for standard chemicals and catalysts (reforming catalyst and water treatment chemicals) are estimated based on discussions with various suppliers and Hatch in-house data. For specialty items such as the cobalt-based FT catalyst, the consumption rate and cost of the catalyst are also estimated using Hatch's in-house data and experience. The annual cost for the cobalt-based FT catalyst is estimated to be approximately 70% of the total chemical and catalyst costs. Confirmation from the selected technology provider is required during future phases of study.

10.1.2 Utilities

As described in section 5.4.1, the GTL facility can export approximately 60 MWe, 119 MWe, and 239 MWe of power for Cases A, B, and C, respectively. Based on discussions with the Matanuska Electric Association (MEA), Chugach Electric Association (CEA) and Golden Valley Electric Association (GVEA), it is estimated that the GTL facility would received a credit of 45 USD/MWh for power exported to the grid in the Port MacKenzie region.

The GVEA estimated that in the Fairbanks region a likely estimate of 60 USD/MWh could be received for power export to the grid. This value is lower than current electricity rates due to the fact that power could be produced at a lower cost with natural gas compared to the current generation fleet, primarily based on coal and oil. Natural gas is currently the primary source of power generation in the Cook Inlet region, and the MEA and CEA confirmed that a wide range of power credits could be available for the GTL facility based primarily on available natural gas price. Historical data shows that the electricity rates in the Cook Inlet are typically about 25% below those of the GVEA; hence a 45 USD/MWh credit was estimated, which falls on the conservative end of the

45-64 USD/MWh range provided (under today's conditions). Details regarding the ability of local grids to handle this power export are to be investigated during future phases of engineering

The other major plant utility is water. Costs associated with water treatment include disposal of effluent from the water reclamation plant, and disposal of waste activated sludge from the biological treatment processes. The water reclamation plant generates a brine stream, which the local utility in Anchorage would likely accept at their wastewater treatment facility for a surcharge of about 5.04 USD/gallon. Waste biosolids from the biological treatment processes are dewatered and disposed of at the local landfill. In Anchorage, the local landfills accept this material for a surcharge of 103 USD/st. The calculated water consumption may be reduced by implementing more capital intensive water treatment configurations. The viability of such systems needs to be investigated at a more detailed phase of the study.

The water treatment facility within the GTL plant is described in section 4.4.4 and only requires a source of groundwater to supply the entire plant water demand. There may be an opportunity to purchase potable water directly from the local publicly owned treatment works. Depending on the plant location, the cost benefits of this should be considered during future stages of engineering, along with the option of drawing all plant water requirements from a publicly owned facility.

10.2 Fixed Costs

Fixed operating costs consist of estimates for operations, maintenance, SG&A (selling, general and administration), and insurance.

10.2.1 Operations

An estimated 260 staff, split in four shifts, are required to maintain operations of the Case B GTL facility. This estimate includes managers/supervisors, plant operators, area specialists, and contracted workers. Average annual salaries for each of these positions were estimated from the Alaska Department of Labor and Workforce Development Occupational Tables, September 2010 [20]. Case A and C require an estimated 160 and 345 personnel, respectively.

Operations expense for each case also includes one general manager for the plant and an allowance for direct overheads, estimated at 75% of personnel cost.

10.2.2 Maintenance

Maintenance costs, including parts and labor, are estimated for each major process area using a percentage factor of area fixed capital investment. The individual factors are presented in Appendix E, but overall maintenance costs total approximately 2% of the fixed capital investment for the GTL facility.

10.2.3 SG&A and Insurance

Selling, general and administrative costs are estimated at 1.0% of plant revenue annually. To present these costs in 2010 USD for the operating costs summary tables, the 1.0% factor is applied to the plant revenue during the first full year of nameplate operating capacity (2021), and discounted back to 2010 using the assumed inflation rate of 3%.

Insurance costs are also a factored estimated, 0.5% of fixed capital investment for the GTL facility.

10.3 Operating Cost Summary

A summary of GTL operating costs for Case B is presented in Table 10-1. Total operating costs are estimated to be approximately 925 Million USD, equivalent to 83.45 USD/bbl of facility production. Detailed breakdown of the operating cost estimate is presented in Appendix E.

Table 10-1: Case B GTL Operating Costs for Port MacKenzie

Operating Costs	Annual Cost [USD]	Unit Cost [USD/bbl]
Variable Costs		
Natural Gas	812,748,000	73.31
Chemicals & Catalysts	36,070,000	3.25
Utilities	(39,355,000)	(3.55)
Total Variable Costs	**809,463,000**	**73.01**
Fixed Costs		
Operations	34,563,000	3.11
Maintenance	56,161,000	5.07
SG&A and Insurance	25,009,000	2.26
Total Fixed Costs	**115,733,000**	**10.44**
TOTAL OPERATING COSTS	**925,196,000**	**83.45**

Using the assumptions described above, natural gas accounts for about 90% of the GTL facility operating costs.

A comparison of costs between Cases A, B and C for the Port MacKenzie location is presented in Table 10-2. Variable expenses are directly related to plant production, and therefore are equal, at this level of accuracy, for all cases on a normalized USD/bbl produced basis. The exception to this is natural gas, which benefits from economies of scale of the pipeline reducing delivered prices to larger plant sizes. Economies of scale are also shown by the differences in fixed operating costs. This results in Case C having the lowest normalized operating costs at approximately 75 USD/bbl.

Table 10-2: Operating Cost Comparison Between Cases A, B and C for Port MacKenzie location

Operating Cost	Unit Cost [USD/bbl]		
	Case A (1 Train)	Case B (2 Trains)	Case C (4 Trains)
Variable Costs			
Natural Gas	84.30	73.31	66.28
Chemicals & Catalysts	3.25	3.25	3.25
Utilities	(3.55)	(3.55)	(3.55)
Total Variable Costs	**84.00**	**73.01**	**65.98**
Fixed Costs			
Operations	3.74	3.11	2.09
Maintenance	5.48	5.07	4.86
SG&A and Insurance	2.36	2.26	2.20
Total Fixed Costs	**11.58**	**10.44**	**9.15**
TOTAL OPERATING COSTS [USD/bbl]	**95.58**	**83.45**	**75.13**
TOTAL ANNUAL OPERATING COSTS [Million USD]	**530**	**925**	**1,666**

11. Cost Variances: Fairbanks vs. Port MacKenzie

11.1 Capital Estimate

As previously discussed, construction and operations are dependent on plant siting, as are their respective costs. Due to the import of major and minor equipment, along with the majority of bulks required for siting at Fairbanks, there are increased costs associated with construction. As previously discussed in section 7 and 7.1.8, access to a deep draft port is important and many of the incremental costs from the base case estimate (Port MacKenzie) to Fairbanks are associated with no such access.

With the base case considered, additional costs associated with construction in the Fairbanks region are expected as a result of:

- Stick-built
- Lower labor productivity
- Additional infrastructure (port, roads)
- Prolonged schedule
- Increased indirects
- Increase labor rates (no low-cost region)

For this study, it is estimated that the 'cost adder' to the base case (Port MacKenzie) CAPEX associated for Fairbanks is 1.25 (accuracy -30%/+40%) times the total Port MacKenzie CAPEX. Further engineering is required in order to quantify the above and other impacts omitted from the current study, in order to increase the level of accuracy of the cost estimate. Further discussion as to the impact of this capital increase is discussed in section 12.3.

