Date: April 11, 2022
To: American Institute of Chemical Engineers
From: Kelsey DePrang, Karissa Huffman, and Aubrey Norton

To whom it may concern,

The AIChE 2021-2022 Student Design Competition was assigned in January 2022 and was submitted April 2022. The purpose of the project was to design three different-sized modular processes that convert natural gas into hydrocarbons that can be repurposed instead of releasing harmful gases into the atmosphere. Once the modular units were designed, the next step was to optimize the deployment of the modules to various wellheads over the 20-year timeframe. The goal was to reduce environmental impacts while making profit off the new hydrocarbon products.

This modular gas-to-liquids synthesis consisted of a Syngas Unit that takes in natural gas, steam, CO_2, and O_2 to form syngas. The syngas was then sent to a Fischer-Tropsch Reactor that converts the syngas into multiple hydrocarbon chains. The reactor was designed as a fixed bed tubular reactor with a Cobalt-based catalyst. These hydrocarbons are sent on to the separation unit to be separated into their respective product streams. The heavy distillates/waxes are sent on to the Hydro-isomerization unit to be converted into lighter products to be sold or repurposed. The products made include tail gas, LPG, naphtha, water, paraffin, and diesel fuel. The products that will be sold are LPG, naphtha, and diesel fuel.

The results of the project show that the present worth cost (PWC) is approximately $4.69 B and the net present value (NPV) is approximately $4.35 B. The discounted cash flow rate of return (DCFROR) is 9.10% with a minimum rate of return of 8%. This indicates that the project is successful but has chances for error. The close range indicates that the project has a potential to have the DCFROR decrease below 8%, however it reduces carbon footprint of harmful gases and should still be pursued by the company.

Sincerely,

Kelsey DePrang, Karissa Huffman, and Aubrey Norton

2021-2022 AIChE Design Competition Report

Modular Distributed Gas-to-Liquids Synthesis

April 11, 2022

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Executive Summary

AIChE has challenged teams with the task of designing three sizes of modular Gas-To-Liquids systems with the purpose of reducing harmful gas emissions into the atmosphere. This task was accomplished by using Fischer-Tropsch synthesis which turns syngas into hydrocarbon chains that can be repurposed and sold for a profit. These products include LPG, naphtha, and diesel fuel. The project also included optimizing the deployment of the different sizes of the modular units. The solution was designed to maximize the NPV and DCFROR as well as mitigate risks to the environment and to a person's health and safety.

Three sizes of modular units are predefined to have natural gas feeds of 500 MSCF/day, 2,500 MSCF/day, and 5,000 MSCF/day. It was assumed that a Syngas Unit, an Air Separation Plant, a Steam Plant, a CO_2 Recovery System were already designed for each modular size and only needed estimates of capital cost and utility costs for the total cost estimate. The focus for this preliminary design was a vertical Fischer-Tropsch reactor and a separations unit including one two-phase separator and two three-phase separators.

This design poses a small risk for explosion or damage to surrounding areas because of its modular size. Units will be transported between multiple well sites and therefore caution should be taken to not improperly assemble units after transportation. Hydrocarbons are fire hazards therefore safety measures were put in place to mitigate the risks.

It was determined that the optimal number of units needed for the project would be 13 small units, 9 medium units, and 5 large units. The capital costs of each small, medium, and large unit were $13.3 MM, $36.9 MM, and $ 50.91 MM, respectively. Annual utilities were estimated to be $1.45 MM, $8.60 MM, and $14.8 MM for the small, medium, and large units, respectively. Annual revenue for the small, medium, and large units were calculated to be $0.54 MM, $3.66 MM, and $5.79 MM, respectively. Four five-year truck contracts will be needed for the duration of the 20-year project life consisting of 10, 9, 8, and 6 trucks for the individual contracts. It was estimated that the modular units would be built January of 2024 and deployed December of 2024 to begin operating in January of 2025. Estimating a build date of January 2022 allowed for 6 to 12 months of detailed design as well as 6 months for bidding of equipment and an additional 6 months to obtain required permits before units are built. A positive net present value of $4.35 B was achieved for this project. An internal rate of return was calculated to be 9.1% which is greater than the 8% hurdle rate given for the project. The payback period was calculated to be 6 years and 9 months which would occur September of 2030. The project NPV and DCFROR show that the project design is economically attractive.

Introduction

Oil production has historically had catastrophic environmental impacts, with flaring of natural gas being one. Natural gas is mainly composed of methane and, when flared, is converted to carbon dioxide and released into the atmosphere. This process occurs, since methane has almost 25 times the greenhouse contribution than carbon dioxide. However, this process has resulted in the accumulation of carbon dioxide in the Earth's atmosphere with associated environmental impacts. The U.S. Energy Information Administration (EIA) reported that approximately 430,000 oil wells produced about 6.5 trillion cubic feet (tcf) of gases in 2019. The flaring of natural gas has contributed to about 111,000 metric tons of atmospheric carbon dioxide per year.

However, an alternative method for natural gas is currently being considered. To minimize the environmental impact, it is possible to convert the natural gases into liquids using Fischer-Tropsch Liquid (FTL) technology. The scope of this project included the conversion of methane into longer hydrocarbon liquids to prevent flaring of natural gas and produce additional profit. A modular approach was taken into consideration to make the system scalable and easy to move. The system consisted of three module sizes capable of a feed of 500 MSCF/day (small unit), 2,500 MSCF/day (medium unit), and 5,000 MSCF/day (large unit) of natural gas. The Syngas Unit, the Hydro-isomerization Unit (HIU), the Air Separation Unit, the CO_2 Separation System, and the Steam System have already been designed. The overall goal of the project included two parts. The first was to optimize the design of the Fischer-Tropsch reactor (FTR) and the separation equipment The second was the optimization of equipment deployed over the project lifetime at various wellhead sites.

Summary

This report covers topics ranging from design specification and practices to a sensitivity analysis of the various costs needed to make this project successful as specified in the preliminary design statement. The main driving factors of this project were to reduce the carbon footprint of CO_2 by converting natural gas into diesel and naphtha that could be sold for a profit instead of being flared. Optimizing time and location of different modules deployed at the wellheads was another driving factor for increasing the project's net present value (NPV). The project timeline spans 20 years, concluding in the year 2044.

Safety and environmental analyses were done to make sure the designed process minimized as much risk as possible. Inherently safe decisions were made throughout the preliminary design process to reduce costs and to minimize risks. A Process Hazard Analysis (PHA) was done which includes: the worst-case scenario of damage, a P&ID of the separation unit, and potential process hazards of the materials being used.

Based on the preliminary design economic results, this project generates a NPV of $4.35 B, breaking even in September of 2030. As such, it should be pursued. The sources of errors were analyzed for the design to clarify potential downfalls to the economic analysis. This included performing two sensitivity analyses with the Monte-Carlo Method and a Tornado Chart.

Project Premises

The entire Fischer-Tropsch (FT) synthesis process requires many specifications to run properly. Each unit had its own set of specifications that needed to be followed as given by AIChE design statement. The goal of the upstream process was to produce syngas. The syngas reactor required a temperature range of 1600-1950 °F and a pressure range of 300-500 psig. The steam and natural gas entering the process needed to be at a minimum of 0.5 mol/mol ratio to prevent coking in the feed preheater. The syngas product was specified to have H_2 and CO at a 2:1 ratio to equal the consumption ratio in the FTR. The maximum feed preheat temperature was set to be no more than 1000 °F. The Air Separation Plant also provided O_2 for the Syngas Unit. The O_2 was supplied at 99% purity with 1% mol nitrogen.

The design portion of the project included the FTR and the separation unit. It was assumed that the vapor phase contained 100% of the H_2, CO, H_2O, CH_4, N_2, and 0.7 moles of C2+ hydrocarbons for every mole of methane made. The maximum allowable pressure drop per reactor was 50 psi. The separator unit could not have the temperature of the distillate and heavier boiling fractions be lower than 250 °F because it could lead to wax crystallization.

The downstream process included the Hydro-Isomerization (HI) Unit where the wax was converted to lighter products and the paraffins are isomerized. This process converted 100% of the hydrocarbons with a boiling point greater than 700 °F to lighter materials with the given overall selectivities of 1.0 wt% methane, 0.5 wt% ethane, 3.5 wt% propane, 3.5 wt% butane, 25 wt% naphtha, and the balanced diesel. The free water could not be above the solubility limit that was fed to the HI unit. The makeup gas could not have a CO content higher than 0.1 mol%.

The FTR units will be deployed to wellheads located various distances from the central plant where the products will be transported for further processing and be blended into finished fuels. The amount of gas produced by each well and the distances from the central plant are listed in Table 1.

Table 1: Wellhead Information

Associated Gas Well ID	Initial Associated Gas Production (MSCF/day)	Distance from Central Plant (Miles)
1A	4,494	62.0
1B	2,989	21.0
1C	8,365	17.0
1D	7,447	73.4
1E	3,290	6.2
1F	14,737	91.0
1G	11,155	59.0
1H	7,081	89.3
2A	9,182	85.8
2B	12,258	84.7
2C	8,742	2.4
2D	3,840	42.0
2E	7,939	98.8
2F	2,874	13.0
2G	6,076	87.0
2H	6,468	36.0

Discussion

a. Design Philosophy

Driving factors for the design of this project included a positive NPV and DCFROR above the 8% hurdle rate, creating modular units that meet specifications, optimize wellhead placement, and a process that can be operated safely as well as sustainably. Major assumptions and key design elements implemented to reach objectives are listed below:

- Maximize the amount of diesel produced
- Optimize product of carbon chains of twenty-one carbons and above out of the FTR
- Optimize the inlet flowrate of carbon chains into the HIU
- Minimize amount of catalyst needed in the FTR
- Minimize equipment size to fit shipping containers for easy transfer
- Remove excess water and carbon dioxide

b. Description of the Process

Natural gas (assumed to be pure methane) is mixed with CO_2 and high-pressure steam. A minimum 2:1 mole ratio is kept between natural gas and steam. The mixture is then sent to a preheater and heated to a temperature of 1000 °F. O_2 from the Air Separation Plant is added after the preheater and supplied at a rate that allows for a 2:1 ratio of hydrogen and carbon dioxide to be produced from the syngas reactors. Next, the stream enters the Syngas Reactor System by first entering the Steam Methane Reforming (SMR) reactor and then into the Water Gas Shift (WGS) reactor. For the purposes of this project, the SMR and WGS reactors were modeled as one reactor. The Syngas Unit operated close to 1600 °F and 450 psi. After the Syngas Unit, the stream was cooled to 212 °F and entered a separator to remove water from the mixture. The stream was then sent to an absorber to remove excess carbon dioxide before being sent to the FTR unit. Removing water and carbon dioxide allows for a smaller FTR to be designed.

The FTR was designed to be a fixed bed tubular reactor that uses a Cobalt-based catalyst. The FTR unit is ran at the max temperature of 450 °F to optimize conversion and the amount of higher carbon chains in the product stream. The product stream from the FTR is then sent to a cooler. The product stream is cooled to 275 °F before entering a two-phase separator. The bottom stream of the two-phase separator is heated to 420 °F by a heater before entering a three-phase separator. This three-phase separator removes wastewater from the bottom and LPG lighter gases out the top so that a heavy carbon chain stream can be sent to the HIU. The LPG product stream is then cooled down to forma liquid. The top vapor stream from the two-phase separator goes into a three-phase separator where tail gas, naphtha, and another wastewater stream are separated.

c. Design Practices

i. Syngas Unit

The Syngas Unit was modeled using Aspen HYSYS. The fluid package used for simulation was CPA. Each size of unit was simulated then optimized using the given specifications. The given natural gas rates of 500 MSCF/day, 2500 MSCF/day, and 5000 MSCF/day were converted to lbmole/hr then put into a stream at the given temperatures and pressures. The natural gas stream was mixed with high pressure (HP) steam and CO_2 which was then heated before adding the purified air stream. The incoming steam and natural gas were kept at a minimum of 0.5 molar ratio to prevent coking in the syngas feed preheater. The mixed stream after the preheater was sent into the Syngas Reactor System which was modeled as a Gibbs reactor in Aspen HYSYS for simplicity [1]. The steam reforming and water gas shift reactions are in equilibrium while the partial oxidation reaction goes to completion. A Ln(Keq) equation was inputted into the reaction setup for the steam reforming and water gas shift reactions [2]. The partial oxidation reaction was modeled as if it went to equilibrium by using an extremely high Keq value (100,000) to favor a forward reaction.

After determining the amount of methane and steam entering the reactor, the O_2 stream flow rate was adjusted to achieve the temperature required for the overall composition of the synthesis gas. The O_2 for the process is purchased from a third-party packaged O_2 plant. The Air Separation Plant provides a 99% pure supply of O_2 with 1 mol% of nitrogen. The O_2 enters the syngas system at 500 psig and 75°F. The flowrate of CO_2 was then supplemented to the mixed stream to meet the specification of a 2:1 molar ratio of H_2 and CO that would be entering the FTR. The flows of O_2 and CO_2 were optimized by using a case study in HYSYS. The molar flow rate of CO_2 and O_2 were set to be the independent variables while the molar flow of H_2 and CO in the syngas were the dependent variables. The results of each case study were exported to Excel and organized to find the cases where the molar ratio of H_2/CO would be 2. The flowrates that had the lowest O_2 rate with the temperature needed (~1600°F) were ultimately chosen as the initial flowrates of O_2 and CO_2. The lower O_2 rate minimize the cost since it was the more expensive reactant. The inlet flowrates to the syngas reactor are listed in Table 2.

Table 2: Inlet Flowrates and into Syngas Reactor

Component	Small Unit Molar Flowrate (lbmol/hr)	Medium Unit Molar Flowrate (lbmol/hr)	Large Unit Molar Flowrate (lbmol/hr)
Natural Gas (Methane)	54.02	270.12	540.20
Steam	27.01	135.05	270.10
Carbon Dioxide	11.00	45.00	105.00
Oxygen	20.79	123.75	216.81
Nitrogen	0.21	1.25	2.19

lists the outlet flowrates of the syngas reactor that were modeled in Aspen HYSYS. Approximately 64.8%, 77.5%, and 67.7% of methane was converted in the syngas reactor for the small, medium, and large units, respectively. The H_2 to CO molar ratio was approximately 2:1 for the small, medium, and large units.

Table 3 lists the outlet flowrates of the syngas reactor that were modeled in Aspen HYSYS. Approximately 64.8%, 77.5%, and 67.7% of methane was converted in the syngas reactor for the small, medium, and large units, respectively. The H_2 to CO molar ratio was approximately 2:1 for the small, medium, and large units.

Table 3: Outlet Flowrates of the Syngas Reactor

Component	Small Unit Molar Flowrate (lbmol/hr)	Medium Unit Molar Flowrate (lbmol/hr)	Large Unit Molar Flowrate (lbmol/hr)
Natural Gas (Methane)	18.99	60.66	174.29
Carbon Monoxide	32.86	196.76	343.33
Hydrogen	65.67	393.57	686.71
Steam	31.39	160.40	315.22
Carbon Dioxide	13.17	57.69	127.59
Nitrogen	0.21	1.25	2.19

After the Gibbs reactor, the syngas stream was approximately 1600°F and was sent to a cooler and cooled down to the boiling point of water (212°F). Most of the water was removed from the stream by being sent to a two-phase separator. The syngas stream also needed to have CO_2 removed before being sent on to the FTR to reduce the size, so a splitter in HYSYS was used to simulate the CO_2 separation unit removing 95% of the CO_2. The removed CO_2 syngas stream was heated up to 450°F and sent to the FTR unit.

ii. Fischer-Tropsch Reactor

The Fischer-Tropsch reactor converts syngas into hydrocarbons and water. The reactants are hydrogen and carbon monoxide and can be described by Equation 1.

$$\mathrm{xH_2 + CO \rightarrow H_2O + (CH_2)_nH_2} \qquad \text{(Equation 1)}$$

It has been determined that the reaction can be modeled by a Langmuir-Hinshelwood form of the overall rate equation. It has been modeled as the rate of conversion of CO and is expressed by Equation 2.

$$-r_{CO} = \frac{(k * T_1 * P_{H_2} * P_{CO})}{(1 + k_2 + T_2 * P_{CO})^2} \qquad \text{(Equation 2)}$$

An ultra-stable Cobalt-based FTR catalyst was specifically designed for this process with a low deactivation rate. The average activity for the life of the catalyst is implemented in the rate of conversion of CO.

The Fischer-Tropsch reactor was modeled in Polymath with the Ordinary Differential Equations Solver as a vertical Packed-Bed Reactor (PBR). A vertical reactor was used because the orientation allows the reactor to fill completely. Furthermore, the catalyst can be installed and replaced with ease. It also takes advantage of gravity-controlled flow of all the fluids.

The reactor was modeled based on the conversion of CO as a differential of weight of catalyst. The rate of methane is dependent on temperature over the range 390 F° – 450 F° and varies slightly. The reactor was designed to operate at a temperature of 450 F° since this allowed for the greatest rate of conversion and more diesel and naphtha to be produced. Ethane, propane, and butane selectivities were then obtained as 4% of methane's selectivity. The formation of water is proportional to the conversion of CO. Hydrogen was estimated to be an average mole ratio of 2.06 for the multiple reactions. CO_2 and N_2 were considered inert gases.

Pressure drop was also considered based on conversion and varied based on the catalyst weight. The partial pressure of CO and H_2 was related to conversion with the ideal gas law and then corrected to consider the mole balance. The vapor phase components were estimated to contain 100% H_2, CO, H_2O, CH_4, N_2 components, and 0.7 moles of C2+ hydrocarbons for every mol of CH_4 made.

Before the reaction was optimized, a Levenspiel Plot was made to determine the required volume of the reactor by taking the integral at desired conversion [3]. Figure 1 shows how conversion is affected by the rate of conversion of CO. Since there is a large jump from 80% to 90% of conversion, the most efficient conversion would be 80% resulting in a smaller reactor. When sized and optimized, the reactor with 80% conversion fit within the size restrictions, however the conversion was increased to 90% to obtain more heavy hydrocarbons and result in a higher profit and still fit within the restrictions.

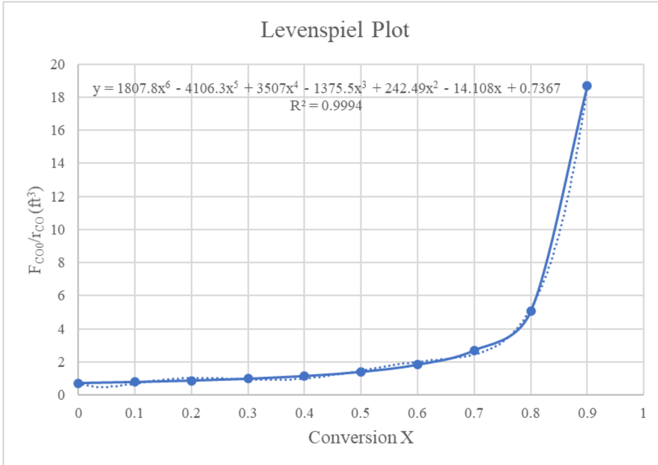

Figure 1: Levenspiel Plot for Fischer-Tropsch Reactors

The Fischer-Tropsch reaction is highly exothermic but was modeled as isothermal and then coolant was provided at a sufficient rate to cool the reactor down to prevent a runaway reaction. The heat of reaction of 70,200 Btu/lbmol of CO converted was multiplied by the amount of CO converted. This calculation provided the total duty occurring within the reactor at 450°F. The duty was then divided by the latent heat of MP pressure steam to calculate the mass flowrate of medium pressure steam needed to maintain a temperature of 450°F.

The Anderson-Schulz Flory (ASF) probability distribution was implemented to determine the hydrocarbon distribution of carbon chains longer than C5. Only carbon chains up to C30 were modeled to be able to replicate the stream into Aspen HYSYS. The relative mass fraction of the C5+ products are listed in Table 4. Based on this distribution, the percentage of the individual C5+ products were determined, and the mass flowrate of these products were obtained by multiplying by the summation of the C1-C5 and water produced during the reaction. A material balance was performed on each of the units and can be found in the Appendix.

Table 4: Relative Mass Fractions of C5+ Products

Component	**Relative Mass Frac.**	**Component**	**Relative Mass Frac.**	**Component**	**Relative Mass Frac.**
C5	0.0332	C14	0.0357	C23	0.0226
C6	0.0359	C15	0.0344	C24	0.0212
C7	0.0376	C16	0.0330	C25	0.0198
C8	0.0386	C17	0.0316	C26	0.0185
C9	0.0391	C18	0.0300	C27	0.0173
C10	0.0391	C19	0.0285	C28	0.0161
C11	0.0386	C20	0.0270	C29	0.0150
C12	0.0379	C21	0.0255	C30	0.0140
C13	0.0369	C22	0.0240		

The Fischer-Tropsch reactor was optimized with four variables in mind: number of tubes (n), catalyst weight (W), temperature (T), and tube diameters (D). These four variables directly affect the products produced and size of the reactor. The products were optimized by maximizing the C5+ components to achieve a greater conversion of heavier carbon chains that result in more revenue sold as naphtha and diesel. The reactors were sized to maintain the constraints of a conventional intermodal shipping container with the dimensions of 8 ft wide x 8 ft 6 in tall x 40 ft long.

After simulating the system in Polymath, the code was translated to MATLAB using the ODE45 differential equation solver. This allows use of MATLAB's built-in optimization toolbox which solves multivariate systems simultaneously. The only requirements for this optimization program are a cost function and constraint functions. The cost function, described in Equation 3, is dependent on the monetary cost of the catalyst and the amount of tubes in each reactor. The "a" term in Equation 3 is the scaling factor for the cost function and was determined to be on the order of 10^{-3} for all reactors. This adjusts the order of magnitude to be a constant offset because of the price fluctuations. The "n" term is the number of tubes in the reactor and the "W_f" term is the catalyst weight in pounds. The cost function accounts for the fact that the catalyst would need to be changed at least every 3 years for the 20-year project life and the catalyst costs $10 per pound. Equation 3 expresses the resulting cost function and was squared to obtain a parabolic curve to determine the minimum.

$$\mathrm{f} = \mathrm{a} * \left(\frac{20}{3} * 10 * \mathrm{n} * \mathrm{W_f}\right)^2 \qquad \text{(Equation 3)}$$

The constraints implemented in the solver are listed in Table 5. The length of the reactor was determined based on the size restrictions of the shipping containers to not exceed a height of 8 feet to allow for some clearance in the container. The inequality function for the number of tubes was determined by an L/D ratio of 3. The maximum number of tubes was determined by calculating how many tubes could be arranged in a circular bundle meeting the ratio constraints [4]. The temperature constraint maintained the reactor temperature such that the selectivity of methane was applicable. Since the reactor was modeled as a tube and shell heat exchanger for the heat transfer, the tube diameters were limited to 2 inches which is equivalent to 0.1667 feet. The overall diameter of the reactor was determined through Engineering Toolbox [4] for sizing the tubes inside of the shell and assuming that the total area of the inner tubes utilizes approximately 72% of the available surface area. Thus, the diameter of the reactor can be expressed by Equation 4.

$$D_{Overall} = 2\sqrt{\frac{n*A}{0.72*\pi}} \qquad \text{(Equation 4)}$$

From there, the L/D ratio was implemented again to restrict the total diameter of the reactor. The pressure drop ratio throughout reactor was limited by the operating pressure of the reactor and a maximum pressure drop of 50 psi. Conversion was set to 94% for the small reactors since the reactors will be operating for the longest period and the medium and large units were set to 90%.

Table 5: Constraint Functions for Optimization

Variables	Constraints
Length (ft)	$L < 8$
Number of Tubes	$1 < n < 1795$
Temperature (°F)	$390 < T < 450$
Diameter of Tubes (ft)	$0 < D < 0.1667$
Diameter of Reactor (ft)	$D_{Reactor} < L / 3$
Pressure Drop Ratio	$p_{Total} < 0.88$
Conversion	$Xf = 0.9$

MATLAB used the constraints given and the cost function to perform various iterations until a local minimum was found satisfying the constraints. The optimization completed the iterations when the objective function stopped decreasing in a feasible direction. The results were then tabulated in Table 6 with the overall diameter rounded up to a quarter foot for ease of construction and the length adjusted to desired L/D ratio.

Table 6: Optimization Results for the FTR

Variables	Small Unit	Medium Unit	Large Unit
Number of Tubes	14.00	29.00	49.00
Tube Diameter (in)	2.00	2.00	2.00
Overall Diameter (ft)	1.00	1.25	1.50
Length (ft)	3.00	3.75	4.50
Volume (ft^3)	2.36	4.60	7.95
Weight of Catalyst (lb)	2.51	3.61	4.43
Conversion	0.94	0.90	0.90

iii. Separation Vessels

After using Polymath, MATLAB, and Excel to find the mass flow rates of the components exiting the FTR, Aspen HYSYS was used to simulate the separation process. The fluid package CPA was once again used for this simulation. A process stream was made with the component mass flow rates leaving the FTR. The process stream leaving the FTR was at 450°F, so it needed to be cooled down for the lighter products to not be vaporized. The stream was sent through a cooler and cooled to a temperature of 275°F. Any process stream prior to isomerization containing C11+ material should not be cooled under 250°F to prevent wax crystallization. The cooled stream was sent to a two-phase separator and split into lighter and heavy products. The lighter product stream was cooled down to a temperature of 250°F with a heat exchanger to allow for even better separation. This stream was sent to a three-phase separator to get the tail gas, naphtha, and wastewater. The bottom stream with the heavier components was heated to 420°F to obtain separation of the lighter products from the heavy components. The streams exiting the three-phase separator included LPG, the feed to the HIU unit with the heavy distillates and waxes, and wastewater. The LPG stream needed to be cooled down to 250°F to condense the C3 and C4 hydrocarbons to sell as liquid. Since LPG can be sold no matter what the composition of the stream is, there are other components mixed into the LPG product stream. The naphtha and LPG were stored and transported without temperature concern. The HIU feed stream had to have CO below 1 mol% and water below the solubility limit. The wastewater from each three-phase separator was combined into one stream and then sent to a Wastewater Treatment Plant.

Each separator was designed after the separation unit was completely simulated. The two-phase separator (V-102) was designed as a horizontal separator because they are preferred for separating vapor from mixtures with a lower vapor/liquid ratio and are less expensive. The separator K value was found with an equation based on a pressure range so the range between 40 and 5,500 psia was applicable. Holdup and surge times were chosen to be 2 min and 1 min, respectively, due to this separator being a feed to another drum or tankage [5]. The diameter was found using an initial L/D estimate in the range from 3-4 because the pressure of the vessel was between 250 and 500 psig. The minimum length required to accommodate the liquid holdup and surge was calculated after the total area, vapor disengagement area, and the low liquid area were found. Both the diameter and length found were rounded up by 0.5 ft. The volume of the vessel was then calculated and was optimized by varying the initial L/D estimate. The final L/D values for the small, medium, and large units were 2, 3, and 3, respectively.

The three-phase separators were designed similarly to the two-phase separator. Many of the same assumptions and correlations were made, including the separator K value and the holdup and surge times. The three-phase separators are horizontal and controlled with a weir. This design maximizes the volume of heavy liquid being separated. A k_s value of 0.163 was chosen due to hydrocarbons being in the light phase and the heavy phase is water [6]. The diameters were found based on an initial L/D estimate and were later optimized to achieve a minimum volume. Length was calculated by adding the minimum length of the light liquid compartment to accommodate holdup and surge with the minimum length to facilitate liquid-liquid separation. The final L/D values for the three-phase separator of the lighter components (V-103) were 2, 3, and 3 for the small, medium, and large units, respectively. For the three-phase separator of the heavy products (V-104), the final L/D values were 2, 3, and 4 for the small, medium, and large unit, respectively.

iv. Heat Exchangers

For each unit size, four heat exchangers were designed to either heat up or cool down streams to optimize the amount of product being separated. The heat exchangers were modeled in Aspen HYSYS as coolers and heaters. For each unit size, the streams being cooled or heated change by the same temperature difference. The location of the first heat exchanger (E-103) is before V-102, and it was cooled from 450°F to 275°F. The next heat exchanger (E-104) is on the light stream separated by V-102 and it was cooled even further to 250°F. On the bottom stream from V-102 with the heavy products, there is a heat exchanger (E-105) that heats up the stream from 275°F to 420°F. The final heat exchanger (E-106) is on the LPG product stream to cool it down from 420°F to 250°F to liquify the stream.

The design portion consisted of taking the heat duty given by HYSYS and using it for finding the needed area of the heat exchanger. The LMTD method was used to find ΔT_{lm} assuming counter-current heat exchangers for the best heat transfer. A 20-degree approach temperature was assumed for all heat exchangers. There were no entering and returning temperatures given for the process cooling water so a heuristic of 90°F and 120°F entering and returning temperatures were used. The overall heat transfer coefficients (U_o) were taken from the GPSA based off the contents of process stream whether process cooling water or HP steam was used. For E-103, a Uo of 160 Btu/(hr•ft^2•°F) was used because the process stream consisted mainly of water and the cooler was using process cooling water. However, for E-104 a Uo of 130 Btu/(hr•ft^2•°F) was used because the stream consisted mainly of hydrocarbons and was cooling with process cooling water. For E-105, Uo was assumed to be 130 Btu/(hr•ft^2•°F) because the heater was using HP steam and the process stream consisted of mainly hydrocarbons. The LMTD correction factor (F) of 0.9 was used because the heat exchangers were designed as 1-shell pass with 2 or more tube passes. After all the necessary variables were found, the area of each heat exchanger was calculated [7].

