Process Design and Economic Analysis

CBE 490

Gas to Liquids (GTL)

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Abstract

The demand for energy worldwide is ever increasing with the expansion of populations and the industrialization of China and India. As the demands increase, there is more and more competition within a finite supply of fossil fuels. With this increased competition comes a corresponding increase in the price for these fuels. Much of this demand for energy is for fuel to power the internal combustion engine, which is the traditional domain of crude oil. As the reserves of crude oil dwindle and the prices continue to climb, an alternative supply of energy to power the world's vehicles is found in natural gas. While plentiful, natural gas is difficult to transport and often found in reserves that are remote from high consumption areas. A solution to this is found in a refinery process known as gas to liquids (GTL). GTL is a process that turns shorter chained hydrocarbons, such as natural gas, into longer chain hydrocarbons that are found in gasoline and diesel fuel. Because the end product is a liquid at standard temperature and pressure, transportation is relatively cheap and easy.

This GTL process design incorporates a Fischer-Tropsch cobalt catalyzed reaction to convert syngas into alkane chains, and a hydro-isomerization unit to convert the waxes into shorter chained alkanes. The objectives of the process design are to convert 500 million standard cubic feet of natural gas per day into final product streams of naphtha and diesel, while keeping the process safe, environmentally benign, and energy efficient.

These criteria were met, and an annual revenue of $147,200,000,000 was determined. Over 644 barrels per hour of naptha and 8930 barrels of diesel per hour was generated. The main challenges of this project included a dealing with a high feed rate and providing adequate temperature control to the highly exothermic Fischer-Tropsch reaction. These necessitated a substantial up front fixed capital investment of $72,300,000. The keys to making this system profitable were using energy conservatively by recycling our wastewater and using our waste steam in order to generate electricity, which both reduced the operating costs and created sources of revenue, ultimately making this process profitable.

Introduction

OBJECTIVE

The task at hand is to design a specified Fischer-Tropsch Reaction Unit (FTR), including reactor effluent separation facilities, as part of a planned GTL plant. This design is to be safe and environmentally clean, as well as cost efficient. Additionally, the designed FTR unit must integrate with the already present specified units within the GTL plant in order to allow for diesel (C11-C20) and naphtha (C5-C10) production.

BACKGROUND

Fischer-Tropsch Reaction and GTL

The History and Definition

Gas to liquids technology or GTL, as the name implies, is an umbrella term for a group of technologies that can create liquid hydrocarbon fuels from a variety of feedstocks. One way to do this is by using a syngas unit to convert methane into hydrogen and carbon monoxide, and using a Fischer-Tropsch synthesis to convert the syngas (hydrogen and carbon monoxide) into hydrocarbons.

At the core of GTL technology is the Fischer-Tropsch Reaction. This reaction was first performed in 1923 by two scientists in Germany using cobalt, rubidium, and iron catalysts. The FTR is usually coupled with the following reactions:

1) Synthesis Gas Formation

$$CH_n + O_2 \rightarrow \frac{1}{2}nH_2 + CO$$

2) Fischer-Tropsch Reaction

$2nH_2 + CO \rightarrow (CH_2)_n + H_2O$

3) Refining

$(CH_2)_n \rightarrow$ fuels, lubricants, etc.

The Market

The world consumes energy from many different sources, including coal, crude oil, nuclear energy, solar energy, wind, water, and natural gas. Natural gas provides over a fifth of the world's energy, and its consumption is on the rise. It is estimated that the world's gas reserves are greater than 6000 trillion cubic feet (Tcf), and that these reserves can cover the world's needs for more than 60 years (al-Shalchi, 6, 2006). Al-Shalchi also states that, "The world consumption of natural gas equal to about 2.5 Tcf, most of it is consumed by the big industrial countries" (al-Shalchi, 6, 2006). This shows a clear demand within the world for natural gas.

Despite that fact that there is currently plenty of natural gas to fuel the world, the location of the natural gas, along with the location of the demand for natural gas, creates difficulties. A substantial portion of known reserves are situated in locations remote from high consumption areas, as the table below demonstrates (Table 1).

World Natural Gas Reserves (tcf) (Samuel, 2, 2003)

Country/Region	Reserve	% share
Former Soviet Union	2000	40.0
Iran	744	14.9
Africa	337	6.7
Asia Pacific	330	6.6
South Africa	204	4.1
Europe	192	3.8
Saudi Arabia	186	3.7
Other (ME countries)	707	14.1
USA	163	3.3
Canada & Mexico	137	2.8
Total (tcf)	*5000*	*100.*

Table 1

Transporting these natural gases by ship, train, or truck is uneconomical, and much more expensive than transporting liquid petroleum. Although pipelines have been used with success in order to transport some natural gas, pipelines are also expensive, and in many cases, geographically unfeasible. The Fischer-Tropsch Reaction gives a way to bypass the larger expense of transporting a gas by transforming this gas into liquid before transporting it.

Also, the demands for light and middle products, gasoline and diesel, respectively, is steadily increasing due to an increasing presence of vehicles in both developed and developing countries. The demand for heavy petroleum products is steadily declining (See Figure 1).

World Demands for Petroleum Products (al-Shalchi, 84, 2006)

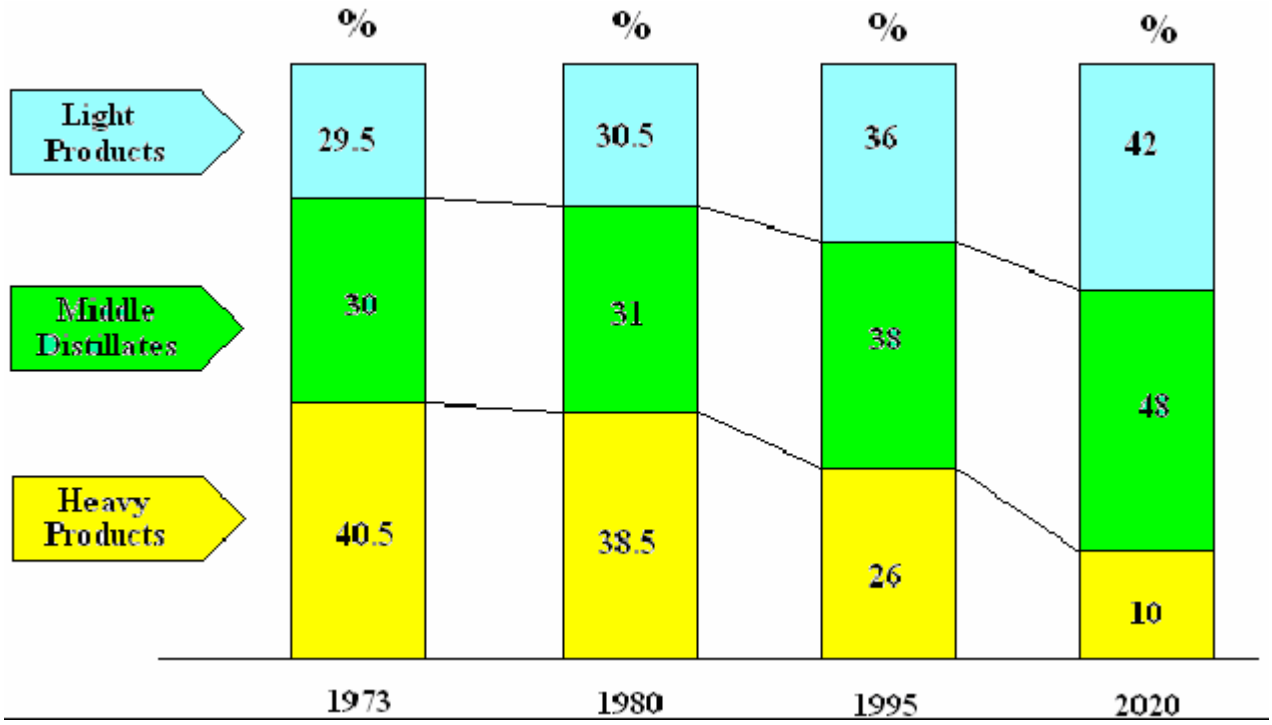


Figure 1

Another benefit of the GTL process is the higher yield of light and middle products, which better meets the demand outlined in Figure 1. Typical refineries yield a significant amount of heavy petroleum products, whereas, the Fischer-Tropsch Reaction/GTL process produce products which more closely correlate with the world's petroleum demands as seen in Figure 2.

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Petroleum Products and the GTL Industry

(al-Shalchi, 85, 2006)

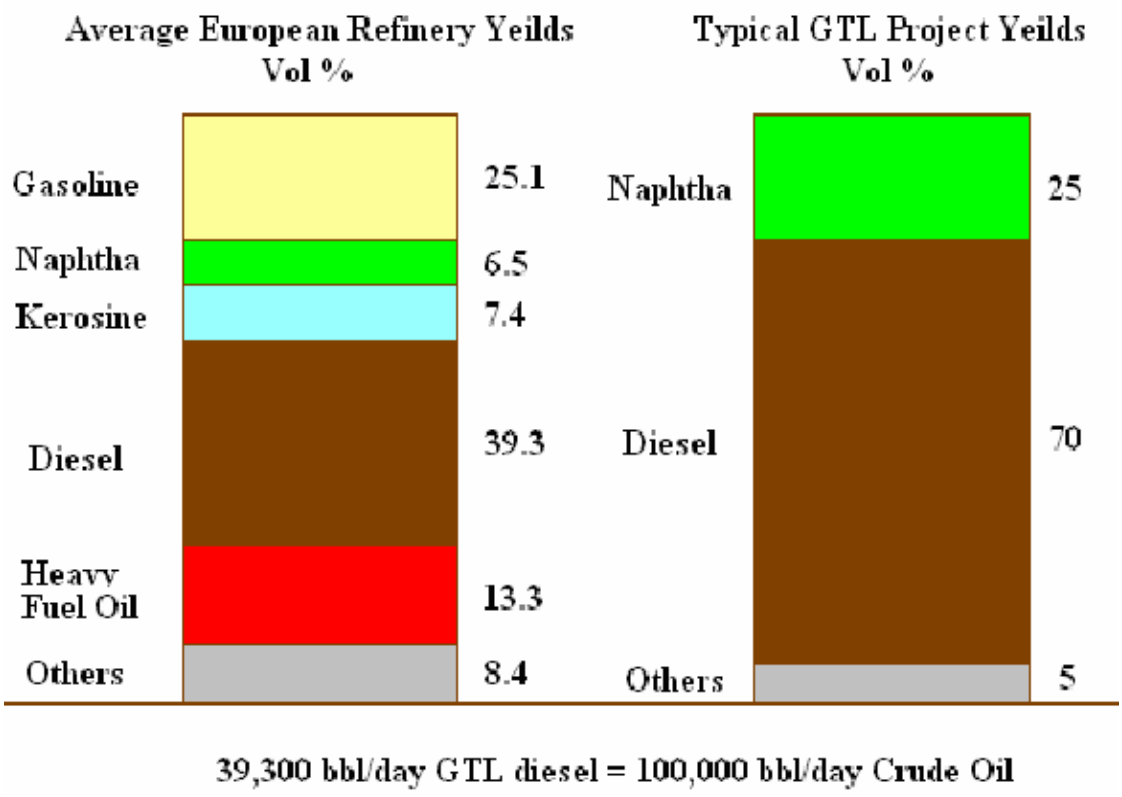


Figure 2

The Fischer-Tropsch Reaction also yields more efficient oil, with less particulate matter upon emission. Reduced sulfur content and a lower aromatic content is generated upon combustion with FTR diesel, which helps to meeting certain health and environmental regulations. Additionally, it produces 45% less carbon dioxide emissions compared with coal to obtain about the same amount of energy (Samuel, 2, 2003). Because of the cleanliness of GTL fuels, they are also an ideal feedstock for fuel cells.