11.2 Operating Cost Variance

In addition to capital cost variances between the two proposed plant locations, operating costs will also vary. Overall, operating cost variances between the proposed Port MacKenzie and Fairbanks locations are dictated by the natural gas price in each region, but the netback FT product price and power credit in Fairbanks will also have an impact on plant economics. These variances are described in subsequent sections.

11.2.1 Natural Gas

Natural gas accounts for the majority of operating costs for the facility, and with the plant located approximately 350 miles closer to the natural gas source (North Slope) in Fairbanks this cost is reduced. Based on the recommendation from AGDC, the cost of natural gas delivered to the plant inlet at Fairbanks for Case B is assumed to be 6.15USD/MMBTU. This results in an annual operating cost of 656,820,000 USD, which represents 155,928,000 USD, or 14.06 USD/bbl in savings compared to the Port MacKenzie plant location.

11.2.2 Netback FT Product Price

A Fairbanks plant location requires overland transportation of FT products in order to reach the coast for export to market. For this study rail transport from Fairbanks to Anchorage has been assumed, and further details of this cost are provided in section 3.1.

The Case B GTL facility in Fairbanks would require approximately 50 operational cars per day (37 for diesel fuel and 13 for naphtha) to ship FT products to Anchorage, which results in the lease of 150 total rail cars per annum. The lease of cars combined with the assumed transport tariff and fuel surcharge result in additional transport costs of about 34.5 Million USD for the Case B facility in Fairbanks compared to Port MacKenzie Case B facility. This added transportation cost has been accounted for in the netback product prices for Fairbanks.

11.2.3 Power Export

As discussed in section 10.1.2, the most likely power credit available for a GTL facility in Fairbanks would be approximately 60 USD/MWh from the GVEA, greater than the expected credit of 45 USD/MWh from MEA in Port MacKenzie. This higher credit would result in approximately 14 Million USD of additional revenue for the Case B facility in Fairbanks, equivalent to 1.30 USD/bbl. The ability of local grids to handle these levels of power will be investigated during future phases of study.

11.2.4 Miscellaneous Variable Operating Costs

Chemical and catalyst costs will increase slightly due to increased transportation required to deliver materials to Fairbanks, typically by rail. From discussions with local transport companies, a value of approximately 0.27 USD/lb is reasonable to assume for material transport of this kind between Anchorage and Fairbanks. The largest catalyst requirement is FT catalyst, and additional transport for this material would result in an additional 112,000 USD per year for Case B, equivalent to 0.01 USD/bbl. Additional expenses can be expected for other materials, but these will have little to no impact on overall operating costs.

The cost for water treatment related utilities will vary in Fairbanks compared to Port MacKenzie, but surcharges from the local utility for effluent discharge are currently unknown. Water treatment accounts for a negligible impact on overall operating cost variance.

11.2.5 Fixed Operating Costs

Estimates for operating personnel are expected to increase by approximately 3% in Fairbanks compared to Port MacKenzie, as provided by AGDC. This small adjustment has little impact on overall plant operating costs.

As discussed in section 10.2, both maintenance and insurance costs are calculated as a factor of fixed capital investment, and therefore will increase proportionately for the Fairbanks plant location.

SG&A costs are assumed to remain relatively constant between the two locations, and any minor changes due to increased transport will not impact the overall plant operating costs.

11.2.6 *Summary of Fairbanks Operating Costs*

A comparison of costs between Cases A, B and C for the Fairbanks plant location is presented in Table 11-1. In addition, Table 11-2 illustrates that the lower operating cost in Fairbanks, a 16% reduction, are directly related to the reduced natural gas price.

Table 11-1: Operating Cost Comparison between Cases A, B and C for Fairbanks location

Operating Cost	Unit Cost [USD/bbl]		
	Case A (1 Train)	**Case B (2 Trains)**	**Case C (4 Trains)**
Variable Costs			
Natural Gas	69.85	59.25	55.30
Chemicals & Catalysts	3.25	3.25	3.25
Utilities	(4.84)	(4.84)	(4.84)
Total Variable Costs	**68.26**	**57.66**	**53.71**
Fixed Costs			
Operations	3.85	3.21	2.15
Maintenance	6.85	6.33	6.08
SG&A and Insurance	2.70	2.57	2.50
Total Fixed Costs	**13.40**	**12.11**	**10.73**
TOTAL OPERATING COSTS [USD/bbl]	**81.66**	**69.77**	**64.43**
TOTAL ANNUAL OPERATING COSTS [Million USD]	**453**	**774**	**1,429**

Table 11-2: Case B Operating Cost Comparison between Port MacKenzie and Fairbanks

Operating Cost	Unit Cost [USD/bbl]		
	Port MacKenzie (CASE B)	**Fairbanks (CASE B)**	**Cost Reduction**
Variable Costs			
Natural Gas	73.31	59.25	19%
Chemicals & Catalysts	3.25	3.25	0%
Utilities	(3.55)	(4.84)	36%
Total Variable Costs	**73.01**	**57.66**	**21%**
Fixed Costs			
Operations	3.11	3.21	(3%)
Maintenance	5.07	6.33	(25%)
SG&A and Insurance	2.26	2.57	(14%)
Total Fixed Costs	**10.44**	**12.11**	**(16%)**
TOTAL OPERATING COSTS [USD/bbl]	**83.45**	**69.77**	**16%**
TOTAL ANNUAL OPERATING COSTS [Million USD]	**925**	**774**	**16%**

12. GTL Economic Analysis

12.1 Framework and Assumptions

In order to assess the economic feasibility of the GTL plant, the project is analyzed using a discounted cash flow (DCF) analysis. Both Free Cash Flows to the Firm (FCFF) (unlevered cash flows) and Free Cash Flows to Equity (FCFE) (levered cash flows) are computed and discounted at their appropriate discount rates of the weighted average cost of capital (WACC) and the cost of equity, respectively.

The valuation horizon is 30 years after commissioning in September 2019 and no residual value is assumed at the end of operations.

The framework and key assumptions for the GTL Economic Analysis are summarized in Table 12-1 and detailed in the subsequent sections below.

Table 12-1: Framework and Key Assumptions for the GTL Economic Analysis

Parameter	Base Case (Port MacKenzie)	Differential Case (Fairbanks)
Product Mix	Base : Diesel + Naphtha Alternative : Diesel + Naphtha + Jet Fuel	Same as Base Case
Target Markets	U.S. West Coast for Diesel and Jet Fuel Japan for Naphtha	Same as Base Case
No. of GTL Trains	2 Variance shown for 1 train (Case A) and 4 trains (Case C)	Same as Base Case
Delivered Natural Gas Price	7.61 USD/MMBTU	6.15 USD/MMBTU
Crude Oil Price Forecast (Figure 12-1)	80 USD/bbl constant in real terms; escalated at 3%/a in nominal terms EIA AEO 2011	Same as Base Case
Power Export Price	45 USD/MWh	60 USD/MWh
Currency	2010 USD	Same as Base Case
CAPEX	Base CAPEX as defined in section 9	Base CAPEX + 25%
Capital Spending Profile	Construction: S-Curve profile Engineering and Development: back-loaded curve	Same as Base Case
Tax Assumptions	See section 12.1.4	Same as Base Case
Debt:Equity	50:50	Same as Base Case
Cost of equity	12% per annum nominal	Same as Base Case
Cost of debt	LIBOR + 6% per annum nominal	Same as Base Case
Working Capital	See section 12.1.6	Same as Base Case

12.1.1 Product Mix and Markets Assumptions

The base case product mix (diesel + naphtha) and target markets have been determined based upon the market analysis results presented in section 3 of this report. These results, which are based on netback prices, suggest that maximizing FT diesel yields higher revenue. The results also indicate that, given the market sizes, growth trends and demand volatility, the U.S. West Coast market is the best target for FT diesel. Note that the economic model is built with the embedded functionality to change both the product mix to include FT jet fuel and the target markets in order to evaluate alternative scenarios for future circumstances. Finally, FT naphtha is assumed to be exported to Japan.