Due to the modular approach to this project, the designed heat exchangers did not fall within many of the size ranges of various heat exchangers. All four of the heat exchangers for the small and medium units were designed to be Teflon tube heat exchangers because they fell within the size range given [8]. For the large unit, E-103 was designed to be a U-Tube heat exchanger because it was more cost-effective and was the largest of the heat exchangers. The remaining heat exchangers in the large unit were designed to be Teflon tube exchangers.

v. Hydro-Isomerization Unit

The Hydro-isomerization Unit did not need to be designed for the purpose of this project, but conversions were given to find the flow rates of the product streams exiting the unit. The distillate and heavier boiling fractions were sent from the separation unit on to this unit to be converted into lighter products to sell. The HIU converted 100% of the material with a boiling point greater than 700°F to material with a boiling point less than 700°F. The selectivity of the feed that will convert was given as 1.0 wt% methane, 0.5 wt% ethane, 3.5 wt% propane, 3.5 wt% butane, 25 wt% naphtha, and the remaining would be diesel. Any material that was below 700°F was assumed to pass through the unit unconverted. The initial and final product flowrates can be seen for the small, medium, and large units in Table 7, Table 8, and Table 9 below, respectively.

Table 7: HIU Small Unit Product Selectivity

Component	Inlet Flowrate (lb/hr)	Converted Product Flowrate (lb/hr)	Outlet Flowrate (lb/hr)
Methane	0.24	0.18	0.42
Ethane	0.01	0.09	0.10
Propane	0.05	0.63	0.68
Butane	0.16	0.63	0.79
Naphtha	68.74	4.53	73.27
Diesel	120.74	12.05	132.79
C11+	18.12	0.00	0.00

Table 8: HIU Medium Unit Product Selectivity

Component	Inlet Flowrate (lb/hr)	Converted Product Flowrate (lb/hr)	Outlet Flowrate (lb/hr)
Methane	1.33	1.08	2.41
Ethane	0.09	0.54	0.63
Propane	0.38	3.77	4.15
Butane	1.33	3.77	5.10
Naphtha	488.16	26.91	515.07
Diesel	729.71	71.58	801.29
C11+	107.64	0.00	0.00

Table 9: HIU Large Unit Product Selectivity

Component	Inlet Flowrate (lb/hr)	Converted Product Flowrate (lb/hr)	Outlet Flowrate (lb/hr)
Methane	2.35	1.88	4.23
Ethane	0.12	0.94	1.06
Propane	0.51	6.59	7.10
Butane	1.81	6.59	8.40
Naphtha	744.53	47.05	791.58
Diesel	1260.3	125.15	1385.45
C11+	188.19	0.00	0.00

Inert material that passed through the HIU are not shown in the table since they did not convert but they include N_2, H_2, CO, CO_2, and water. Everything except the water had negligible amounts. The catalyst in the HIU was said to be sensitive to water and CO, so it was made sure that there was no free water above the solubility limit in the feed. The specification for CO was that the content in the HIU could be no greater than 0.1 mol%. The small unit HIU feed contains 0.04 mol% CO, the medium unit contains 0.05 mol% CO, and the large unit contains 0.05 mol% CO so the design is within that specification.

Sources of Error

There are numerous potential errors when the preliminary design was done for this process. Some could stem from assumptions made, the differences between modular and regular plants, or from human error. The heuristics used for the sake of this project were designed for conventional equipment and provided uncertainty for the module units.

When modeling the Syngas Unit in Aspen HYSYS, a singular Gibbs reactor was used to simulate the syngas being made. Since the steam reforming and water gas shift reactions are equilibrium and the partial oxidation reaction goes to completion, there were some difficulties during simulation. To get the simulation to work properly, the partial oxidation reaction was modeled as an equilibrium reaction using an extremely high Keq value to simulate it as if it was going to completion. This assumption could lead to some potential error in the output of this simulated reactor.

The Fischer-Tropsch reactor was designed using Polymath, MATLAB, and Excel. Once completely modeled using Polymath and MATLAB, the ASF distribution was used in Excel to determine the relative weight fraction and the relative mole fractions of the C5+ products. There was a percent error in the mass balance due to this distribution. From calculations, there was more material leaving the FTR than there was entering and that is due to this error.

In the designed separation unit, a two-phase separator was used to make an initial separation of products. The two separated streams were then sent to two separate three-phase separators. This gives decent separation of products but could potentially be better with a distillation column.

Heat and Material Balance

For the heat and material balance, a stream table was put together to show the molar, mass, and actual volume flow rate as well as the mass enthalpy of each stream. There was a percent error for the material balance between what is entering the FTR and what is leaving the FTR. This error is due to the ASF probability distribution which was not exact to the mass flow rate coming in. The small unit had a percent error of 7.54%, the medium unit had a percent error of 7.76%, and the large unit had a percent error of 1.52%. The stream tables are based off the simplified process flow diagram of the whole unit from the Syngas Unit to the HIU. The stream table for the small unit can be seen in Table 10 below. The stream tables for the medium and large unit can be found in the appendix.

Overall Process Flow Diagram

The overall process flow diagram (PFD) spans from the Syngas Unit all the way through the HIU. The stream table shown in Table 10 below gives a complete summary of each labeled stream in this PFD. For each control valve, a pressure drop of 10 psi was assumed. For each heat exchanger, a pressure drop of 4 psi was assumed [8]. The overall PFD for the small unit can be seen in Figure 2 below. The PFDs for the medium and large units can be found in the appendix.

Table 10: Small Unit Stream Table

	Stream Number		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
	Description		Oxygen	Natural Gas	HP Steam	Carbon Dioxide	H-101 Feed	H-101 Heated Stream	R-101 Feed	E-101 Feed	V-101 Feed	CO2 Separation Unit Feed	V-101 Waste Water	Removed CO2	E-102 Feed	R-102 Feed	E-103 Feed	V-102 Feed
	Vapor Fraction		1.00	1.00	0.00	1.00	0.74	1.00	1.00	1.00	0.83	1.00	0.00	1.00	1.00	1.00	1.00	0.52
	Temperature	°F	75.00	100.00	490.00	100.00	233.23	1000.00	890.84	1606.73	212.00	212.00	212.00	189.99	217.97	450.00	450.00	275.00
	Pressure	psia	514.70	514.70	614.70	514.70	504.70	500.70	500.70	500.70	496.70	496.70	496.70	486.70	486.70	472.70	445.00	431.00
	Total Molar Flow	lbmole/hr	21.00	54.02	27.01	11.00	92.03	92.03	113.03	162.30	162.30	135.47	26.83	12.50	122.97	122.97	66.64	66.64
	Total Mass Flow	lb/hr	671.16	866.64	486.59	484.11	1837.33	1837.33	2508.49	2508.49	2508.49	2024.84	483.65	549.91	1474.93	1474.93	1535.46	1535.46
	Actual Vol. Flow	barrel/day	970.66	2524.52	42.22	446.06	4099.64	12099.84	13744.58	30313.09	8267.78	8232.90	34.89	657.99	7574.99	10296.68	5915.98	2634.30
	Mass Enthalpy	BTU/lb	-4.85	-2010.79	-6371.27	-3857.52	-3652.18	-2993.28	-2193.71	-2193.70	-3182.26	-2345.61	-6685.02	-3833.73	-1790.78	-1643.13	-3027.61	-3456.82
	Oxygen	lb/hr	665.28	0.00	0.00	0.00	0.00	0.00	665.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Water	lb/hr	0.00	0.00	486.59	0.00	486.59	486.59	486.59	565.58	565.58	82.61	482.97	0.00	82.61	82.61	620.15	620.15
	Nitrogen	lb/hr	5.88	0.00	0.00	0.00	0.00	0.00	5.88	5.88	5.88	5.88	0.00	0.00	5.88	5.88	2.94	2.94
	Carbon Monoxide	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	920.46	920.46	920.42	0.04	0.00	920.42	920.42	84.89	84.89
	Carbon Dioxide	lb/hr	0.00	0.00	0.00	484.11	484.11	484.11	484.11	579.49	579.49	578.86	0.63	549.91	28.94	28.94	28.94	28.94
	Hydrogen	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	132.39	132.39	132.39	0.01	0.00	132.39	132.39	4.26	4.26
	Methane	lb/hr	0.00	866.64	0.00	0.00	866.64	866.64	866.64	304.68	304.68	304.68	0.00	0.00	304.68	304.68	359.91	359.91
	Ethane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.15	4.15
	Propane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.08	6.08
	n-Butane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	8.01	8.01
	n-Pentane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	55.86	55.86
	n-Hexane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	51.93	51.93
Naphtha	n-Heptane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	46.93	46.93
	n-Octane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	41.55	41.55
	n-Nonane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	36.21	36.21
	n-Decane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	31.17	31.17
	n-C11	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	26.56	26.56
	n-C12	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	22.45	22.45
	n-C13	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	18.84	18.84
	n-C14	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	15.71	15.71
	n-C15	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	13.04	13.04
Diesel	n-C16	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	10.78	10.78
	n-C17	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	8.87	8.87
	n-C18	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	7.28	7.28
	n-C19	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	5.95	5.95
	n-C20	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.85	4.85
	n-C21	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.95	3.95
	n-C22	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.20	3.20
	n-C23	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	2.59	2.59
	n-C24	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	2.10	2.10
	n-C25	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.69	1.69
	n-C26	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.36	1.36
	n-C27	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.10	1.10
	n-C28	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.88	0.88
	n-C29	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.71	0.71
	n-C30	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.57	0.57

	Stream Number		17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32
	Description		E-104 Feed	V-103 Feed	Tail Gas	Top Naphtha	Top Water Product	E-105 Feed	V-104 Feed	E-106 Feed	Cooled LPG	HIU Feed	Bottom Water Product	Top & Bottom Water Products	Waste Water Disposal	Paraffin	Bottom Naphtha	Diesel Fuel
	Vapor Fraction		1.00	0.96	1.00	0.00	0.00	0.00	0.04	1.00	0.11	0.00	0.00	0.00	0.00	0.21	0.00	0.00
	Temperature	°F	275.00	250.00	250.00	250.00	250.00	275.00	420.00	420.00	250.00	420.00	420.00	412.41	321.48	250.00	250.00	250.00
	Pressure	psia	431.00	417.00	417.00	417.00	417.00	431.00	417.00	417.00	433.00	417.00	417.00	407.00	407.00	407.00	407.00	407.00
	Total Molar Flow	lbmole/hr	34.47	34.47	32.95	0.09	1.43	32.17	32.17	1.29	1.29	1.53	29.35	30.79	57.62	0.307	0.113898	0.06057
	Total Mass Flow	lb/hr	752.06	752.06	715.78	10.45	25.83	783.41	783.41	41.93	41.93	212.55	528.93	554.76	1038.40	6.08	73.27	132.78
	Actual Vol. Flow	barrel/day	2566.81	2395.69	2392.63	1.16	1.90	67.49	170.98	100.65	13.46	27.17	43.15	44.99	79.21	0.42	5.10	9.24
	Mass Enthalpy	BTU/lb	-1967.78	-2014.18	-1864.21	-849.21	-6640.86	-4886.27	-4730.07	-2771.92	-3272.32	-789.65	-6468.78	-6476.79	-6570.26	-5210.93	-859.82	-767.18
	Oxygen	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Water	lb/hr	69.88	69.88	44.04	0.03	25.81	550.27	550.27	17.13	17.13	4.44	528.70	554.52	1037.49	4.44	0.00	0.00
	Nitrogen	lb/hr	2.93	2.93	2.93	0.00	0.00	0.01	0.01	0.01	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Carbon Monoxide	lb/hr	84.65	84.65	84.64	0.01	0.00	0.24	0.24	0.22	0.22	0.02	0.00	0.01	0.04	0.02	0.00	0.00
	Carbon Dioxide	lb/hr	28.64	28.64	28.62	0.01	0.00	0.30	0.30	0.24	0.24	0.05	0.02	0.02	0.65	0.05	0.00	0.00
	Hydrogen	lb/hr	4.25	4.25	4.25	0.00	0.00	0.01	0.01	0.01	0.01	0.00	0.00	0.00	0.01	0.00	0.00	0.00
	Methane	lb/hr	357.87	357.87	357.76	0.09	0.02	2.04	2.04	1.72	1.72	0.24	0.08	0.10	0.10	0.00	0.00	0.00
	Ethane	lb/hr	4.10	4.10	4.09	0.00	0.00	0.05	0.05	0.04	0.04	0.01	0.00	0.00	0.00	0.10	0.00	0.00
	Propane	lb/hr	5.92	5.92	5.91	0.01	0.00	0.16	0.16	0.11	0.11	0.05	0.00	0.00	0.00	0.68	0.00	0.00
	n-Butane	lb/hr	7.61	7.61	7.58	0.03	0.00	0.40	0.40	0.24	0.24	0.16	0.00	0.00	0.00	0.80	0.00	0.00
	n-Pentane	lb/hr	50.65	50.65	50.29	0.36	0.00	5.21	5.21	2.50	2.50	2.68	0.03	0.03	0.03	0.00		0.00
	n-Hexane	lb/hr	43.16	43.16	42.54	0.63	0.00	8.77	8.77	3.23	3.23	5.51	0.03	0.03	0.03	0.00		0.00
Naphtha	n-Heptane	lb/hr	33.99	33.99	33.04	0.95	0.00	12.94	12.94	3.61	3.61	9.31	0.02	0.02	0.02	0.00	73.27	0.00
	n-Octane	lb/hr	24.21	24.21	22.92	1.29	0.00	17.34	17.34	3.56	3.56	13.76	0.02	0.02	0.02	0.00		0.00
	n-Nonane	lb/hr	15.51	15.51	13.99	1.52	0.00	20.71	20.71	3.02	3.02	17.67	0.01	0.01	0.01	0.00		0.00
	n-Decane	lb/hr	9.06	9.06	7.51	1.55	0.00	22.11	22.11	2.30	2.30	19.80	0.01	0.01	0.01	0.00		0.00
	n-C11	lb/hr	4.82	4.82	3.46	1.36	0.00	21.74	21.74	1.54	1.54	20.19	0.00	0.00	0.00	0.00	0.00	
	n-C12	lb/hr	2.44	2.44	1.41	1.04	0.00	20.00	20.00	0.98	0.98	19.02	0.00	0.00	0.00	0.00	0.00	
	n-C13	lb/hr	1.24	1.24	0.54	0.70	0.00	17.60	17.60	0.63	0.63	16.97	0.00	0.00	0.00	0.00	0.00	
	n-C14	lb/hr	0.57	0.57	0.16	0.41	0.00	15.14	15.14	0.35	0.35	14.79	0.00	0.00	0.00	0.00	0.00	
	n-C15	lb/hr	0.28	0.28	0.05	0.23	0.00	12.77	12.77	0.21	0.21	12.56	0.00	0.00	0.00	0.00	0.00	
Diesel	n-C16	lb/hr	0.13	0.13	0.01	0.11	0.00	10.65	10.65	0.12	0.12	10.53	0.00	0.00	0.00	0.00	0.00	132.78
	n-C17	lb/hr	0.07	0.07	0.01	0.07	0.00	8.80	8.80	0.08	0.08	8.72	0.00	0.00	0.00	0.00	0.00	
	n-C18	lb/hr	0.03	0.03	0.00	0.03	0.00	7.25	7.25	0.04	0.04	7.21	0.00	0.00	0.00	0.00	0.00	
	n-C19	lb/hr	0.01	0.01	0.00	0.01	0.00	5.94	5.94	0.02	0.02	5.91	0.00	0.00	0.00	0.00	0.00	
	n-C20	lb/hr	0.01	0.01	0.00	0.01	0.00	4.84	4.84	0.01	0.01	4.83	0.00	0.00	0.00	0.00	0.00	
	n-C21	lb/hr	0.00	0.00	0.00	0.00	0.00	3.94	3.94	0.01	0.01	3.93	0.00	0.00	0.00	0.00	0.00	0.00
	n-C22	lb/hr	0.00	0.00	0.00	0.00	0.00	3.20	3.20	0.01	0.01	3.20	0.00	0.00	0.00	0.00	0.00	0.00
	n-C23	lb/hr	0.00	0.00	0.00	0.00	0.00	2.59	2.59	0.00	0.00	2.59	0.00	0.00	0.00	0.00	0.00	0.00
	n-C24	lb/hr	0.00	0.00	0.00	0.00	0.00	2.10	2.10	0.00	0.00	2.09	0.00	0.00	0.00	0.00	0.00	0.00
	n-C25	lb/hr	0.00	0.00	0.00	0.00	0.00	1.69	1.69	0.00	0.00	1.69	0.00	0.00	0.00	0.00	0.00	0.00
	n-C26	lb/hr	0.00	0.00	0.00	0.00	0.00	1.36	1.36	0.00	0.00	1.36	0.00	0.00	0.00	0.00	0.00	0.00
	n-C27	lb/hr	0.00	0.00	0.00	0.00	0.00	1.10	1.10	0.00	0.00	1.10	0.00	0.00	0.00	0.00	0.00	0.00
	n-C28	lb/hr	0.00	0.00	0.00	0.00	0.00	0.88	0.88	0.00	0.00	0.88	0.00	0.00	0.00	0.00	0.00	0.00
	n-C29	lb/hr	0.00	0.00	0.00	0.00	0.00	0.71	0.71	0.00	0.00	0.71	0.00	0.00	0.00	0.00	0.00	0.00
	n-C30	lb/hr	0.00	0.00	0.00	0.00	0.00	0.57	0.57	0.00	0.00	0.57	0.00	0.00	0.00	0.00	0.00	0.00

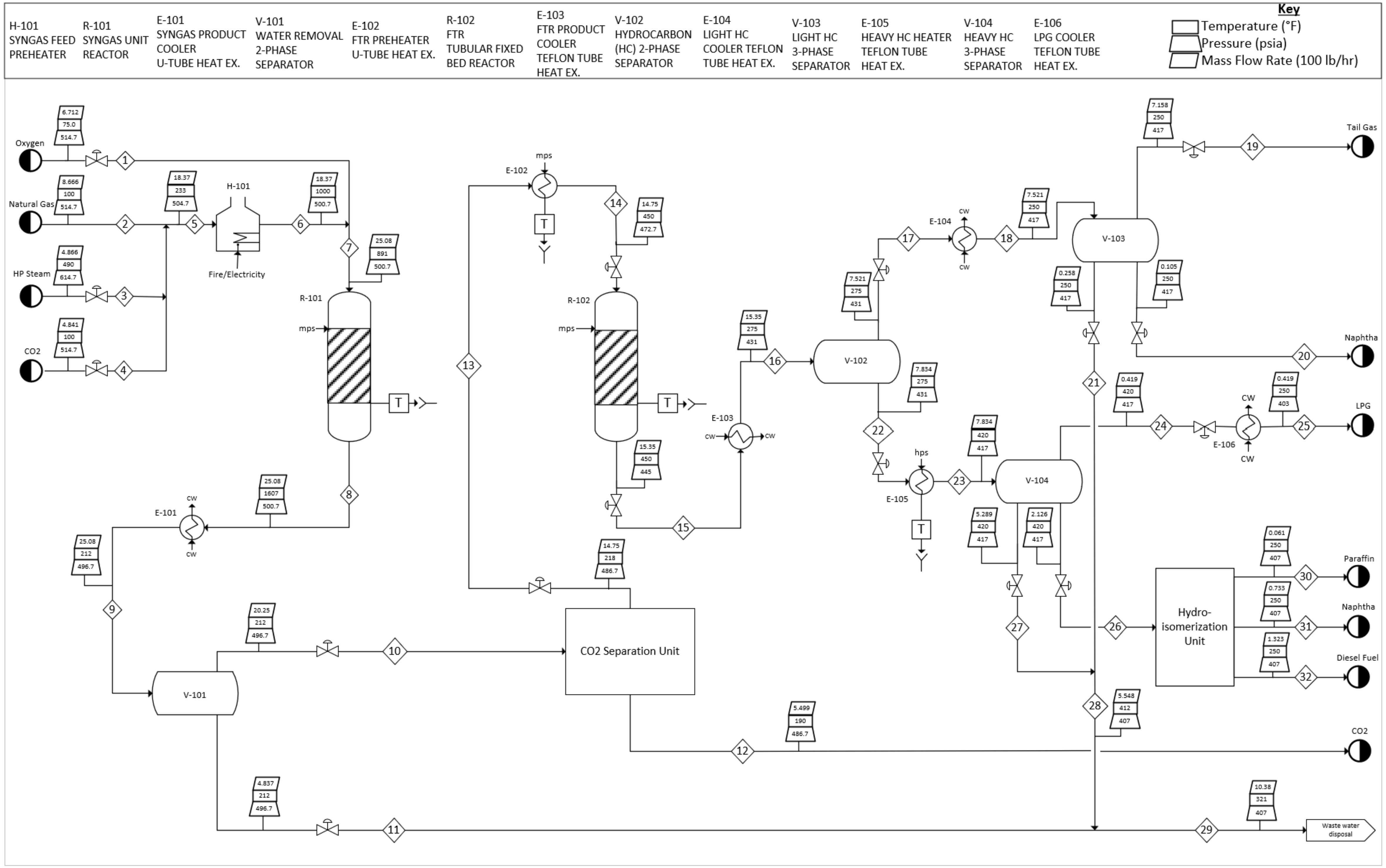

Figure 2: Small Unit Overall PFD

Safety/Environmental Summary

a. Inherent Safety Evaluation

Being inherently safe in the preliminary design stage was extremely important to the overall value of the project. There are numerous features of this project that were designed to be inherently safe including the materials of construction and the size of equipment. One feature that will make this process inherently safe will be making incorrect assembly impossible for dismantling and reattaching pieces of equipment during maintenance. This will be done by having specific labels for the pieces of equipment that will need to be replaced or maintained.

An example of minimization in the design was using a continuous-flowing reactor instead of a batch reactor. The FTR is a vertical fixed bed tubular reactor, which means it has more uniform mixing and less material in the reactor at a given time. It was determined to be a vertical reactor for ease of catalyst replacement. Another example of minimization is the trucks taking product back to the central plant periodically. The storage tanks will not hold more than 5,000 gallons of liquid at a given time. This reduces the amount of product that is kept at the wellheads. Also, since this whole process is modular, the pieces of equipment are smaller than normal equipment. This minimizes risk of potential large-scale damage.

For heat exchangers and the FTR, medium or high-pressure steam was substituted in place of fuel gas. This reduces risk of fire and the probability of an incident. The inherent safety practice of simplification was used in the separation system by using less pieces of equipment than what could be required. For completely perfect separation of each stream, multiple distillation columns would need to be used. The designed separator system only consists of three separators and four heat exchangers. All equipment was designed using stainless steel to have stronger vessels to handle the operating conditions. Stainless steel has an acceptable corrosion characteristic for gasoline while carbon steel is cautionary. Although gasoline is not a product, the intermediates contain a similar composition that could potentially result in corrosive behaviors. Even though carbon steel is significantly cheaper than stainless steel, a substitution of stainless steel will provide an inherently safer design.

b. Process Safety Management

This process was designed to mitigate risk and prioritize safety so that minimal accidents occur. If an accident occurs, it would be because the operating conditions have deviated outside the permitted range. During more detailed design, a more in-depth process hazard analysis should be done. Safety culture will also be important to maintain this process. Operators and other staff will be encouraged to say something or act if something seems off about the process. If workers remain proactive, the risk of hazards should decrease significantly. Proper training on how to operate the pieces of equipment and how they work will be vital to making sure that accidents are limited. Lockout tagout procedures should be implemented during maintenance or servicing to prevent mistakes from occurring during that time. Audits will also occur periodically to ensure that safety practices are maintained. An emergency response plan should be made in case of various emergencies including fires and loss of containment. Implementing a management of change plan will need to be created as well to prepare for various changes with the internal and external processes.

i. Process Hazards

There are numerous process hazards for this GTL process. Many of the components required for this process to work are flammable and could lead to fires and explosions if ignited. The components that mix to make syngas have many hazards that need to be monitored as seen in Figure 3 obtained from Cameo Chemicals [9].

	METHANE	WATER	CARBON MONOXIDE	HYDROGEN	NITROGEN	OXYGEN
WATER	Compatible					
CARBON MONOXIDE	Compatible	Compatible				
HYDROGEN	Compatible	Compatible	Compatible			
NITROGEN	Compatible	Compatible	Compatible	Compatible		
OXYGEN	Incompatible Flammable Generates gas Generates heat Intense or explosive reaction Toxic	Incompatible Corrosive Generates gas Generates heat Toxic	Incompatible Explosive Flammable Generates gas Generates heat Intense or explosive reaction Unstable when heated	Incompatible Explosive Flammable Generates gas Generates heat Intense or explosive reaction Unstable when heated	Compatible	
CARBON DIOXIDE	Compatible	Caution Corrosive Generates heat	Compatible	Compatible	Compatible	Compatible

Figure 3: Material Compatibility Chart

The lighter hydrocarbons that are being formed in the FTR are flammable and need to be kept safe from potential ignition sources. If something were to happen which caused the FTR to run outside of operating conditions, a runaway reaction could possibly occur. All the hydrocarbons formed are compatible together as opposed to the O_2 in the Syngas Unit which could lead to potential fires or explosions.

High pressures and temperatures are used throughout the entire process, so overheating or overpressure events are possible. There is potential for overpressure in the FTR as well as the separator equipment. It will be important that the FTR has a rupture disk as well as a pressure relief valve on it to indicate an overpressure event. Pressure relief valves will also be on every separator being used in the units. Steam and cooling water are used to cool down or heat up different parts of the process. Medium pressure (MP) steam cools down the extremely exothermic reaction happening in the FTR and HP steam is used as a utility for one of the heat exchangers in the separator unit. Cooling water is used for three heat exchangers in each unit. These utilities are essential for ensuring temperatures do not fall out of their setpoint.

Human error is also a potential process hazard if proper procedures are not done correctly. If proper lockout tagout procedures are not followed, it could lead to failures in the system and potential injuries to people in the plant. Extensive training will be done on how to operate the equipment

ii. P&ID of the Major Fractionator

Many safety features were taken into consideration when designing the controls system of the major fractionator. The major fractionator was assumed to be the separation unit part of the process, so control systems were only put on that section of the process. There were pressure relief valves and control loops put in place to prevent overpressure or other potential hazards. The full P&ID of the small unit can be found in Figure 4 below. The P&IDs for the medium and large unit can be found in the Appendix.

To prevent overpressure, pressure relief valves were placed in various places in the separation unit. The first valve is on the cooling water stream out of E-103. This was put in place in case the cooling water is heated too high and causes the tubing in the heat exchanger to expand and burst [8]. Other pressure relief valves were put on the cooling water streams coming out of E-104 and E-106 for the same exact reason. Pressure relief valves were also put on each separator because they are pressure vessels and could easily overpressure.

There were 11 different control loops designed to control different parts of the process to minimize the risks of this process as much as possible. On each of the process streams from the heat exchangers, a temperature control loop was put in place to control either the temperature of the cooling water entering or the HP steam entering the exchanger. This was done to ensure that the exiting process stream was not too hot or too cold for the next part of the unit. On V-102, a pressure control loop was needed to ensure the exiting vapor stream's pressure was not too high. Also, on V-102, there was a level control loop designed to ensure the liquid level in the vessel did not get too low or too high. On the three-phase separators, V-103 and V-104, two level control loops were designed. Since there were three process streams leaving the vessel, two of them being heavy and light liquids, two liquid level controls were needed to control the heavy and light liquid levels. For the exiting vapor streams on both, there was pressure control loop to control the vapor pressure.

For the fail open and fail closed positions of the control valves, a decision was made based on what the safest position would be for that valve. The valves on the cooling water streams into the heat exchangers were made to be fail closed to ensure that the cooling water does not keep cooling the process stream for the risk of the temperature falling below 250°F. The control valve on the HP steam inlet to E-105 was chosen as fail closed to prevent the process stream from overheating and potentially causing a runaway reaction. For the pressure control loops on the vapor outlet of each separator, the position chosen was fail open to ensure there would not be a gas buildup and possibly overpressure the vessel. For the level control loops on the light and heavy liquid outlets of the separators, the fail closed position was chosen for the valve so that the liquid leaving does not continue moving on to the next part of the process. This helps stop the process in case a potential hazard arises.

Pipe diameters were calculated using heuristics given in the 1997 ASHRAE Fundamentals Handbook for gas flow [10]. For liquid and two-phase flow streams, equation heuristics that were used can be found in the Appendix [11]. It was assumed that for the modular system there would be 20 feet between each equipment. This is based on the heuristic that 50 feet is between equipment for conventional process plants [8].

E-103	V-102	E-104	V-103	E-105	V-104	E-106
FTR COOLER	HYDROCARBON	LIGHT HC COOLER	LIGHT HC	HEAVY HC HEATER	HEAVY HC	LPG COOLER
TEFLON TUBE	(HC) 2-PHASE	TEFLON TUBE	3-PHASE	TEFLON TUBE	3-PHASE	TEFLON TUB
HEAT EX.	SEPARATOR	U-TUBE HEAT EX.	SEPARATOR	HEAT EX.	SEPARATOR	HEAT EX.