THE PROCESS

The process for converting natural gas to liquid hydrocarbons consists of three primary steps: synthesis gas production, GTL synthesis, and product work up. Generally, the feed gas is treated to remove any sulfur containing components in order to avoid poisoning the catalyst, or causing corrosion and other environmental problems. Figure 3 (below) gives a general outline of the GTL synthesis.

Overall GTL Process Schematic (al-Schalchi, 11, 2006)

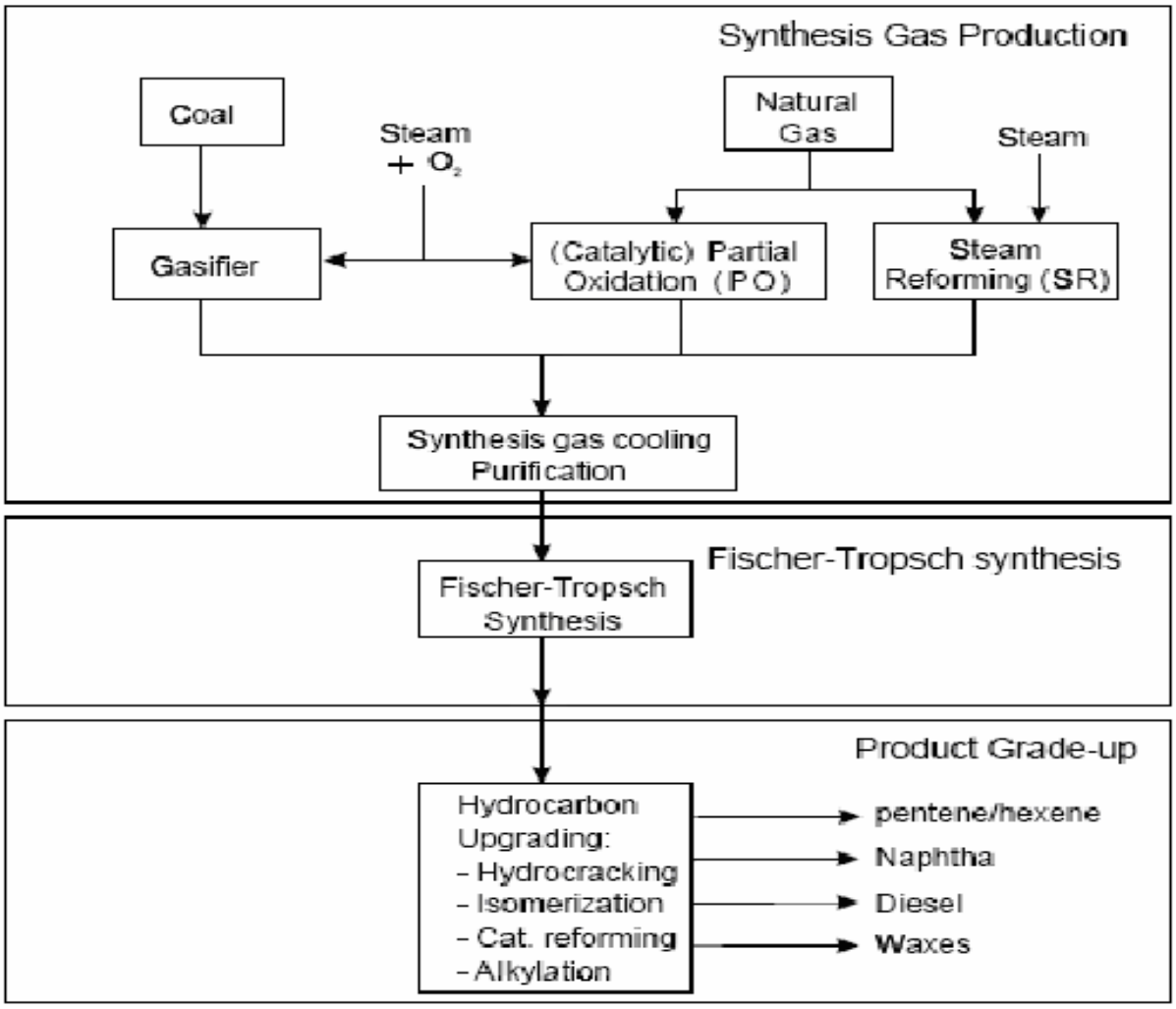


Figure 3

Synthesis Gas Production

When using the Fischer-Tropsch reaction within the GTL process, a specific molar ratio of hydrogen to carbon monoxide must be used as feed, depending upon the product desired. The syngas unit houses a series of reactions which ultimately convert clean methane into syngas. In order to meet the desired ratio of hydrogen to carbon monoxide leaving the syngas unit and being fed to the Fischer-Tropsch reactor, steam and carbon dioxide are required as feeds to drive the reforming reaction (Long, 3, 2009). The following are the three primary reactions used for syngas production:

1) Steam Reforming

$$CH_4 + H_2O \Leftrightarrow CO + 3H_2$$

2) Partial Oxidation

$$CH_4 + 3/2O_2 \rightarrow CO + 2H_2O$$

3) Shift Reaction

$$CO + H_2O \Leftrightarrow CO_2 + H_2$$

The partial oxidation runs to completion; whereas, the other two reactions meet equilibrium. Generally, a heat balance is used to determine the amount of oxygen supplied. The combination of these reactions determines the amount and composition of the product sent from the reformer and to the Fischer-Tropsch reactor. The composition of the synthesis gas can be altered to some extent by changing the operating pressure, temperature, and feed composition (Long, 3, 2009). A schematic of the syngas unit is outlined in Figure 4.

Syngas Unit Basic Flowsheet (Long, 4, 2009)

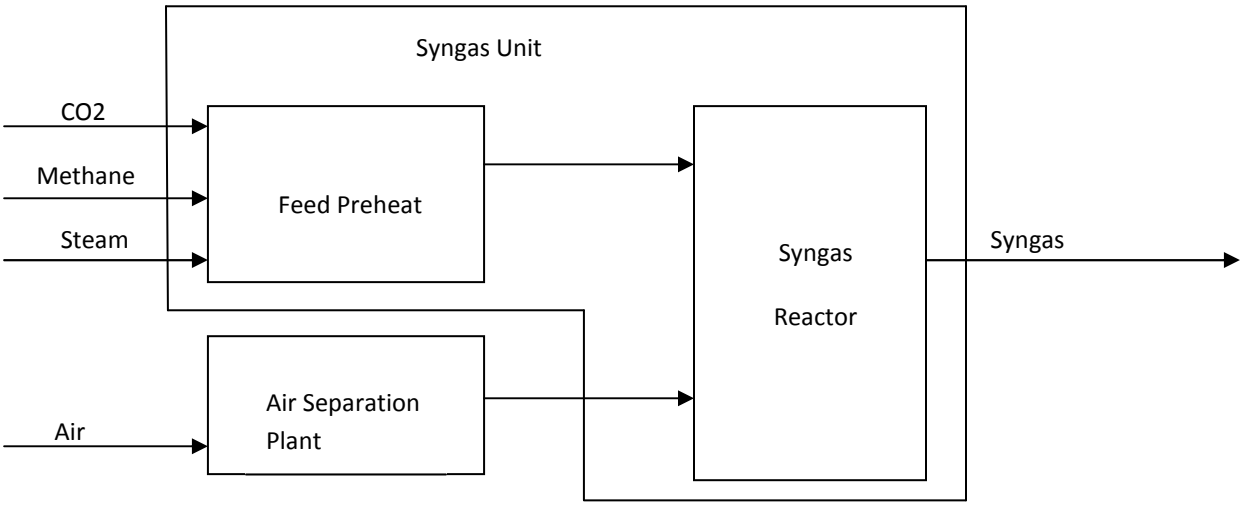


Figure 4

GTL Fischer-Tropsch Reaction

In the second step of GTL synthesis, syngas is converted into hydrocarbons of varying chain length. The Fischer-Tropsch reaction generally uses an iron, nickel, or cobalt based catalyst, and is carried out at moderate temperature (200-300˚C) and pressure(10-40 bar). The basic reaction is:

$$xH_2 + CO \Rightarrow H_2O + (CH_2)_nH_2$$

There are also side reactions that produce olefins and alcohols, but the desired product lies within the hydrocarbons (Samuel, 4, 2003).

The exact mechanism of this reaction is still under debate; however, simply stated, the following occurs: initiation of carbon chain formation, chain elongation by successive carbon building blocks, and termination of chain growth by desorption and hydrogenation of saturated species, desorption of unsaturated species, or hydrogenation, hydrolysis, and desorption of oxygenated species (al-Shalchi, 24, 2006).

Selectivity is controlled by temperature, synthesis gas composition, reactor resistance time, and catalyst formulation. For example, high temperatures favor gasoline production and lower temperatures favor the production of high molar compounds.

Product Work-Up

As mentioned earlier, the products from the FTR synthesis contain hydrocarbons of varying lengths, mostly paraffins and alpha olefins when the desired product is a middle petroleum product. Although this mixture can be shipped as a feedstock for refineries, syncrude, transportation may not be feasible. Frequently, rather than shipping the syncrude, it is separated and further processed at the processing site. This produces fuels, fuel blending components, waxes, lube bases, and other specialty products. A separation unit is not always an integral part of a GTL plant, but is used in many cases when economics and marketing call for it.

Process and Controls

PROCESS

FTR Design

Design Equations

The Fischer-Tropsch Reaction is a heterogeneous catalytic reaction described by the following rate equation.

$$-r_{CO} = \frac{kT_1 P_{H2} P_{CO}}{(1 + k_2 T_2 P_{CO})^2}$$

$$T_1 = e^{\left(-4492*(1/T - 1/473)\right)}, T\ in \deg K$$

$$T_2 = e^{\left(8237*(1/T - 1/473)\right)}, T\ in \deg K$$

$$k = 0.0173\ gmol\ CO/hr\ , cc\ cat, atm^2$$

$$catalyst\ bulk\ density = 0.8\ g/cc$$

$$k_2 = 4.512\ atm^{-1}$$

$$P_{H2}\ ,\ P_{CO} = partial\ pressures\ in\ vapor\ phase, atm$$

$$T = Reactor\ operating\ temperature$$

The product selectivity was predicted using the following probability distributions. It was assumed that all hydrocarbon products produced in the Fischer-Tropsch Reaction were straight-chain alkanes.