The base case product mix is therefore comprised of 24,680 bbl/day of FT diesel, equivalent to 74% of the total plant output, and 8,580 bbl/day of FT naphtha, equivalent to 26% of the total plant output.

The alternative product mix is comprised of 13,780 bbl/day of FT diesel, 12,555 bbl/day of FT jet fuel and 6,660 bbl/day of FT naphtha, equivalent to 42%, 38% and 20% of the total plant output, respectively.

Summary economic results are presented for Cases A (1 FT train) and C (4 FT trains) corresponding to 16,630 bbl/day and 66,520 bbl/day of product respectively.

12.1.2 Key Price Assumptions

Crude oil and natural gas prices are the two most important drivers of a GTL plant's economics. Specifically, a large spread between feedstock (natural gas) prices and output (FT products relative to crude oil) prices improves the business case. Price escalation, to be applied to both revenue and costs, is assumed at 3% per annum based on the U.S. historical inflation rate.

The GTL plant revenue from the FT products are calculated relative to the crude oil price. The market analysis derives the historical, statistically significant, relationships between crude oil (defined as Cushing, OK WTI) and the best proxy products to FT liquids for which data is available. Therefore, determining the appropriate WTI projection is a key step in this analysis as it determines the GTL plant's revenue stream. AGDC has determined that 80 USD/bbl (constant in real terms and escalated at 3% per annum in nominal terms) is the appropriate projection to be used in this assessment. An alternative projection, provided by U.S. Energy Information Administration (EIA), is presented in Figure 12-1 and shown alongside the AGDC forecast. Summary economic results are presented for both crude oil projections.

No selling price premium is assumed for FT liquids as there is only anecdotal evidence or unverifiable data on potential FT liquid premiums over and above the proxy products used in this study. Furthermore, it is assumed that any potential premium will be offset by the required blending prior to final sale.

Natural gas prices are comprised of wellhead costs at the North Slope plus transportation costs to the GTL plant, henceforth described as the 'delivered natural gas price.' AGDC has determined that a 7.61 USD/MMBTU and a 6.15 USD/MMBTU delivered natural gas price should be assumed for the purpose of this study for the Port MacKenzie and Fairbanks locations, respectively. Note that the natural gas transportation cost is levelized in nominal terms, and therefore only the wellhead cost of 1.00 USD/MMBTU is escalated at 3% per annum, while the pipeline tariff is left constant over the valuation horizon.

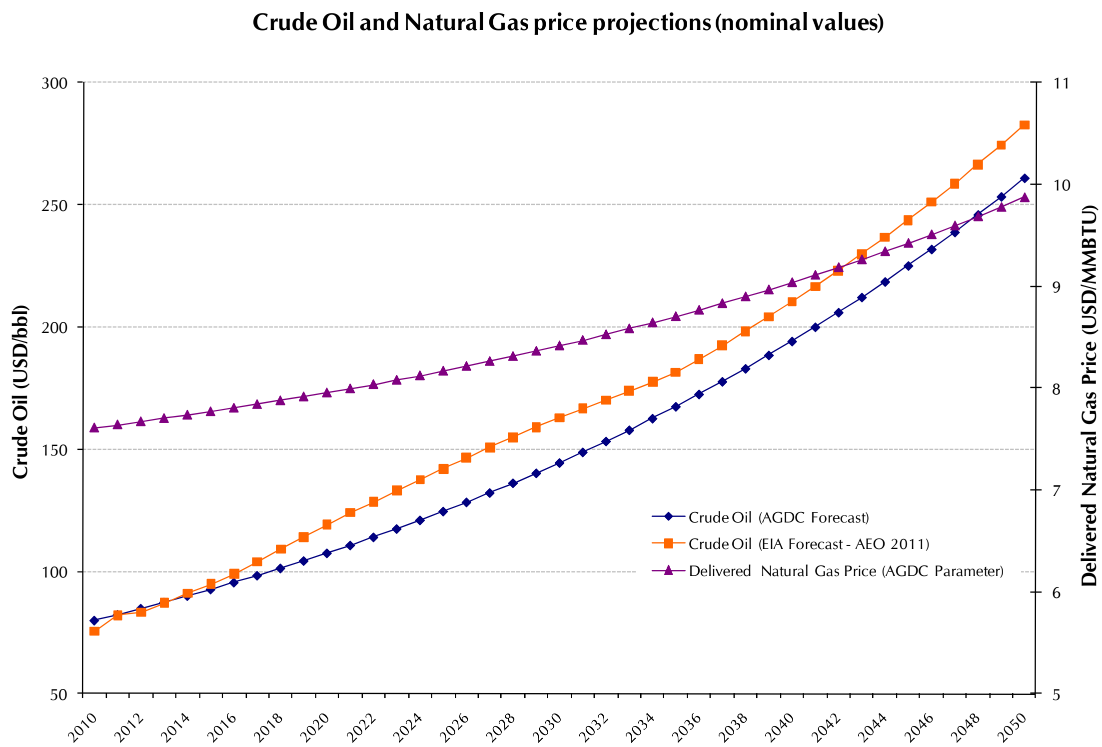


Figure 12-1: Crude oil and delivered natural gas to Port MacKenzie projections (in nominal terms)

An additional source of revenue for the GTL plant is from power export, which amounts to 119 MWe and 107 MWe for Case B (base case) producing the base product mix and the alternative product mix, respectively. Power export price, in 2010 USD, and in consideration of a new supply-demand equilibrium after the GTL plant is commissioned, is estimated at 45 USD/MWh and 60 USD/MWh at Port MacKenzie and Fairbanks, respectively.

12.1.3 Capital Spending Profile, Depreciation and Sustaining Capital Assumptions

Capital spending is assumed to follow an S-Curve profile and is split between engineering and development and plant construction costs. Following the project execution schedule presented in section 8 of this report, engineering and development is materialized from 2011 to Q4-2018, whereas plant construction occurs from Q2-2017 to Q4-2019. Engineering and development spending is assumed to follow a back-loaded S-Curve, with 92% being spent during detailed engineering onward. Costs are escalated on an annual basis at 3% and a financing and banking fee of 3% is added to the total installed cost.

For income tax calculation purposes, fixed asset depreciation is split between buildings and GTL equipment. An approximate split based on CAPEX breakdown is 35% for buildings and 65% for GTL equipment. Depreciation rules are based on the U.S. Internal Revenue Service (IRS) Publication 946 and the advice of KPMG Anchorage, AK office. Buildings are depreciated over 30 years using a straight-line method, whereas GTL equipment is depreciated over 7 years using the Modified Accelerated Cost Recovery System (MACRS) method.

For financial purposes, in the preparation of pro-forma income statements, total installed costs are depreciated over 30 years using the straight-line method.

Sustaining capital, that is, fixed asset investments required to keep production and efficiency levels, are assumed to be 4% of the direct capital costs (total installed costs excluding indirect costs and contingency) per annum. For tax depreciation purposes, sustaining capital is depreciated following the rules for GTL equipment. The sustaining capital assumption is based on the average value applicable to crude oil refineries. No difference in sustaining capital is assumed between Fairbanks and Port MacKenzie.