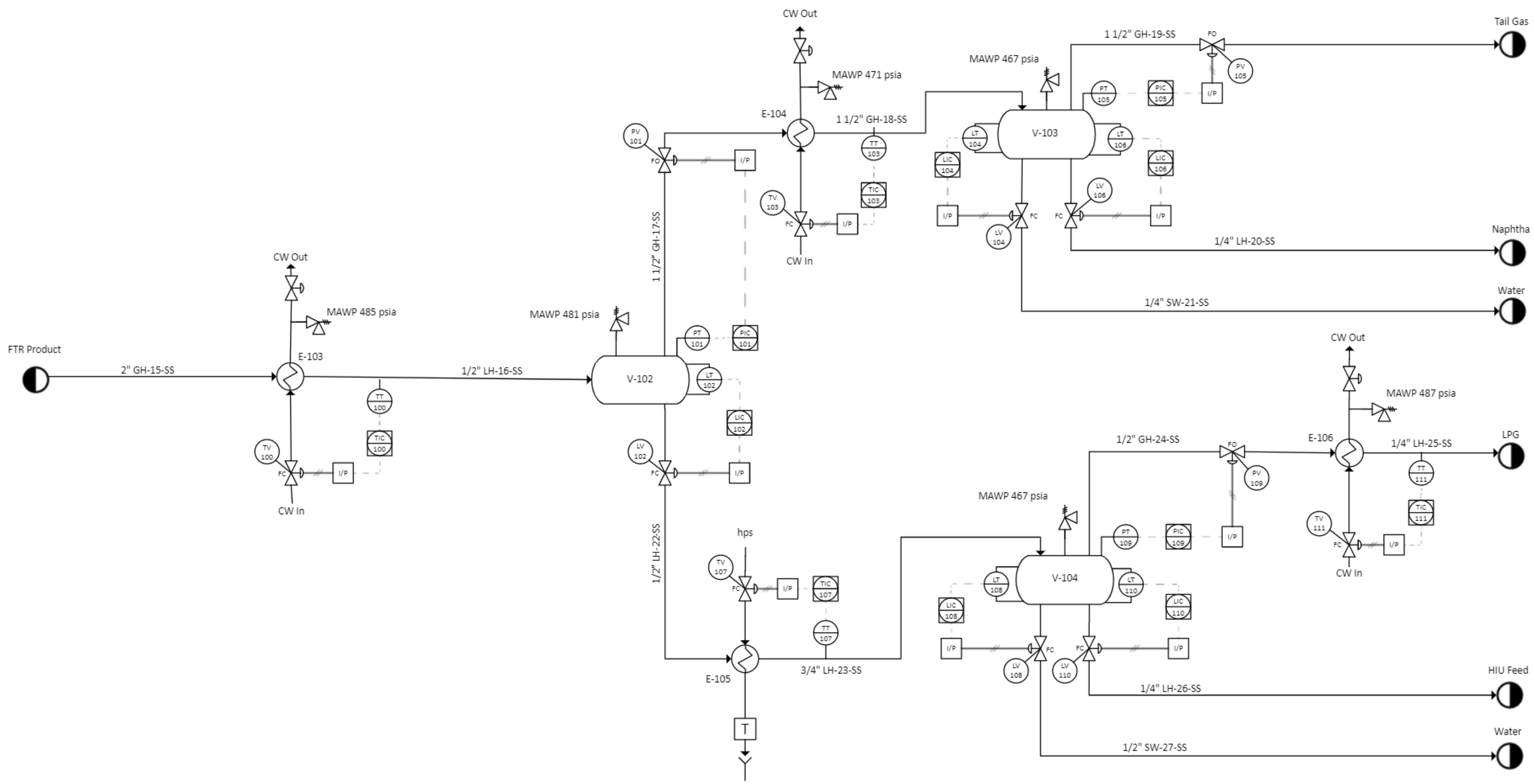

Figure 4: Small Unit P&ID

iii. Uncongested Vapor Cloud Deflagration

In a deflagration, the pressure front moves away from the reaction front, which is moving at less than the speed of sound. The potential explosions of this process were measured on an equivalent TNT basis. Based on that equivalent TNT value, the overpressure of the potential leak from the equipment was found and potential damage determined [12]. Due to this being a modular process with smaller pieces of equipment, the damage from a potential explosion is significantly less than a regular-sized process. All scenarios were calculated at 10 m (32.81 ft) from the explosion source. An explosion of efficiency of 2% was assumed and all required values for calculation were taken from Aspen HYSYS.

The small unit's worst-case scenario would occur at the FTR where the hydrocarbons are being formed. The damage for the small unit would be occasionally breaking of large glass windows already under strain. The worst-case scenario for the medium unit would also occur at the FTR. The damage for this unit would be a loud noise (143 dB), sonic boom, and/or glass failure. The worst-case scenario that could occur in the large unit would be in V-103, the first three-phase separator. The damage would be the breakage of small windows under strain. Table 11 below shows the information needed to calculate the overpressure to determine the damage that could potentially happen.

Table 11: Vapor Deflagration Values

	Small	**Medium**	**Large**
	R-102	R-102	V-103
Mass of vapor released (lb)	0.10	0.18	0.33
Heat of combustion (Btu/lb)	10315.89	9276.43	10066.33
Mass of TNT (lb)	0.01	0.02	0.04
Distance (ft)	32.81	32.81	32.81
Overpressure (kPa)	0.24	0.38	0.81

c. Potential Health Impacts

Based on the entire safety analysis, it was also important to assess and mitigate potential health impacts. High temperatures are used throughout the process, so hot steam or surfaces could lead to burns. If a fire or explosion occurs, severe burns or death could be a result. The cobalt catalyst that is used for the FTR has the potential to be toxic to workers who replace it so it will be important to mitigate the risk involved in switching out the catalyst [13]. A summary of potential health risks and how they are mitigated can be seen below in Table 12.

Table 12: Health Risks and Mitigation

Risk	Health Impact	Mitigation
Steam	Severe burns	Proper training and storage
Runaway reactions	Chemical burn; death	Strong vessels and relief systems
Flammable chemicals	Severe burns	Away from ignition sources
Toxic chemicals	Poisoning; death	Wear proper PPE; proper training; shower & eyewash stations
Cobalt catalyst	Harmful to eyes, skin, heart, and lungs	Wear proper PPE; store safely
Hot surfaces	Skin burn	Wear proper PPE; proper training

d. Safety Summary

The main risks of this GTL process are the flammability of the lighter hydrocarbons, the mixture of O_2 in the Syngas Unit, the high temperatures and pressures, the potential for loss of containment, vapor cloud deflagration of the units, and lack of understanding of the process. The best way to prevent safety risks is to design the system with inherent safety practices and have proper training on how to operate the equipment. An additional safety feature was making the design pressures of each piece of equipment 50 psi more than the operating pressure [8]. The safety features that were put in place significantly minimize the environmental and human health risks.

e. Environmental Summary

The main goal for this project was minimizing the amount of greenhouse gases produced during oil extraction. This was successfully accomplished by converting methane into naphtha and diesel. Table 13 illustrates the methane that has been converted due to the Fischer-Tropsch synthesis. The small unit converts 58.8% of the methane, medium unit 70.2%, and the large unit by 61.3%. The remaining methane in the tail gas will be flared resulting in a much smaller environmental impact than conventional methods of flaring the natural gas initially.

Table 13: Converted Methane

	Small	Medium	Large
Starting Methane (lb/hr)	864.3	4,321.7	8,643.4
Methane in Tail Gas (lb/hr)	357.8	1,289.2	3,347.8
Converted Methane (lb/hr)	508	3,032.5	5,295.6

As the tail gas will be flared later in the process, Table 14 lists the mass flows and quantity of gas that will be flared. This information is required in order to obtain operating permits from the regulatory authorities. The application for Best Available Control Technology (BACT) will be utilized to treat waste prior to discharging into the environment [14].

Table 14: Tail Gas Composition and Quantity

	Mass Flow (lb/hr)	Mass Flow (lb/hr)	Mass Flow (lb/hr)
	Small Unit	**Medium Unit**	**Large Unit**
Water	44.0	187.3	429.1
Nitrogen	2.9	17.4	30.6
Carbon Monoxide	84.6	543.3	933.9
Carbon Dioxide	28.6	124.6	277.1
Hydrogen	4.3	28.6	48.7
Methane	357.8	1289.2	3347.8
Ethane	4.1	24.1	42.4
Propane	5.9	34.6	61.3
Butane	7.6	45.6	78.4
Pentane	50.3	281.1	517.2
Hexane	42.5	226.8	433.9
Heptane	33.0	165.1	333.2
Octane	22.9	105.4	227.5
Nonane	14.0	58.9	136.3
Decane	7.5	29.1	71.9
C11	3.5	12.4	32.6
C12	1.4	4.7	13.0
C13	0.5	1.7	5.0
C14	0.2	0.5	1.5
C15	–	0.1	0.4
C16	–	–	0.1
Total Mass Flow (lb/hr)	715.8	3178.7	7022.0

The Fischer-Tropsch synthesis results in by-product of water that is separated out of the main product streams through multiple three-phase separators. The resulting wastewater is combined into a singular stream and sent to the wastewater treatment plant. Table 15 shows the wastewater mass flow rates of each unit.

Table 15: Wastewater Mass Flow Rates

	Small	Medium	Large
Wastewater (lb/hr)	1038.4	5729.2	10606.0

Since the catalyst is replaced almost every three years, it will need to be properly disposed. For this project, the CRI-MET process will be incorporated. Table 16 shows the total amount of catalyst that will need to be recycled over the course of the project life [15].

Table 16: Total Amount of Catalyst

	Small	Medium	Large
Catalyst (lb)	125.3	144.4	97.4

Equipment Information Summary

Below are the Equipment Summary Tables for the small, medium, and large units:

Table 17: Small Unit FTR and Separation Equipment List

Reactor	**R-102**			
Type	Fixed Bed Tubular Reactor			
Pressure (psig)	430.30			
Temperature (°F)	450.00			
Orientation	Vertical			
Number of Tubes	14.00			
Diameter of Tubes (ft)	0.17			
Diameter (ft)	1.00			
Height (ft)	2.50			
Volume (ft^3)	1.96			
MOC	SS			
Vessel	**V-102**	**V-103**	**V-104**	
Type	Two-Phase	Three-Phase	Three-Phase	
Pressure (psig)	416.30	402.00	402.00	
Temperature (°F)	275.00	250.00	420.00	
Orientation	Horizontal	Horizontal	Horizontal	
Diameter (ft)	1.50	0.50	1.50	
Length (ft)	2.00	0.50	1.50	
Volume (ft^3)	3.53	0.10	2.65	
MOC	SS	SS	SS	
Heat Exchanger	**E-103**	**E-104**	**E-105**	**E-106**
Type	Teflon Tube	Teflon Tube	Teflon Tube	Teflon Tube
Area (ft^2)	19.48	1.54	7.28	0.81
Duty (Btu/hr)	659,039.02	34,891.13	122,369.43	20,980.00
Shell				
Temp. In (°F)	450.00	275.00	275.00	420.00
Temp. Out (°F)	275.00	250.00	420.00	250.00
Pressure (psig)	420.30	406.30	406.30	392.30
Phase	Vapor	Vapor	Liquid	Vapor
MOC	SS	SS	SS	SS
Tube				
Temp. In (°F)	90.00	90.00	490.00	90.00
Temp. Out (°F)	120.00	120.00	490.00	120.00

Pressure (psig)	35.30	35.30	600.00	35.30
Phase	Liquid	Liquid	Vapor	Liquid
MOC	SS	SS	SS	SS

Table 18: Medium Unit FTR and Separation Equipment List

Reactors	**R-102**			
Type	Fixed Bed Tubular Reactor			
Pressure (psig)	430.30			
Temperature (°F)	450.00			
Orientation	Vertical			
Number of Tubes	29.00			
Diameter of Tubes (ft)	0.17			
Diameter (ft)	1.25			
Height (ft)	3.50			
Volume (ft^3)	4.29			
MOC	SS			
Vessel	**V-102**	**V-103**	**V-104**	
Type	Two-Phase	Three-Phase	Three-Phase	
Pressure (psig)	416.30	402.00	402.00	
Temperature (°F)	275.00	250.00	420.00	
Orientation	Horizontal	Horizontal	Horizontal	
Diameter (ft)	2.00	1.00	2.00	
Length (ft)	4.50	1.00	2.00	
Volume (ft^3)	14.14	0.79	6.28	
MOC	SS	SS	SS	
Heat Exchanger	**E-103**	**E-104**	**E-105**	**E-106**
Type	Teflon Tube	Teflon Tube	Teflon Tube	Teflon Tube
Area (ft^2)	114.68	6.65	47.34	6.87
Duty (Btu/hr)	3,879,048.97	150,701.92	795,251.66	179,000.00
Shell				
Temp. In (°F)	450.00	275.00	275.00	420.00
Temp. Out (°F)	275.00	250.00	420.00	250.00
Pressure (psig)	420.30	406.30	406.30	392.30
Phase	Vapor	Vapor	Liquid	Vapor
MOC	SS	SS	SS	SS
Tube				
Temp. In (°F)	90.00	90.00	490.00	90.00
Temp. Out (°F)	120.00	120.00	490.00	120.00
Pressure (psig)	35.30	35.30	600.00	35.30
Phase	Liquid	Liquid	Vapor	Liquid
MOC	SS	SS	SS	SS

Table 19: Large Unit FTR and Separation Equipment List

Reactors	**R-102**			
Type	Fixed Bed Tubular Reactor			
Pressure (psig)	430.30			
Temperature (°F)	450.00			
Orientation	Vertical			
Number of Tubes	49.00			
Diameter of Tubes (ft)	0.17			
Diameter (ft)	1.50			
Height (ft)	4.25			
Volume (ft^3)	7.51			
MOC	SS			
Vessel	**V-102**	**V-103**	**V-104**	
Type	Two-Phase	Three-Phase	Three-Phase	
Pressure (psig)	416.30	402.00	402.00	
Temperature (°F)	275.00	250.00	420.00	
Orientation	Horizontal	Horizontal	Horizontal	
Diameter (ft)	2.00	1.00	2.00	
Length (ft)	7.00	4.00	3.00	
Volume (ft^3)	21.99	3.14	9.42	
MOC	SS	SS	SS	
Heat Exchangers	**E-103**	**E-104**	**E-105**	**E-106**
Type	U-Tube	Teflon Tube	Teflon Tube	Teflon Tube
Area (ft^2)	181.73	13.52	85.47	8.93
Duty (Btu/hr)	6,829,817.44	340,712.46	1,292,192.59	232,700.00
Shell				
Temp. In (°F)	450.00	275.00	275.00	420.00
Temp. Out (°F)	275.00	250.00	420.00	250.00
Pressure (psig)	420.30	406.30	406.30	392.30
Phase	Vapor	Vapor	Liquid	Vapor
MOC	SS	SS	SS	SS
Tube				
Temp. In (°F)	90.00	90.00	490.00	90.00
Temp. Out (°F)	120.00	120.00	490.00	120.00
Pressure (psig)	35.30	35.30	600.00	35.30
Phase	Liquid	Liquid	Vapor	Liquid
MOC	SS	SS	SS	SS

Unit Control and Instrumentation Description

Even though the FTR was not included in the P&ID, it will require different controls and pressure relief instruments. For pressure relief, a rupture disk in series with a pressure relief valve was designed to prevent overpressure. The rupture disk would be in place to protect against toxicity or corrosion and the spring-loaded relief valve would close and minimize losses [12]. To control the pressure of the reactor, a pressure control loop will be needed. A pressure transmitter would be directly attached to the reactor and send a signal to the feed stream to manipulate the flowrate of that stream. A temperature control loop would be needed for the FTR product stream due to the highly exothermic nature of the reaction. The flowrate of the medium-pressure steam entering the reactor would be manipulated to ensure the temperature does not go above 450°F. A simplified P&ID of what the FTR should look like can be seen in Figure 5 below.

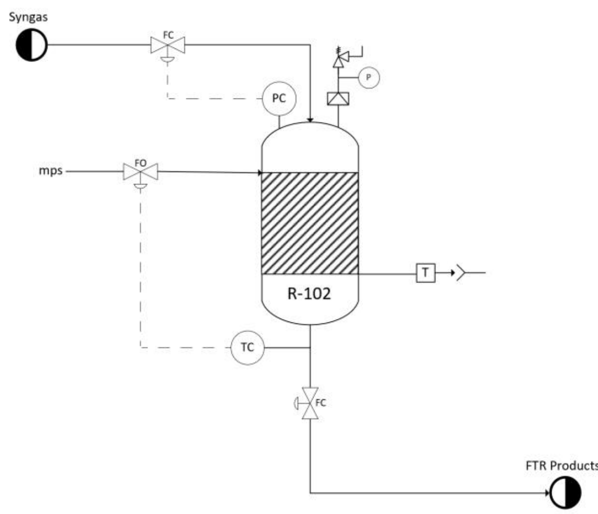

Figure 5: Simplified P&ID of FTR

For the separation unit, there were three kinds of control loops added: temperature, pressure, and level. The separation unit P&ID can be seen above in Figure 4. The temperature controls were necessary for the heat exchangers that either cooled or heated up process streams. The pressure control loops were needed for the outlet vapor streams from the separators. Level control loops were put in place to adjust the heavy and light liquid levels in the separation vessels. Table 20 below shows a brief description of what each control loop's controlled and manipulated variables are.

Table 20: P&ID Controlled and Manipulated Variables

Loop Number	Controlled Variable	Manipulated Variable
100	E-103 outlet stream temperature	Cooling water flowrate into E-103
101	V-102 pressure	Outlet vapor flowrate from V-102
102	V-102 liquid level	Outlet liquid flowrate from V-102
103	E-104 outlet stream temperature	Cooling water flowrate into E-104
104	V-103 heavy liquid level	Outlet heavy liquid flowrate from V-103
105	V-103 pressure	Outlet vapor flowrate from V-103
106	V-103 light liquid level	Outlet light liquid flowrate of V-103
107	E-105 outlet stream temperature	High-pressure steam flowrate into E-105
108	V-104 heavy liquid level	Outlet heavy liquid level from V-104
109	V-104 pressure	Outlet vapor flowrate from V-104
110	V-104 heavy liquid level	Outlet heavy liquid flowrate from V-104
111	E-106 outlet stream temperature	Cooling water flowrate into E-106

Expected Plot Layout for the Wellhead Site System

The wellheads are located at various distances from the central plant with varying production rates of natural gas. A parallel modular manufacturing approach was taken to maximize the process without bottlenecking the wellhead. The small, medium, and large units were utilized throughout the field. The following cartesian coordinate plot, Figure 6, allowed access to visually observe the distances between each wellhead from the central plant and from each other.

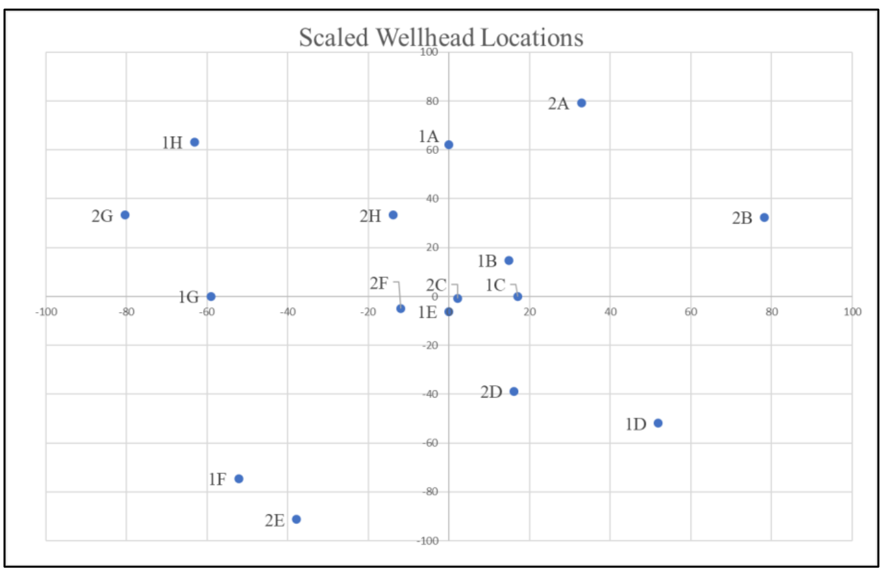

Figure 6: Plot Layout of the Wellhead Site System

Wellsite Plan and Modular Unit Deployment

Since each well will have a sharp decline of production of natural gas after the first two years by 35%, the units will be moved between wellheads to ensure maximum efficiency of the units. Figure 7 illustrates how each wellhead will decrease in production when put into operation. The central plant can also only process 30,000 MSCF of natural gas per day feed equivalent from the wells regardless of the products formed. This emphasizes the efficiency of exchanging units and turning on the wellheads at various times throughout the project life.

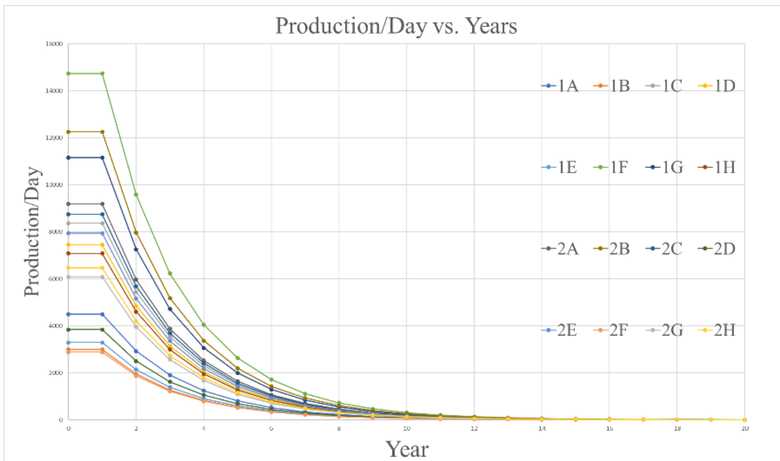

Figure 7: Declining Production Rate of Wellheads

The modular network optimization was approached on three premises. Firstly, each unit will be transported from the central plant to the desired wellhead. Secondly, the well sites will be turned on at different years throughout the project life to maximize the product that the central plant can process. Thirdly, when a unit is no longer being utilized it will be transported back to the central plant to be decommissioned.

Based on the three premises stated above, the modular network optimization began by starting the process at the well sites farthest from the central plant transferring units to the well sites closest to the central plant to eliminate the cost of decommission transportation costs. It also consisted of moving the units to close by well sites instead of transporting new units to a well site to only transport back at the end of its life. All these conditions were optimized while ensuring the largest possible gas production rate of 30,000 MSF of natural gas per day and striving to maximize gas production rate at the beginning of the project life. This was done to obtain the greatest revenue in the first five years since time value of money will increase. Table 21 depicts the gas production rate of all the well sites at a given year.

Table 21: Gas production rate per year of deployment plan

Year	2024	2025	2026	2027	2028	2029	2030
Gas Production Rate (MSCF)	0	27,199	27,199	29,937	29,914	29,966	29,478
Year	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**	**2037**
Gas Production Rate (MSCF)	28,107	29,275	25,929	29,159	25,754	17,562	26,061
Year	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**
Gas Production Rate (MSCF)	22,029	22,258	17,177	10,996	7,061	4,510	2,760

The cost of redeploying a module will be $3.00 per MCF capacity per mile transported and will require a month of downtime. This results in a loss of production during redeployment. Also, the catalyst in the FTR will require to be changed every three years and result in a downtime of one month. To limit the loss of production from these downtimes, the catalyst will sometimes be exchanged a year before requirement and changed the same month as the redeployment. The downtime will also be scheduled to occur in January of every year.

The number of units needed at a given period was determined based on the capacity of the wells. Since wells with the highest production rate declined the fastest, large units were used sparingly and replaced with medium units when needed to lengthen their capabilities. For this project, 13 small units, 9 medium units and 5 large units will be used. Table 22 shows the sizes and number of units needed for each well site. At the bottom of Table 22, the number of units need to be deployed in the field are listed and will not begin to be manufactured until one year before required. The numbers marked red indicate the units that are at the well site but not operating at the time.

Table 22: Unit Distribution between Wellheads

	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044
Well ID	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day	Gas Production MSCF/day
1A	4494	4494	2921	1899	1234	802	521	339	220	143											
Small			1	1	3	1,2	1,2	1	1	1	0										0
Medium			1	1	1	0															0
Large	1	1	1	1	1	0															0
1B											2989	2989	1943	1263	821	534	347	225	147		
Small											1	1	1	1	2	2	1,1	1,1	1,1	2	2
Medium											1	1	1	1	1	1	1	1	1	1	1
Large																					0
1C					8365	8356	5431	3530	2295	1492	970	630	410	266	173	112					
Small							1	1	1	1	2	2	1,1	1	1	1	1	1	1	1	1
Medium					2	2	2	2	1	1	0										0
Large					1	1	1	0													0
1D	7447	7447	4841	3146	2045	1329	864	562	365	237	154										
Small						3	2	2	1,1	1,1	1,1	2	2	2	0						0
Medium	1	1	1	1	1	1	1	1	0												0
Large	1	1	1	1	1	1	0														0
1E				3290	3290	2139	1390	904	587	382	248	161									
Small								2	2	1,1	1	1	1	1	1	1	1	1	1	1	1
Medium				2	2	1,1	1,1	2	2	0											0
Large																					0
1F													14737	14737	9579	6226	4047	2631	1710	1111	722
Small																					0
Medium													2	2	2	1,1	2	2	1,1	1,1	1,1
Large													2	2	1	1	1	1	1	1	1
1G										11155	11155	7251	4713	3063	1991	1294	841	547	355	231	150
Small																	2	2	1,1	1,1	1,1
Medium										3	3	3	2	2	1	1	1	1	1	1	1
Large										1	1	1	0								0
1H						7081	7081	4603	2992	1945	1264	822	534	347	226	147					
Small									1	1	1	2	2	1,1	1,1	1,1	0				0
Medium						1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Large						1	1	1	1	0											0
2A	9182	9182	5968	3879	2522	1639	1065	692	450	293	190										
Small			2	2	2	2	2	2	1,1	1	1	0									0
Medium																					0
Large	2	2	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
2B			12258	12258	7968	5179	3366	2188	1422	924	601	391	254	165							
Small					1	1	1	1	1	2	2	1,1	1,1	1	0						0
Medium			1	1	1	1	1	1	1	0											0
Large			2	2	1	1	1	0													0
2C								8742	8742	5682	3693	2401	1561	1014	659	429	279	181	118		
Small															2	1,1	1,1	1,1	1,1	2	2
Medium										1	1	1	1	1	1	1	1	1	1	1	1
Large								2	2	1,1	1,1	2	2	2	2	2	2	2	2	2	2
2D									3840	3840	2496	1622	1055	685	446	290	188	122			
Small														2	1	1	1	1	1	1	1
Medium									2	2	1,1	1,1	1,1	2	2	2	2	2	2	2	2
Large																					0
2E															7939	7939	5160	3354	2180	1417	921
Small															1	1	1	1	1	1	1
Medium															1	1	1	1	1	1	1
Large															1	1	1	1	1	1	1
2F				2874	2874	1868	1214	789	513	333	217	141									
Small				1	1	1	1	2	2	1,1	1,1	1,1	1	1	1	1	1	1	1	1	1
Medium				1	1	1	1	1	1	0											0
Large																					0
2G	6076	6076	3949	2567	1669	1085	705	458	298												
Small				1	1	1	2	1,1	1	0											0
Medium	3	3	1,2	2,1	1	1	0														0
Large																					0
2H							6468	6468	4204	2733	1776	1155	750	488	317	206	134				
Small										1	1	1	2	1,1	1,1	1,1	1,1	2	2	2	2
Medium							1	1	1	1	1	1	0								0
Large							1	1	1	1	1	1	0								0
Summation	27207	27207	29948	29930	29988	29498	28126	29298	25950	29175	25770	17580	25975	22052	22174	17196	11017	7082	4527	2781	1815
Difference	2793	2793	52	70	12	502	1874	702	4050	825	4230	12420	4025	7948	7826	12804	18983	22918	25473	27219	28185
					Yr 5	173778				Yr 10	312096				Yr 15	417073				Yr 20	444294
Small	0	0	3	5	8	11	12	13	13	13	13	13	13	13	13	13	13	13	13	13	13
Medium	4	4	6	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9
Large	4	4	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5

After determining which well sites would be operating at a given period, the redeployment of the units was determined by proximity. Since the redeployment of the wells was used in the optimization of the initial plan, it allowed easy transition to determining where to send the units no longer needed at a certain well. Each unit was broken down by number to assign how the unit would be transferred from wells.

Table 23 and Table 24 illustrate the units' deployment for each well site broken down by the first 10 years and the last 10 years. The tables also analyze the redeployment downtimes and catalyst downtimes to minimize the total amount of downtime required for the entire project. The blue boxes depict when the units will be required to shut down for redeployment. The red boxes represent the years needed to replace the catalyst in the FTR. The purple boxes represent the years that the catalyst will be replaced during redeployment. The grey boxes represented the period of time that the units will not be in service. Minimizing the shutdown periods is critical to maximize the revenue. The wellhead will not need to be shut down for a month if a unit is being disconnected from the process or if the unit was previously connected and is still on site.