The distribution of C_{5+} products was predicted using the Anderson-Shulz-Flory (ASF) probability distribution.

$$W_n/n = [\,(1-\alpha)^2/\alpha]\alpha^n$$

$$M_n/n = (1-\alpha)\alpha^{n-1}$$

$$\alpha = (0.93)T_4$$

$$T_4 = e^{\left(250(1/T - 1/473)\right)}$$

$$\alpha = ASF\ chain\ growth\ parameter$$

$$W_n = relative\ weight\ fraction\ of\ carbon\ number\ n$$

$$M_n = relative\ mol\ fraction\ of\ carbon\ number\ n$$

The selectivity of methane and the light ends (C_2-C_4) were calculated using the following equations.

$$S_{CH_4} = r_{CH_4}/-r_{CO} = (0.03)T_3$$

$$T_3 = e^{\left(-1000\left(^1/_T - ^1/_{473}\right)\right)}$$

$$S_{C_n} = (0.04)S_{CH_4}\ , where\ n = 2, 3, 4$$

These equations were used to construct an Excel spreadsheet, which takes the reactor mass feed rate (syngas effluent flow rate) and predicts the mass flow rates of all the reactor products. These were then converted to molar flow rates by multiplying the components by their respective molar masses. Using the CO molar feed rate as a basis, each of the reactant and product feed rates was divided by the CO molar feed rate to derive a stoichiometric equation for the Fischer Trophsh reaction as a function of the average reactor temperature. Using the feed conditions, rate equation and stoichiometric equation and specifying the reactor length, tube count, tube diameter, and the reactor cooling rate, the reactor was fully characterized and able to be simulated in Hysys. Reactor pressure drop was calculated automatically using the Ergun equation which is embedded in the Hysys PFR solver utility.

Design Goals

The Fischer-Tropsch, packed-bed reactor was designed to meet three general, interrelated design criteria: reasonable reaction yield, thermal stability, and a pressure drop below 50 psi. How these goals were met through this design is discussed below.

The most challenging design criterion to meet was effective temperature control. The Fischer-Tropsch reaction is extremely exothermic. Using Hysys's ability to monitor the reactor temperature along the length of the reactor, it was found that in early iterations of the design, the temperature would spike as high as 1200° F within the first foot of the reactor. Attempts to raise the cooling rate only resulted in a linear decrease in reaction temperature throughout the reactor but failed to alleviate the temperature spike which occurred near the reactor entrance. This problem was solved by two methods.

First, it was determined that diluting the reactor feed would be an effective method of suppressing the initial temperature spike. Since the Fischer-Tropsch reaction is not an equilibrium reaction, and therefore Le Chatelier's Principle was not an issue, it was determined that a cost effective way to dilute the reactor feed was to add a recycle loop

to mix the fresh reactor feed with a portion of the reactor effluent. Since hydrocarbons have high heat capacities, this proved to be an effective method of temperature control.

The second method was to split the feed stream into multiple reaction trains. After 42 iterations, it was determined that effective temperature control, which kept the reaction temperature below 600°F, could be achieved with 20 reaction trains with two reactors in each train, designated PFR-100 and PFR-100-2. In addition the reactor's heat transfer performance was manipulated by changing the tube count and tube diameter (Figures 5 and 6).

Temperature as a Function of Reactor Length PFR-100

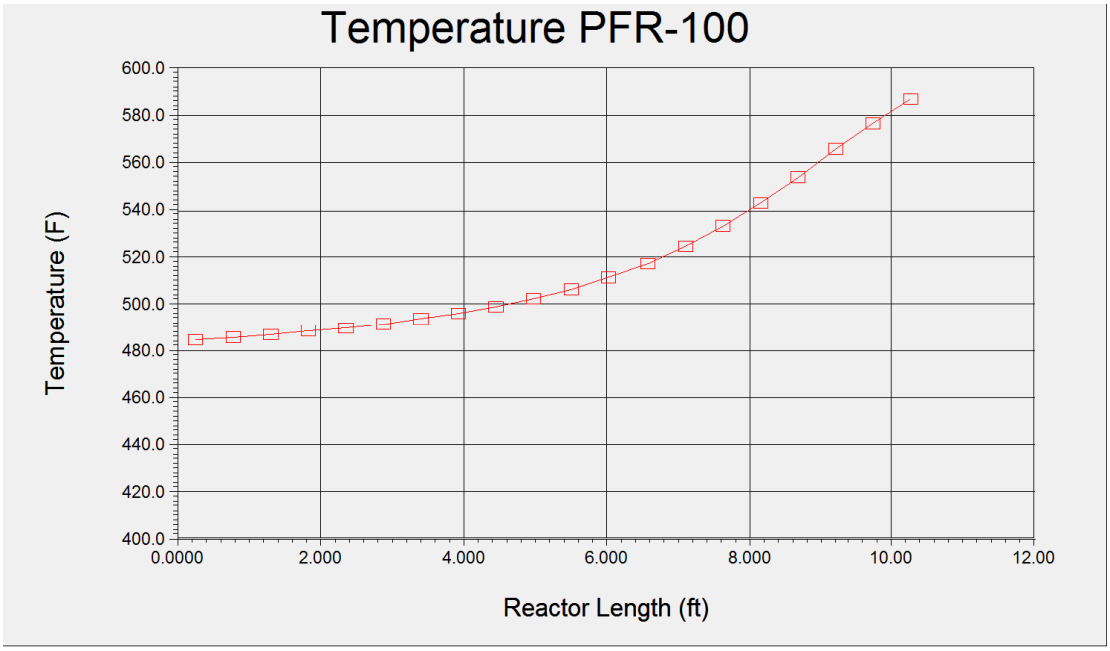


Figure 5

Temperature as a Function of Reactor Length PFR 100-2

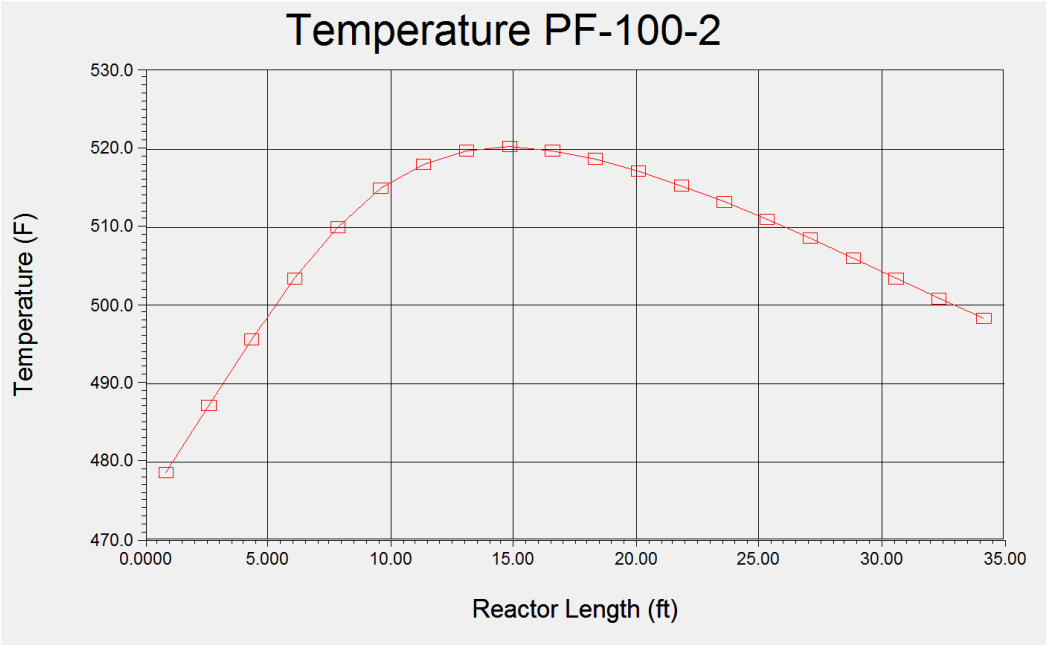


Figure 6

The second most challenging criterion to meet was keeping the pressure drop below 50 psi. Fortunately, attempts to control the temperature also resulted in lowering the pressure drop. The most effective method was splitting the feed stream. In addition, the pressure drop was lowered by decreasing the reactor length by using two reactors in series per reaction train instead of one long reactor and increasing the tube count. Moreover, manipulating the heat transfer rate also affected the vapor density in the reactor, which manipulated the pressure drop. Graphs demonstrating the correlation between pressure and reactor length are found in figures 7 and 8.

Pressure as a Function of Reactor Length (PFR-100)

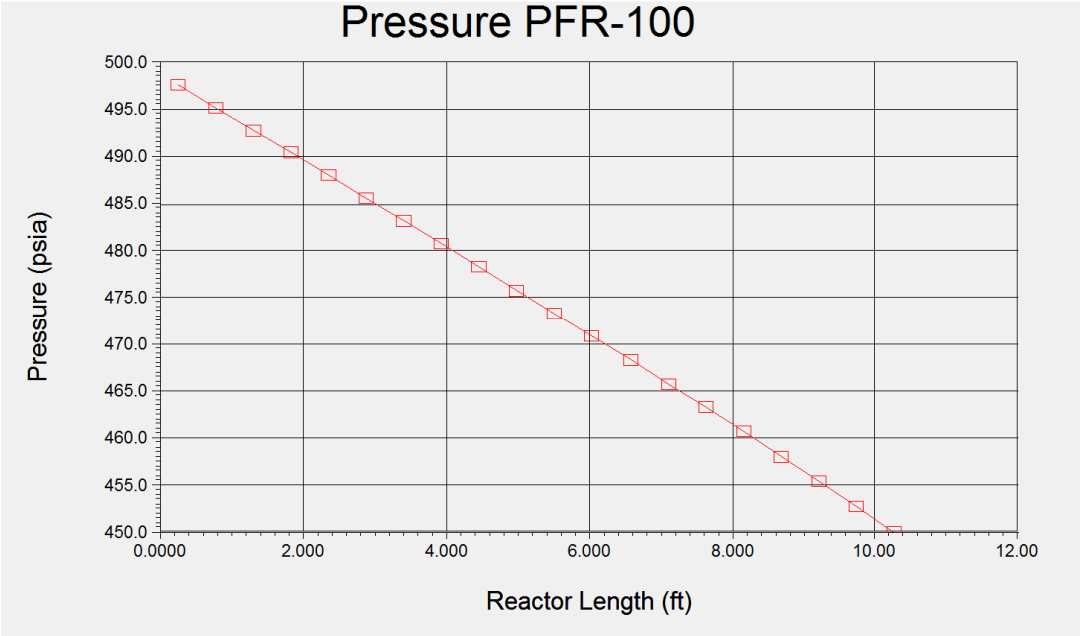


Figure 7

Pressure as a Function of Reactor Length (PFR-100-2)