12.1.4 Tax Assumptions

U.S. federal income tax rate is assumed at 35% and the state of Alaska oil and gas corporation net income tax rate of 9.4% is assumed for state income tax. This results in a blended statutory tax rate of 41.1%.

Net Operating Loss Carry Forward (NOLCF) are modeled in order to appropriately reflect the deferred taxes that arise during the years where the GTL plant operates at a loss. Net operating losses are assumed to be carried forward for a maximum period of 20 years.

Borough property taxes are also modeled since they represent a material expense. These taxes are incurred during construction and are applied to the work in progress once the equipment and material purchased outside of Alaska enters the state. Property taxes paid during construction can be capitalized and then depreciated over time.

Once the GTL plant is commissioned, the property tax is applied to the assessed fair market value and improvements, to be determined on an annual basis between the owner and the applicable Borough. For the purpose of this study, it is assumed that a proxy for the fair market value is the accounting book value, that is, plant cost plus sustaining capital minus accumulated depreciation.

For calculation purposes, this study uses the Matanuska-Susitna Borough mill rate, which amounts to 9.956 USD (area wide) plus 0.394 USD (outside of built up areas), as well as a service fee for the Port MacKenzie area of 7.33 USD for a total mill rate of 17.68 USD. For every 1,000 USD of assessed fair market value, 17.68 USD in property tax is due to the Matanuska-Susitna Borough. No assumptions are made regarding negotiations between the GTL plant owner and the Borough for a lower mill rate, although this scenario is conceivable. No property tax distinction is made between Fairbanks and Port MacKenzie at this level of study.

12.1.5 Capital Structure, Required Returns and Debt Term Assumptions

Capital structure is assumed at 50:50 Debt:Equity to finance FEL4 and construction costs. This assumption is based on a DOE recommendation for GTL plants [21]. It is important to point out that, depending on the crude oil and delivered natural gas price scenario, this capital structure may not be attainable since a minimum required debt service coverage ratio may not be met.

Cost of debt is assumed at LIBOR + 6% per annum nominal, following the DOE recommendation for GTL plants [21]. Cost of equity is assumed at 12% per annum nominal as suggested by AGDC in order to make meaningful comparisons to the other two industrial user opportunities under evaluation, namely, LNG export and NGL sales.

Debt term is assumed to be a 20-year bank loan with interest capitalized during the construction period and initial year of operations.

12.1.6 Working Capital Assumptions

Working capital is defined as the required investments in inventories and accounts receivable deducted from the supplier's implicit credit though accounts payable. Working capital investments are incurred on a yearly basis and are recovered at the end of the project life. Therefore, working capital essentially represents a financial cost.

It is assumed that the GTL plant does not hold natural gas inventory and that the inventory for chemicals and catalysts, maintenance supplies, and FT liquids (finished products) are 45 days, 30 days and 30 days, respectively. Finished product inventories are valued at cost.

Accounts payable and accounts receivable are assumed to be paid in 30 days.

12.2 Economic Results

12.2.1 Base Case at Port MacKenzie

The economic results for the Base Case GTL facility at Port MacKenzie are shown in Table 12-2 for three GTL train sizes (defined by the number of FT trains in the plant): 1 FT train, 2 FT trains and 4 FT trains. The subsequent discussion is mostly focused on the base case (2 FT trains), unless otherwise stated. Table 12-3 below also shows the economic results based on the EIA crude oil forecast.

Table 12-2: Summary of Base Case (Port MacKenzie) Economic Results Using AGDC Crude Oil Forecast

	1 FT Train 16,630 bbl/day	**2 FT Trains 33,260 bbl/day**	**4 FT Trains 66,520 bbl/day**
NPV (12%) to Equity [USD million]	(597)	(695)	(808)

Table 12-3: Summary of Base Case (Port MacKenzie) Economic Results Using EIA Crude Oil Forecast

	1 FT Train 16,630 bbl/day	2 FT Trains 33,260 bbl/day	4 FT Trains 66,520 bbl/day
NPV (12%) to Equity [USD million]	(370)	(288)	(46)

From an operating perspective, the base case results in a steady-state (when plant nameplate capacity is reached) EBITDA (Earnings before interest, tax, depreciation and amortization) margin of 26% when the plant is located at Port MacKenzie. Of the total plant revenue, 60% is consumed in feedstock natural gas. Figure 12-2 below shows how revenue is spent in variable and fixed costs for Port MacKenzie.

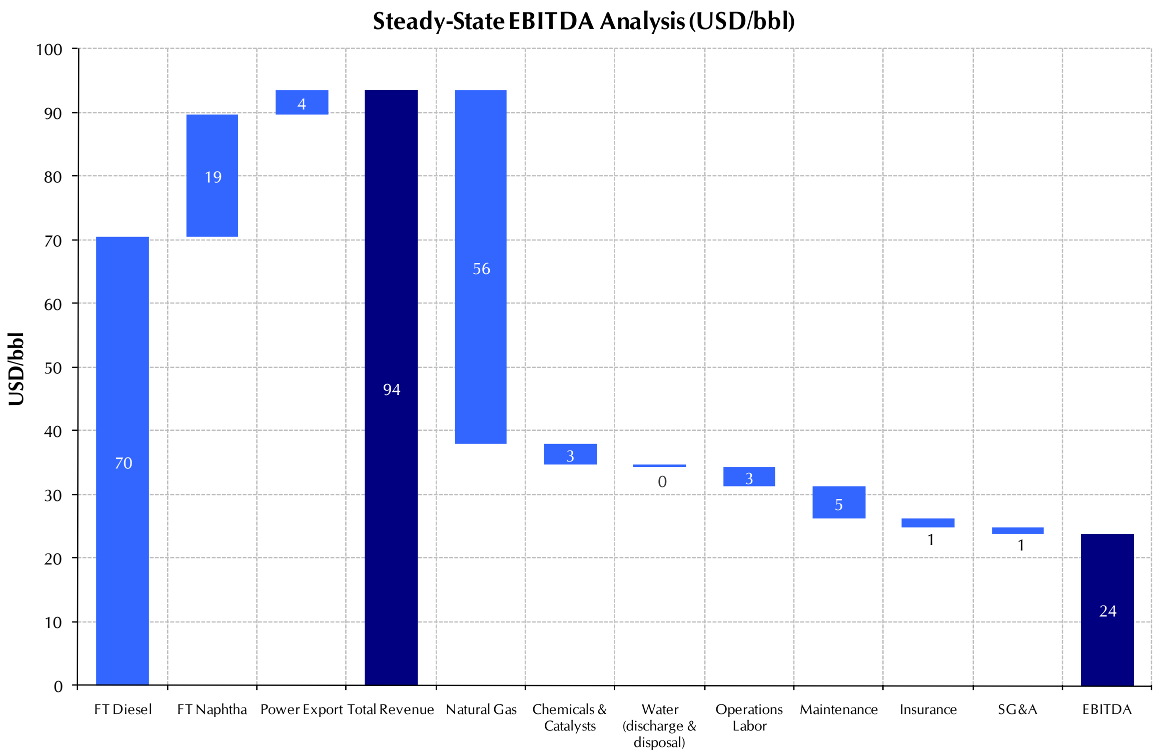


Figure 12-2: EBITDA Analysis (Port MacKenzie)

These negative NPV values indicate that, given the assumed crude oil price, delivered natural gas price, and other assumptions, the GTL plant does not appear to represent a positive value proposition at Port MacKenzie. However, it is important to expand the analysis to include breakeven prices and sensitivities in order to understand what would be required for a GTL plant to represent an economically feasible industrial opportunity at Port MacKenzie.