Table 23: First 10 Years of Redeployment and Shutdown Periods

Small units										
	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034
1	-	-	1A	1A	1A	1A	1A	1E	1E	1E
2	-	-	2A	2A	2A	2A	2A	2A	2A	2B
3	-	-	2A	2A	2A	2A	2A	2A	2A	2A
4	-	-	-	2F	2F	2F	2F	2F	2F	2F
5	-	-	-	2G	2G	2G	2G	2G	1H	1H
6	-	-	-	-	2B	2B	2B	2B	2B	2B
7	-	-	-	-	1A	1A	1A	1A	1A	1A
8	-	-	-	-	1A	1A	1A	2F	2F	2F
9	-	-	-	-	-	1D	1D	1D	1D	1D
10	-	-	-	-	-	1D	1D	1D	1D	1D
11	-	-	-	-	-	1D	1C	1C	1C	1C
12	-	-	-	-	-	-	2G	2G	2G	2H
13	-	-	-	-	-	-	-	1E	1E	1E
Medium units										
	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034
1	1D	1D	1D	1D	1D	1D	1D	1D	2D	2D
2	2G	2G	2G	2G	2G	2G	2H	2H	2H	2H
3	2G	2G	2G	2G	1C	1C	1C	1C	2D	2D
4	2G	2G	2G	2G	1C	1C	1C	1C	1C	1C
5	-	-	1A	1A	1A	1H	1H	1H	1H	1H
6	-	-	2B	2B	2B	2B	2B	2B	2B	2C
7	-	-	-	1E	1E	1E	1E	1E	1E	1G
8	-	-	-	1E	1E	1E	1E	1E	1E	1G
9	-	-	-	2F	2F	2F	2F	2F	2F	1G
Large units										
	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034
1	1A	1A	1A	1A	1A	1H	1H	1H	1H	1G
2	1D	1D	1D	1D	1D	1D	2H	2H	2H	2H
3	2A	2A	2B	2B	1C	1C	1C	2C	2C	2C
4	2A	2A	2A	2A	2A	2A	2A	2A	2A	2A
5	-	-	2B	2B	2B	2B	2B	2C	2C	2C

Table 24: Last 10 Years of Redeployment and Shutdown Periods

Small units										
	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
1	1E	1E	1E	1E	1E	1E	1E	1E	1E	1E
2	2B	2B	2D	2C	2C	2C	2C	2C	2C	2C
3	1H	1H	1H	1H	1H	1G	1G	1G	1G	1G
4	2F	2H	2H	2H	2H	2H	2H	2H	2H	2H
5	1H	1H	1H	1H	1H	1G	1G	1G	1G	1G
6	2B	2B	2B	1B	1B	1B	1B	1B	1B	1B
7	1B	1B	1B	1B	1B	1B	1B	1B	1B	1B
8	2F	2F	2F	2F	2F	2F	2F	2F	2F	2F
9	1D	1D	1D	2E	2E	2E	2E	2E	2E	2E
10	1D	1D	1D	2C	2C	2C	2C	2C	2C	2C
11	1C	1C	2D	2D	2D	2D	2D	2D	2D	2D
12	2H	2H	2H	2H	2H	2H	2H	2H	2H	2H
13	1C	1C	1C	1C	1C	1C	1C	1C	1C	1C

Medium units										
	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
1	2D	2D	2D	2D	2D	2D	2D	2D	2D	2D
2	2H	1F	1F	1F	1F	1F	1F	1F	1F	1F
3	2D	2D	2D	2D	2D	2D	2D	2D	2D	2D
4	1B	1B	1B	1B	1B	1B	1B	1B	1B	1B
5	1H	1H	1H	1H	1H	1H	1H	1H	1H	1H
6	2C	2C	2C	2C	2C	2C	2C	2C	2C	2C
7	1G	1F	1F	1F	1F	1F	1F	1F	1F	1F
8	1G	1G	1G	1G	1G	1G	1G	1G	1G	1G
9	1G	1G	1G	2E	2E	2E	2E	2E	2E	2E

Large units										
	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
1	1G	1F	1F	1F	1F	1F	1F	1F	1F	1F
2	2H	1F	1F	2E	2E	2E	2E	2E	2E	2E
3	2C	2C	2C	2C	2C	2C	2C	2C	2C	2C
4	2A	2A	2A	2A	2A	2A	2A	2A	2A	2A
5	2C	2C	2C	2C	2C	2C	2C	2C	2C	2C

As mentioned before, the cost of redeployment will cost $3 per MSCF per mile traveled. With transferring units to close by well sites, this minimizes the cost. The initial costs are accounted for by the initial deployment of the equipment into the field from the central plant. The distance between wells was calculated by the distance in miles of each well from the central plant and the angle between each well. Each ring of wells has a 45° angle between the well sites and to relate the two ring of wells it was determined that the inner ring would be half of the angle of the outer ring. For example, well 2A and 2B have a 45° angle from the central plant, while well 1B has a 22.5° angle from 2A and 2B in respect to the central plant. The total distance between each wellsite was calculated using the law of cosines. These distances were then used to calculate the cost of transportation of redeployment. Table 25 also includes the transportation cost of transferring the units back to the central plant for decommissioning before being removed.

Table 25: Transportation Cost of Redeployment

Cost of Transportation of Units Between Wells																					
Cost per MSCF per mile																					
$3																					
Small units	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
500																					
1	-	-	$93,000	-	-	-	-	$102,300	-	-	-	-	$9,300	-	-	-	-	-	-	-	-
2	-	-	$128,700	-	-	-	-	-	-	$130,891.26	-	-	-	$141,812.21	$60,507.99	$3,600	-	-	-	-	-
3	-	-	$128,700	-	-	-	-	-	-	-	-	$145,985.80	-	-	-	-	$94,920.77	-	$88,500	-	-
4	-	-	-	$19,500	-	-	-	-	-	-	-	-	$57,412.98	-	-	-	-	$54,000	-	-	-
5	-	-	-	$130,500	-	-	-	-	$51,702.48	-	-	-	-	-	-	-	$94,920.77	-	-	-	$88,500
6	-	-	-	-	$127,050	-	-	-	-	-	-	-	-	-	$98,686.79	-	-	-	-	-	$31,500
7	-	-	-	-	$93,000	-	-	-	-	-	$74,150.60	-	-	-	-	-	$31,500	-	-	-	-
8	-	-	-	-	$93,000	-	-	$102,064.90	-	-	-	-	$19,500	-	-	-	-	-	-	-	-
9	-	-	-	-	-	$110,100	-	-	-	-	-	-	-	-	$146,958.83	-	-	$148,200	-	-	-
10	-	-	-	-	-	$110,100	-	-	-	-	-	-	-	-	$106,782.92	-	-	-	-	-	$3,600
11	-	-	-	-	-	$110,100	$93,817.83	-	-	-	-	-	-	$58,221.03	-	-	-	-	$63,000	-	-
12	-	-	-	-	-	-	$130,500	-	-	$99,901.39	-	-	-	-	-	-	-	$54,000	-	-	-
13	-	-	-	-	-	-	-	$9,300	-	-	$27,142.95	-	-	-	-	-	-	-	-	-	$25,500
Medium units	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
2500																					
1	$550,500	-	-	-	-	-	-	-	$284,752.15	-	$315,000	-	-	-	-	-	-	-	-	-	-
2	$652,500	-	-	-	-	-	$499,506.93	-	-	-	-	-	$824,466.03	-	-	-	-	-	-	-	$682,500
3	$652,500	-	-	-	$771,838.39	-	-	-	$291,105.14	-	-	-	-	$315,000	-	-	-	-	-	-	-
4	$652,500	-	-	-	$771,838.39	-	-	-	-	-	$112,531.44	-	-	-	$157,500	-	-	-	-	-	-
5	-	-	$465,000	-	-	$473,663	-	-	-	-	-	$669,750	-	-		-	-	-	-	-	-
6	-	-	$635,250	-	-	-	-	-	-	$622,652.18	-	-	-	-	$18,000	-	-	-	-	-	-
7	-	-	-	$46,500	-	-	-	-	-	$444,936.51	-	-	-	-	-	-	-	-	$442,500	-	-
8	-	-	-	$46,500	-	-	-	-	-	$444,936.51	-	-	-	-	-	-	$442,500	-	-	-	-
9	-	-	-	$97,500	-	-	-	-	-	$444,936.51	-	-	-	-	$702,800.96	-	-	-	-	-	$741,000
Large	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
5000																					
1	$930,000	-	-	-	-	$947,325.14	-	-	-	-	-	-	$1,912,455.29	-	$1,365,000	-	-	-	-	-	-
2	$1,101,000	-	-	-	-	-	$1,610,210.33	-	-	-	-	-	$1,648,932.05	-	$567,152.32	-	-	-	$1,482,000	-	-
3	$1,287,000	-	$978,831.59	-	$1,039,501.27	-	-	$222,167.89	-	$36,000	-	-	-	-	-	-	-	-	-	-	-
4	$1,287,000	-	-	-	-	-	-	$1,287,000	-	-	-	-	-	-	-	-	-	-	-	-	-
5	-	-	$1,270,500	-	-	-	-	$1,245,304.36	-	-	-	-	-	$36,000	-	-	-	-	-	-	-
Total Transportation Cost	$7,113,000	$0	$3,699,982	$340,500	$2,896,228	$1,751,288	$2,334,035	$2,968,137	$627,560	$2,224,254	$528,825	$815,736	$4,472,066	$551,033	$3,223,390	$3,600	$663,842	$256,200	$2,076,000	$0	$1,572,600

The products from the units will be collected from each well site and transported to the central plant for further processing. Each truck has a maximum capacity of 5,000 gallons of liquid. LPG, naphtha, and diesel will require separate trucks for each trip with cleaning in between routes to eliminate cross contamination. Table 26 shows the number of trips that will be required for the trucks to take per year for each product. This was determined by calculating the gallons produced in a year given the mass flow rate and density and then dividing them by the maximum capacity of the trucks. Each product will be collected at least once from a given well per year to empty the storage tank and collect a revenue for that product in that year regardless of the quantity in the tank. Some trucks will be carrying a smaller quantity of product for each route, while others are carrying the maximum capacity.

Table 26: Number of Trips per Year Required per Product

Year	2024	2025	2026	2027	2028	2029	2030
LPG	0	14	14	15	18	19	18
Naphtha	0	1238	1238	1272	1348	1360	1315
Diesel	0	1567	1567	1715	1750	1774	1706
Year	2031	2032	2033	2034	2035	2036	2037
LPG	19	22	21	22	21	16	19
Naphtha	1250	1331	1195	1313	1202	842	1169
Diesel	1646	1712	1528	1734	1508	1065	1543
Year	2038	2039	2040	2041	2042	2043	2044
LPG	16	19	15	10	8	7	3
Naphtha	1032	1115	802	402	313	226	139
Diesel	1295	1448	1027	506	409	282	174

The total number of trucks required was calculated by summing the number of trucks needed for all products in one year. The trucks will be required to make one trip per day considering the service factor of 80%. From this assumption, the total number of trips for each well can be divided by the days that the process is operating. Table 27 shows the total number of trucks that will be making trips throughout the year.

Table 27: Number of Trucks Required per Year

Year	2024	2025	2026	2027	2028	2029	2030
Trucks Required	0	10	8	9	9	9	9
Year	2031	2032	2033	2034	2035	2036	2037
Trucks Required	8	9	8	9	8	6	8
Year	2038	2039	2040	2041	2042	2043	2044
Trucks Required	7	8	6	3	2	2	1

The vehicles and drivers will be supplied on a 5-year contract. Due to this, the project life was divided into four timeframes of 5 years. In those five years, the maximum number of trucks needed at one time from that period will be the number of trucks contracted for the entire five years. Table 28 illustrates the number of trucks that will be contracted and the period that the contracts will begin and end.

Table 28: Number of Trucks Contracted

Contract Period	Trucks Contracted
2025 – 2029	10
2030 – 2024	9
2035 – 2039	8
2040 – 2044	6

The contract with the drivers states that it will cost $1.25/mile driven for the vehicle and driver for transportation of product to the central plant. Each truck will travel from the central plant to a desired well site to collect the product from the storage tanks on site. After the product is loaded onto the trucks, they will then travel directly back to the central plant for the product to be further processed. The total cost of transportation was calculated using the number of trips that a well site will require and the distance to and from the central plant. Table 29 and Table 30 show the transportation cost of the product from the well sites to the central plant broken down in the first 10 years and the last 10 years.

Table 29: Product Transportation Cost for the First 10 Years

		Transportation Cost of Product										
Well ID	distance (mile)	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034
1A	62.0	$ -	$ 67,890.00	$ 67,890.00	$ 49,600.00	$ 34,875.00	$ 17,205.00	$ 11,780.00	$ 7,750.00	$ 4,960.00	$ 3,410.00	$ 2,325.00
1B	21.0	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
1C	17.0	$ -	$ -	$ -	$ -	$ -	$ 36,040.00	$ 37,485.00	$ 25,670.00	$ 17,637.50	$ 11,517.50	$ 7,182.50
1D	73.4	$ -	$ 142,029.00	$ 142,029.00	$ 86,428.50	$ 54,316.00	$ 44,223.50	$ 21,836.50	$ 14,863.50	$ 9,725.50	$ 6,422.50	$ 4,404.00
1E	6.2	$ -	$ -	$ -	$ -	$ 5,766.00	$ 6,014.00	$ 3,906.00	$ 2,449.00	$ 1,271.00	$ 852.50	$ 573.50
1F	91.0	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
1G	59.0	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ 171,690.00
1H	89.3	$ -	$ -	$ -	$ -	$ -	$ -	$ 157,168.00	$ 163,195.75	$ 120,555.00	$ 73,002.75	$ 51,347.50
2A	85.8	$ -	$ 191,548.50	$ 191,548.50	$ 119,047.50	$ 81,081.00	$ 52,767.00	$ 33,033.00	$ 22,522.50	$ 14,157.00	$ 9,223.50	$ 6,006.00
2B	84.7	$ -	$ -	$ -	$ 252,194.25	$ 261,934.75	$ 167,917.75	$ 106,510.25	$ 69,454.00	$ 54,631.50	$ 35,574.00	$ 17,787.00
2C	2.4	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ 4,920.00	$ 5,100.00	$ 3,174.00
2D	42.0	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ 45,675.00	$ 47,460.00
2E	98.8	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
2F	13.0	$ -	$ -	$ -	$ -	$ 10,302.50	$ 10,692.50	$ 7,182.50	$ 4,485.00	$ 2,405.00	$ 1,592.50	$ 1,072.50
2G	87.0	$ -	$ 155,295.00	$ 155,295.00	$ 101,137.50	$ 63,292.50	$ 42,847.50	$ 27,840.00	$ 14,137.50	$ 9,570.00	$ 6,307.50	$ -
2H	36.0	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ 58,950.00	$ 61,290.00	$ 41,400.00	$ 27,360.00
Total Transportation Cost		$ -	$ 556,762.50	$ 556,762.50	$ 608,407.75	$ 511,567.75	$ 377,707.25	$ 406,741.25	$ 383,477.25	$ 301,122.50	$ 240,077.75	$ 340,382.00

Table 30: Product Transportation Cost for the Last 10 Years

Transportation Cost of Product											
Well ID	distance (mile)	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**
1A	62.0	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
1B	21.0	$ 14,910.00	$ 15,487.50	$ 11,602.50	$ 7,875.00	$ 3,937.50	$ 2,677.50	$ 1,785.00	$ 1,155.00	$ 787.50	$ -
1C	17.0	$ 3,697.50	$ 2,550.00	$ 1,657.50	$ 1,062.50	$ 765.00	$ 510.00	$ -	$ -	$ -	$ -
1D	73.4	$ 3,119.50	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
1E	6.2	$ 372.00	$ 263.50	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
1F	91.0	$ -	$ -	$ 334,652.50	$ 347,392.50	$ 299,390.00	$ 159,932.50	$ 108,290.00	$ 70,297.50	$ 44,362.50	$ 30,030.00
1G	59.0	$ 178,180.00	$ 120,802.50	$ 81,715.00	$ 53,100.00	$ 33,482.50	$ 22,567.50	$ 11,357.50	$ 7,670.00	$ 5,015.00	$ 3,392.50
1H	89.3	$ 33,487.50	$ 16,743.75	$ 11,385.75	$ 7,590.50	$ 4,911.50	$ 3,348.75	$ -	$ -	$ -	$ -
2A	85.8	$ 4,075.50	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
2B	84.7	$ 12,069.75	$ 8,046.50	$ 5,293.75	$ 3,599.75	$ -	$ -	$ -	$ -	$ -	$ -
2C	2.4	$ 2,160.00	$ 1,698.00	$ 1,062.00	$ 720.00	$ 360.00	$ 246.00	$ 156.00	$ 114.00	$ 78.00	$ -
2D	42.0	$ 30,870.00	$ 19,320.00	$ 13,125.00	$ 6,615.00	$ 4,515.00	$ 3,045.00	$ 1,890.00	$ 1,365.00	$ -	$ -
2E	98.8	$ -	$ -	$ -	$ -	$ 195,130.00	$ 202,540.00	$ 67,184.00	$ 77,805.00	$ 63,726.00	$ 39,767.00
2F	13.0	$ 682.50	$ 487.50	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
2G	87.0	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
2H	36.0	$ 18,900.00	$ 12,330.00	$ 6,120.00	$ 4,230.00	$ 2,790.00	$ 1,800.00	$ 1,350.00	$ -	$ -	$ -
Total Transportation Cost		$ 302,524.25	$ 197,729.25	$ 466,614.00	$ 432,185.25	$ 545,281.50	$ 396,667.25	$ 192,012.50	$ 158,406.50	$ 113,969.00	$ 73,189.50

Once the units are no longer needed, they will be transported back to the central plant. The equipment will be retired at the central plant to be decommissioned at the end of the project life. At this time, the equipment will be salvaged with and removal cost of $25,000 per unit.

<u>Economics</u>

An important objective of the design was to maximize the NPV. Maximizing the NPV was done by optimizing the product revenue while keeping capital cost, utility cost, and other manufacturing cost at a minimum. There is a tradeoff between the cost of equipment and manufacturing cost, as well as the revenue made from the product. The design reflects optimization and balance of these variables. It was estimated that the modular units would be built January of 2024 and deployed December of 2024 to begin operating in January of 2025. Estimating a built date of January 2022 allowed for 6 to 12 months of detailed design as well as 6 months for bidding of equipment and an additional 6 months to obtain required permits before units are built. A positive net present value of $4.35 B was achieved for this project. The DCFROR was calculated to be 9.1% which is greater than the 8% hurdle rate given for the project. The project NPV and IRR show that the project design is economically attractive [16].

a. Capital Cost Estimates

The capital cost was estimated for the fractionator and FTR by using the module costing technique. Since the equipment is modular there is expected error within the cost estimate since modular variables such as reactor or heat exchanger area do not fall within the module costing correlations. To limit the modular bare module cost error, the purchase cost of base conditions was multiplied by 4.8 instead of calculating for pressure factor and material of construction factor. The 4.8 multiplier is given in the AIChE design problem statement and accounts for the direct costs, indirect costs, and working capital for the modular equipment. The cost estimate equation constants are from a 2001 price index, therefore a 2001 CEPCI value of 397 and 2022 CEPCI value of 776.9 were applied to account for the inflation between years 2001 and 2022 [17]. The capital costs were then escalated again to year 2024 with an inflation rate of 3% for two years. Guthrie's Method was used to estimate the total modular cost to include 18% of contingency and fees. The Syngas Unit, CO_2 Recovery System, Air Separation Plant, and HIU capital cost were estimated by equipment cost attributes and given scaling factors. Table 31 shows the bare module cost of the FTR and separator, the capital cost of the Syngas Unit, CO_2 Recovery System, and HIU, and the total capital cost for each small, medium, and large unit.

Table 31: Capital Cost of Units

	Small Unit		**Medium Unit**		**Large Unit**	
	Bare Module Cost	**Capital Cost**	**Bare Module Cost**	**Capital Cost**	**Bare Module Cost**	**Capital Cost**
FTR	$1,140	$1,350	$1,480	$1,740	$1,830	$2,160
Separator	$214,000	$252,000	$487,000	$574,000	$650,000	$768,000
Syngas Unit	–	$1,980,000	–	$5,890,000	–	$9,290,000
CO_2 Recovery System	–	$398,000	–	$897,000	–	$1,390,000
HI Unit	–	$10,700,000	–	$29,600,000	–	$39,000,000
Total Capital Cost	$13.3 MM		$37.0 MM		$50.5 MM	

Another economic factor considered in the design is the cost of modular manufacturing. There is a cost reduction for every unit made after the first unit built. This first-of-a-kind (FOAK) concept is specified in the project statement. The economy of mass production decreases the unit cost by a "p" factor when number of units are doubled as well as 20% learning rate which decreases the cost by 20% when number of units are doubled. A summary of the FOAK cost is given in Table 32 , Table 33, and Table 34 for the small, medium, and large units, respectively.

Table 32: Small Unit FOAK Cost

Unit Number	**FOAK Cost**
Small Unit	
1	$13.3 MM
2	$10.6 MM
3	$9.3 MM
4	$8.5 MM
5	$7.9 MM
6	$7.4 MM
7	$7.1 MM
8	$6.8 MM
9	$6.6 MM
10	$6.3 MM
11	$6.2 MM
12	$6.0 MM
13	$5.8 MM

Table 33: Medium Unit FOAK Cost

Unit Number	FOAK Cost
Medium Unit	
1	$37.0 MM
2	$29.6 MM
3	$26.0 MM
4	$23.7 MM
5	$22.0 MM
6	$20.8 MM
7	$19.8 MM
8	$18.9 MM
9	$18.2 MM

Table 34: Large Unit FOAK Cost

Unit Number	FOAK Cost
Large Unit	
1	$50.5 MM
2	$40.4 MM
3	$35.5 MM
4	$32.3 MM
5	$30.0 MM

b. Revenue Estimates

Revenue includes the LPG, naphtha, and diesel produced and maximum revenue made for each is given in Table 35. A 11/12 factor was also used to account for the one-month turnaround that occurs when needed. The turnaround period is marked in green.

Table 35: Product Revenue

										Product Revenue												
	distance (mile)	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
1A	62	$4,494	$4,494	$2,921	$1,899	$1,234	$802	$521	$339	$220	$143											
LPG		$3,026	$3,026	$2,895	$2,006	$1,472	$484	$315	$205	$133	$86											
Naphtha		$5,751	$5,751	$4,570	$3,089	$745	$957	$622	$404	$263	$171											
Diesel		$9,039	$9,039	$6,628	$4,417	$2,379	$1,546	$1,005	$653	$425	$276											
1B	21											$2,989	$2,989	$1,943	$1,263	$821	$534	$347	$225	$147		
LPG												$2,936	$2,936	$2,052	$1,334	$496	$322	$209	$136	$88		
Naphtha												$4,651	$4,651	$3,161	$2,055	$979	$636	$414	$269	$175		
Diesel												$6,400	$6,400	$4,520	$2,938	$956	$621	$404	$262	$171		
1C	17					$8,365	$8,356	$5,431	$3,530	$2,295	$1,492	$970	$630	$410	$266	$173	$112					
LPG						$6,921	$6,912	$5,542	$3,729	$2,424	$1,576	$586	$381	$247	$161	$105	$68					
Naphtha						$11,873	$11,858	$8,649	$5,744	$3,733	$2,427	$1,157	$752	$489	$318	$206	$134					
Diesel						$17,885	$17,864	$12,464	$8,214	$5,339	$3,470	$1,869	$1,215	$790	$513	$334	$217					
1D	73	$7,447	$7,447	$4,841	$3,146	$2,045	$1,329	$864	$562	$365	$237	$154										
LPG		$5,952	$5,952	$3,259	$2,119	$2,160	$803	$522	$339	$220	$143	$93										
Naphtha		$10,379	$10,379	$6,194	$4,026	$3,327	$1,586	$1,031	$670	$435	$283	$184										
Diesel		$15,750	$15,750	$6,556	$4,261	$4,758	$2,563	$1,666	$1,083	$704	$457	$297										
1E	6				$3,290	$3,290	$2,139	$1,390	$904	$587	$382	$248	$161									
LPG					$3,476	$3,476	$2,259	$1,468	$546	$355	$231	$150	$97									
Naphtha					$5,353	$5,353	$3,479	$2,261	$1,078	$701	$455	$296	$192									
Diesel					$7,654	$7,654	$4,975	$3,234	$1,742	$1,132	$736	$478	$311									
1F	91													$14,737	$14,737	$9,579	$6,226	$4,047	$2,631	$1,710	$1,111	$722
LPG														$11,738	$11,738	$8,204	$6,577	$4,275	$2,779	$1,806	$1,174	$763
Naphtha														$20,503	$20,503	$13,848	$10,130	$6,584	$4,280	$2,782	$1,808	$1,175
Diesel														$31,134	$31,134	$20,710	$14,486	$9,416	$6,120	$3,978	$2,586	$1,681
1G	59										$11,155	$11,155	$7,251	$4,713	$3,063	$1,991	$1,294	$841	$547	$355	$231	$150
LPG											$9,869	$9,869	$7,660	$4,979	$3,236	$2,104	$1,367	$508	$330	$215	$140	$91
Naphtha											$16,412	$16,412	$11,797	$7,668	$4,984	$3,240	$2,106	$1,004	$652	$424	$276	$179
Diesel											$24,376	$24,376	$16,869	$10,965	$7,127	$4,633	$3,011	$1,622	$1,054	$685	$445	$289
1H	89						$7,081	$7,081	$4,603	$2,992	$1,945	$1,264	$822	$534	$347	$226	$147					
LPG							$5,565	$5,565	$4,862	$2,938	$2,054	$1,335	$496	$323	$210	$136	$89					
Naphtha							$9,784	$9,784	$7,488	$4,654	$3,164	$2,056	$980	$637	$414	$269	$175					
Diesel							$14,898	$14,898	$10,708	$6,764	$4,524	$2,941	$1,584	$1,029	$669	$435	$283					
2A	86	$9,182	$9,182	$5,968	$3,879	$2,522	$1,639	$1,065	$692	$450	$293	$190										
LPG		$6,183	$6,183	$3,952	$2,612	$1,698	$1,104	$717	$418	$272	$177	$115										
Naphtha		$11,750	$11,750	$7,553	$4,964	$3,227	$2,097	$1,363	$826	$537	$349	$227										
Diesel		$18,468	$18,468	$11,923	$7,803	$5,072	$3,297	$2,143	$1,335	$868	$564	$367										
2B	85			$12,258	$12,258	$7,968	$5,179	$3,366	$2,188	$1,422	$924	$601	$391	$254	$165							
LPG				$9,119	$9,119	$6,290	$3,367	$2,267	$2,312	$1,502	$558	$363	$236	$153	$100							
Naphtha				$16,470	$16,470	$11,023	$6,398	$2,901	$3,761	$2,444	$666	$433	$281	$183	$119							
Diesel				$25,366	$25,366	$16,774	$10,057	$5,834	$8,749	$5,687	$1,284	$834	$542	$353	$229							
2C	2								$8,742	$8,742	$5,682	$3,693	$2,401	$1,561	$1,014	$659	$429	$279	$181	$118		
LPG									$5,886	$5,886	$3,779	$2,487	$2,536	$1,648	$1,072	$398	$259	$168	$109	$71		
Naphtha									$11,187	$11,187	$7,212	$4,726	$3,906	$2,539	$1,650	$786	$511	$332	$216	$140		
Diesel									$17,583	$17,583	$11,372	$7,429	$5,585	$3,631	$2,360	$1,271	$826	$537	$349	$227		
2D	42									$3,840	$3,840	$2,496	$1,622	$1,055	$685	$446	$290	$188	$122			
LPG										$4,057	$4,057	$2,637	$1,714	$1,114	$414	$269	$175	$114	$74			
Naphtha										$6,247	$6,247	$4,061	$2,640	$1,716	$818	$531	$345	$225	$146			
Diesel										$8,934	$8,934	$5,807	$3,775	$2,453	$1,321	$859	$558	$363	$236			
2E	99															$7,939	$7,939	$5,160	$3,354	$2,180	$1,417	$921
LPG																$6,273	$6,273	$1,954	$2,259	$2,303	$1,497	$973
Naphtha																$10,989	$10,989	$3,607	$4,292	$3,547	$2,306	$1,499
Diesel																$16,719	$16,719	$5,546	$6,746	$5,072	$3,297	$2,143
2F	13				$2,874	$2,874	$1,868	$1,214	$789	$513	$333	$217	$141									
LPG					$2,867	$2,867	$1,973	$1,283	$477	$310	$201	$131	$85									
Naphtha					$4,513	$4,513	$3,039	$1,976	$941	$612	$398	$259	$168									
Diesel					$6,537	$6,537	$4,346	$2,825	$1,521	$989	$643	$418	$272									
2G	87	$6,076	$6,076	$3,949	$2,567	$1,669	$1,085	$705	$458	$298												
LPG		$6,419	$6,419	$4,172	$2,681	$1,763	$1,146	$426	$277	$180												
Naphtha		$9,885	$9,885	$6,425	$4,284	$2,715	$1,765	$841	$547	$355												
Diesel		$14,136	$14,136	$9,188	$9,255	$3,882	$2,523	$1,359	$883	$574												
2H	36							$6,468	$6,468	$4,204	$2,733	$1,776	$1,155	$750	$488	$317	$206	$134				
LPG								$5,313	$5,313	$3,788	$3,989	$1,876	$1,220	$453	$295	$191	$124	$81				
Naphtha								$9,145	$9,145	$6,248	$6,731	$2,890	$1,878	$895	$582	$378	$246	$160				
Diesel								$13,797	$13,797	$3,428	$10,120	$4,133	$2,686	$1,447	$940	$611	$397	$258				
Product Revenue per day																						
LPG		$21,579	$21,579	$23,397	$24,879	$26,647	$23,613	$23,418	$24,364	$22,066	$26,720	$22,578	$17,361	$22,708	$18,558	$18,176	$15,254	$7,309	$5,687	$4,484	$2,811	$1,827
Naphtha		$37,765	$37,765	$41,213	$42,700	$42,777	$40,963	$38,573	$41,790	$37,417	$44,515	$37,351	$27,245	$37,790	$31,442	$31,228	$25,273	$12,326	$9,855	$7,068	$4,389	$2,853
Diesel		$57,392	$57,392	$59,662	$65,294	$64,942	$62,070	$59,225	$66,269	$52,426	$66,757	$55,349	$39,238	$56,321	$47,232	$46,527	$37,119	$18,146	$14,768	$10,133	$6,328	$4,113
All Product per day		$116,736	$116,736	$124,271	$132,873	$134,366	$126,646	$121,216	$132,423	$111,909	$137,992	$115,277	$83,844	$116,819	$97,232	$95,930	$77,646	$37,781	$30,311	$21,685	$13,528	$8,794
All Product per year		$34,086,983	$34,086,983	$34,134,264	$37,411,251	$37,401,047	$35,965,974	$33,677,487	$37,639,998	$31,826,785	$37,740,287	$33,190,643	$23,326,265	$32,057,221	$28,305,636	$26,802,686	$21,894,971	$10,913,504	$8,510,887	$6,123,642	$3,756,616	$2,567,706

c. Operating Expense Estimates

The utility cost of the heat exchangers, the FTR unit, and the Syngas Unit are shown in Table 36. Fuel gas was needed for the syngas feed preheat furnace and accounted for 80% of the furnace duty. The remanding amount of furnace duty was split between electricity and 125 psig steam. A service factor of 0.8 was applied to the calculation.

Table 36: Utility Cost of Heat Exchanger, FTR Unit, and Syngas Unit

Equipment	Utility Cost ($/yr)		
	Small Unit	**Medium Unit**	**Large Unit**
E-103	$77,100	$454,000	$799,000
E-104	$4,530	$17,600	$39,900
E-105	$5,890	$38,300	$62,200
E-106	$2,460	$20,900	$27,200
FTR Unit	$70,200	$502,000	$589,000
Syngas Unit	$98,500	$507,000	$990,000

An additional utility of cooling water was required for the Air Separation Plant at a rate of 400 gpm per short ton/day. For the Air Separation Plant, it was found that the electricity option was significantly less expensive per year than the steam option when combined with the utility cost of O_2 feed. Wastewater treatment was also included in manufacturing cost and includes the water stream removed from the Syngas Unit and the water streams removed from the separator system. A summary of the wastewater treatment cost is shown in Table 37.