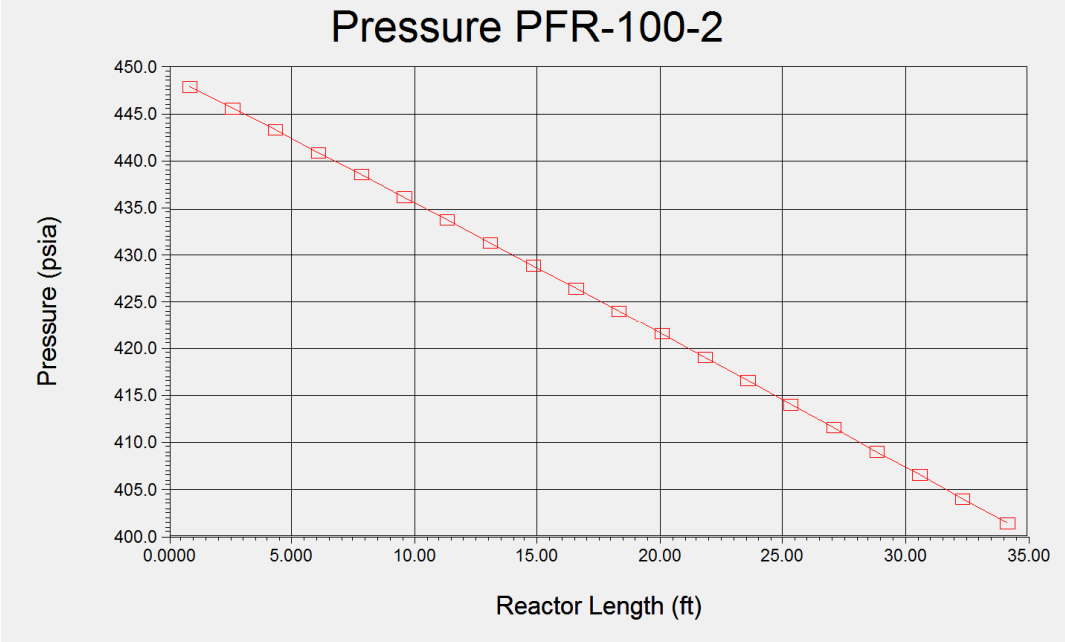
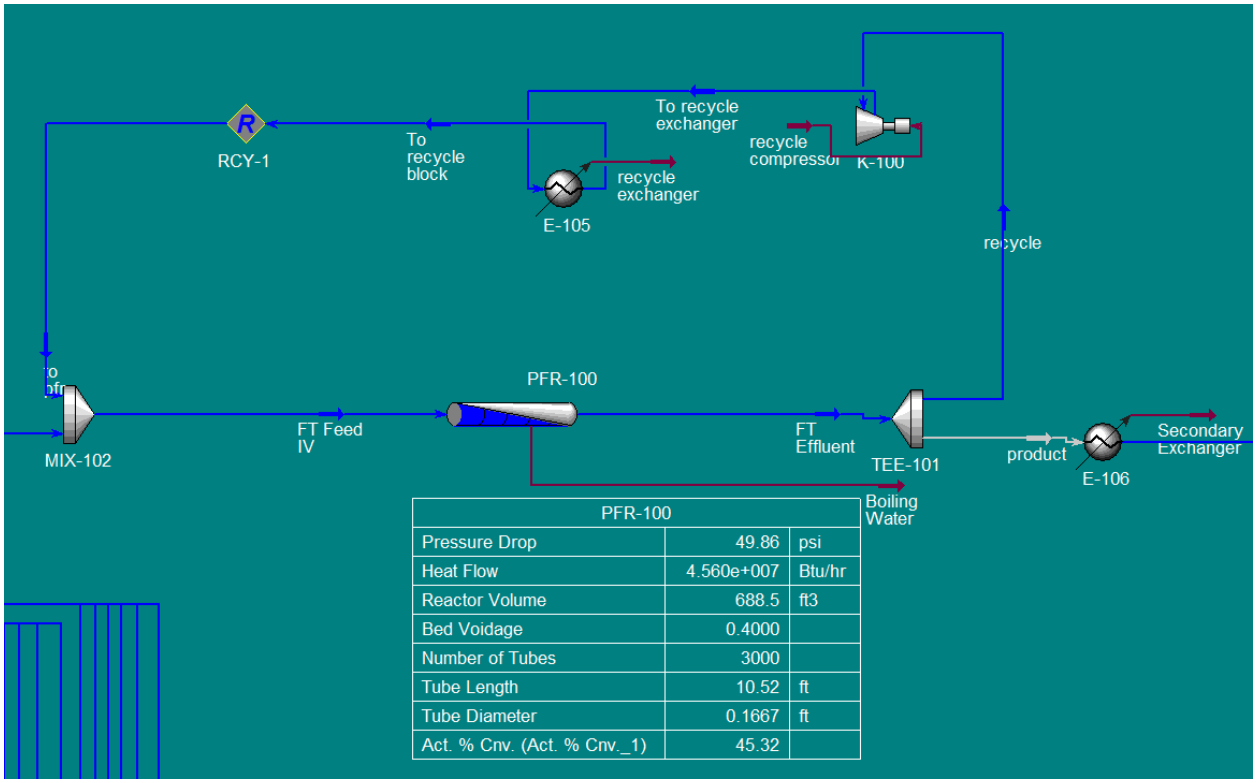


Figure 8

However, attempts to control the temperature and pressure came partially at the expense of the reaction yield. Diluting the feed decreased the reaction rate, lowering the yield. This was another reason why a second reactor was added to each reaction train. Over 644 barrels per hour of naptha and 8930 barrels of diesel per hour was generated. A summary of the reactor specifications can be found in figures 9 and 10 and tables 2 and 3 below.

PFR-100 Reactor Specifications



PFR-100		
Pressure Drop	49.86	psi
Heat Flow	4.560e+007	Btu/hr
Reactor Volume	688.5	ft3
Bed Voidage	0.4000	
Number of Tubes	3000	
Tube Length	10.52	ft
Tube Diameter	0.1667	ft
Act. % Cnv. (Act. % Cnv._1)	45.32	

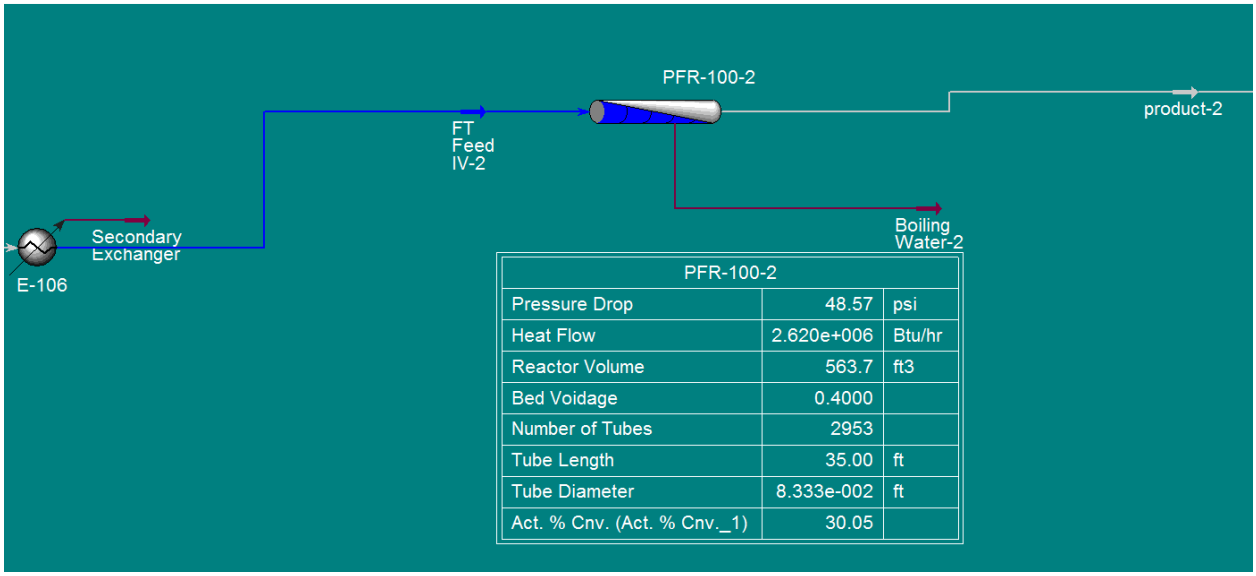
Figure 9

PFR-100 Reactor Specifications

PFR-100	
Pressure Drop	49.86 PSI
Heat Flow	4.560e+07
Tube Count	3000
Tube Length	10.52 ft.
Tube Diameter	2 in.
Shell Size	20 ft.
% Conversion	45.32

Table 2

PFR-100-2 Reactor Specifications



PFR-100-2		
Pressure Drop	48.57	psi
Heat Flow	2.620e+006	Btu/hr
Reactor Volume	563.7	ft3
Bed Voidage	0.4000	
Number of Tubes	2953	
Tube Length	35.00	ft
Tube Diameter	8.333e-002	ft
Act. % Cnv. (Act. % Cnv._1)	30.05	

Figure 10

PFR 100-2 Reactor Specifications

PFR-100	
Pressure Drop	48.57 PSI
Heat Flow	2.620e+06
Tube Count	2953
Tube Length	35 ft.
Tube Diameter	2 in.
Shell Size	10 ft.
% Conversion	30.05

Table 3

Separations

The reactor effluent was separated into a final product stream consisting of naphtha, and a stream for further processing containing diesel and paraffin waxes. The presence of tail gas and water in the effluent necessitated separation of these into respective streams as well. The water is purified for reuse while the tail gas, consisting in large part of burnable hydrocarbons, was recycled into the furnace to heat the feed stream of the syngas reactor.

The first step in the process was the removal of the water from the reactor effluent. This is accomplished by a series of 3-phase separators. The reactor effluent is cooled to a temperature of 280° F, and then fed into the first 3-phase separator, as shown in Figure 11. The effluent is cooled in an effort to condense as much of the water and the diesel as is possible, but is not cooled below280° because the paraffin waxes in the effluent stream can solidify below 250 °F.