12.2.2 Breakeven Analysis

The breakeven analysis discussion assumes that different views can be adopted in terms of both crude oil and delivered natural gas prices, showing the price combinations that will make the NPV to equity to breakeven, that is, the price combinations that will earn equity holders a 12% rate of return. The results of this analysis for the Port MacKenzie option, both the base case CAPEX and a worst-case CAPEX assuming a 40% increase, are shown on Figure 12-3 below.

This analysis shows that, as expected, higher crude oil prices resulting in higher revenues can absorb higher feedstock prices. Any price combination to the right of the breakeven points are NPV positive alternatives. At 80.00 USD/bbl, the GTL plant built at the base CAPEX breaks even with a delivered natural gas price of 4.42 USD/MMBTU. In addition, if the worst-case CAPEX scenario materializes, that is, a 40% increase as assumed by the CAPEX accuracy range, the GTL plant breaks even at 1.80 USD/MMBTU delivered natural gas price.

Table 12-4 below presents the corresponding breakeven delivered natural gas prices for the 1, 2 and 4 FT train cases for Port MacKenzie. The economies of scale effect of a larger plant translate into higher breakeven feedstock prices.

Table 12-4: Delivered natural gas price breakeven analysis in 2010 USD (Port MacKenzie)

	1 FT Train 16,630 bbl/day	2 FT Trains 33,260 bbl/day	4 FT Trains 66,520 bbl/day
Delivered Natural Gas Price [USD/MMBTU]	3.52	4.42	4.97

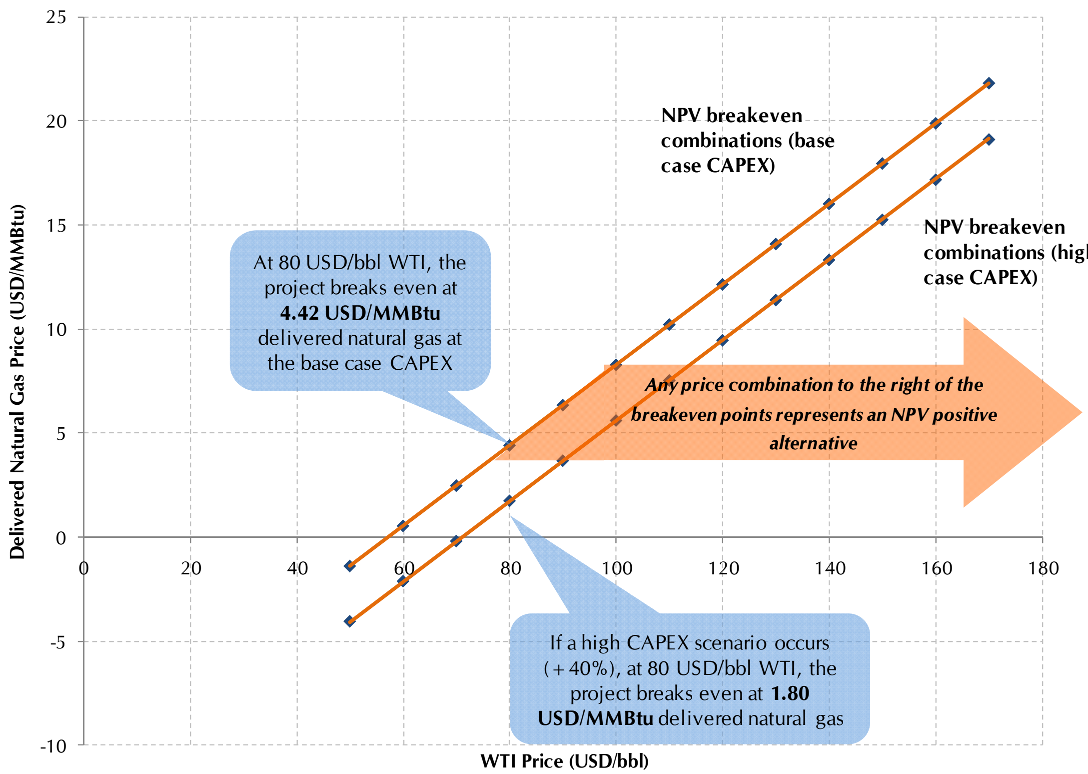


Figure 12-3: Breakeven Analysis in 2010 USD (Port MacKenzie)

12.2.3 *Sensitivity Analysis*

A sensitivity analysis on the base case at Port MacKenzie is conducted in order to identify and rank the project's key drivers and the effect of their variations on the base case. Figure 12-4 below shows the base case sensitivities to variations from the base case parameters. It can be seen that the GTL plant is most sensitive to crude oil price, delivered natural gas price and CAPEX, in that order. In addition, variable and fixed OPEX changes have only a marginal effect on the project's economic results.

Figure 12-4 also shows that, all else constant, an increase of approximately 20% in the assumed crude oil price or, alternatively, a decrease of approximately 40% in the assumed delivered natural gas price, will allow equity holders to obtain a 12% rate of return. It can also be seen that, all else constant, a significant decrease in CAPEX, of about 50% would make the plant economically viable under these conditions. The latter is obviously an unrealistic target.

A more focused analysis on delivered natural gas sensitivity is provided in Figure 12-5 below, which shows the breakeven delivered natural gas price of 4.42 USD/MMBTU derived as part of the breakeven analysis and the levered NPV results obtained at different prices. An alternative representation is shown on Figure 12-6, which depicts the IRR to equity sensitivity to the delivered natural gas.