Table 37: Wastewater Treatment Cost

	Small	Medium	Large
Cost ($/yr)	$5,240	$28,900	$53,500

Reactant costs are shown in Table 38 and include the H_2 for the HIU and the O_2, and MP steam supplied to the Syngas Unit. The CO_2 is recycled from the absorber to the syngas feed after the FTR and will not require a supplemental CO_2 stream when operating at steady state.

Table 38: Reactant Costs

Reactant	Small	Medium	Large
MP Steam	$3,900	$19,200	$38,800
Oxygen	$235,000	$1,400,000	$2,450,000
Hydrogen	$24,000	$153,000	$254,000

Another manufacturing cost includes the cost of catalyst for each of the small, medium, and large FTR unit shown in Table 39.

Table 39: Cost of Catalyst

	Small	**Medium**	**Large**
Catalyst Weight (lb)	$25	$36	$44

Transportation costs were an operating cost considered in the design and are summarized in Table 25. Other yearly operating costs were found by multiplying the utilities cost by 3% and estimates extra costs such as plant overhead cost, administrative costs, distribution and marketing, as well as research and development. A removal cost of units also had to be implemented which is equivalent to $25,000 for every unit removed.

d. Discounted Cash Flow

To calculate the PWC, NPV, and IRR a cash flow table was constructed seen in Table 40 and Table 41. The cash flow is modeled as an expense project over a 20-year project life. Fixed capital cost was depreciated with a 7-year straight line depreciation method.

Table 40: Cash Flow Table

Project Title	Modular Distributed Gas-to-Liquids (GTL) Synthesis								
Corporate Financial Situation	Expense								
Minimum ROR	8%	or	0.08						
Inflation	3%	or	0.03						
Real ROR	5%		0.05						
$1=$1000									
Actual Year	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**
End of Year	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**
+Revenue		34,087	34,087	34,134	37,411	37,401	35,966	33,677	37,640
+Salvage Value									
Net Revenue	0	34,087	34,087	34,134	37,411	37,401	35,966	33,677	37,640
-Utilities		(84,757)	(84,757)	(87,469)	(91,377)	(91,499)	(89,753)	(82,695)	(90,046)
-Transportation		(303)	(812)	(1,111)	(1,172)	(1,198)	(903)	(764)	(725)
-Yearly Expenses (Fixed Charges)		(2,543)	(2,543)	(2,624)	(2,741)	(2,745)	(2,693)	(2,481)	(2,701)
-Removal Cost									
-Depreciation		(114,655,840)	(114,655,840)	(114,766,854)	(114,845,866)	(114,869,609)	(114,891,380)	(114,898,418)	(242,578)
Taxable Income	0	(114,709,356)	(114,709,865)	(114,823,924)	(114,903,744)	(114,927,651)	(114,948,763)	(114,950,681)	(298,411)
-Tax @ 20%	0	22,941,871	22,941,973	22,964,785	22,980,749	22,985,530	22,989,753	22,990,136	59,682
Net Income	0	(91,767,485)	(91,767,892)	(91,859,139)	(91,922,996)	(91,942,121)	(91,959,010)	(91,960,545)	(238,729)
+Depreciation		114,655,840	114,655,840	114,766,854	114,845,866	114,869,609	114,891,380	114,898,418	242,578
+Writeoff									
-Fixed Capital (Equipment)	(116,317,519)		(112,623)	(80,157)	(24,088)	(22,086)	(7,140)	0	0
Cash Flow	(116,317,519)	22,888,355	22,775,326	22,827,558	22,898,782	22,905,403	22,925,229	22,937,874	3,850
Discount factor (P/F)	1.00	0.93	0.86	0.79	0.74	0.68	0.63	0.58	0.54
Discounted Cash Flow	(116,317,519)	21,192,922	19,526,171	18,121,252	16,831,289	15,589,032	14,446,783	13,384,029	2,080
PWC	4,694,170								
NPV	4,346,454								
DCFROR	9.096%								

Table 41: Continued Cash Flow Table

2033	**2034**	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**
9	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**
31,827	37,740	33,191	23,326	32,057	28,306	26,803	21,895	10,914	8,511	6,124	3,757
											11,656,361
31,827	37,740	33,191	23,326	32,057	28,306	26,803	21,895	10,914	8,511	6,124	11,660,118
(80,672)	(93,279)	(81,771)	(56,260)	(78,450)	(69,786)	(66,783)	(53,737)	(26,851)	(21,243)	(14,692)	(8,908)
(581)	(464)	(582)	(543)	(373)	(924)	(898)	(1,190)	(909)	(462)	(400)	(302)
(2,420)	(2,798)	(2,453)	(1,688)	(2,353)	(2,094)	(2,003)	(1,612)	(806)	(637)	(441)	(267)
											(675)
(242,578)	(131,564)	(52,553)	(28,809)	(7,038)	0	0	0	0	0	0	0
(294,424)	(190,366)	(104,168)	(63,974)	(56,157)	(44,498)	(42,882)	(34,645)	(17,652)	(13,832)	(9,409)	11,649,965
58,885	38,073	20,834	12,795	11,231	8,900	8,576	6,929	3,530	2,766	1,882	(2,329,993)
(235,539)	(152,293)	(83,334)	(51,179)	(44,926)	(35,598)	(34,306)	(27,716)	(14,122)	(11,066)	(7,527)	9,319,972
242,578	131,564	52,553	28,809	7,038	0	0	0	0	0	0	0
0	0	0	0	0	0	0	0	0	0	0	0
7,039	(20,728)	(30,782)	(22,370)	(37,887)	(35,598)	(34,306)	(27,716)	(14,122)	(11,066)	(7,527)	9,319,972
0.50	0.46	0.43	0.40	0.37	0.34	0.32	0.29	0.27	0.25	0.23	0.21
3,521	(9,601)	(13,202)	(8,884)	(13,931)	(12,120)	(10,815)	(8,090)	(3,817)	(2,769)	(1,744)	1,999,583

Units were built a year before they were put in operation, so units were depreciated the year after they were built. Units were also inflated to the year they were built by using the 3% inflation rate given in the project statement. A salvage value was calculated with the AIChE method in the project statement where salvage values is assumed to be equivalent to 10% of the fixed capital cost. Working capital or write-off was not needed because it was assumed to be accounted for in the 4.8 multiplier for fixed capital costs. A tax rate of 20% was given and no loss forward was required for the expense cash flow and was assumed to be absorbed through other projects. No amortization or depletion was needed since no minerals, timbers, or patents were needed. All costs were brought to the present at an 8% hurdle rate.

<u>NPV and Sensitivity Analysis</u>

An NPV of $4.35 B was achieved for this project. An internal rate of return was calculated to be 9.1%, which is greater than the 8% hurdle rate given for the project. The project NPV and IRR show that the project designed is economically attractive. However, with any preliminary design cost estimates there are errors and uncertainties present. Since the design cost estimates were based on heuristic methods for conventional process plants, there are uncertainties existing for the estimated modular processing plant.

a. Payback Period

The cumulative cash position is shown in Figure 8 and a payback period was determined to be 6 years and 9 months which would occur September of 2030.

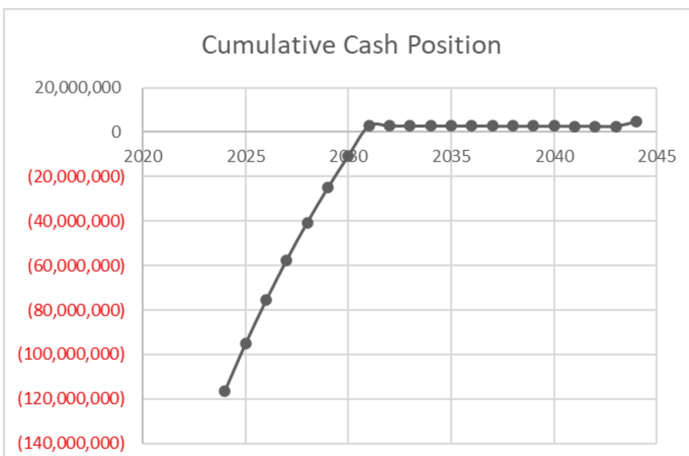

Figure 8: Cumulative Cash Position

b. Monte-Carlo Sensitivity Analysis

A propagation of uncertainty analysis was accomplished using the Monte-Carlo method. The Monte-Carlo method determines the probability of a parameter to occur. For the designed project, the probability of NPV was determined by varying the revenue, utilities, and fixed capital costs and applying a function of dependent of the variables. A distribution is then found by measuring the frequency of each NPV occurring out of 22,000 iterations. The Monte-Carlo

Method Analysis NPV frequency is shown in Figure 9. The probability of NPV occurring is then calculated from the frequencies and is shown in Figure 10.

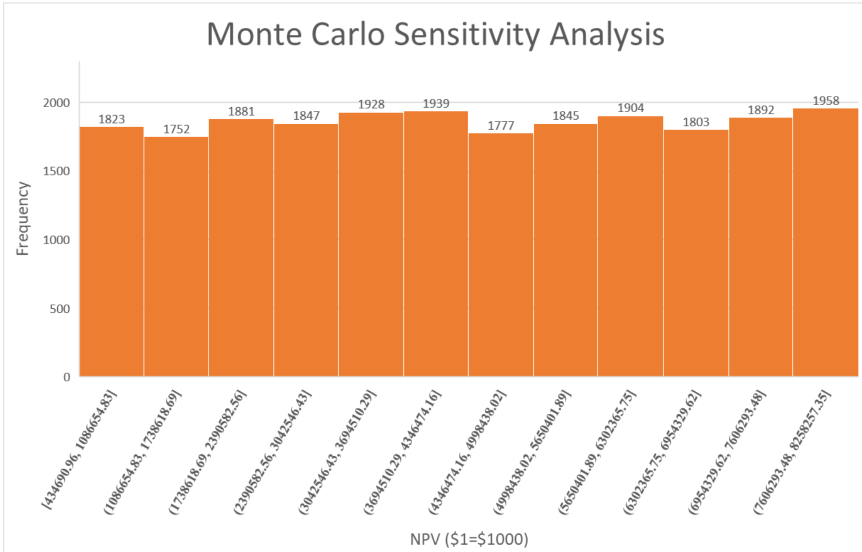

Figure 9: Monte-Carlo Sensitivity Analysis

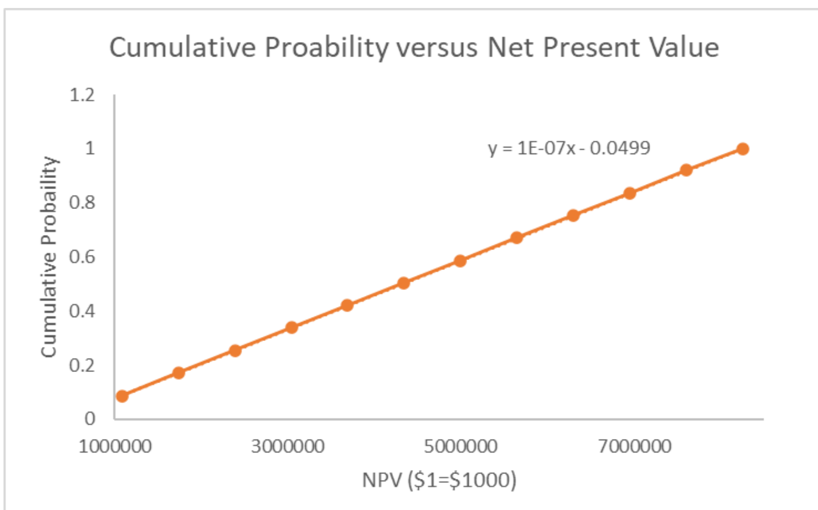

Figure 10: Monte-Carlo Cumulative Probability

The cumulative probability distribution in Figure 10 is interpreted as a uniform probability function [8]. The probability does not increase or decrease with increasing or decreasing NPV. The cumulative probability also shows that there is a 30% probability of the NPV to be less than or equal to the NPV calculated for the design project.

c. Single-Variable Sensitivity Analysis

A tornado chart was constructed for a single-variable analysis. Revenue, utility cost, and transportation costs were parameters varied because of their impact on the NPV as well as the uncertainties possible with these parameters. Other operating costs were not varied because they were directly calculated from the cost of utilities or fixed capital cost. Best-case and worst-case predictions used for the single-variable analysis shown in Table 42 were from *Table 10.1 Range of Variation of Factors Affecting the Profitability of a Chemical Process* in Turton, et.al [8]. However, these variables are from a 10-year project life, and it should be noted that the change in predictions could be minimal to the actual risks present. Figure 11 shows the tornado chart results.

Table 42: Single Variable Predictions

Variable	Prediction
Utility Cost	-10% to +100%
Revenue	+5 to –20%
Transportation	-10% to =100%

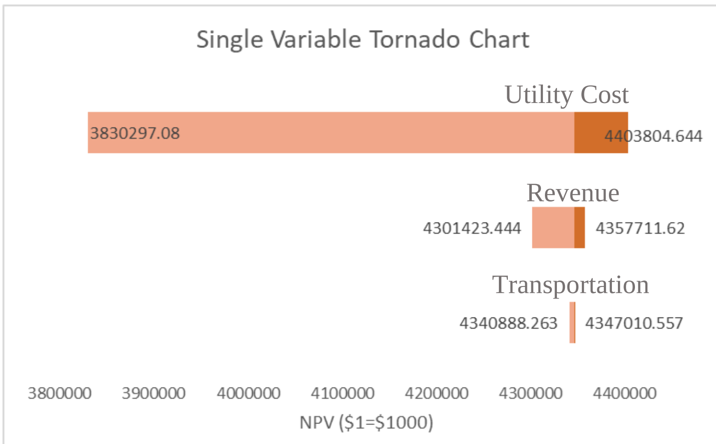

Figure 11: Single Variable Tornado Chart

It can be seen from the tornado chart that utility costs have the biggest impact on the project NPV compared to the other two variables and impose a larger risk on the success of the project. Capital costs were also varied but not shown because it had the biggest impact on NPV than other variables. However, it made utilities, transportation, and revenue almost negligible and their effects on NPV could not easily analyzed.

Conclusions

The conversion of methane into higher molecular weight hydrocarbons was designed for a Gas-To-Liquid plant by utilizing Fischer-Tropsch Synthesis with three modular sized units capable of handling 500 MMSF/day, 2500 MMSF/day, and 5,000 MMSF/day: small, medium, and large units, respectively. The modular designs are a more efficient design than conventional units since they can be redeployed on multiple wells. In total, 13 small units, 9 medium units, and 5 large units were required throughout the project life to maximize the product. A different number of trucks were contracted for each quarter of the project life, 10 during the first, 9 during the second, 8 for the third, and 6 in the last quarter.

The results of the project show that the PWC is approximately $4.69 B and the net NPV is approximately $4.35 B. The DCFROR is 9.10% with a hurdle rate of 8%. Based on the uncertainty analysis provided by the Monte Carlo Method the project has a 70% chance of producing a higher profit, however, it has a 30% probability of the NPV to be less than or equal to the NPV calculated. If the project were to decrease in profit, there could be a chance that the company would lose money through this project. However, the environmental impacts are less intense than flaring the methane into the atmosphere and therefore this project should be pursued.

Recommendations

Even though the conclusions of the project results in a reduced carbon footprint, there is still natural gas produced. A recycle stream of methane back to the Syngas Unit would allow the environmental hazards to be reduced even further.

Another recommendation would be to design a modular distillation column to further improve separation and maximize the products. This could potentially increase the NPV and make the project more desirable. A further way of saving money would be through heat integration. Heat integration using process streams for heat exchangers would cause there to be less cost of utilities. For the future of modular manufacturing, heuristics for modular designs would make the preliminary design stage even more accurate and cause less sources of error.

Additionally, the separation process could be improved if additional research was performed to find out if the heavier hydrocarbons form wax crystallization when cooled below 250 °F. It is standard practice to cool naphtha to 100 °F but this design limited that cooling due to the 250 °F restriction. The amount of heavier hydrocarbons that are in this stream are very little and could have the potential to have little to no affect. If the stream was cooled to 100 °F, this would result in an increase in Naphtha production, thus increasing revenue.

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Calculations and Simulation Outputs

Small Unit MATLAB Simulation

FTR Small Unit Optimization Program

This script optimizes the FTR differential program with respect to a constructed objective function

Include Parameters or Data

Set the initial point `x0` and scale a for the optimization.

```
clear all;
close all;

x0 = [50;50;390;0.1667];   % n; W; T; D
a = 0.001;        % Scaling factor for cost of n
```

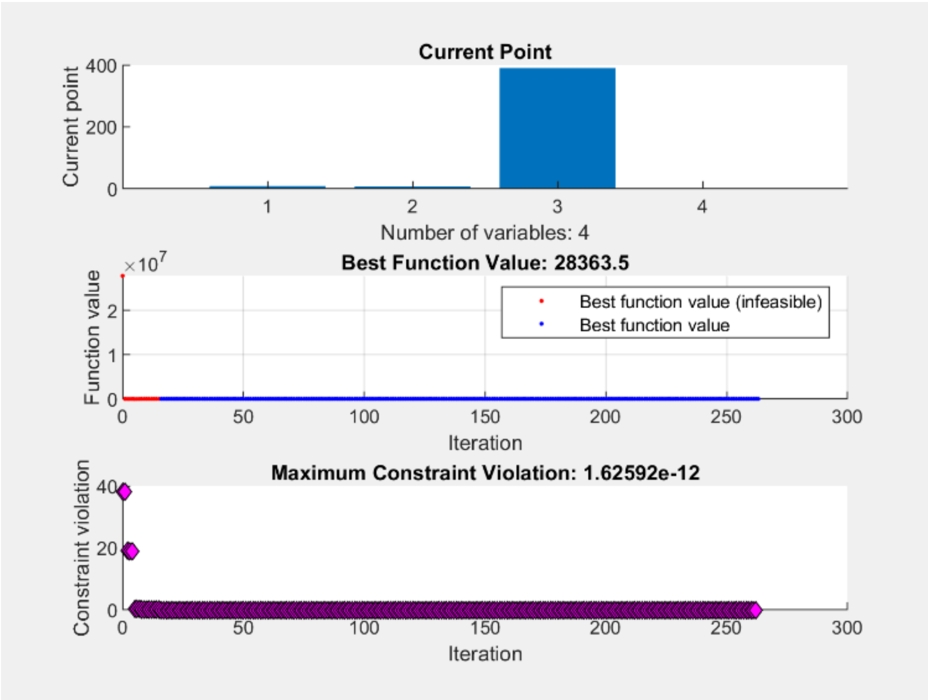
Place the `x0` value and any other problem data into the workspace by running this section before proceeding.

Optimize Live Editor Task

```
% Pass fixed parameters to objfun
objfun = @(x)objectiveFcn(x,a);

% Set nondefault solver options
options = optimoptions('fmincon','PlotFcn',{'optimplotx',...
    'optimplotfvalconstr','optimplotconstrviolation'});

% Solve
[solution,objectiveValue] = fmincon(objfun,x0,[],[],[],[],[],[],...
    @constraintFcn,options);
```



<stopping criteria details>

```
% Clear variables
clearvars objfun options
```

Results

View these variables.

```
sol.n = solution(1);
sol.w = solution(2);
sol.T = solution(3);
sol.D = solution(4);
disp(sol)
```

n: 9.3047

w: 8.5855

T: 390

D: 0.1667

```
objectiveValue
```

objectiveValue = 2.8363e+04

View the nonlinear constraint function values at the solution.

```
[ccons,ceqcons] = constraintFcn(solution)
```

ccons = 1×9

10^3 ×

```
   -0.0083   -1.7857         0         0   -0.0001   -0.0000   -0.0002 ⋯
```

ceqcons = 1.6259e-12

```
D  = 0.1667;
Ac = 3.14/4*D^2;
V  = solution(2)*454/0.8/28217;
L  = V/Ac
```

L = 7.9155

```
[w,X] = FTR_eq(solution);7
Xf = X(end,1)
```

Xf = 0.9000

```
[w,X] = FTR_eq(solution);
Pf = X(end,2)
```

Pf = 1.0000

```
solution(1)*solution(2)
```

ans = 79.8861

Helper Functions — Local Functions

The following code creates the objective function. Modify this code for your problem.

```
function f = objectiveFcn(x,a)
% Run simulation
[w, X] = FTR_eq(x);

% We care about w, n, &, Xf
Xf = X(end,1);
Wf = w(end);


f = a*((20/3*10*x(1)*Wf)^2); %-(10000*Xf)^2);
end
```

The following code creates the constraint function. Modify this code for your problem.

```
function [c,ceq] = constraintFcn(x)
[w,X] = FTR_eq(x);
Wf = w(end);
Xf = X(end,1);
Pf = X (end,2);

D   = 0.1667;
Ac  = 3.14/4*D^2;
V   = Wf*454/0.8/28217;
L   = V/Ac;
PT  = 450*Pf;
Total_P = 450 - PT;
Allowable_P = 1.5*Total_P;


bigPipeDiameter = 2*sqrt(x(1)*Ac/(.65*pi));

c(1) = 1-x(1);              % n > 1
c(2) = x(1)-1795;           % n < 1795
%c(3) = x(2)-100;           % Wf < 100
c(3) = x(3)-390;            % T < 450
c(4) = 390-x(3);            % 390 < T
c(5) = L - 8;               % L < 8ft
c(6) = x(4) - 0.1667;       % D < 0.1667 ft
c(7) = 0 - x(4);            % D > 0
c(8) = (bigPipeDiameter*3) - L;
c(9) = Allowable_P - 0.88;               % pT < 0.88 for PT/P0 and liquid
ceq  = [Xf - .9];                  % Xf = 0.75 (conversion ratio)
end
```

```
function dXdW = FTR_differential(~, y, params)

% Extract Parameters
X       = y(1);
p       = y(2);

n       = params(1);
T       = params(2);
k       = params(3);
k2      = params(4);
delta   = params(5);
P0      = params(6);
```

```
D       = params(7);
rho0    = params(8);
mu      = params(9);
phi     = params(10);
rhob    = params(11);
gc      = params(12);
Dp      = params(13);

% Explicit Equations
FT0 = 122.97/n;
FCO0 = 32.86/n;

yCO0 = FCO0/FT0;
e = yCO0*delta;
PCO0 = P0*yCO0;
Ac = 3.14/4*D^2;
G = FT0/Ac;

rhoc = rhob/(1-phi);

B0 = (G*(1-phi))/(rho0*gc*Dp*phi^3)*(((150*(1-
phi)*mu)/Dp)+1.75*G)/144^2/3600^2;
alpha = (2*B0)/(Ac*rhoc*(1-phi)*P0);
PCO = PCO0*((1-X)/(1+e*X))*p;
PH2 = PCO0*((2-2.06*X)/(1+e*X))*p;
PCOSI = PCO/14.7;
PH2SI = PH2/14.7;
TK = (T-32)*(5/9)+273;
T1 = exp(-4492*(1/TK-1/473));
T2 = exp(8237*(1/TK-1/473));
T3 = exp(-10000*(1/TK-1/473));

rCOSI = -(k*T1*PH2SI*PCOSI)/(1+k2+T2*PCOSI)^2;
rCO = rCOSI*28316.8/454;

rH2O = -rCO;
rCH4 = -rCO*0.03*T3;

rH2 = 2.06*rCO;
rC2 = 0.04*rCH4;
rC3 = 0.04*rCH4;
rC4 = 0.04*rCH4;


% ODEs
dXdW = [-rCO / FCO0;                % dX/dW
```

```
        (-alpha/(2*p))*(1+e*X);     % dp/dW
        rCH4;                       % dFCH4/dW
        rC2;                        % dFC2/dW
        rC3;                        % dFC3/dW
        rC4;                        % dFC4/dW
        rH2O;                       % dFH2O/dW
        rH2;                        % dFH2/dW
        ];
end
```

```
function [w,X] = FTR_eq(x)
% Setup parameters
n       = x(1);
T       = x(3);
k       = 0.0173;
k2      = 4.512;
delta   = -2;
P0      = 450;
D       = x(4);
rho0    = 0.5919;
mu      = 0.04196;
phi     = 0.4;
rhob    = 49.9424;
gc      = 32.2;
Dp      = 0.006667;


% Initial Conditions
X0      = 0;
p0      = 1;
FCH40   = 18.99/n;
FC20    = 0/n;
FC30    = 0/n;
FC40    = 0/n;
FH2O0   = 4.59/n;
FH20    = 65.67/n;


IC = [X0 p0 FCH40 FC20 FC30 FC40 FH2O0 FH20];
params = [n T k k2 delta P0 D rho0 mu phi rhob gc Dp];

% Span of catalyst (low high)
Wspan = [0 x(2)];
```

```
% Run simulation
[w, X] = ode45(@(t,y) FTR_differential(t,y,params), Wspan, IC);
end
```


Small Unit Polymath Report

POLYMATH Report	No Title
Ordinary Differential Equations	28-Mar-2022

Calculated values of DEQ variables

	Variable	Initial value	Minimal value	Maximal value	Final value
1	Ac	0.0218143	0.0218143	0.0218143	0.0218143
2	Allowablepressuredrop	0	0	0.0018409	0.0018409
3	alpha	9.47E-07	9.47E-07	9.47E-07	9.47E-07
4	B0	0.0002321	0.0002321	0.0002321	0.0002321
5	D	0.1667	0.1667	0.1667	0.1667
6	delta	-2.	-2.	-2.	-2.
7	Dp	0.006667	0.006667	0.006667	0.006667
8	e	-0.5344393	-0.5344393	-0.5344393	-0.5344393
9	FC2	0	0	0.0034218	0.0034218
10	FC3	0	0	0.0034218	0.0034218
11	FC4	0	0	0.0034218	0.0034218
12	FCH4	3.286	3.286	3.371546	3.371546
13	FCH40	3.286	3.286	3.286	3.286
14	FCO	3.286	0.2889359	3.286	0.2889359
15	FCO0	3.286	3.286	3.286	3.286

16	FCO20	0.06576	0.06576	0.06576	0.06576
17	FH2	6.5669	0.3929479	6.5669	0.3929479
18	FH20	6.5669	6.5669	6.5669	6.5669
19	FH2O	0.4585	0.4585	3.455564	3.455564
20	FH2O0	0.4585	0.4585	0.4585	0.4585
21	FN20	0.021	0.021	0.021	0.021
22	FT0	12.297	12.297	12.297	12.297
23	G	563.7133	563.7133	563.7133	563.7133
24	gc	32.2	32.2	32.2	32.2
25	k	0.0173	0.0173	0.0173	0.0173
26	k2	4.512	4.512	4.512	4.512
27	mu	0.04186	0.04186	0.04186	0.04186
28	n	10.	10.	10.	10.
29	p	1.	0.9999973	1.	0.9999973
30	P0	450.	450.	450.	450.
31	PCO	120.2488	20.62882	120.2488	20.62882
32	PCO0	120.2488	120.2488	120.2488	120.2488
33	PCOSI	8.180193	1.403321	8.180193	1.403321
34	PH2	240.4977	28.41896	240.4977	28.41896
35	PH2SI	16.36039	1.933263	16.36039	1.933263
36	phi	0.4	0.4	0.4	0.4
37	PT	450.	449.9988	450.	449.9988
38	rC2	0.0008185	6.72E-05	0.0008185	6.72E-05
39	rC3	0.0008185	6.72E-05	0.0008185	6.72E-05
40	rC4	0.0008185	6.72E-05	0.0008185	6.72E-05

41	rCH4	0.0204636	0.00168	0.0204636	0.00168
42	rCO	-0.716937	-0.716937	-0.0588572	-0.0588572
43	rCO2	0	0	0	0
44	rCOSI	-0.0114946	-0.0114946	-0.0009437	-0.0009437
45	rH2	-1.47689	-1.47689	-0.1212459	-0.1212459
46	rH2O	0.716937	0.0588572	0.716937	0.0588572
47	rho0	0.6123	0.6123	0.6123	0.6123
48	rhob	49.9424	49.9424	49.9424	49.9424
49	rhoc	83.23733	83.23733	83.23733	83.23733
50	rN2	0	0	0	0
51	T	390.	390.	390.	390.
52	T1	0.9778869	0.9778869	0.9778869	0.9778869
53	T2	1.041856	1.041856	1.041856	1.041856
54	T3	0.9514385	0.9514385	0.9514385	0.9514385
55	T4	1.001245	1.001245	1.001245	1.001245
56	TK	471.8889	471.8889	471.8889	471.8889
57	Totalpressuredrop	0	0	0.0012273	0.0012273
58	TR	849.4	849.4	849.4	849.4
59	W	0	0	8.5855	8.5855
60	X	0	0	0.9120706	0.9120706
61	yCO0	0.2672196	0.2672196	0.2672196	0.2672196

Differential equations

1 d(X)/d(W) = -rCO / FCO0

2 d(p)/d(W) = (-alpha/(2*p))*((1+e*X))

3 d(FCH4)/d(W) = rCH4

4 d(FC2)/d(W) = rC2

5 d(FC3)/d(W) = rC3

6 d(FC4)/d(W) = rC4

7 d(FH2O)/d(W) = rH2O

8 d(FH2)/d(W) = rH2

Explicit equations

1 n = 10

number of tubes

2 FT0 = 122.97/n

3 FCO0 = 32.86/n

4 k2 = 4.512

5 k = 0.0173

6 delta = -2

7 yCO0 = FCO0/FT0

8 P0 = 450

9 e = yCO0*delta

10 PCO0 = P0*yCO0

11 PT = p*P0

12 Totalpressuredrop = P0-PT

13 Allowablepressuredrop = 1.5*Totalpressuredrop

14 D = .1667

ft

15 Ac = 3.14/4 *D^2

ft

```
16 G = FT0/Ac
superficial mass velocity
17 rho0 = 0.6123
aspen lb/ft^3
18 mu = 0.04186
aspen lb/ft*hr
19 phi = 0.4
20 rhob = 49.9424
lb/ft^3 based on the bulk density 0.8 g/cm^3 converted
21 rhoc = (rhob)/(1-phi)
22 gc = 32.2
lb*ft/lbf*s^2
23 Dp = 0.006667
ft based on catalyst diameter of 1/16 in
24 B0 = (G*(1-phi))/(rho0*gc*Dp*phi^3)*(((150*(1-phi)*mu)/Dp)+1.75*G)/144^2/3600^2
25 alpha = (2*B0)/(Ac*rhoc*(1-phi)*P0)
rhoc is density of catalyst
26 PCO = PCO0*((1-X)/(1+e*X))*p
27 PH2 = PCO0*((2-2.06*X)/(1+e*X))*p
28 PCOSI = PCO/14.7
29 PH2SI = PH2/14.7
30 T = 390
31 TK = (T-32)*(5/9)+273
32 T1 = exp(-4492*(1/TK-1/473))
33 T2 = exp(8237*(1/TK-1/473))
34 T3 = exp(-10000*(1/TK-1/473))
```

35	T4 = exp(250*(1/TK-1/473))
36	rCOSI = -(k*T1*PH2SI*PCOSI)/(1+k2+T2*PCOSI)^2
37	rCO = rCOSI*28316.8/(454)
38	TR = TK*1.8
39	FCH40 = 3.286
40	FH2O0 = 0.4585
41	FH20 = 6.5669
42	FCO20 = 0.6576/n
43	FN20 = 0.21/n
44	FCO = FCO0 - FCO0*X
45	rH2O = -rCO
46	rCH4 = -rCO*0.03*T3
47	rN2 = 0
48	rCO2 = 0
49	rH2 = 2.06*rCO
50	rC2 = 0.04*rCH4
51	rC3 = 0.04*rCH4
52	rC4 = 0.04*rCH4

General

Total number of equations	60
Number of differential equations	8
Number of explicit equations	52
Elapsed time	1.157 sec
Solution method	RKF_45
Step size guess. h	0.000001

Truncation error tolerance. eps	0.000001

Medium Unit MATLAB Simulation

FTR Medium Unit Optimization Program

This script optimizes the FTR differential program with respect to a constructed objective function

Include Parameters or Data

Set the initial point `x0` and scale a for the optimization.

```
clear all;
close all;

x0 = [10;50;450;0.1667];   % n; W; T; D
a = 0.001;         % Scaling factor for cost of n
```

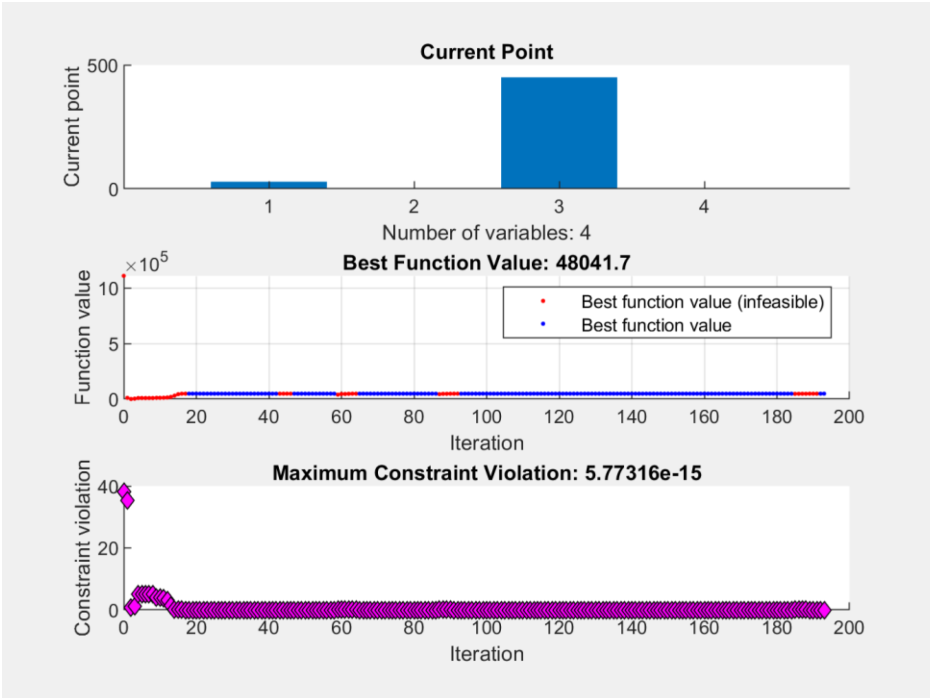
Place the `x0` value and any other problem data into the workspace by running this section before proceeding.