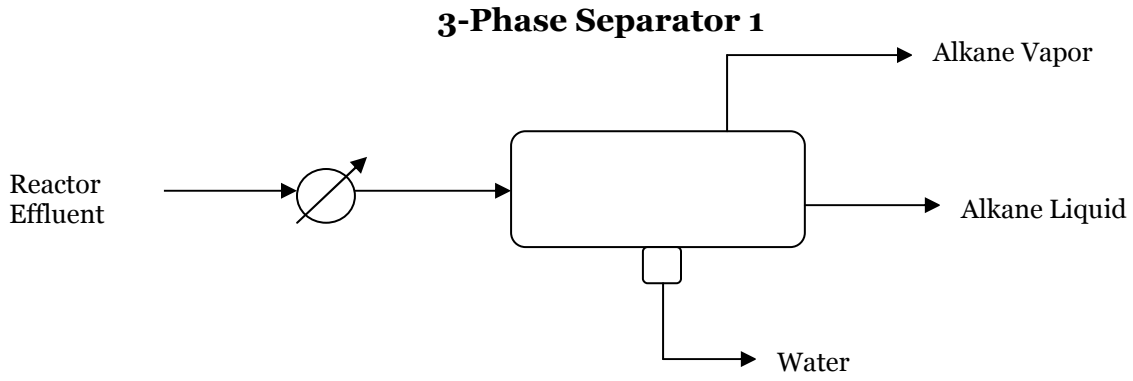


Figure 11

The alkane vapor stream from 3-phase separator 1 contains the remainder of the water from the effluent stream, and none of the paraffin wax from the effluent stream. The alkane liquid stream consists mostly of C6 and higher alkanes. The pressure on the alkane vapor stream is lowered by use of a throttle valve to 100 psi in order to vaporize the more volatile alkanes, and the stream is cooled 120° F to condense the water vapor. The alkane vapor stream is then fed into another 3-phase separator as seen in Figure 12.

3-Phase Separator 2

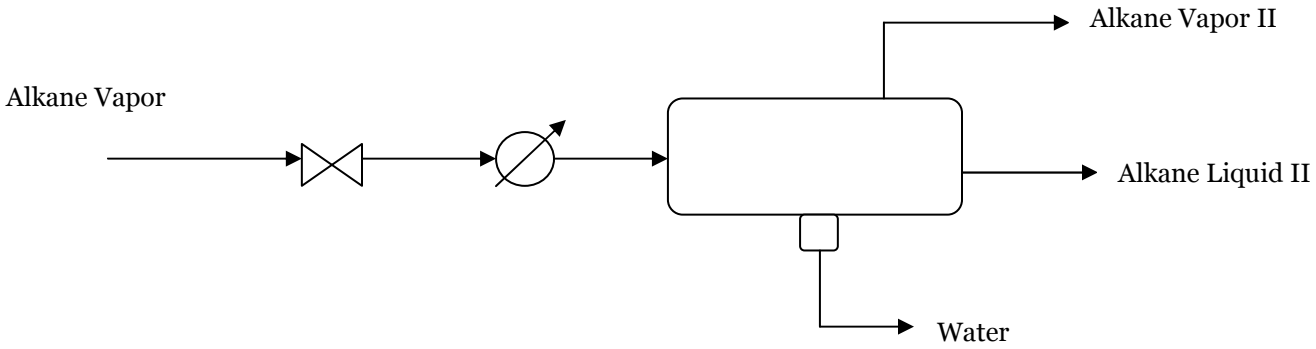


Figure 12

The alkane vapor II stream consists of the remaining water vapor, carbon dioxide, carbon monoxide, and the more volatile alkanes (C1-C6). The alkane liquid II stream consists of the less volatile (C6 +) alkanes. To remove the remainder of the water from the alkane vapor II stream, the stream is cooled by refrigeration to 35° F and fed into the last 3-phase separator, as seen in Figure 13. With this, the last of the water is removed from the alkane streams.

3-Phase Separator 3

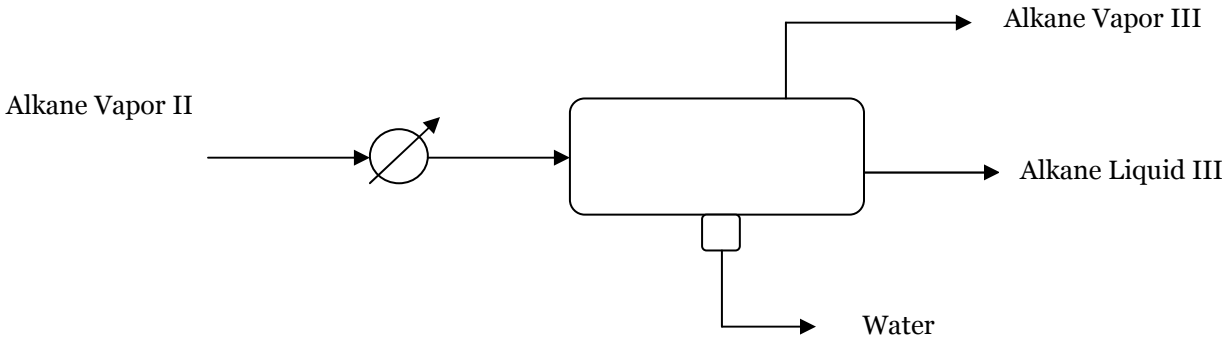


Figure 13

In order to separate the most volatile alkanes, C1-C4, it is necessary to cool the stream to -20° F. This was not done before the last 3-phase separator because the water would have frozen. The alkane vapor III and alkane liquid III streams are then combined into one stream and cooled to -20° F. This combined stream is fed into a 2-phase separator in which the most volatile alkanes (C1-C4) are separated from the less volatile alkanes (C5-C10) as in Figure 14.

2-Phase Separator

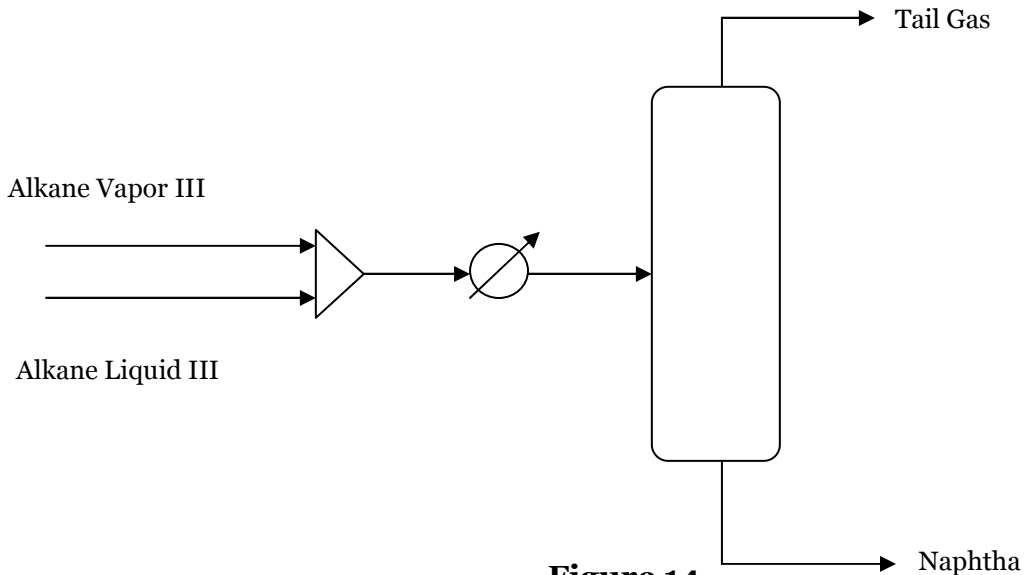


Figure 14

The tail gas consists of carbon dioxide, carbon monoxide, a small amount of hydrogen gas, and alkanes C1-C5. This stream is fed to the furnace used to heat the feed stream for the syngas reactor. The liquid stream from the 2-phase separator is naphtha, and can be sold as a final product. The water streams from the 3-phase separators is purified, and used to cool other streams.

The Alkane Liquid stream from the first separator, and the Alkane Liquid II stream from the second separator, are combined and used as a feed to a distillation tower as in Figure 15. The distillation tower specifications are in Table 4. The distillate from the tower is naphtha ready for resale, and the bottoms product is diesel and paraffin waxes that are ready for further processing. A complete flow sheet copied from the Hysys screen can be seen in Figure 16.

Distillation Tower

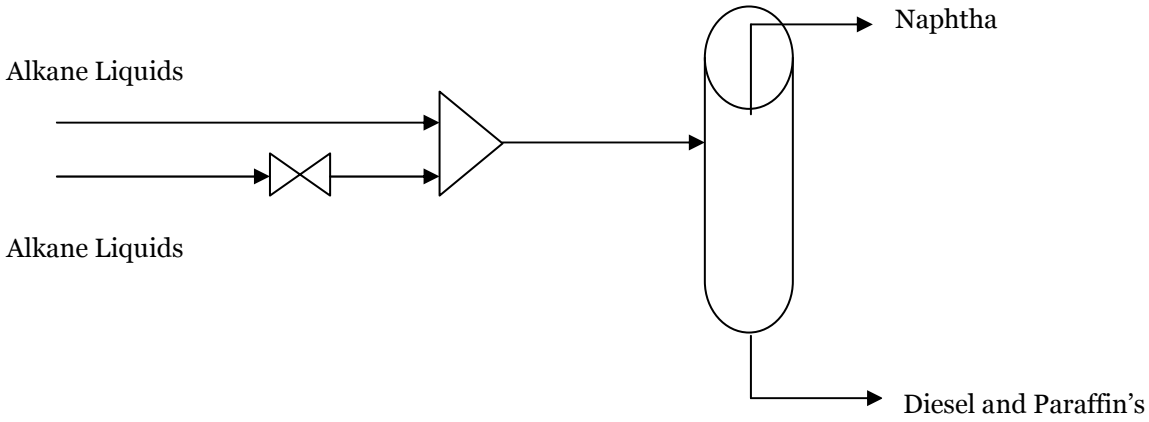


Figure 15

Distillation Tower Specifications

Diameter	4.921 ft
Number of Trays	40
Tray Spacing	1.804 ft
Reboiler Duty	1.141×10^8 btu/hr
Condenser Type	Full Reflux
Reflux Ratio	1.07
Reboil Ratio	4.264
HoldUp	3.12 ft^3