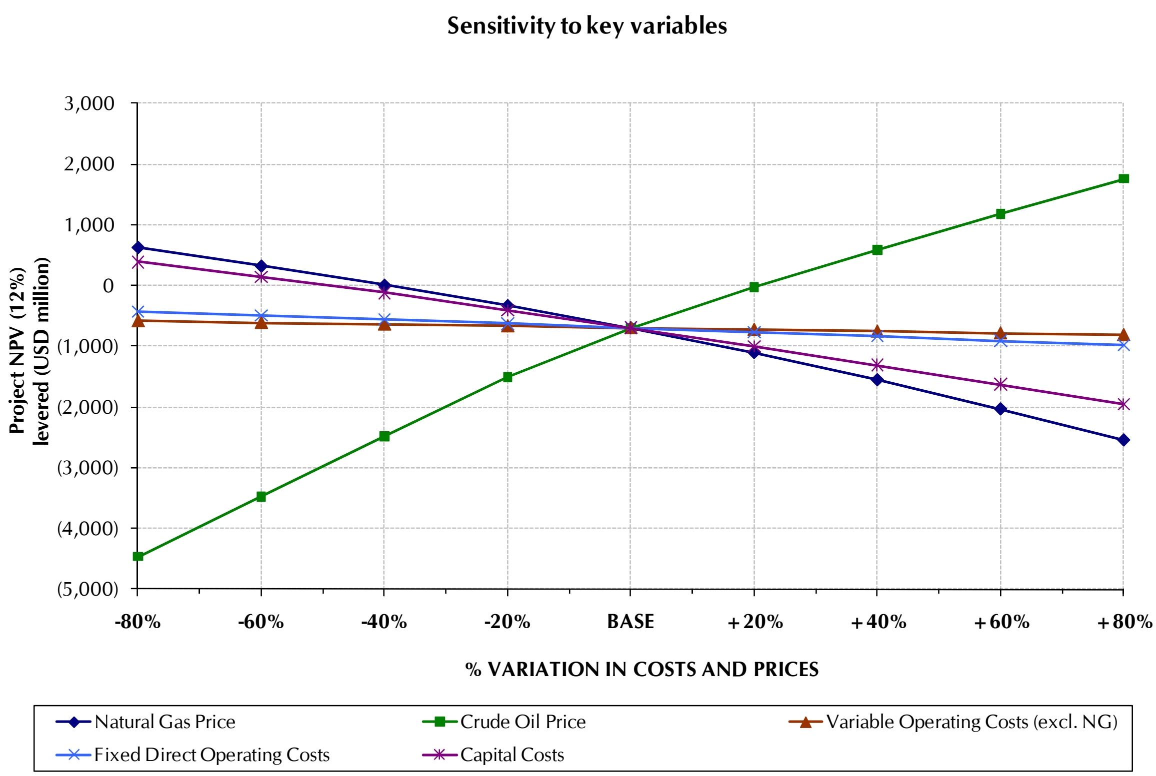


Figure 12-4: NPV sensitivity to key variables

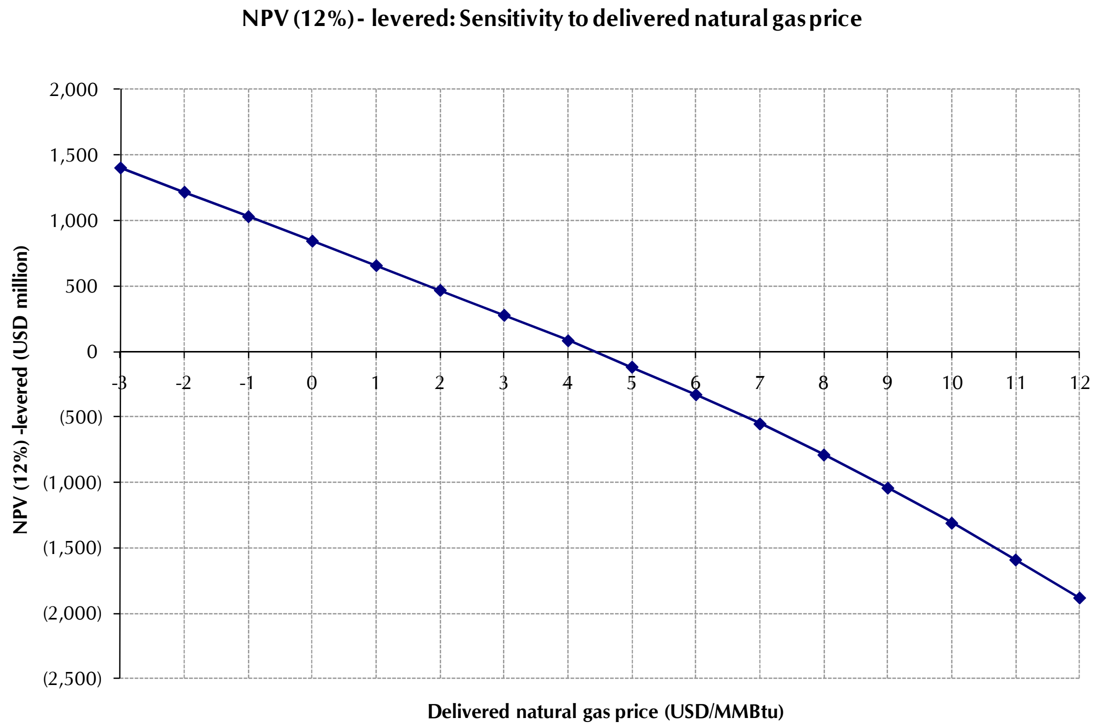


Figure 12-5: NPV sensitivity to delivered natural gas (2010 USD) at Port MacKenzie

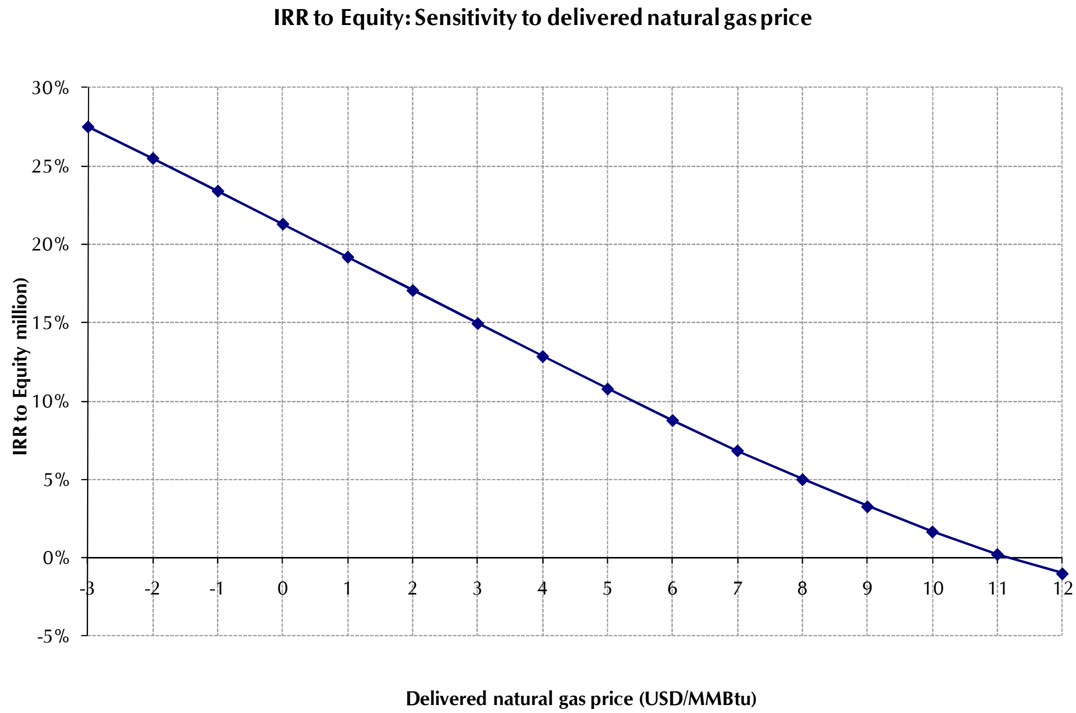


Figure 12-6: IRR to Equity sensitivity to delivered natural gas (2010 USD) at Port MacKenzie

Appendix F to this report contains the detailed financial projections, including pro-forma income statement and statement of cash flows and the resulting FCFF and FCFE for the Port MacKenzie option.

12.3 Differential Case at Fairbanks Compared to Port MacKenzie

The economics of siting the GTL plant in Fairbanks was compared to the base case (Port MacKenzie) economics using a delivered natural gas price to Fairbanks of 6.15 USD/MMBTU as given by AGDC. The basis for this comparison is summarized in Table 12-1. The economic results for the differential case at Fairbanks are compared to the Base Case GTL facility at Port MacKenzie in Table 12-5 (using AGDC crude oil forecast) and Table 12-6 (using EIA crude oil forecast).

Table 12-5: NPV Compared – Fairbanks vs. Port MacKenzie Using AGDC Crude Oil Forecast

NPV (12%) to Equity [USD million]	1 FT Train 16,630 bbl/day	2 FT Trains 33,260 bbl/day	4 FT Trains 66,520 bbl/day
Fairbanks	(688)	(861)	(1,220)
Port MacKenzie	(597)	(695)	(808)

Table 12-6: NPV Compared – Fairbanks vs. Port MacKenzie Using EIA Crude Oil Forecast

NPV (12%) to Equity [USD million]	1 FT Train 16,630 bbl/day	2 FT Trains 33,260 bbl/day	4 FT Trains 66,520 bbl/day
Fairbanks	(460)	(450)	(437)
Port MacKenzie	(370)	(288)	(46)

When the plant is located at Fairbanks and has achieved nameplate capacity, the EBITDA margin improves, from 26% for the GTL plant in Port MacKenzie to 34% due to the lower feedstock price. Of the total plant revenue, 49% is consumed in feedstock natural gas. Figure 12-7 below shows how revenue is spent in variable and fixed costs for Fairbanks.