Optimize Live Editor Task

```
% Pass fixed parameters to objfun
objfun = @(x)objectiveFcn(x,a);

% Set nondefault solver options
options = optimoptions('fmincon','PlotFcn',{'optimplotx',...
    'optimplotfvalconstr','optimplotconstrviolation'});

% Solve
[solution,objectiveValue] = fmincon(objfun,x0,[],[],[],[],[],[],...
    @constraintFcn,options);
```



<stopping criteria details>

```
% Clear variables
clearvars objfun options
```

Results

View these variables.

```
sol.n = solution(1);
sol.w = solution(2);
sol.T = solution(3);
sol.D = solution(4);
disp(sol)
```

n: 28.8016

w: 3.6098

T: 450.0000

D: 0.1667

```
objectiveValue
```

objectiveValue = 4.8042e+04

View the nonlinear constraint function values at the solution.

```
[ccons,ceqcons] = constraintFcn(solution)
```

ccons = 1×9

10^3 ×

```
-0.0278   -1.7662   -0.0000   -0.0600   -0.0047   -0.0000   -0.0002 ⋯
```

ceqcons = -5.7732e-15

```
D  = 0.1667;
Ac = 3.14/4*D^2;
V  = solution(2)*454/0.8/28217;
L  = V/Ac
```

L = 3.3281

```
[w,X] = FTR_eq(solution);
Xf = X(end,1)
```

Xf = 0.9000

```
[w,X] = FTR_eq(solution);
Pf = X(end,2)
```

Pf = 1.0000

```
solution(1)*solution(2)
```

ans = 103.9682

Helper Functions — Local Functions

The following code creates the objective function. Modify this code for your problem.

```
function f = objectiveFcn(x,a)
% Run simulation
[w, X] = FTR_eq(x);

% We care about w, n, &, Xf
Xf = X(end,1);
Wf = w(end);


f = a*((20/3*10*x(1)*Wf)^2); %-(10000*Xf)^2);
end
```

The following code creates the constraint function. Modify this code for your problem.

```
function [c,ceq] = constraintFcn(x)
[w,X] = FTR_eq(x);
Wf = w(end);
Xf = X(end,1);
Pf = X (end,2);

D    = 0.1667;
Ac   = 3.14/4*D^2;
V    = Wf*454/0.8/28217;
L    = V/Ac;
PT   = 450*Pf;
Total_P = 450 - PT;
Allowable_P = 1.5*Total_P;


bigPipeDiameter = 2*sqrt(x(1)*Ac/(.65*pi));

c(1) = 1-x(1);              % n > 1
c(2) = x(1)-1795;           % n < 1795
%c(3) = x(2)-100;           % Wf < 100
c(3) = x(3)-450;            % T < 450
c(4) = 390-x(3);            % 390 < T
c(5) = L - 8;               % L < 8ft
c(6) = x(4) - 0.1667;       % D < 0.1667 ft
c(7) = 0 - x(4);            % D > 0
c(8) = (bigPipeDiameter*3) - L;
c(9) = Allowable_P - 0.88;               % pT < 0.88 for PT/P0 and liquid
ceq  = [Xf - .9];                  % Xf = 0.75 (conversion ratio)
end
```

```
function dXdW = FTR_differential(~, y, params)

% Extract Parameters
X       = y(1);
p       = y(2);

n       = params(1);
T       = params(2);
k       = params(3);
k2      = params(4);
delta   = params(5);
P0      = params(6);
```

```
D       = params(7);
rho0    = params(8);
mu      = params(9);
phi     = params(10);
rhob    = params(11);
gc      = params(12);
Dp      = params(13);

% Explicit Equations
FT0 = 679.9355/n;
FCO0 = 196.757/n;

yCO0 = FCO0/FT0;
e = yCO0*delta;
PCO0 = P0*yCO0;
Ac = 3.14/4*D^2;
G = FT0/Ac;

rhoc = rhob/(1-phi);

B0 = (G*(1-phi))/(rho0*gc*Dp*phi^3)*(((150*(1-
phi)*mu)/Dp)+1.75*G)/144^2/3600^2;
alpha = (2*B0)/(Ac*rhoc*(1-phi)*P0);
PCO = PCO0*((1-X)/(1+e*X))*p;
PH2 = PCO0*((2-2.06*X)/(1+e*X))*p;
PCOSI = PCO/14.7;
PH2SI = PH2/14.7;
TK = (T-32)*(5/9)+273;
T1 = exp(-4492*(1/TK-1/473));
T2 = exp(8237*(1/TK-1/473));
T3 = exp(-10000*(1/TK-1/473));

rCOSI = -(k*T1*PH2SI*PCOSI)/(1+k2+T2*PCOSI)^2;
rCO = rCOSI*28316.8/454;

rH2O = -rCO;
rCH4 = -rCO*0.03*T3;

rH2 = 2.06*rCO;
rC2 = 0.04*rCH4;
rC3 = 0.04*rCH4;
rC4 = 0.04*rCH4;


% ODEs
dXdW = [-rCO / FCO0;                % dX/dW
```

```
        (-alpha/(2*p))*(1+e*X);     % dp/dW
        rCH4;                       % dFCH4/dW
        rC2;                        % dFC2/dW
        rC3;                        % dFC3/dW
        rC4;                        % dFC4/dW
        rH2O;                       % dFH2O/dW
        rH2;                        % dFH2/dW
        ];
end
```

```
function [w,X] = FTR_eq(x)
% Setup parameters
n       = x(1);
T       = x(3);
k       = 0.0173;
k2      = 4.512;
delta   = -2;
P0      = 450;
D       = x(4);
rho0    = 0.5919;
mu      = 0.04196;
phi     = 0.4;
rhob    = 49.9424;
gc      = 32.2;
Dp      = 0.006667;


% Initial Conditions
X0      = 0;
p0      = 1;
FCH40   = 60.6622/n;
FC20    = 0/n;
FC30    = 0/n;
FC40    = 0/n;
FH2O0   = 24.8352/n;
FH20    = 393.5494/n;


IC = [X0 p0 FCH40 FC20 FC30 FC40 FH2O0 FH20];
params = [n T k k2 delta P0 D rho0 mu phi rhob gc Dp];

% Span of catalyst (low high)
Wspan = [0 x(2)];
```

```
% Run simulation
[w, X] = ode45(@(t,y) FTR_differential(t,y,params), Wspan, IC);
end
```


Medium Unit Polymath Report

Calculated values of DEQ variables

	Variable	Initial value	Minimal value	Maximal value	Final value
1	Ac	0.0218143	0.0218143	0.0218143	0.0218143
2	Allowablepressuredrop	0	0	0.0022624	0.0022624
3	alpha	2.946E-06	2.946E-06	2.946E-06	2.946E-06
4	B0	0.0007222	0.0007222	0.0007222	0.0007222
5	D	0.1667	0.1667	0.1667	0.1667
6	delta	-2.	-2.	-2.	-2.
7	Dp	0.006667	0.006667	0.006667	0.006667
8	e	-0.578752	-0.578752	-0.578752	-0.578752
9	FC2	0	0	0.0282493	0.0282493
10	FC3	0	0	0.0282493	0.0282493
11	FC4	0	0	0.0282493	0.0282493
12	FC5up	0	0	3.883317	3.883317
13	FCH4	2.0918	2.0918	2.798032	2.798032
14	FCH40	2.0918	2.0918	2.0918	2.0918
15	FCO	6.784724	0.6720297	6.784724	0.6720297
16	FCO0	6.784724	6.784724	6.784724	6.784724
17	FCO20	0.099369	0.099369	0.099369	0.099369
18	FH2	13.5707	0.9785495	13.5707	0.9785495

19	FH20	13.5707	13.5707	13.5707	13.5707
20	FH2O	0.8564	0.8564	6.969094	6.969094
21	FH2O0	0.8564	0.8564	0.8564	0.8564
22	FN20	0.0431	0.0431	0.0431	0.0431
23	Fproducts	0	0	10.78699	10.78699
24	FT	23.44609	15.52824	23.44609	15.52824
25	FT0	23.44605	23.44605	23.44605	23.44605
26	G	1074.803	1074.803	1074.803	1074.803
27	gc	32.2	32.2	32.2	32.2
28	k	0.0173	0.0173	0.0173	0.0173
29	k2	4.512	4.512	4.512	4.512
30	mu	0.04196	0.04196	0.04196	0.04196
31	n	29.	29.	29.	29.
32	p	1.	0.9999966	1.	0.9999966
33	P0	450.	450.	450.	450.
34	PCO	130.2192	26.95138	130.2192	26.95138
35	PCO0	130.2192	130.2192	130.2192	130.2192
36	PCOSI	8.858448	1.833427	8.858448	1.833427
37	PH2	260.4384	39.19399	260.4384	39.19399
38	PH2SI	17.7169	2.666258	17.7169	2.666258
39	phi	0.4	0.4	0.4	0.4
40	PT	450.	449.9985	450.	449.9985
41	rC2	0.0201838	0.0011943	0.0201838	0.0011943
42	rC3	0.0201838	0.0011943	0.0201838	0.0011943
43	rC4	0.0201838	0.0011943	0.0201838	0.0011943

44	rCH4	0.5045961	0.0298575	0.5045961	0.0298575
45	rCO	-4.367463	-4.367463	-0.2584276	-0.2584276
46	rCO2	0	0	0	0
47	rCOSI	-0.070023	-0.070023	-0.0041433	-0.0041433
48	rH2	-8.996973	-8.996973	-0.5323608	-0.5323608
49	rH2O	4.367463	0.2584276	4.367463	0.2584276
50	rho0	0.5919	0.5919	0.5919	0.5919
51	rhob	49.9424	49.9424	49.9424	49.9424
52	rhoc	83.23733	83.23733	83.23733	83.23733
53	rN2	0	0	0	0
54	T	450.	450.	450.	450.
55	T1	1.832519	1.832519	1.832519	1.832519
56	T2	0.3293417	0.3293417	0.3293417	0.3293417
57	T3	3.851177	3.851177	3.851177	3.851177
58	T4	0.9668524	0.9668524	0.9668524	0.9668524
59	TK	505.2222	505.2222	505.2222	505.2222
60	Totalpressuredrop	0	0	0.0015083	0.0015083
61	TR	909.4	909.4	909.4	909.4
62	W	0	0	3.6098	3.6098
63	X	0	0	0.9009496	0.9009496
64	yC5up	0	0	0.2500809	0.2500809
65	yCO0	0.289376	0.289376	0.289376	0.289376

Differential equations

1 d(X)/d(W) = -rCO / FCO0

2 d(p)/d(W) = (-alpha/(2*p))*((1+e*X))

3 d(FCH4)/d(W) = rCH4

4 d(FC2)/d(W) = rC2

5 d(FC3)/d(W) = rC3

6 d(FC4)/d(W) = rC4

7 d(FH2O)/d(W) = rH2O

8 d(FH2)/d(W) = rH2

Explicit equations

1 n = 29

number of tubes

2 FT0 = 679.9355/n

3 FCO0 = 196.757/n

4 k2 = 4.512

5 k = 0.0173

6 delta = -2

7 yCO0 = FCO0/FT0

8 P0 = 450

9 e = yCO0*delta

10 PCO0 = P0*yCO0

11 PT = p*P0

12 Totalpressuredrop = P0-PT

13 Allowablepressuredrop = 1.5*Totalpressuredrop

14 D = .1667

ft

15 Ac = 3.14/4 *D^2

```
ft
16 G = FT0/Ac
superficial mass velocity
17 rho0 = 0.5919
aspen lb/ft^3
18 mu = 0.04196
aspen lb/ft*hr
19 phi = 0.4
20 rhob = 49.9424
lb/ft^3 based on the bulk density 0.8 g/cm^3 converted
21 rhoc = (rhob)/(1-phi)
22 gc = 32.2
lb*ft/lbf*s^2
23 Dp = 0.006667
ft based on catalyst diameter of 1/16 in
24 B0 = (G*(1-phi))/(rho0*gc*Dp*phi^3)*(((150*(1-phi)*mu)/Dp)+1.75*G)/144^2/3600^2
25 alpha = (2*B0)/(Ac*rhoc*(1-phi)*P0)
rhoc is density of catalyst
26 PCO = PCO0*((1-X)/(1+e*X))*p
27 PH2 = PCO0*((2-2.06*X)/(1+e*X))*p
28 PCOSI = PCO/14.7
29 PH2SI = PH2/14.7
30 T = 450
31 TK = (T-32)*(5/9)+273
32 T1 = exp(-4492*(1/TK-1/473))
33 T2 = exp(8237*(1/TK-1/473))
```

```
T3 = exp(-10000*(1/TK-1/473))
T4 = exp(250*(1/TK-1/473))
rCOSI = -(k*T1*PH2SI*PCOSI)/(1+k2+T2*PCOSI)^2
rCO = rCOSI*28316.8/(454)
TR = TK*1.8
FCH40 = 2.0918
FH2O0 = 0.8564
FH20 = 13.5707
FCO20 = 2.8817/n
FN20 = 1.2499/n
FCO = FCO0 - FCO0*X
rH2O = -rCO
rCH4 = -rCO*0.03*T3
rN2 = 0
rCO2 = 0
rH2 = 2.06*rCO
rC2 = 0.04*rCH4
rC3 = 0.04*rCH4
rC4 = 0.04*rCH4
Fproducts = ((FH2O-FH2O0)+(FCH4-FCH40)+FC2+FC3+FC4)/.64
FC5up = Fproducts*.360
FT = FH2+FCO+FH2O+FCH4+FN20+FC2+FC3+FC4+FC5up+FCO20
yC5up = FC5up/FT
```

Large Unit MATLAB Simulation

FTR Large Unit Optimization Program

This script optimizes the FTR differential program with respect to a constructed objective function

Include Parameters or Data

Set the initial point `x0` and scale a for the optimization.

```
clear all;
close all;

x0 = [15;50;450;0.1667];    % n; W; T; D
a = 0.001;          % Scaling factor for cost of n
```

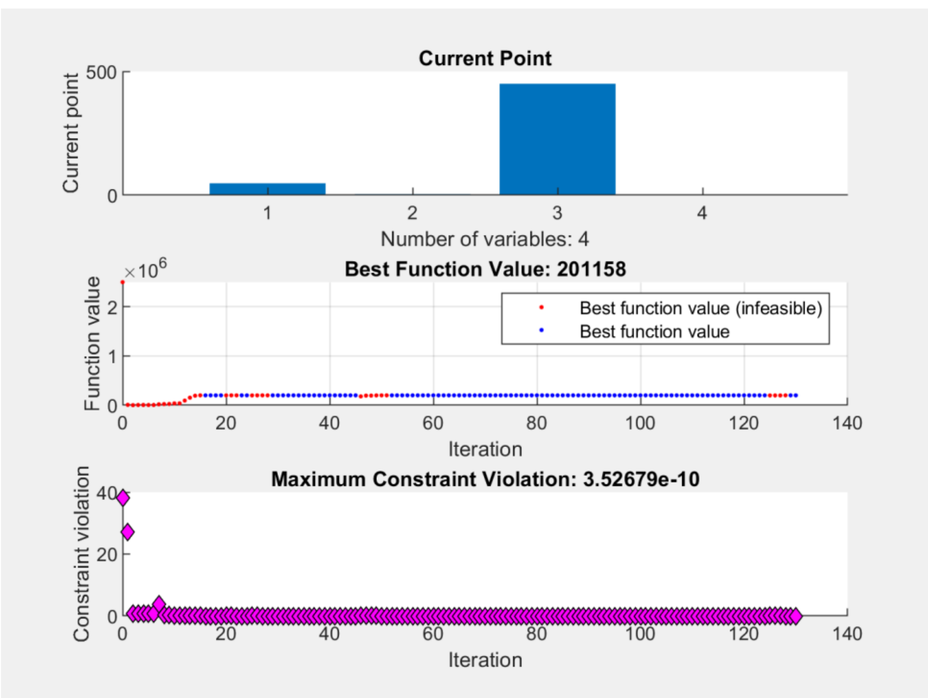
Place the `x0` value and any other problem data into the workspace by running this section before proceeding.

Optimize Live Editor Task

```
% Pass fixed parameters to objfun
objfun = @(x)objectiveFcn(x,a);

% Set nondefault solver options
options = optimoptions('fmincon','PlotFcn',{'optimplotx',...
    'optimplotfvalconstr','optimplotconstrviolation'});

% Solve
[solution,objectiveValue] = fmincon(objfun,x0,[],[],[],[],[],[],...
    @constraintFcn,options);
```



Local minimum found that satisfies the constraints.

Optimization completed because the objective function is non-decreasing in feasible directions, to within the value of the optimality tolerance, and constraints are satisfied to within the value of the constraint tolerance.

<stopping criteria details>

```
% Clear variables
clearvars objfun options
```

Results

View these variables.

```
sol.n = solution(1);
sol.w = solution(2);
sol.T = solution(3);
sol.D = solution(4);
disp(sol)
```

n: 48.0313

w: 4.4293

T: 450.0000

D: 0.1667

```
objectiveValue
```

objectiveValue = 2.0116e+05

View the nonlinear constraint function values at the solution.

```
[ccons,ceqcons] = constraintFcn(solution)
```

ccons = 1×9

10^3 ×

-0.0470 -1.7470 -0.0000 -0.0600 -0.0039 -0.0000 -0.0002 ⋯

ceqcons = -3.5268e-10

```
D  = 0.1667;
Ac = 3.14/4*D^2;
V  = solution(2)*454/0.8/28217;
L  = V/Ac
```

L = 4.0837

```
[w,X] = FTR_eq(solution);
Xf = X(end,1)
```

Xf = 0.9000

```
[w,X] = FTR_eq(solution);
Pf = X(end,2)
```

Pf = 1.0000

```
solution(1)*solution(2)
```

ans = 212.7452

Helper Functions — Local Functions

The following code creates the objective function. Modify this code for your problem.

```
function f = objectiveFcn(x,a)
% Run simulation
[w, X] = FTR_eq(x);

% We care about w, n, &, Xf
Xf = X(end,1);
Wf = w(end);


f = a*((20/3*10*x(1)*Wf)^2); %-(10000*Xf)^2);
end
```

The following code creates the constraint function. Modify this code for your problem.

```
function [c,ceq] = constraintFcn(x)
[w,X] = FTR_eq(x);
Wf = w(end);
Xf = X(end,1);
Pf = X (end,2);
```

```
D    = 0.1667;
Ac   = 3.14/4*D^2;
V    = Wf*454/0.8/28217;
L    = V/Ac;
PT   = 450*Pf;
Total_P = 450 - PT;
Allowable_P = 1.5*Total_P;


bigPipeDiameter = 2*sqrt(x(1)*Ac/(.72*pi));

c(1) = 1-x(1);                 % n > 1
c(2) = x(1)-1795;              % n < 1795
%c(3) = x(2)-100;              % Wf < 100
c(3) = x(3)-450;               % T < 450
c(4) = 390-x(3);               % 390 < T
c(5) = L - 8;                  % L < 8ft
c(6) = x(4) - 0.1667;          % D < 0.1667 ft
c(7) = 0 - x(4);               % D > 0
c(8) = (bigPipeDiameter*3) - L;
c(9) = Allowable_P - 0.88;                  % pT < 0.88 for PT/P0 and liquid
ceq  = [Xf - .9];                     % Xf = 0.75 (conversion ratio)
end
```

```
function dXdW = FTR_differential(~, y, params)

% Extract Parameters
X        = y(1);
p        = y(2);

n        = params(1);
T        = params(2);
k        = params(3);
k2       = params(4);
delta    = params(5);
P0       = params(6);
D        = params(7);
rho0     = params(8);
mu       = params(9);
phi      = params(10);
rhob     = params(11);
gc       = params(12);
Dp       = params(13);
```

```
 % Explicit Equations
 FT0 = 1259.56/n;
 FCO0 = 343.3128/n;

 yCO0 = FCO0/FT0;
 e = yCO0*delta;
 PCO0 = P0*yCO0;
 Ac = 3.14/4*D^2;
 G = FT0/Ac;

 rhoc = rhob/(1-phi);

 B0 = (G*(1-phi))/(rho0*gc*Dp*phi^3)*(((150*(1-
phi)*mu)/Dp)+1.75*G)/144^2/3600^2;
 alpha = (2*B0)/(Ac*rhoc*(1-phi)*P0);
 PCO = PCO0*((1-X)/(1+e*X))*p;
 PH2 = PCO0*((2-2.06*X)/(1+e*X))*p;
 PCOSI = PCO/14.7;
 PH2SI = PH2/14.7;
 TK = (T-32)*(5/9)+273;
 T1 = exp(-4492*(1/TK-1/473));
 T2 = exp(8237*(1/TK-1/473));
 T3 = exp(-10000*(1/TK-1/473));

 rCOSI = -(k*T1*PH2SI*PCOSI)/(1+k2+T2*PCOSI)^2;
 rCO = rCOSI*28316.8/454;

 rH2O = -rCO;
 rCH4 = -rCO*0.03*T3;

 rH2 = 2.06*rCO;
 rC2 = 0.04*rCH4;
 rC3 = 0.04*rCH4;
 rC4 = 0.04*rCH4;


 % ODEs
 dXdW = [-rCO / FCO0;                  % dX/dW
         (-alpha/(2*p))*(1+e*X);       % dp/dW
         rCH4;                         % dFCH4/dW
         rC2;                          % dFC2/dW
         rC3;                          % dFC3/dW
         rC4;                          % dFC4/dW
         rH2O;                         % dFH2O/dW
         rH2;                          % dFH2/dW
         ];
```

```
end
```

```
function [w,X] = FTR_eq(x)
% Setup parameters
n       = x(1);
T       = x(3);
k       = 0.0173;
k2      = 4.512;
delta   = -2;
P0      = 450;
D       = x(4);
rho0    = 0.6017;
mu      = 0.04186;
phi     = 0.4;
rhob    = 49.9424;
gc      = 32.2;
Dp      = 0.006667;


% Initial Conditions
X0      = 0;
p0      = 1;
FCH40   = 174.2847/n;
FC20    = 0/n;
FC30    = 0/n;
FC40    = 0/n;
FH2O0   = 46.7159/n;
FH20    = 686.6826/n;


IC = [X0 p0 FCH40 FC20 FC30 FC40 FH2O0 FH20];
params = [n T k k2 delta P0 D rho0 mu phi rhob gc Dp];

% Span of catalyst (low high)
Wspan = [0 x(2)];


% Run simulation
[w, X] = ode45(@(t,y) FTR_differential(t,y,params), Wspan, IC);
end
```

Large Unit Polymath Report

Calculated values of DEQ variables

	Variable	Initial value	Minimal value	Maximal value	Final value
1	Ac	0.0218143	0.0218143	0.0218143	0.0218143
2	Allowablepressuredrop	0	0	0.003285	0.003285
3	alpha	3.411E-06	3.411E-06	3.411E-06	3.411E-06
4	B0	0.0008361	0.0008361	0.0008361	0.0008361
5	D	0.1667	0.1667	0.1667	0.1667
6	delta	-2.	-2.	-2.	-2.
7	Dp	0.006667	0.006667	0.006667	0.006667
8	e	-0.5451313	-0.5451313	-0.5451313	-0.5451313
9	FC2	0	0	0.0292248	0.0292248
10	FC3	0	0	0.0292248	0.0292248
11	FC4	0	0	0.0292248	0.0292248
12	FC5up	0	0	4.017417	4.017417
13	FCH4	3.5568	3.5568	4.28742	4.28742
14	FCH40	3.5568	3.5568	3.5568	3.5568
15	FCO	7.006384	0.6826033	7.006384	0.6826033
16	FCO0	7.006384	7.006384	7.006384	7.006384
17	FCO20	0.1300531	0.1300531	0.1300531	0.1300531
18	FH2	14.0139	0.9869124	14.0139	0.9869124
19	FH20	14.0139	14.0139	14.0139	14.0139
20	FH2O	0.9534	0.9534	7.27718	7.27718
21	FH2O0	0.9534	0.9534	0.9534	0.9534

22	FN20	0.0446918	0.0446918	0.0446918	0.0446918
23	Fproducts	0	0	11.15949	11.15949
24	FT	25.70523	17.51395	25.70523	17.51395
25	FT0	25.70531	25.70531	25.70531	25.70531
26	G	1178.371	1178.371	1178.371	1178.371
27	gc	32.2	32.2	32.2	32.2
28	k	0.0173	0.0173	0.0173	0.0173
29	k2	4.512	4.512	4.512	4.512
30	mu	0.04186	0.04186	0.04186	0.04186
31	n	49.	49.	49.	49.
32	p	1.	0.9999951	1.	0.9999951
33	P0	450.	450.	450.	450.
34	PCO	122.6545	23.52397	122.6545	23.52397
35	PCO0	122.6545	122.6545	122.6545	122.6545
36	PCOSI	8.343847	1.60027	8.343847	1.60027
37	PH2	245.3091	33.97208	245.3091	33.97208
38	PH2SI	16.68769	2.311026	16.68769	2.311026
39	phi	0.4	0.4	0.4	0.4
40	PT	450.	449.9978	450.	449.9978
41	rC2	0.0186493	0.0009267	0.0186493	0.0009267
42	rC3	0.0186493	0.0009267	0.0186493	0.0009267
43	rC4	0.0186493	0.0009267	0.0186493	0.0009267
44	rCH4	0.4662327	0.0231665	0.4662327	0.0231665
45	rCO	-4.035413	-4.035413	-0.2005145	-0.2005145
46	rCO2	0	0	0	0

47	rCOSI	-0.0646993	-0.0646993	-0.0032148	-0.0032148
48	rH2	-8.312951	-8.312951	-0.4130599	-0.4130599
49	rH2O	4.035413	0.2005145	4.035413	0.2005145
50	rho0	0.6017	0.6017	0.6017	0.6017
51	rhob	49.9424	49.9424	49.9424	49.9424
52	rhoc	83.23733	83.23733	83.23733	83.23733
53	rN2	0	0	0	0
54	T	450.	450.	450.	450.
55	T1	1.832519	1.832519	1.832519	1.832519
56	T2	0.3293417	0.3293417	0.3293417	0.3293417
57	T3	3.851177	3.851177	3.851177	3.851177
58	T4	0.9668524	0.9668524	0.9668524	0.9668524
59	TK	505.2222	505.2222	505.2222	505.2222
60	Totalpressuredrop	0	0	0.00219	0.00219
61	TR	909.4	909.4	909.4	909.4
62	W	0	0	4.4293	4.4293
63	X	0	0	0.9025741	0.9025741
64	yC5up	0	0	0.2293838	0.2293838
65	yCO0	0.2725657	0.2725657	0.2725657	0.2725657