Table 4

Hysys Flow-Sheet for Separations

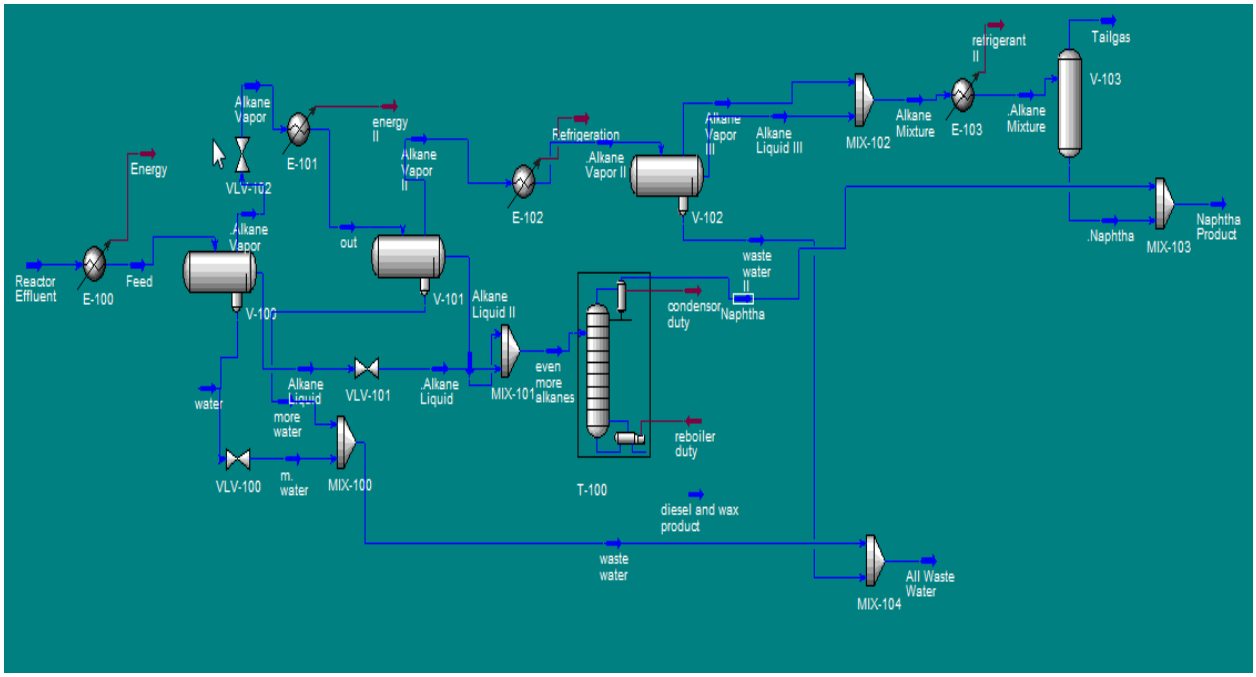


Figure 16

CONTROLS

Syngas and FTR Unit

The level should be controlled on the ERV-100 column by adjusting a valve on the liquid stream (bottoms product). The temperatures of all three streams entering the column ERV-100 should be controlled using the heat exchangers associated with each stream. A composition controller should be used to control the stream exiting the mixer "MIX-100" and should correspond to flow controllers on the CO_2 and O_2 streams which need to be in proper ratio. The three streams that directly enter the ERV-100 column should be kept in proper ratio with each other by the use of an appropriate ratio controller which will connect with basic flow rate controllers on each stream. The level of the V-100 column should be controlled by adjusting the FT feed stream coming out the top of the column. The temperature of the FT feed stream going into the column V-100 will be controlled by the exchanger "E-103". The temperature going into the splitter "TEE-100" will be maintained by the "E-104" heat exchanger.

PFR-100

The temperature of the PRF-100 should be controlled by how much of the product is being drawn out in the recycle stream from the TEE-100. Also, the flow going into the PFR will be kept under control by adjusting the flow of the stream going into the splitter TEE-100. Since the rest of the plant to this point will have controls, only some of which are mentioned here, the flow into the PFR will ultimately be controlled by the flow of the initial feed streams. Therefore, in a situation when the flow into the PFR must be decreased for a long period of time, there would eventually be a decrease of the feed streams into the plant.

The pressure will be controlled in part by the rate of boiling water that is being removed. Also, by controlling the temperature, as discussed above, the pressure will be controlled.

PFR-100-2:

The pressure in this PFR will partly be controlled by that rate the boiling water is being removed. Additionally, the temperature and pressure can be further controlled by the secondary heat exchanger "E-106". The flow can be increased or decreased by the rate at which the product is being drawn out of the PFR.

Separations

For the separations portion of the plant, as with the other portion, all of the flows, pressures, and temperatures, along with the compositions, which are closely linked with the flows, should be kept under control with the use of sensors and valves. Some of the more prominent control strategy points will be discussed as follows.

The temperature going into “V-100” will be controlled by the heat exchanger “E-100” and the pressure of “V-100” will be controlled with the flow of the vapor coming out the top of it. The composition of the alkanes stream which goes into mixer “MIX-101” will be controlled by adjusting the flow of the waste which mixes with water in “MIX-100” before being taken out as waste water. The temperature of the stream coming out the top of “V-100” which will go to “V-101” will be controlled by the heat exchanger “E-101” to bring it to the optimal temperature before entering “V-101”. The pressure of “V-101” should be controlled by the flowrate of the vapors coming out of the top. The temperature of the vapor stream which exits “V-101” and enters “V-102” will be brought to optimum specification by the heat exchanger “E-102”. The pressure of “V-102” should be controlled by the tops and bottoms products coming out of it. The composition of “T-100” should be controlled by the flowrate of the bottoms. The flowrate of the naptha product coming out of “MIX-103” should be controlled by the flowrate of the naptha product coming out of the top of “T-100”. The composition of the naptha product will be controlled in part by keeping the two streams entering “MIX-103” in the proper ratio with one another. The flow of the naptha stream is controlled as stated above, and the flow of the naptha III stream should be controlled by the flowrate of the tailgas that exits “MIX-102”.

Costing and Economics

Costing of the equipment was carried out using data from the HYSYS program in conjunction with costing tables from Gael Ulrich's *Chemical Engineering: Process Design and Economics, A Practical Guide.*

EQUIPMENT COSTS

Heat Exchangers and Plug Flow Reactors

Heat exchangers were priced using the following methods. First, the bare module cost C_{BM} was determined by multiplying the purchase cost C_P by the bare module cost factor, which is based on the material F_{BM}. As shown in Figure 17, in order to find the C_P, the surface area for the heat exchanger is needed. This area is calculated using the equation $Q = UA\Delta T$. The heat flow, Q, and the change in temperature, ΔT, are obtained from the HYSYS simulation. The overall heat transfer coefficient, U, is obtained from a table listing heat transfer coefficient values for various types of materials in a Shell and Tube heat exchanger (Ulrich, 205-208, 2004). Using these values, the surface area was calculated, and purchase cost C_P was obtained.

Shell and Tube Heat Exchangers Purchased Equipment Cost ($)

(Ulrich, 383, 2004)

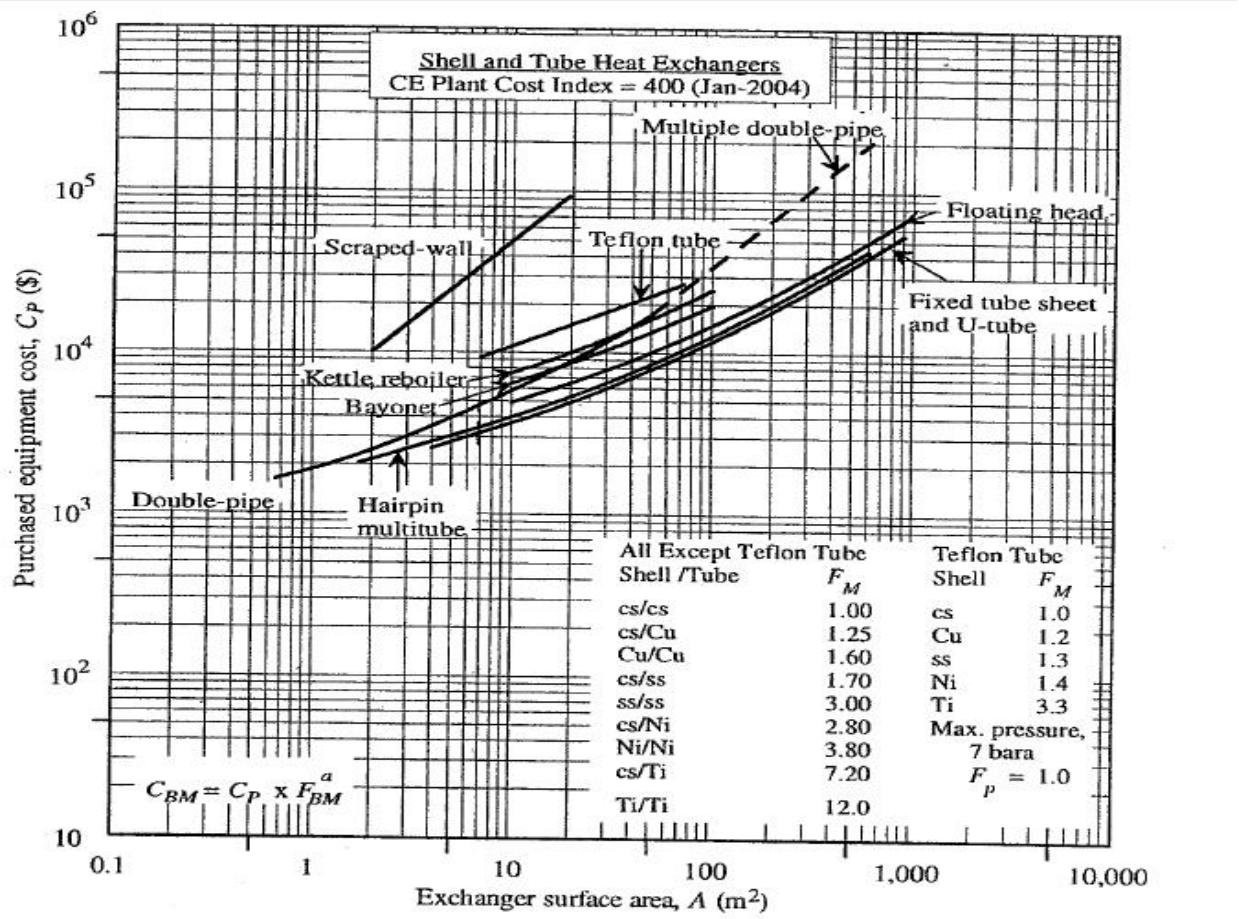


Figure 17

In order to cost a heat exchanger, the bare module cost factor, F_{BM}, is calculated as a product of the pressure factor, F_P, and the materials factor, F_M. Using the pressure of the stream going through the heat exchanger, the pressure factor is found with the chart in Figure 18.