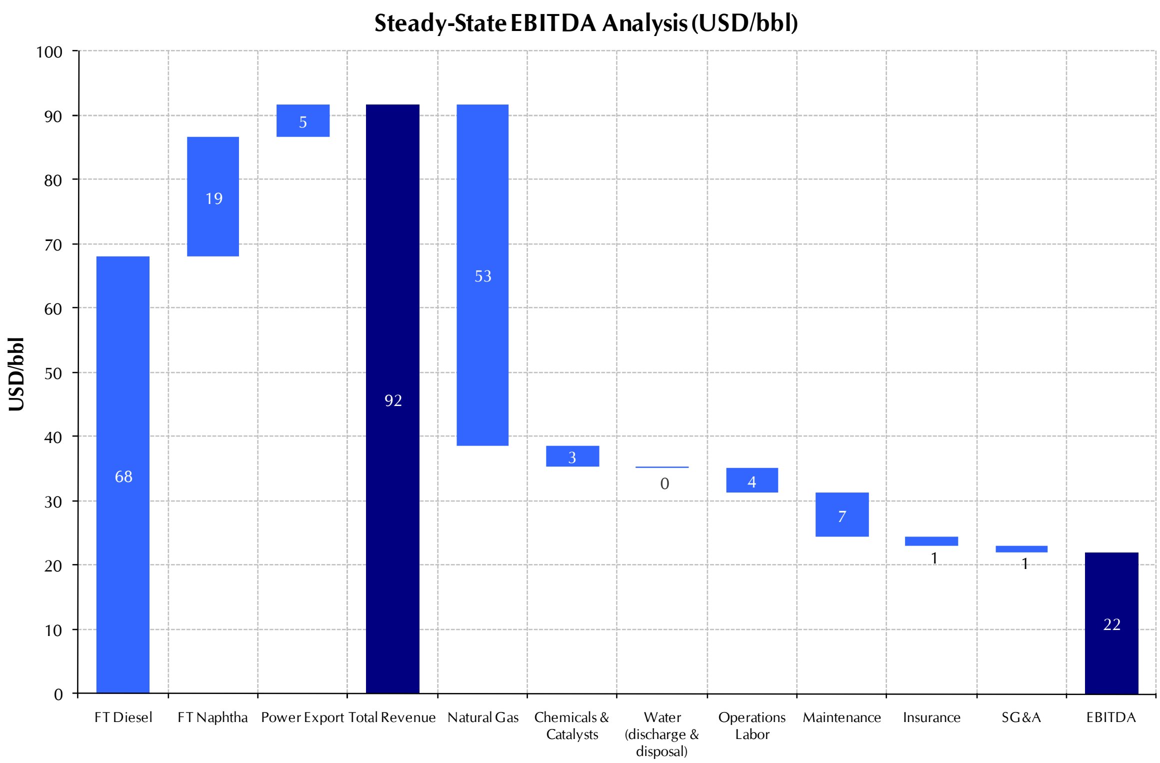


Figure 12-7: EBITDA Analysis (Fairbanks)

As with the base case economics at Port MacKenzie, these negative NPV values indicate that, given the assumed crude oil price, delivered natural gas price, and other assumptions, the GTL plant does not appear to represent a positive value proposition at Fairbanks as well. However, breakeven prices and sensitivities are analyzed below to understand what would be required for a GTL plant to represent an economically feasible industrial opportunity at Fairbanks.

12.3.1 Advantage of a Lower Natural Gas Price

By locating the GTL plant in Fairbanks, it is expected that there will be an increase in capital cost due to the higher complexity associated with constructing the plant in Fairbanks, as detailed in section 11.1. The NPV for the GTL project at Fairbanks is calculated for various CAPEX increases relative to the Port MacKenzie CAPEX (base case), showing the effect of increasing the capital cost on the project economics (Table 12-7). The differential NPV to Equity is also calculated to show that the Fairbanks GTL project NPV is higher than that of Port MacKenzie up until a 14% increase in CAPEX. Therefore, the GTL project at Fairbanks can absorb up to a 14% increase in capital cost while still remaining economically equivalent to the GTL project at Port Mackenzie, all as a result of a lower delivered natural gas price to Fairbanks. Comparing the economics of siting the GTL plant in Port MacKenzie and in Fairbanks shows that locating the plant closer to the natural gas source is economically beneficial to the project. It is important to note that while a 25% cost adder was used to estimate the CAPEX for the GTL plant in Fairbanks relative to Port MacKenzie, this is a conservative cost adder and must be refined with the completion of transportation and constructability studies.

Table 12-7: NPV (12%) to Equity Differential Analysis

CAPEX increase with respect to Port MacKenzie (%)	NPV (12%) to Equity at Fairbanks (USD million)	Δ NPV (12%) to Equity with respect to Port MacKenzie (USD million)
10%	(638)	57
14%	(695)	0
15%	(711)	(16)
20%	(786)	(91)
25%	(861)	(166)

Finally, sensitivity to delivered natural gas price at Fairbanks was conducted assuming different scenarios for capital cost increase with respect to Port MacKenzie using $80/bbl for crude oil. These results, presented in Table 12-8 below, show the lower feedstock prices that can be absorbed in light of the likely capital cost increases due to construction complexity.

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Table 12-8: Delivered natural gas price breakeven analysis (Fairbanks)

CAPEX increase for Fairbanks Case with respect to Port MacKenzie (%)	Breakeven Delivered Natural Gas at Fairbanks (USD/MMBTU)
10%	3.17
14%	2.91
15%	2.84
20%	2.52
25%	2.19

13. Conclusions and Recommendations

A gas-to-liquids plant converts natural gas into liquid fuels, and was evaluated as a potential industrial anchor for the proposed Alaska Stand Alone Gas Pipeline/***ASAP***. An economic feasibility study was carried out for this GTL facility for the Cook Inlet and Fairbanks regions in Alaska.

The facility was designed to process pipeline-quality natural gas delivered by the proposed pipeline; to produce diesel and naphtha via syngas production and FT synthesis; to generate power using steam produced by the plant; and to capture CO_2 from flue gas streams. Three GTL facility sizes were considered to evaluate the effect of facility size on the project economics: Case A (1 FT train), Case B (2 FT trains) and Case C (4 FT trains). The number of trains for a GTL facility are based on the number of FT synthesis reactors required, and a 17,000 bbl/day FT reactor was considered for this study, based on the capacity of current commercial reactors.

A base product mix of diesel and naphtha was defined for this GTL facility based on netback prices for the product options in target markets, as analyzed in the market study (section 3). For the economic analysis, an alternative product mix of jet fuel, diesel and naphtha was also assessed.

The GTL plant performance for all cases were calculated using Aspen Plus® (process simulation software) as summarized below.