Differential equations

1 d(X)/d(W) = -rCO / FCO0

2 d(p)/d(W) = (-alpha/(2*p))*((1+e*X))

3 d(FCH4)/d(W) = rCH4

4 d(FC2)/d(W) = rC2

5 d(FC3)/d(W) = rC3

6 d(FC4)/d(W) = rC4

7 d(FH2O)/d(W) = rH2O

8 d(FH2)/d(W) = rH2

Explicit equations

1 n = 49

number of tubes

2 FT0 = 1259.56/n

3 FCO0 = 343.3128/n

4 k2 = 4.512

5 k = 0.0173

6 delta = -2

7 yCO0 = FCO0/FT0

8 P0 = 450

9 e = yCO0*delta

10 PCO0 = P0*yCO0

11 PT = p*P0

12 Totalpressuredrop = P0-PT

13 Allowablepressuredrop = 1.5*Totalpressuredrop

14 D = .1667

ft

15 Ac = 3.14/4 *D^2

ft

16 G = FT0/Ac

superficial mass velocity

17 rho0 = 0.6017

aspen lb/ft^3

18 mu = 0.04186

aspen lb/ft*hr

19 phi = 0.4

20 rhob = 49.9424

lb/ft^3 based on the bulk density 0.8 g/cm^3 converted

21 rhoc = (rhob)/(1-phi)

22 gc = 32.2

lb*ft/lbf*s^2

23 Dp = 0.006667

ft based on catalyst diameter of 1/16 in

24 B0 = (G*(1-phi))/(rho0*gc*Dp*phi^3)*(((150*(1-phi)*mu)/Dp)+1.75*G)/144^2/3600^2

25 alpha = (2*B0)/(Ac*rhoc*(1-phi)*P0)

rhoc is density of catalyst

26 PCO = PCO0*((1-X)/(1+e*X))*p

27 PH2 = PCO0*((2-2.06*X)/(1+e*X))*p

28 PCOSI = PCO/14.7

29 PH2SI = PH2/14.7

30 T = 450

31 TK = (T-32)*(5/9)+273

32 T1 = exp(-4492*(1/TK-1/473))

33 T2 = exp(8237*(1/TK-1/473))

34 T3 = exp(-10000*(1/TK-1/473))

35 T4 = exp(250*(1/TK-1/473))

36 rCOSI = -(k*T1*PH2SI*PCOSI)/(1+k2+T2*PCOSI)^2

```
rCO = rCOSI*28316.8/(454)
TR = TK*1.8
FCH40 = 3.5568
FH2O0 = 0.9534
FH20 = 14.0139
FCO20 = 6.3726/n
FN20 = 2.1899/n
FCO = FCO0 - FCO0*X
rH2O = -rCO
rCH4 = -rCO*0.03*T3
rN2 = 0
rCO2 = 0
rH2 = 2.06*rCO
rC2 = 0.04*rCH4
rC3 = 0.04*rCH4
rC4 = 0.04*rCH4
Fproducts = ((FH2O-FH2O0)+(FCH4-FCH40)+FC2+FC3+FC4)/.64
FC5up = Fproducts*.360
FT = FH2+FCO+FH2O+FCH4+FN20+FC2+FC3+FC4+FC5up+FCO20
yC5up = FC5up/FT
```

Appendix

Medium Unit Stream Table

	Stream Number		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
	Description		Oxygen	Natural Gas	HP Steam	Carbon Dioxide	H-101 Feed	H-101 Heated Stream	R-101 Feed	E-101 Feed	V-101 Feed	CO2 Separation Unit Feed	V-101 Waste Water	Removed CO2	E-102 Feed	R-102 Feed	E-103 Feed	V-102 Feed
	Vapor Fraction		1.00	1.00	1.00	1.00	0.74	1.00	1.00	1.00	0.84	1.00	0.00	1.00	1.00	1.00	1.00	0.43
	Temperature	°F	75.00	100.00	490.00	100.00	235.60	1000.00	870.40	1699.00	212.00	212.00	212.00	192.10	216.78	450.00	450.00	275.00
	Pressure	psia	514.70	514.70	614.70	514.70	504.70	500.70	500.70	500.70	496.70	496.70	496.70	486.70	486.70	472.70	445.00	431.00
	Total Molar Flow	lbmole/hr	125.00	270.10	135.10	45.00	450.20	450.20	575.20	870.30	870.30	734.69	135.65	54.75	679.94	679.94	343.53	343.53
	Total Mass Flow	lb/hr	3995.00	4334.00	2433.00	1980.00	8747.00	8747.00	12740.00	12740.00	12740.00	10296.83	2445.03	2409.65	7887.18	7887.18	8243.74	8243.74
	Actual Vol. Flow	barrel/day	12623.51	991.10	211.09	1824.81	20012.75	59183.96	68872.12	169796.10	44938.49	44762.12	176.37	2896.72	41865.88	56955.63	30021.14	11355.06
	Mass Enthalpy	BTU/lb	-4.85	-2011.00	-6371.00	-3858.00	-3642.00	-2962.00	-2035.00	-2035.00	-3085.00	-2230.07	-6685.68	-3833.20	-1740.29	-1591.53	-3130.07	-3600.60
	Oxygen	lb/hr	3995.00	0.00	0.00	0.00	0.00	3960.00	3960.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Water	lb/hr	0.00	0.00	2433.00	0.00	2432.94	2432.94	2432.94	2889.61	2889.61	447.41	2442.20	0.00	447.41	447.41	3641.91	3641.91
	Nitrogen	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	35.02	35.02	35.02	35.01	0.00	0.00	35.01	35.01	17.51	17.51
	Carbon Monoxide	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	5511.55	5511.55	5511.34	0.21	0.00	5511.34	5511.34	545.88	545.88
	Carbon Dioxide	lb/hr	0.00	0.00	0.00	45.00	1980.44	1980.44	1980.44	2539.06	2539.06	2536.47	2.59	2409.65	126.82	126.82	126.82	126.82
	Hydrogen	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	793.43	793.43	793.40	0.03	0.00	793.40	793.40	28.66	28.66
	Ethane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	24.63	24.63
	Propane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	36.13	36.13
	Methane	lb/hr	0.00	4334.00	0.00	0.00	4333.51	4333.51	4333.51	973.20	973.20	973.20	0.00	0.00	973.20	973.20	1301.53	1301.53
	n-Butane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	47.61	47.61
	n-Pentane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	332.00	332.00
	n-Hexane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	308.63	308.63
Naphtha	n-Heptane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	278.93	278.93
	n-Octane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	246.94	246.94
	n-Nonane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	215.21	215.21
	n-Decane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	185.24	185.24
	n-C11	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	157.85	157.85
	n-C12	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	133.39	133.39
	n-C13	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	111.95	111.95
	n-C14	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	93.39	93.39
Diesel	n-C15	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	77.51	77.51
	n-C16	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	64.05	64.05
	n-C17	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	52.72	52.72
	n-C18	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	43.24	43.24
	n-C19	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	35.36	35.36
	n-C20	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	28.83	28.83
	n-C21	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	23.45	23.45
	n-C22	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	19.03	19.03
	n-C23	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	15.41	15.41
	n-C24	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	12.46	12.46
	n-C25	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	10.05	10.05
	n-C26	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	8.10	8.10
	n-C27	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.52	6.52
	n-C28	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	5.23	5.23
	n-C29	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.20	4.20
	n-C30	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.37	3.37

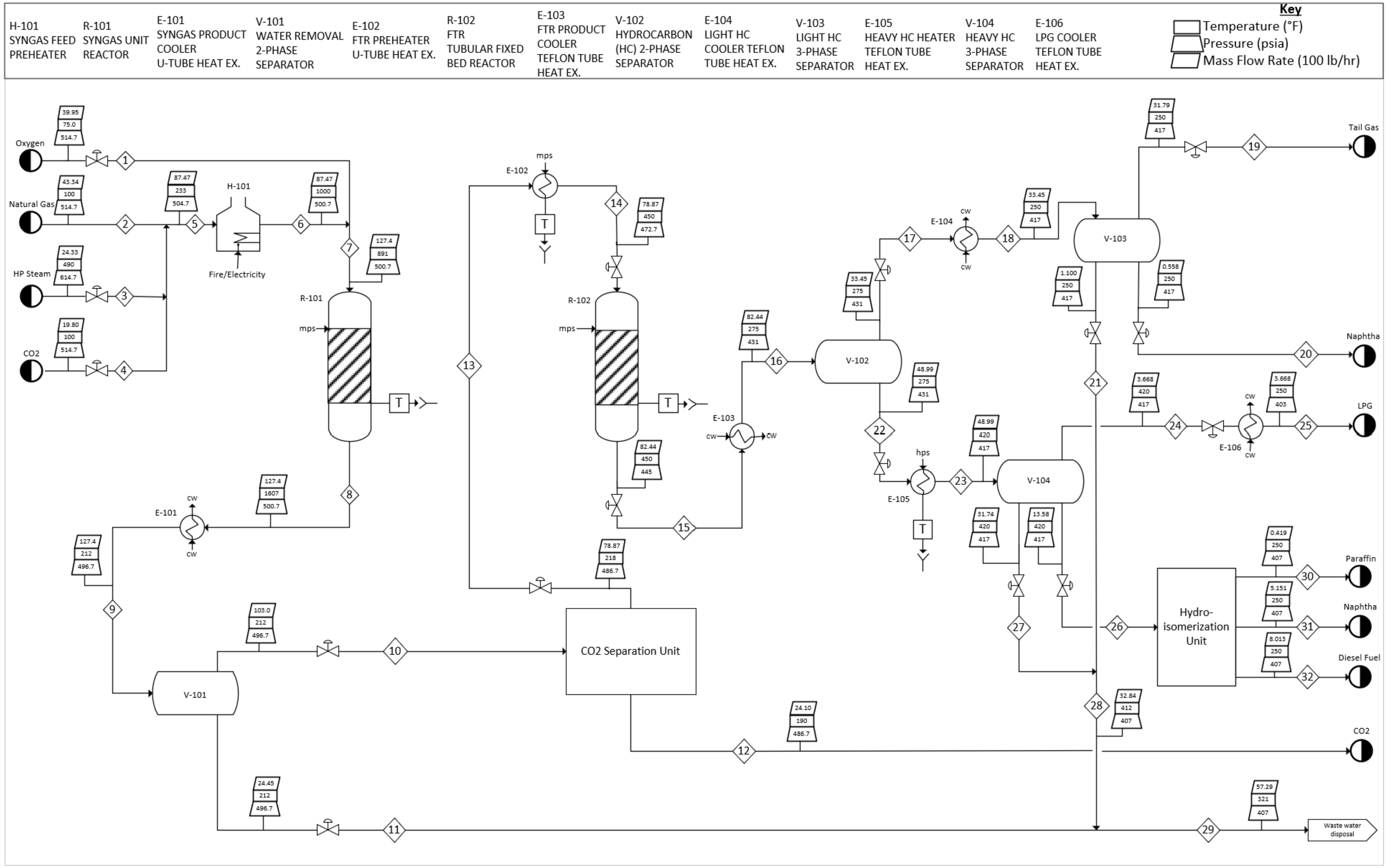
	Stream Number		17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32
	Description		E-104 Feed	V-103 Feed	Tail Gas	Top Naphtha	Top Water Product	E-105 Feed	V-104 Feed	LPG	Cooled LPG	HIU Feed	Bottom Water Product	Top & Bottom Water Products	Waste Water Disposal	Paraffin	Bottom Naphtha	Diesel Fuel
	Vapor Fraction		1.00	0.96	1.00	0.00	0.00	0.00	0.05	1.00	0.08	0.00	0.00	0.00	0.00	0.19	0.00	0.00
	Temperature	°F	275.00	250.00	250.00	250.00	250.00	275.00	420.00	420.00	250.00	420.00	420.00	414.69	331.60	250.00	250.00	250.00
	Pressure	psia	431.00	417.00	417.00	417.00	417.00	431.00	417.00	417.00	433.00	417.00	417.00	407.00	407.00	407.00	407.00	407.00
	Total Molar Flow	lbmole/hr	146.59	146.59	140.00	0.49	6.11	196.94	196.94	10.71	10.71	10.07	176.16	182.26	317.91	1.98	0.800722	0.365511
	Total Mass Flow	lb/hr	3344.53	3344.53	3178.66	55.84	110.04	4899.21	4899.21	366.84	366.84	1358.24	3174.13	3284.16	5729.20	41.88	515.07	801.29
	Actual Vol. Flow	barrel/day	10928.58	10195.30	10180.96	6.26	8.08	426.48	1266.42	832.62	94.39	174.86	258.95	274.68	444.50	2.91	35.85	55.77
	Mass Enthalpy	BTU/lb	-1883.54	-1928.60	-1784.31	-856.37	-6641.05	-4772.79	-4610.47	-2653.33	-3141.24	-796.31	-6468.77	-6463.19	-6558.14	-4998.58	-859.82	-767.18
	Oxygen	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Water	lb/hr	297.42	297.42	187.32	0.15	109.95	3344.49	3344.49	142.29	142.29	29.37	3172.83	3282.78	5724.98	29.37	0.00	0.00
	Nitrogen	lb/hr	17.42	17.42	17.42	0.00	0.00	0.09	0.09	0.08	0.08	0.01	0.00	0.00	0.00	0.01	0.00	0.00
	Carbon Monoxide	lb/hr	543.42	543.42	543.32	0.09	0.01	2.46	2.46	2.27	2.27	0.15	0.03	0.04	0.25	0.15	0.00	0.00
	Carbon Dioxide	lb/hr	124.71	124.71	124.62	0.08	0.02	2.11	2.11	1.75	1.75	0.26	0.09	0.10	2.69	0.26	0.00	0.00
	Hydrogen	lb/hr	28.57	28.57	28.57	0.00	0.00	0.09	0.09	0.08	0.08	0.00	0.00	0.00	0.04	0.00	0.00	0.00
	Ethane	lb/hr	24.12	24.12	24.09	0.02	0.00	0.51	0.51	0.42	10.34	0.09	0.01	0.01	0.01	0.62	0.00	0.00
	Propane	lb/hr	34.63	34.63	34.56	0.08	0.00	1.50	1.50	1.10	0.42	0.38	0.02	0.02	0.02	4.14	0.00	0.00
	Methane	lb/hr	1289.71	1289.71	1289.21	0.45	0.06	11.82	11.82	10.34	1.10	1.13	0.35	0.40	0.40	2.21	0.00	0.00
	n-Butane	lb/hr	43.79	43.79	43.59	0.20	0.00	3.82	3.82	2.45	2.45	1.33	0.03	0.03	0.03	5.10	0.00	0.00
	n-Pentane	lb/hr	283.79	283.79	281.11	2.67	0.00	48.21	48.21	26.03	26.03	21.99	0.20	0.20	0.20	0.00		0.00
	n-Hexane	lb/hr	231.25	231.25	226.83	4.42	0.00	77.38	77.38	32.98	32.98	44.21	0.19	0.19	0.19	0.00		0.00
Naphtha	n-Heptane	lb/hr	171.41	171.41	165.14	6.28	0.00	107.52	107.52	35.64	35.64	71.73	0.14	0.14	0.14	0.00	515.07	0.00
	n-Octane	lb/hr	113.29	113.29	105.44	7.85	0.00	133.66	133.66	33.30	33.30	100.26	0.10	0.10	0.10	0.00		0.00
	n-Nonane	lb/hr	67.30	67.30	58.86	8.44	0.00	147.91	147.91	26.70	26.70	121.15	0.07	0.07	0.07	0.00		0.00
	n-Decane	lb/hr	36.96	36.96	29.06	7.90	0.00	148.27	148.27	19.40	19.40	128.83	0.04	0.04	0.04	0.00		0.00
	n-C11	lb/hr	18.79	18.79	12.40	6.39	0.00	139.05	139.05	12.53	12.53	126.50	0.02	0.02	0.02	0.00	0.00	
	n-C12	lb/hr	9.24	9.24	4.70	4.54	0.00	124.15	124.15	7.82	7.82	116.32	0.01	0.01	0.01	0.00	0.00	
	n-C13	lb/hr	4.61	4.61	1.71	2.90	0.00	107.34	107.34	4.96	4.96	102.38	0.00	0.00	0.00	0.00	0.00	
	n-C14	lb/hr	2.12	2.12	0.49	1.62	0.00	91.27	91.27	2.78	2.78	88.50	0.00	0.00	0.00	0.00	0.00	
Diesel	n-C15	lb/hr	1.01	1.01	0.15	0.87	0.00	76.50	76.50	1.63	1.63	74.87	0.00	0.00	0.00	0.00	0.00	801.29
	n-C16	lb/hr	0.47	0.47	0.04	0.43	0.00	63.58	63.58	0.92	0.92	62.66	0.00	0.00	0.00	0.00	0.00	
	n-C17	lb/hr	0.27	0.27	0.02	0.25	0.00	52.45	52.45	0.61	0.61	51.84	0.00	0.00	0.00	0.00	0.00	
	n-C18	lb/hr	0.11	0.11	0.00	0.11	0.00	43.13	43.13	0.32	0.32	42.82	0.00	0.00	0.00	0.00	0.00	
	n-C19	lb/hr	0.05	0.05	0.00	0.05	0.00	35.31	35.31	0.18	0.18	35.12	0.00	0.00	0.00	0.00	0.00	
	n-C20	lb/hr	0.03	0.03	0.00	0.02	0.00	28.81	28.81	0.10	0.10	28.70	0.00	0.00	0.00	0.00	0.00	
	n-C21	lb/hr	0.01	0.01	0.00	0.01	0.00	23.44	23.44	0.07	0.07	23.37	0.00	0.00	0.00	0.00	0.00	0.00
	n-C22	lb/hr	0.01	0.01	0.00	0.01	0.00	19.03	19.03	0.04	0.04	18.98	0.00	0.00	0.00	0.00	0.00	0.00
	n-C23	lb/hr	0.00	0.00	0.00	0.00	0.00	15.41	15.41	0.02	0.02	15.39	0.00	0.00	0.00	0.00	0.00	0.00
	n-C24	lb/hr	0.00	0.00	0.00	0.00	0.00	12.46	12.46	0.01	0.01	12.44	0.00	0.00	0.00	0.00	0.00	0.00
	n-C25	lb/hr	0.00	0.00	0.00	0.00	0.00	10.05	10.05	0.01	0.01	10.05	0.00	0.00	0.00	0.00	0.00	0.00
	n-C26	lb/hr	0.00	0.00	0.00	0.00	0.00	8.10	8.10	0.00	0.00	8.10	0.00	0.00	0.00	0.00	0.00	0.00
	n-C27	lb/hr	0.00	0.00	0.00	0.00	0.00	6.52	6.52	0.00	0.00	6.51	0.00	0.00	0.00	0.00	0.00	0.00
	n-C28	lb/hr	0.00	0.00	0.00	0.00	0.00	5.23	5.23	0.00	0.00	5.23	0.00	0.00	0.00	0.00	0.00	0.00
	n-C29	lb/hr	0.00	0.00	0.00	0.00	0.00	4.20	4.20	0.00	0.00	4.20	0.00	0.00	0.00	0.00	0.00	0.00
	n-C30	lb/hr	0.00	0.00	0.00	0.00	0.00	3.37	3.37	0.00	0.00	3.37	0.00	0.00	0.00	0.00	0.00	0.00

Large Unit Stream Table

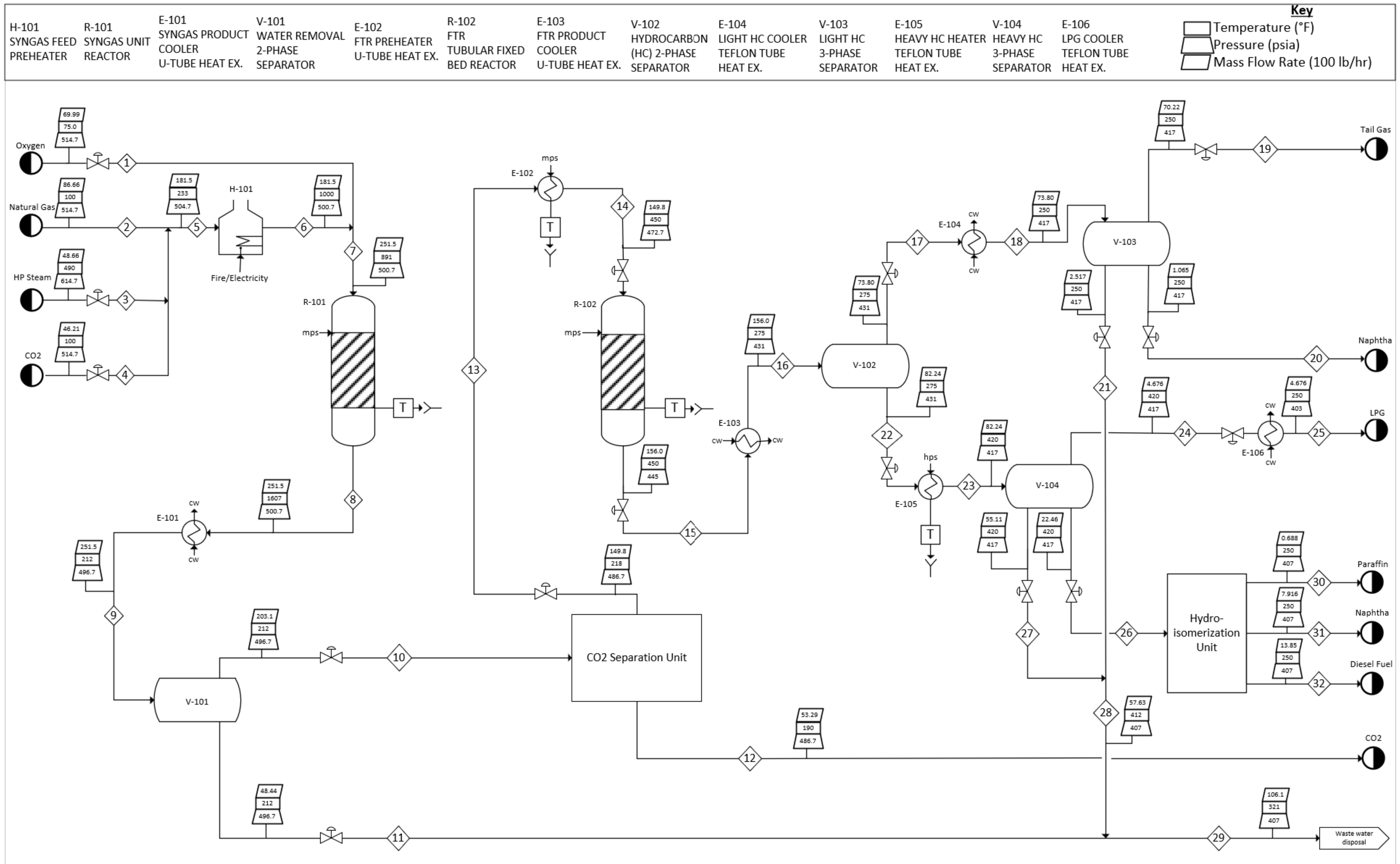
	Stream Number		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
	Description		Oxygen	Natural Gas	HP Steam	Carbon Dioxide	H-101 Feed	H-101 Heated Stream	R-101 Feed	E-101 Feed	V-101 Feed	CO2 Separation Unit Feed	V-101 Waste Water	Removed CO2	E-102 Feed	R-102 Feed	E-103 Feed	V-102 Feed
	Vapor Fraction		1.00	1.00	0.00	1.00	0.74	1.00	1.00	1.00	0.84	1.00	0.00	1.00	1.00	1.00	1.00	0.50
	Temperature	°F	75.00	100.00	490.00	100.00	233.82	1000.00	886.18	1626.92	212.00	212.00	212.00	190.46	217.67	450.00	450.00	275.00
	Pressure	psia	514.70	514.70	614.70	514.70	504.70	500.70	500.70	500.70	496.70	496.70	496.70	486.70	486.70	472.70	445.00	431.00
	Total Molar Flow	lbmole/hr	219.00	540.20	270.10	105.00	915.30	915.30	1134.30	1649.32	1649.32	1380.64	268.68	121.08	1259.56	1259.56	672.12	672.12
	Total Mass Flow	lb/hr	6999.27	8666.37	4865.88	4621.02	18153.27	18153.27	25152.54	25152.47	25152.47	20308.94	4843.54	5328.66	14980.28	14980.28	15604.41	15604.41
	Actual Vol. Flow	barrel/day	10122.55	25245.16	422.19	4257.89	40753.58	120339.45	137451.34	311045.78	84311.27	83961.90	349.37	6382.53	77580.39	105476.42	59504.81	25733.44
	Mass Enthalpy	BTU/lb	-4.85	-2010.79	-6371.27	-3857.52	-3649.69	-2985.65	-2156.18	-2156.16	-3158.72	-2317.69	-6685.19	-3833.61	-1778.45	-1630.49	-3047.50	-3485.18
	Oxygen	lb/hr	6937.92	0.00	0.00	0.00	0.00	0.00	6937.92	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Water	lb/hr	0.00	0.00	4865.88	0.00	4865.88	4865.88	4865.88	5678.66	5678.66	841.59	4837.07	0.00	841.59	841.59	6425.60	6425.60
	Nitrogen	lb/hr	61.35	0.00	0.00	0.00	0.00	0.00	61.35	61.35	61.35	61.35	0.00	0.00	61.35	61.35	30.68	30.68
	Carbon Monoxide	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	9616.89	9616.89	9616.50	0.39	0.00	9616.50	9616.50	936.87	936.87
	Carbon Dioxide	lb/hr	0.00	0.00	0.00	4621.02	4621.02	4621.02	4621.02	5615.13	5615.13	5609.12	6.01	5328.66	280.46	280.46	280.46	280.46
	Hydrogen	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1384.41	1384.41	1384.35	0.06	0.00	1384.35	1384.35	48.84	48.84
	Ethane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	43.06	43.06
	Propane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	63.15	63.15
	Methane	lb/hr	0.00	8666.37	0.00	0.00	8666.37	8666.37	8666.37	2796.04	2796.04	2796.03	0.00	0.00	2796.03	2796.03	3369.74	3369.74
	n-Butane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	83.23	83.23
Naphtha	n-Pentane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	580.32	580.32
	n-Hexane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	539.46	539.46
	n-Heptane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	487.55	487.55
	n-Octane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	431.64	431.64
	n-Nonane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	376.17	376.17
	n-Decane	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	323.79	323.79
Diesel	n-C11	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	275.91	275.91
	n-C12	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	233.16	233.16
	n-C13	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	195.67	195.67
	n-C14	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	163.24	163.24
	n-C15	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	135.49	135.49
	n-C16	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	111.96	111.96
	n-C17	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	92.15	92.15
	n-C18	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	75.58	75.58
	n-C19	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	61.80	61.80
	n-C20	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	50.40	50.40
	n-C21	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	40.99	40.99
	n-C22	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	33.27	33.27
	n-C23	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	26.94	26.94
	n-C24	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	21.78	21.78
	n-C25	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	17.57	17.57
	n-C26	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	14.16	14.16
	n-C27	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	11.39	11.39
	n-C28	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	9.15	9.15
	n-C29	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	7.34	7.34
	n-C30	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	5.88	5.88