Heat Exchangers Pressure Factor

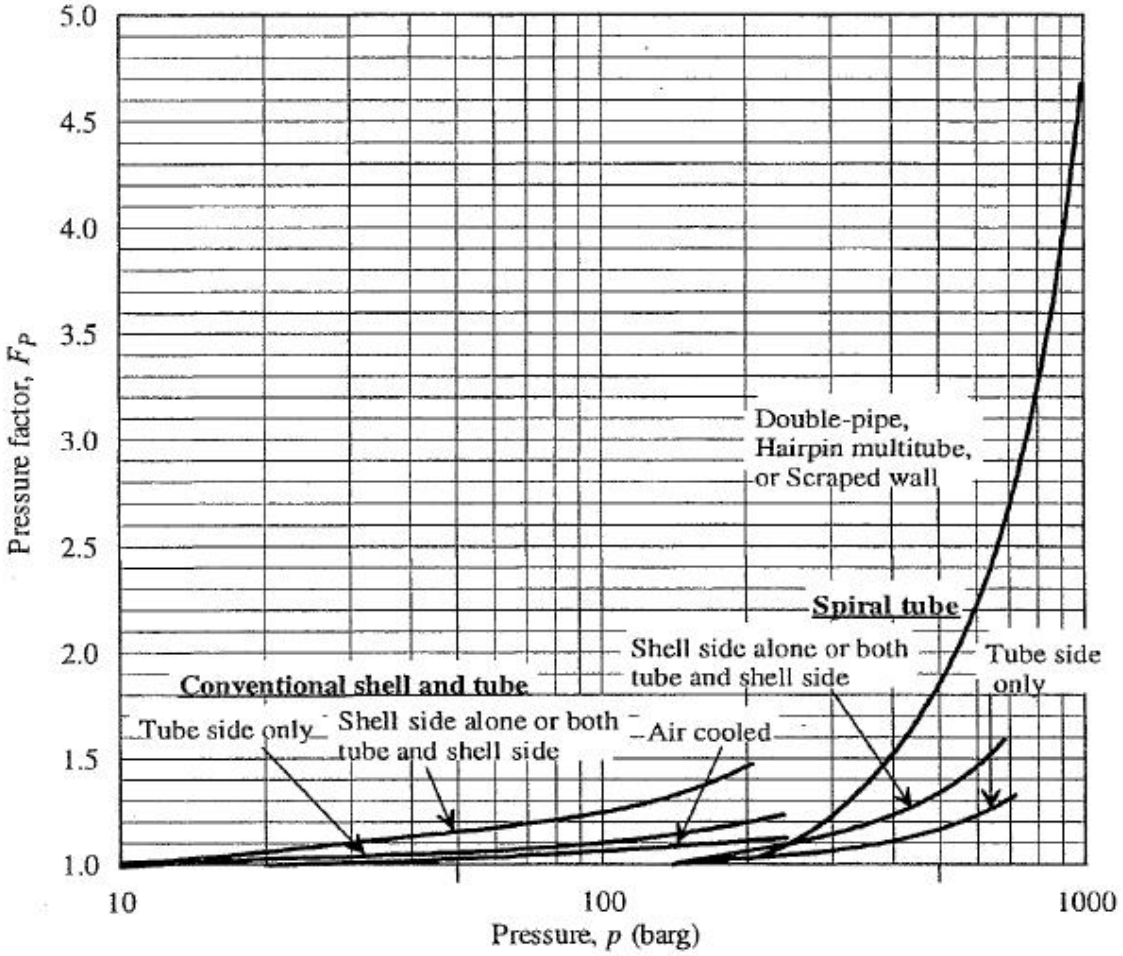


Figure 18

Using the value for the product of F_P and F_M, Figure 19, below, is used to determine the bare module cost factor F_{BM} of the heat exchanger.

The C_{BM} can be calculated by multiplying the F_{BM} and the C_P.

Heat Exchangers Bare Module Factor

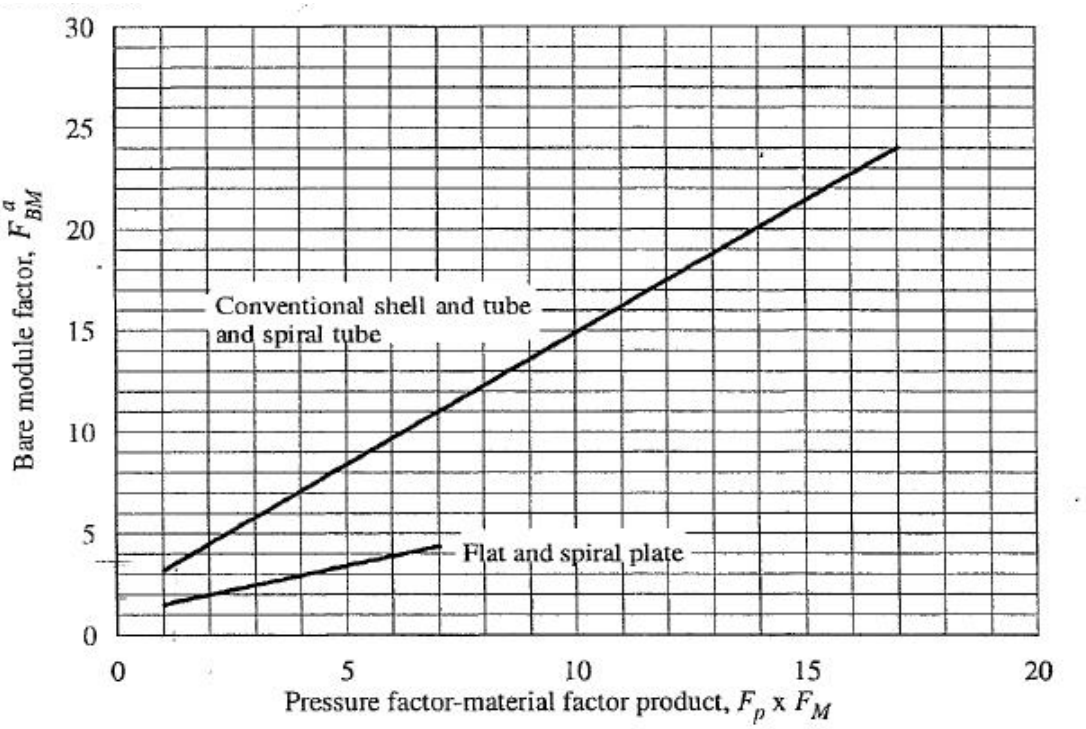


Figure 19

Plug flow reactors were costed in the same manner, with the addition of the catalyst cost. This was done by subtracting the void volume from the total volume in order to get the catalyst volume. The catalyst volume was then multiplied by the given bulk density, resulting in the catalyst mass. The mass was then multiplied by the given cost for the catalyst, $10.00/lb.

Process Vessels

Process vessels were costed by dividing the mass flow rate by the density in order to obtain the volume. This volume was then put into the HYSYS file in order to obtain the height and density of the vessel. The purchased equipment cost was then obtained from figure 20, below.

Process Vessels Purchased Equipment Cost ($)

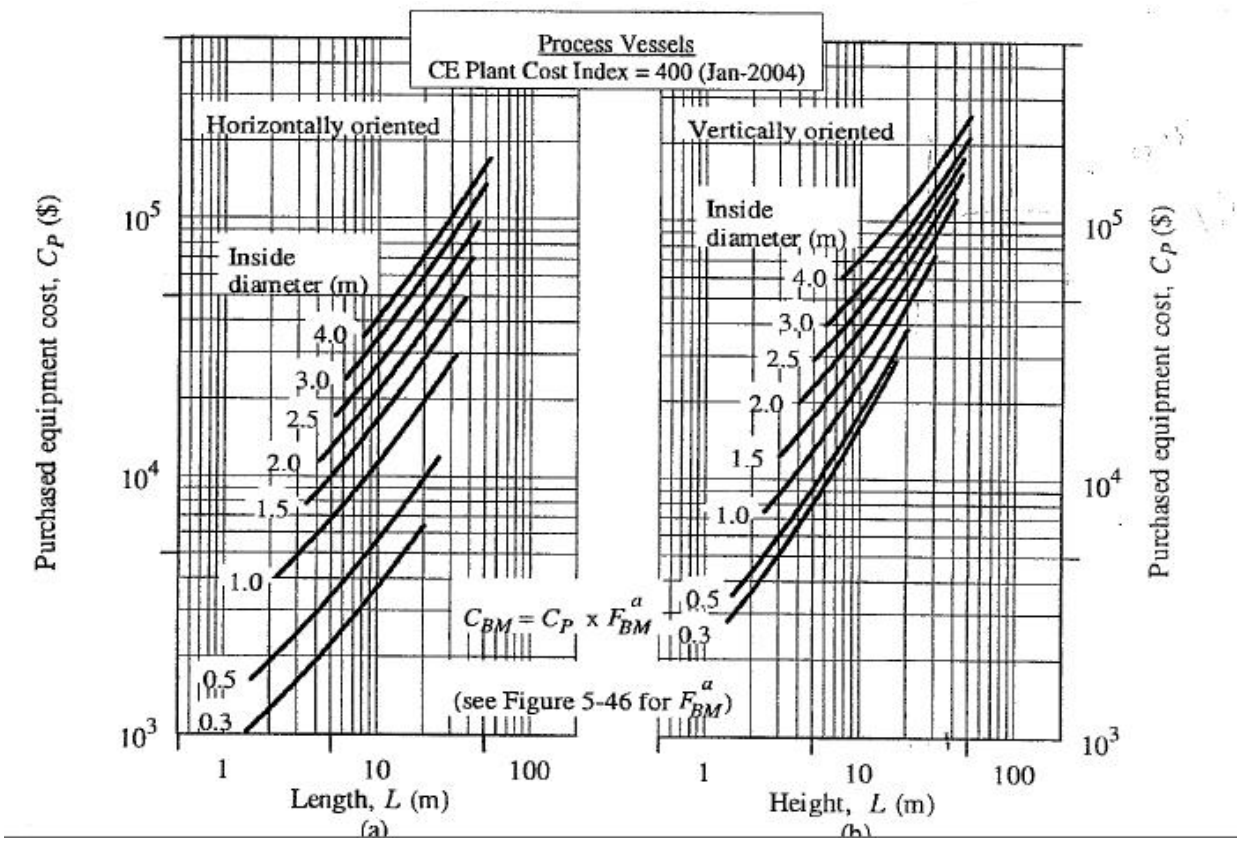


Figure 20

The pressure factor was obtained from figure 21, and the bare module factor was obtained from figure 22. The bare module cost was then found by multiplying the purchased cost and the bare module factor, similar to the process for obtaining the cost of heat exchangers and PFRs.