	Case A (1 Train)	**Case B – Base Case (2 Trains)**	**Case C (4 Trains)**
Pipeline-quality Natural Gas Required [MMSCFD]	154	309	617
FT Products: Diesel + Naphtha [bbl/day]	16,630	33,260	66,520
Power Export [MWe]	60	119	239
Raw Water Intake [st/h]	228	445	910

The GTL facility design also includes process units to capture carbon dioxide (CO_2) from purge gas. In the Cook Inlet region, the recommended end use for the captured CO_2 is for enhanced oil recovery (EOR) in local oilfields, with a secondary option being to sequester the CO_2 in depleted gas reservoirs or saline aquifers. In the Fairbanks region, the captured CO_2 is recommended to be sequestered in deep, unmineable coal seams, which are located near the Healy coal fields.

Potential locations for this GTL facility were investigated for the Fairbanks and Cook Inlet regions. The base case facility requires a footprint of 140 acres for all major process units, utilities, offsites and administration. For both regions, considerations were taken for lot size; distance from residential, commercial and military zones; transportation access; natural gas, power and water utilities access; topography and zoning. In the Cook Inlet region, land in the Port MacKenzie Master Plan, located in the Matanuska-Susitna borough was identified as a potential site suitable for this GTL facility. In the

Fairbanks region, land owned by the Fairbanks North Star Borough, located north of Interstate A2 at Rentals Street, was identified as a potential site. The Port MacKenzie site is located near a deep water port, and the FNS Borough site is landlocked.

Subarctic conditions of both the Anchorage and Fairbanks areas add complexity to the execution of major capital projects, with a relatively short construction season and with labor productivity adversely affected by the low temperatures. In order to maintain schedule and minimize cost overruns, the plant can be built in modules offsite in low-cost yards and then shipped to site for assembly.

As a basis for this study, modularization construction and the need to transport the large FT reactors by ship were taken into account, when using the Port Mackenzie site for development of the project execution schedule, along with the capital and operating cost estimates. The base case GTL facility is projected, based on previously completed and current GTL projects under construction, to begin start-up and commissioning in Q4 2019, achieving nameplate capacity in the third quarter of 2020 if permitting and pre-feasibility engineering begin in 2012.

A capital cost estimate was prepared for the base case (Case B) in 2010 USD assuming a fully integrated gas-to-liquids facility based in Port MacKenzie. An order-of-magnitude or AACE level 4 estimate was completed with an accuracy of -30%, +40%. The estimate includes the use of equipment quotations, factored equipment costs and parametric models. The fixed capital investment for the base case GTL facility (33,260 bbl/day of diesel and naphtha) is estimated to be USD 2.93 Billion or 88,000 USD/(bbl/day). This capital cost is benchmarked (shown in Appendix D) against previously constructed (or under construction) GTL facility estimates, escalated to the current southern Alaskan market (using CEPCI index and location factor). Estimates for Cases A and C were scaled from Case B on a per train basis. It was assumed that a savings of 15% is common on subsequent trains. The resulting savings per unit of production from Case A to Case B and to Case C are 7.5% and 11.3%, respectively. The savings are attributable due to duplication of design; however, the majority of the costs are associated with procurement and construction tasks.

For the capital cost estimate of the GTL plant in Fairbanks, 25% was added to the base case CAPEX for Port MacKenzie in order to account for the complexity of construction and transportation of equipment to the Fairbanks facility. This brings the capital cost of the base case GTL facility at Fairbanks up to USD 3.66 Billion.

The operating cost of the Case B facility considers fixed (operations, maintenance, SG&A, insurance) and variable (raw materials, utilities) costs, and is estimated in 2010 USD to be USD 925 Million per year or 83 USD/bbl for the Port MacKenzie GTL facility, and USD 774 Million per year or 70 USD/bbl for the Fairbanks GTL facility. Natural gas accounts for approximately 90% of the GTL facility operating costs, driving the lower costs in Fairbanks since it is closer to the feedstock source. Variable expenses are directly related to plant production, and therefore for Cases A, B and C, the operating cost on a normalized USD/bbl basis are practically equal at this level of accuracy with the exception of natural gas. Since each case represents the anchor tenant for pipelines of different sizes, economies of scale for the pipeline are transferred to the GTL facility by the natural gas transportation tariff, resulting in the largest GTL plant (Case C) having the lowest natural gas price. Economies of scale are also shown by the differences in fixed operating costs. Note that the natural gas transportation cost is levelized in nominal terms, and therefore only the wellhead cost of 1.00 USD/MMBTU is escalated at 3% per annum, while the pipeline tariff is left constant over the valuation horizon.

An economic analysis was performed for the GTL facility considering an 80 USD/bbl crude oil price and a natural gas price of 7.61 USD/MMBTU delivered to the base case facility at Port MacKenzie and 6.15 USD/MMBTU delivered to the base case facility at Fairbanks. Other key assumptions include a 3% per year escalation rate; debt to equity ratio of 50:50; a required rate of return on equity of 12%; and the analysis was carried out in 2010 USD.

The economic analysis shows that for both sites, given the base case assumptions, the GTL plant would not meet the equity holders' required rate of return in order to result in a positive value proposition. However, the breakeven analysis reveals that a delivered natural gas price of 4.42 USD/MMBTU to Port MacKenzie and 2.19 USD/MMBTU to Fairbanks, all else constant, would yield the required 12% to its equity holders. There is an economy-of-scale effect such that, the larger the plant, the higher the allowable delivered natural gas price.

The sensitivity analysis shows that the crude oil price, delivered natural gas price and CAPEX, in that order, are the most important drivers of the GTL plant's economics. It can also be concluded that a significant decrease in the delivered natural gas price or a significant increase in the crude oil price greatly improves the economic attractiveness of the project. However, significant reduction in CAPEX, all else constant, has a smaller effect on the project economics.

The sensitivity analysis also shows that under the stated assumptions, a reduction in the natural gas price of USD 1.46/MMBTU delivered to Fairbanks results in the ability of the Fairbanks GTL plant to absorb up to a 14% increase in CAPEX while remaining economically equivalent to the Port MacKenzie GTL plant. Therefore, despite the logistical and locational advantages embodied by the Port MacKenzie site, the advantage of locating a facility nearer to the natural gas source, as embodied by Fairbanks, gives significant headroom for increased capital expenditures to overcome the locational and logistical disadvantage. The source of this advantage is that natural gas is a more expensive product to transport than the denser liquid output products, bearing in mind that the volume of the gaseous feedstock is reduced over 1500 times when converted to liquid products. Based on the siting considerations researched and the economic analysis, both the Port MacKenzie and Fairbanks sites are viable locations for the GTL plant.

It can also be concluded that improved economic viability of the project is best achieved through increases in carbon efficiency and through improved yields in higher value products. Hatch assumed a conventional design for this study to be in-line with AGDC's request to focus on proven technology. However, it must be noted that significant advances are currently being pursued in the GTL field, and are expected to produce higher diesel yields and higher carbon efficiencies. Carbon efficiency is especially important when the feed gas price is high as is the case with this study; therefore, it would be worth spending additional capital to increase efficiency. Advances are also being made in lowering the capital costs per unit production, for example, increasing FT reactor throughput. These improvements would also improve the economics, although to a lesser extent relative to a more efficient process or a higher value product slate.

This study qualifies as a conceptual or FEL1 study in terms of the Hatch guidelines for the deliverables required for this level of work. In a future phase of engineering, both transportation and constructability studies are required to further detail each potential plant location and to associate definitive costs to overcoming the logistical challenges of Fairbanks. Upon completion of these studies, a plant location can be selected. Process optimization and refinement of the cost estimate accuracy would also be performed in the next phase of engineering to determine whether the economics of this GTL facility would improve as an anchor tenant for the proposed Alaska Stand Alone Gas Pipeline/***ASAP***. GTL technology licensor input, for both process and cost information, would be a requirement in the next stage of engineering.

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