Process Vessel Pressure Factor

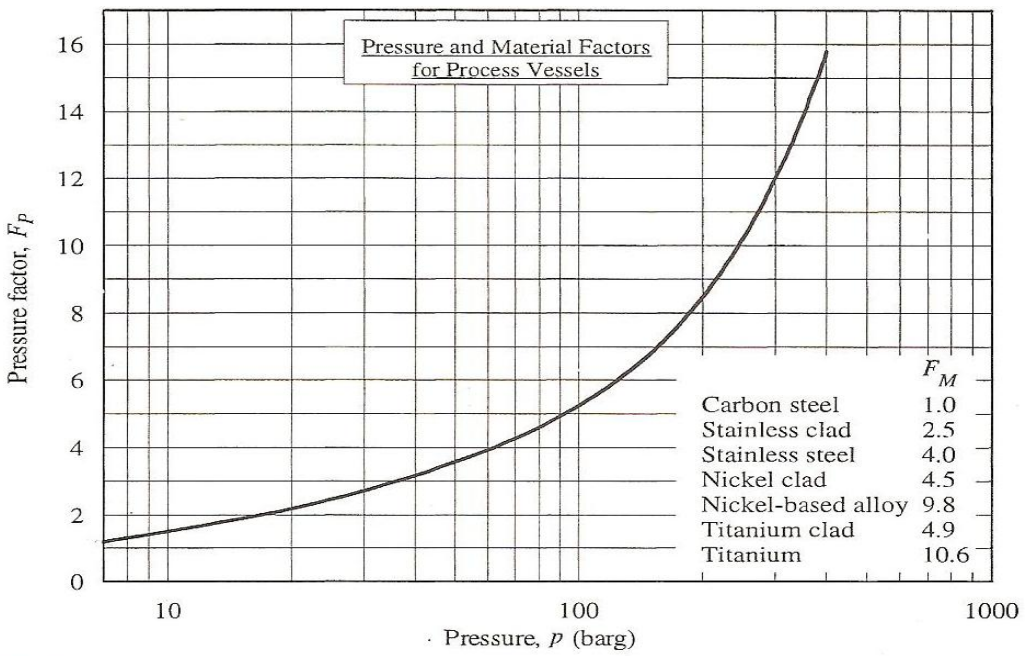


Figure 21

Process Vessel Bare Module Factor

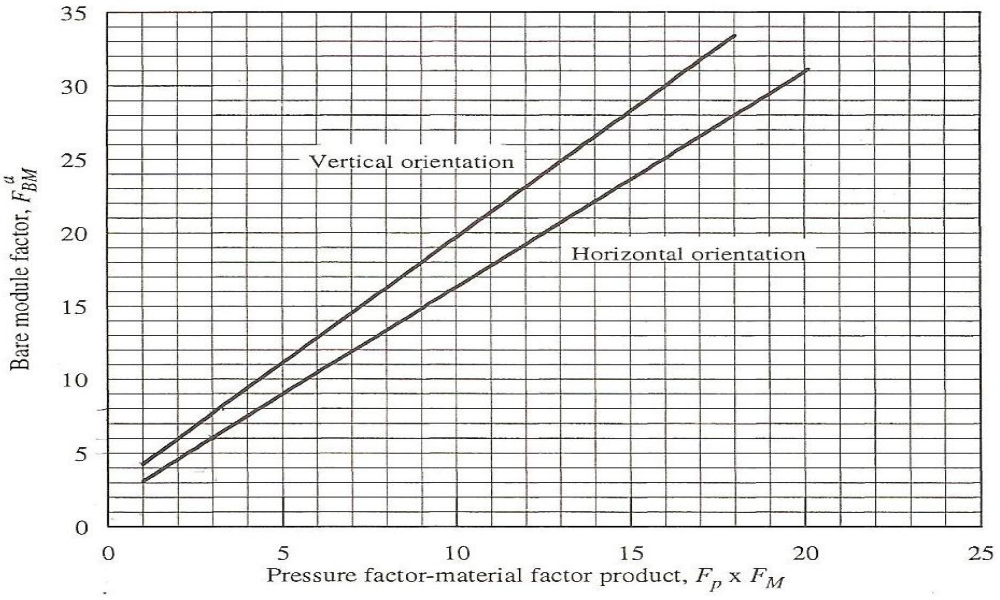


Figure 22

Compressor

The compressor was also costed in a manner similar to the way heat exchangers were costed. The purchased cost was determined from figure 23. The fluid power was found with the equation located below the x-axis. The type used was centrifugal, due to its cost and durability. The bare module factor was then multiplied by the purchased cost in order to obtain the bare module cost.

Compressor Purchased Cost

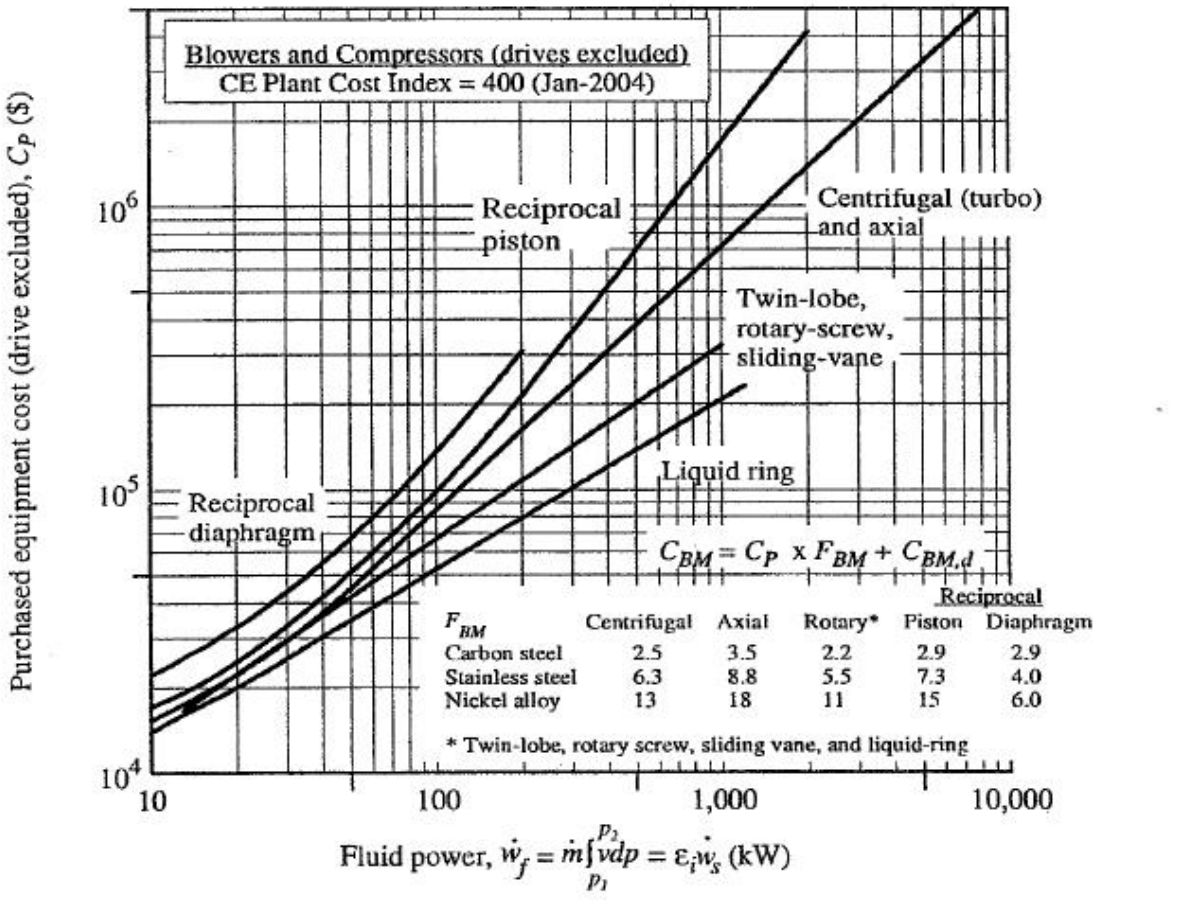


Figure 23

A summary of the total costs associated with the syngas unit and Fischer-Tropsch reactor, along with the separations unit can be found in Appendix A.

Material and Utility Costs

The four components going into the syngas unit are methane, steam, carbon dioxide, and oxygen.

The cost of the methane feed is $2000/MSCF (million standard cubic feet). The flow of the methane stream into the system is 1,334.40 cubic meters/hr, or 47,124.89 cubic feet/hr, resulting in a cost of $94.25 per hour.

The cost for 100% carbon dioxide feed is $400/MSCF. The flow of carbon dioxide is 266.06 cubic meters/hr, or 9395.82 cubic feet/hr. This results in a cost of $3.76 per hour.

The cost of oxygen is $100/short ton (2000lb). The flow of oxygen into the system is 672,000 lb/hr, which is 336 short tons/hr. This results in a cost of $33,600 per hour.

Because there are 20 reaction trains, each with its own compressor, electricity is a major cost factor within the plant. However, using an expander powered by waste steam, electricity demands of the plant were met, and excess electricity can be sold.

All steam demands were met using 20 pound steam for both heating process streams and cooling the packed bed reactors. The steam was generated by using our waste water to cool various process streams. Therefore, no 20 pound steam was purchased. However, in order to meet the air separation unit requirements, 3.4 million lbs/hr of 600 pound steam was purchased to run the compressor.

The majority of cooling water needs was met using water recovered from the separations unit. Make-up cooling water was purchased to meet additional cooling needs and to generate additional 20 pound steam in order to cool the PFRs and to sell as revenue.

Fuel gas needs were met by recovering the tail gas from the separations unit. By using the lower heating value of this fuel, the needs of the feed preheat furnace were met, and the surplus fuel gas was sold as an additional revenue stream.

A summary of the total costs associated with utilities can be found in Appendix A.

Overall Cost/Total Revenue

The total fixed cost of the entire plant is $72,331,000. This consists of all equipment costs. Straight line depreciation was assumed using the specified 15 year project life and a salvage value of zero. Because the catalyst must be replaced every four years, and given an inflation rate of 3%, the three turnaround costs were annualized using the following formula:

$$P_j = \frac{F_j}{(1+i)^n}$$

where j=1,2, and 3 and n=4,8, and 12.

The annual revenue before taxes was calculated by taking the value of the product, plus utility credits, minus the feed stock costs, the yearly operating expenses, the annualized turnaround costs and depreciation. Taxes were calculated as specified, 33% of this value.

Conclusions and Recommendations

All project criteria were met using a Fischer-Tropsch Reactor system with 20 trains, containing a total of 40 packed bed reactors. Over 644 barrels per hour of naptha and 8930 barrels of diesel per hour was generated. Total fixed costs for this project were $72,300,000 and annual revenue generated was $147,200,000,000. The main challenges of this project included the highly exothermic nature of the Fischer-Tropsch reaction, as well as the criteria stipulated in the project description of a reactor diameter of less than 20 feet, reactor length of no greater than 60 feet, and a pressure drop of no greater than 60 psi. The combination of these constraints, the highly exothermic nature of the reaction and the high feed rate, necessitated the splitting of the feed stream into 20 identical reaction trains containing a total of 40 reactors. This led to a high fixed capital cost, as well as significant energy consumption.

In order for this process to be profitable, it was necessary to utilize as much waste as possible, namely using the steam created from wastewater, harnessing the tail gas as fuel, and utilizing waste steam to generate electricity. It was through these methods rather than the value of the desired products, naptha and diesel, that this plant was ultimately profitable.

The chief recommendations in order to lower the fixed capital cost and energy consumption in the project would be to eliminate the size restrictions on the FTR reactor and to allow for a larger pressure drop. These would allow for larger vessels with more effective heat transfer and could reduce the number of reactor units, thus greatly reducing both fixed and operating costs.

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