	Stream Number		17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32
	Description		E-104 Feed	V-103 Feed	Tail Gas	Top Naphtha	Top Water Product	E-105 Feed	V-104 Feed	LPG	Cooled LPG	HIU Feed	Bottom Water Product	Top & Bottom Water Products	Waste Water Disposal	Paraffin	Bottom Naphtha	Diesel Fuel
	Vapor Fraction		1.00	0.96	1.00	0.00	0.00	0.00	0.04	1.00	0.10	0.00	0.00	0.00	0.00	0.21	0.00	0.00
	Temperature	°F	275.00	250.00	250.00	250.00	250.00	275.00	420.00	420.00	250.00	420.00	420.00	413.08	324.62	250.00	250.00	250.00
	Pressure	psia	431.00	417.00	417.00	417.00	417.00	431.00	417.00	417.00	433.00	417.00	417.00	407.00	407.00	407.00	407.00	407.00
	Total Molar Flow	lbmole/hr	335.83	335.83	320.96	0.90	13.97	336.29	336.29	14.18	14.18	16.27	305.84	319.81	588.50	3.25	1.230576	0.631977
	Total Mass Flow	lb/hr	7380.20	7380.20	7022.01	106.46	251.74	8224.20	8224.20	467.64	467.64	2245.80	5510.76	5762.50	10606.03	68.76	791.58	1385.45
	Actual Vol. Flow	barrel/day	25023.48	23354.71	23324.36	11.87	18.49	709.94	1846.12	1108.99	143.67	287.54	449.59	483.42	820.39	4.79	55.09	96.43
	Mass Enthalpy	BTU/lb	-1948.93	-1995.09	-1845.89	-850.51	-6640.92	-4863.78	-4706.66	-2745.52	-3243.13	-791.11	-6468.78	-6465.06	-6565.59	-4926.10	-859.82	-767.18
	Oxygen	lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Water	lb/hr	680.96	680.96	429.15	0.27	251.53	5744.64	5744.64	188.89	188.89	47.27	5508.48	5760.01	10597.07	47.27	0.00	0.00
	Nitrogen	lb/hr	30.58	30.58	30.57	0.00	0.00	0.10	0.10	0.09	0.09	0.01	0.00	0.00	0.01	0.01	0.00	0.00
	Carbon Monoxide	lb/hr	934.00	934.00	933.85	0.12	0.02	2.87	2.87	2.62	2.62	0.21	0.05	0.07	0.46	0.21	0.00	0.00
	Carbon Dioxide	lb/hr	277.24	277.24	277.06	0.14	0.04	3.22	3.22	2.58	2.58	0.47	0.17	0.20	6.21	0.47	0.00	0.00
	Hydrogen	lb/hr	48.74	48.74	48.73	0.00	0.00	0.10	0.10	0.10	0.10	0.01	0.00	0.00	0.07	0.01	0.00	0.00
	Ethane	lb/hr	42.46	42.46	42.42	0.03	0.00	0.60	0.60	0.47	17.77	0.12	0.01	0.01	0.01	1.06	0.00	0.00
	Propane	lb/hr	61.37	61.37	61.27	0.11	0.00	1.78	1.78	1.24	0.47	0.51	0.03	0.03	0.03	7.10	0.00	0.00
	Methane	lb/hr	3348.84	3348.84	3347.77	0.93	0.14	20.90	20.90	17.77	1.24	2.35	0.78	0.92	0.92	4.24	0.00	0.00
	n-Butane	lb/hr	78.65	78.65	78.35	0.29	0.00	4.58	4.58	2.73	2.73	1.81	0.05	0.05	0.05	8.40	0.00	0.00
Naphtha	n-Pentane	lb/hr	521.15	521.15	517.22	3.92	0.00	59.18	59.18	28.96	28.96	29.92	0.29	0.29	0.29	0.00	791.58	0.00
	n-Hexane	lb/hr	440.71	440.71	433.95	6.76	0.00	98.76	98.76	37.24	37.24	61.24	0.28	0.28	0.28	0.00		0.00
	n-Heptane	lb/hr	343.30	343.30	333.17	10.13	0.00	144.25	144.25	41.43	41.43	102.61	0.22	0.22	0.22	0.00		0.00
	n-Octane	lb/hr	241.06	241.06	227.50	13.56	0.00	190.58	190.58	40.38	40.38	150.04	0.17	0.17	0.17	0.00		0.00
	n-Nonane	lb/hr	152.02	152.02	136.32	15.69	0.00	224.16	224.16	33.85	33.85	190.19	0.11	0.11	0.11	0.00		0.00
	n-Decane	lb/hr	87.65	87.65	71.94	15.71	0.00	236.14	236.14	25.54	25.54	210.53	0.07	0.07	0.07	0.00		0.00
Diesel	n-C11	lb/hr	46.16	46.16	32.62	13.53	0.00	229.75	229.75	16.93	16.93	212.78	0.04	0.04	0.04	0.00	0.00	1385.45
	n-C12	lb/hr	23.20	23.20	13.03	10.16	0.00	209.97	209.97	10.72	10.72	199.22	0.02	0.02	0.02	0.00	0.00	
	n-C13	lb/hr	11.71	11.71	4.95	6.76	0.00	183.97	183.97	6.84	6.84	177.12	0.00	0.00	0.00	0.00	0.00	
	n-C14	lb/hr	5.41	5.41	1.47	3.94	0.00	157.83	157.83	3.85	3.85	153.98	0.00	0.00	0.00	0.00	0.00	
	n-C15	lb/hr	2.60	2.60	0.45	2.15	0.00	132.89	132.89	2.25	2.25	130.64	0.00	0.00	0.00	0.00	0.00	
	n-C16	lb/hr	1.21	1.21	0.12	1.09	0.00	110.74	110.74	1.28	1.28	109.47	0.00	0.00	0.00	0.00	0.00	
	n-C17	lb/hr	0.69	0.69	0.05	0.64	0.00	91.46	91.46	0.85	0.85	90.61	0.00	0.00	0.00	0.00	0.00	
	n-C18	lb/hr	0.28	0.28	0.01	0.27	0.00	75.30	75.30	0.43	0.43	74.87	0.00	0.00	0.00	0.00	0.00	
	n-C19	lb/hr	0.13	0.13	0.00	0.13	0.00	61.67	61.67	0.25	0.25	61.42	0.00	0.00	0.00	0.00	0.00	
	n-C20	lb/hr	0.06	0.06	0.00	0.06	0.00	50.33	50.33	0.14	0.14	50.19	0.00	0.00	0.00	0.00	0.00	
	n-C21	lb/hr	0.03	0.03	0.00	0.03	0.00	40.96	40.96	0.09	0.09	40.87	0.00	0.00	0.00	0.00	0.00	0.00
	n-C22	lb/hr	0.02	0.02	0.00	0.02	0.00	33.25	33.25	0.06	0.06	33.19	0.00	0.00	0.00	0.00	0.00	0.00
	n-C23	lb/hr	0.01	0.01	0.00	0.01	0.00	26.93	26.93	0.03	0.03	26.90	0.00	0.00	0.00	0.00	0.00	0.00
	n-C24	lb/hr	0.00	0.00	0.00	0.00	0.00	21.77	21.77	0.02	0.02	21.76	0.00	0.00	0.00	0.00	0.00	0.00
	n-C25	lb/hr	0.00	0.00	0.00	0.00	0.00	17.57	17.57	0.01	0.01	17.56	0.00	0.00	0.00	0.00	0.00	0.00
	n-C26	lb/hr	0.00	0.00	0.00	0.00	0.00	14.16	14.16	0.01	0.01	14.15	0.00	0.00	0.00	0.00	0.00	0.00
	n-C27	lb/hr	0.00	0.00	0.00	0.00	0.00	11.39	11.39	0.00	0.00	11.39	0.00	0.00	0.00	0.00	0.00	0.00
	n-C28	lb/hr	0.00	0.00	0.00	0.00	0.00	9.15	9.15	0.00	0.00	9.15	0.00	0.00	0.00	0.00	0.00	0.00
	n-C29	lb/hr	0.00	0.00	0.00	0.00	0.00	7.34	7.34	0.00	0.00	7.34	0.00	0.00	0.00	0.00	0.00	0.00
	n-C30	lb/hr	0.00	0.00	0.00	0.00	0.00	5.88	5.88	0.00	0.00	5.88	0.00	0.00	0.00	0.00	0.00	0.00

Medium Unit Process Flow Diagram

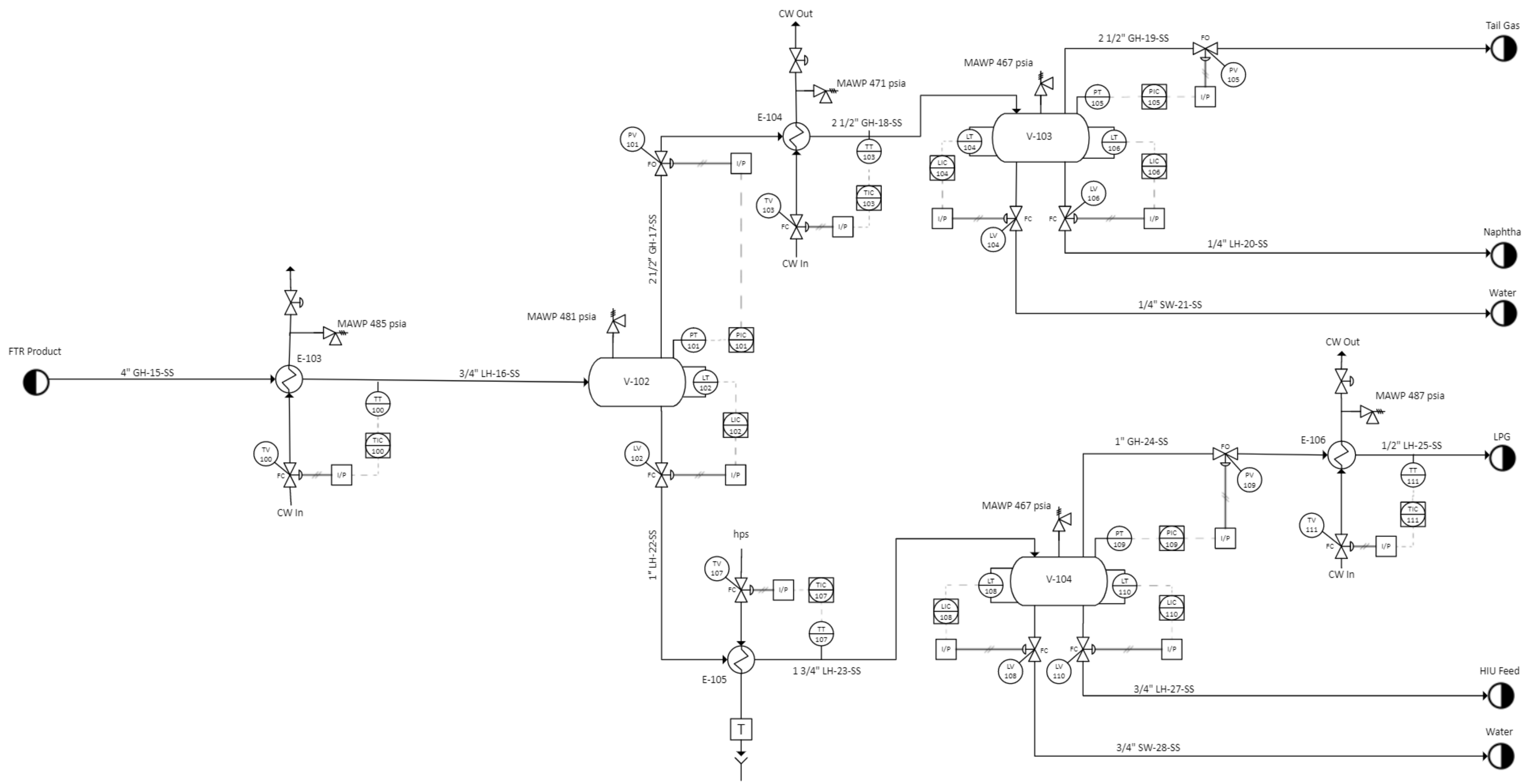


Large Unit Process Flow Diagram



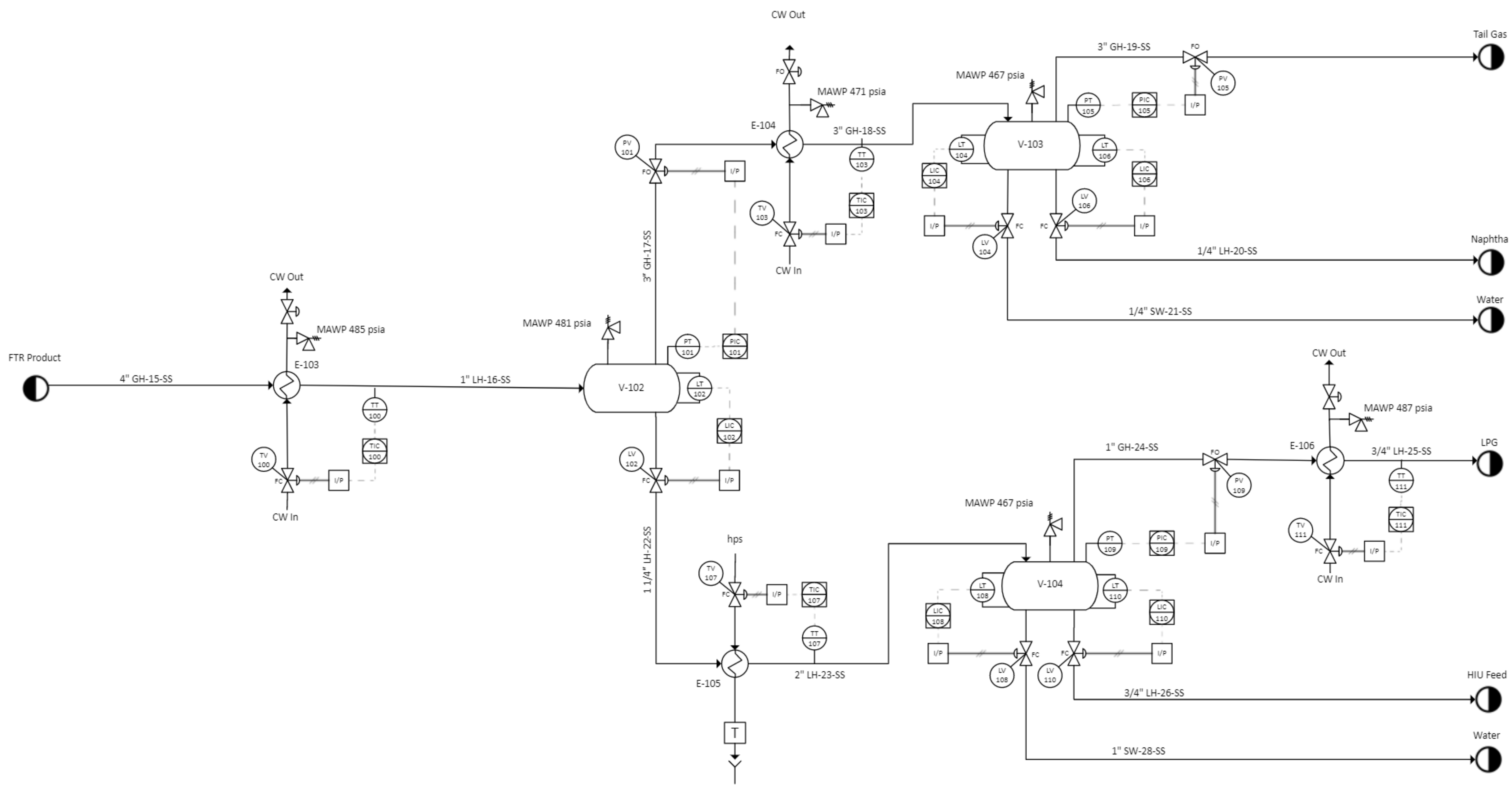
Medium Unit P&ID

E-103	V-102	E-104	V-103	E-105	V-104	E-106
FTR COOLER	HYDROCARBON	LIGHT HC COOLER	LIGHT HC	HEAVY HC HEATER	HEAVY HC	LPG COOLER
TEFLON TUBE	(HC) 2-PHASE	TEFLON TUBE	3-PHASE	TEFLON TUBE	3-PHASE	TEFLON TUB
HEAT EX.	SEPARATOR	U-TUBE HEAT EX.	SEPARATOR	HEAT EX.	SEPARATOR	HEAT EX.



Large Unit P&ID

E-103	V-102	E-104	V-103	E-105	V-104	E-106
FTR COOLER	HYDROCARBON	LIGHT HC COOLER	LIGHT HC	HEAVY HC HEATER	HEAVY HC	LPG COOLER
TEFLON TUBE	(HC) 2-PHASE	TEFLON TUBE	3-PHASE	TEFLON TUBE	3-PHASE	TEFLON TUB
HEAT EX.	SEPARATOR	U-TUBE HEAT EX.	SEPARATOR	HEAT EX.	SEPARATOR	HEAT EX.



Syngas Reactions

	Reaction 1 (Equilibrium)			
	CH4	H2O	CO	H2
Stoich	1	1	1	3
	Reaction 2 (Completion)			
	CH4	O2	CO	H2O
Stoich	1	1.5	1	2
	Reaction 3 (Equilibrium)			
	CO	H2O	CO2	H2
Stoich	1	1	1	1

Flow Rate of Natural Gas into the System

Molar Flow of Methane Into Syngas Unit						CO2	Steam	O2	CO2+Steam+O2	Total
MSCF/day	lbmole/hr	lb/hour	kg/hr	kg/day	kg/s	kg/hr	kg/hr	kg/hr	kg/s	kg/s
500	54.0211472	864.3383552	392	9409.4	0.108905	219.6	220.7	304.4	0.206861111	0.315766
2500	270.105736	4321.691776	1960	47046.8	0.544524	898.3	1104	1812	1.059527778	1.604051
5000	540.211472	8643.383552	3921	94093.7	1.089047	2096	2207	3175	2.077222222	3.166269

Capital Cost of Syngas Reactor

Capital Cost of Unit		
Ca	21620000	$
scaling factor	0.67	
Base Capacity	12.2	kg/s
Small Unit	1868827.58	$
Medium Unit	5552529.06	$
Large Unit	8757127.37	$

Utility Cost of Syngas Reactor

Total Utilty Cost of Syngas		
Small	14.0496415	$/hr
Medium	72.32035452	$/hr
Large	141.277803	$/hr

Cost of Steam into Syngas Reactor

Reactant Cost		
	Steam (kg/hr)	lb/hr
Small	220.7	486.5606095
Medium	1104	2433.90536
Large	2207	4865.606095

Utilities of Small Unit Syngas Reactor

Small Unit		
	1211000	
Total RequiredDuty	1211000	Btu/hr
Total Actual Duty	1424705.88	Btu/hr
Duty from Fire	1139764.71	Btu/hr
	1.13976471	MBtu/hr
Cost from Fire	3.41929412	$/hr
Duty from Electricity	142470.588	Btu/hr
	41.7539978	kW
Cost of Electricity	1.67015991	$/hr
Duty from 125 psig Steam	142470.588	Btu/hr
mass flowrate of 125psig stear	139.267437	lb/hr
	0.13926744	klb/hr
Cost of 125 psig Steam	0.55706975	$/hr
Total PreHeater Utility	5.64652378	$/hr
Cooler		
Duty	2480000	Btu/hr
mass flowrate of PCW	82749.4161	lb/hr
	9.92199234	kgal/hr
Utility Cost of Cooler	4.96099617	$/hr

Small Unit		
Duty	217800	Btu/hr
mass flowrate of HP Steam	299.175824	lb/hr
	0.29917582	
Cost of HP Steam	1.49587912	$/hr

Utilities of Medium Unit Syngas Reactor

Medium Unit		
	5950000	
Total RequiredDuty	5950000	Btu/hr
Total Actual Duty	7000000	Btu/hr
Duty from Fire	5600000	Btu/hr
	5.6	MBtu/hr
Cost from Fire	16.8	$/hr
Duty from Electricity	700000	Btu/hr
	205.1497	kW
Cost of Electricity	8.205988	$/hr
Duty from 125 psig Steam	700000	Btu/hr
mass flowrate of 125psig stear	684.261975	lb/hr
	0.68426197	klb/hr
Cost of 125 psig Steam	2.7370479	$/hr
Total Preheater Utility	27.7430359	$/hr
Cooler		
Duty	13390000	Btu/hr
mass flowrate of PCW	446780.113	lb/hr
	53.570757	kgal/hr
Utility Cost of Cooler	26.7853785	$/hr

Medium Unit		
Duty	1173000	Btu/hr
mass flowrate of HP Steam	1611.26374	lb/hr
	1.61126374	
Cost of HP Steam	8.05631868	$/hr

Utilities of Large Unit Syngas Reactor

Large Unit		
	12050000	
Total RequiredDuty	12050000	Btu/hr
Total Actual Duty	14176470.6	Btu/hr
Duty from Fire	11341176.5	Btu/hr
	11.3411765	MBtu/hr
Cost from Fire	34.0235294	$/hr
Duty from Electricity	1417647.06	Btu/hr
	415.471241	kW
Cost of Electricity	16.6188496	$/hr
Duty from 125 psig Steam	1417647.06	Btu/hr
mass flowrate of 125psig stear	1385.77425	lb/hr
	1.38577425	klb/hr
Cost of 125 psig Steam	5.543097	$/hr
Total PreHeater Utility	56.1854761	$/hr
Cooler		
Duty	25200000	Btu/hr
mass flowrate of PCW	840840.841	lb/hr
	100.820245	kgal/hr
Utility Cost of Cooler	50.4101224	$/hr

Large Unit		
Duty	2216000	Btu/hr
mass flowrate of HP Steam	3043.95604	lb/hr
	3.04395604	
Cost of HP Steam	15.2197802	$/hr

CO2 Recovery Capital Cost

CO2 Recovery Captital Cost			
Base Cost	5300000	$	
Base Capacity	8.54	kg/s	
scaling factor	0.55		
	CO2 kg/hr	CO2 kg/s	$
Small Unit	249.4	0.069278	375232.7
Medium Unit	1093	0.303611	845763.1
Large Unit	2417	0.671389	1308609

Anderson-Schulz-Flory Probability Distribution

Product Selectivity		450	(F)
	T	505.2222222	(K)
	T1	1.832519396	(K)
	T2	0.329341661	(K)
	T3	3.851176787	(K)
	T4	0.966852364	(K)
	alpha	0.899172699	

Hydrocarbon	#C	Wn	Mn
Pentane	5	0.033227582	0.065909891
Hexane	6	0.035852801	0.059264374
Heptane	7	0.037610837	0.053288907
Octane	8	0.038649872	0.047915931
Nonane	9	0.039097023	0.043084697
Decane	10	0.039061084	0.038740583
11-C	11	0.038634927	0.034834475
12-C	12	0.037897605	0.031322209
13-C	13	0.036916199	0.028164075
14-C	14	0.035747426	0.025324367
15-C	15	0.034439046	0.02277098
16-C	16	0.033031093	0.020475043
17-C	17	0.031556948	0.0184106
18-C	18	0.030044273	0.016554309
19-C	19	0.028515823	0.014885182
20-C	20	0.026990157	0.01338435
21-C	21	0.025482253	0.012034842
22-C	22	0.024004039	0.010821401
23-C	23	0.022564857	0.009730309
24-C	24	0.021171865	0.008749228
25-C	25	0.019830378	0.007867067
26-C	26	0.018544172	0.007073852
27-C	27	0.017315736	0.006360614
28-C	28	0.016146498	0.005719291
29-C	29	0.015037008	0.00514263
30-C	30	0.013987104	0.004624113
31-C	31	0.012996049	0.004157876
32-C	32	0.01206265	0.003738648
33-C	33	0.011185356	0.003361691
34-C	34	0.010362342	0.00302274
35-C	35	0.00959158	0.002717966
36-C	36	0.008870901	0.00244392
37-C	37	0.00819804	0.002197507
38-C	38	0.007570682	0.001975938
39-C	39	0.006986492	0.001776709
40-C	40	0.006443141	0.001597569
41-C	41	0.005938334	0.00143649
42-C	42	0.005469822	0.001291653
43-C	43	0.005035417	0.001161419
44-C	44	0.004633005	0.001044316
45-C	45	0.004260551	0.000939021
46-C	46	0.003916104	0.000844342
47-C	47	0.003597802	0.000759209
48-C+	48	0.003303876	0.00068266
	Sum	0.88177875	0.647602988
			0.352397012

C21-C25	C26-C29	C30-C35	C36-C47
0.049202846	0.024296	0.007342078	0.017468092

Levenspiel Plot Data

	Graph indepentent of temperature			
X	Rate of CC	FCO0/rCO	PCO (atm)	PH2 (atm)
0	3.269415	0.717898	7.154	14.308
0.1	3.0043	0.781249	6.754	13.464
0.2	2.714725	0.864584	6.313	12.532
0.3	2.398904	0.978408	5.825	11.499
0.4	2.054881	1.142211	5.28	10.348
0.5	1.682093	1.395349	4.668	9.056
0.6	1.283392	1.828831	3.976	7.596
0.7	0.868768	2.701651	3.189	5.933
0.8	0.463055	5.068749	2.285	4.022
0.9	0.125686	18.67434	1.235	1.803

Flow Rate into Small Unit FTR

Small Unit			Number of Tubes	14
Out of Syngas Reactor			Flow rate per channel	
Molar Flow Overall	1.0000	122.97	8.7838	
Molar Fraction H2O	0.0373	4.59	0.3275	
Molar Fraction Nitrogen	0.0017	0.21	0.0150	
Molar Fraction CO	0.2672	32.86	2.3471	
Molar Fraction CO2	0.0053	0.66	0.0470	
Molar Fraction Hydrogen	0.5340	65.67	4.6906	
Molar Fraction Methane	0.1544	18.99	1.3566	

Variables for Small Unit FTR

Diamter of each tube	0.1667	ft
Number of tubes	14	
Overall minimum diameter	1	ft
Length	3	ft
Volume	2.35619449	ft3
Weight of catalyst	2.505	lb
Overall pressure drop	0.0001357	psi
Conversion	0.94	

Mass Balance of Small FTR

Product Flow Rates from FTR								
	Relative Mass Frac	Molar Flow for each tube	Total Molar Flow		Mass Flows		Total Inlet Mass Flow	
Molar Flow Overall	1475		1475		1586.258691		1475	
FC2		0.0102284	0.1431976		4.305951832			
FC3		0.0102284	0.1431976		6.31501416		Carbon Flow Rate	
FC4		0.0102284	0.1431976		8.322644512		635.3965	
FCH4		1.61231	22.57234		362.0603336			
FCO		0.1338822	1.8743508		52.50056591		Percent error	
FCO2		0.0469714	0.6575996		28.9409584		-7.542962074	
FH2		0.130683	1.829562		1.84785762			
FH2O		2.543261	35.605654		641.6138851			
FN2		0.015	0.21		2.9421			
FC5	0.033227582				21.11269024			
FC6	0.035852801				22.7807456			
FC7	0.037610837				23.89779525			
FC8	0.038649872				24.55799435			
FC9	0.039097023				24.84211282			
FC10	0.039061084				24.81927736			
FC11	0.038634927				24.54849827			
FC12	0.037897605				24.08000667			
FC13	0.036916199				23.45642497			
FC14	0.035747426				22.71379056			
FC15	0.034439046				21.88245038			
FC16	0.033031093				20.9878421			
FC17	0.031556948				20.05117554			
FC18	0.030044273				19.09002666			
FC19	0.028515823				18.11885473			
FC20	0.026990157				17.14945211			
FC21	0.025482253				16.1913351			
FC22	0.024004039				15.25208298			
FC23	0.022564857				14.33763192			
FC24	0.021171865				13.45252924			
FC25	0.019830378				12.60015315			
FC26	0.018544172				11.78290226			
FC27	0.017315736				11.0023588			
FC28	0.016146498				10.25942883			
FC29	0.015037008				9.554462531			
FC30	0.013987104				8.887357097			

Flow Rate into Medium Unit FTR

Medium Unit			Number of Tubes	29
Out of Syngas Reactor				Flow rate per channel
Molar Flow Overall	1.0000	679.94		23.4461
Molar Fraction H2O	0.0365	24.84		0.8564
Molar Fraction Nitrogen	0.0018	1.25		0.0431
Molar Fraction CO	0.2894	196.76		6.7847
Molar Fraction CO2	0.0042	2.88		0.0994
Molar Fraction Hydrogen	0.5788	393.55		13.5707
Molar Fraction Methane	0.0892	60.66		2.0918

Variables for Medium Unit FTR

Diamter of each tube	0.1667	ft
Number of tubes	29	
Overall minimum diameter	1.25	ft
Length	3.75	ft
Volume	4.601942364	ft3
Weight of catalyst	3.6098	lb
Overall pressure drop	0.0022624	psi
Conversion	0.9009	

Mass Balance of FTR

Product Flow Rates from FTR							
	Relative Mass Fra	Molar Flow for each tube	Total Molar Flow	Mass Flows		Total Inlet Mass Flow	
Molar Flow Overall (lbmol/hr	7887		7887	8499.031234		7887	
FC2		0.0282493	0.8192297	24.63423708			
FC3		0.0282493	0.8192297	36.12802977		Product Flow Rate	
FC4		0.0282493	0.8192297	47.61363016		3631.211	
FCH4		2.798032	81.142928	1301.532565			
FCO		0.6720297	19.4888613	545.883005		Percent error	
FCO2		0.099369	2.881701	126.823661		-7.76	
FH2		0.9785495	28.3779355	28.66171486			
FH2O		6.969094	202.103726	3641.909143			
FN2		0.0431	1.2499	17.511099			
FC5	0.033227582			120.6563512			
FC6	0.035852801			130.1890764			
FC7	0.037610837			136.5728737			
FC8	0.038649872			140.3458279			
FC9	0.039097023			141.969529			
FC10	0.039061084			141.8390273			
FC11	0.038634927			140.2915591			
FC12	0.037897605			137.6141889			
FC13	0.036916199			134.0504984			
FC14	0.035747426			129.8064368			
FC15	0.034439046			125.055433			
FC16	0.033031093			119.9428599			
FC17	0.031556948			114.5899291			
FC18	0.030044273			109.097085			
FC19	0.028515823			103.5469604			
FC20	0.026990157			98.00694721			
FC21	0.025482253			92.53142979			
FC22	0.024004039			87.16372286			
FC23	0.022564857			81.93775084			
FC24	0.021171865			76.87950111			
FC25	0.019830378			72.00827969			
FC26	0.018544172			67.33779436			
FC27	0.017315736			62.8770873			
FC28	0.016146498			58.6313366			
FC29	0.015037008			54.60254349			
FC30	0.013987104			50.79012041			

Flow Rate into Large Unit FTR

Large Unit			Number of Tubes	49			
Out of Syngas Reactor				Flow rate per channel			
Molar Flow Overall	1.0000	1259.5600		25.7053			
Molar Fraction H2O	0.0371	46.7159		0.9534			
Molar Fraction Nitrogen	0.0017	2.1899		0.0447			
Molar Fraction CO	0.2726	343.3128		7.0064			
Molar Fraction CO2	0.0051	6.3726		0.1301			
Molar Fraction Hydrogen	0.5452	686.6826		14.0139			
Molar Fraction Methane	0.1384	174.2847		3.5568			

Variables for Large Unit FTR

Diamter of each tube	0.1667	ft
Number of tubes	49	
Overall minimum diameter	1.5	ft
Length	4.5	ft
Volume	7.952156404	ft3
Weight of catalyst	4.4293	lb
Overall pressure drop	0.003285	psi
Conversion	0.9026	

Mass Balance of FTR

Product Flow Rates from FTR								
	Relative Mas	Molar Flow for each tube	Total Molar Flow		Mass Flows		Total Mass Flow Rate	
Molar Flow Overall (lbmol/hr	14980		14980		15207.44		14980.2767	
FC2		0.0292248	1.4320152	FC2	43.0607			
FC3		0.0292248	1.4320152	FC3	63.15187		Product Flow Rate	
FC4		0.0292248	1.4320152	FC4	83.22872		6347.16191	
FCH4		4.28742	210.08358	FCH4	3369.741			
FCO		0.06826033	3.34475617	FCO	93.68662		Percent error	
FCO2		0.1300531	6.3726019	FCO2	280.4582		-1.5163888	
FH2		0.9869124	48.3587076	FH2	48.84229			
FH2O		7.27718	356.58182	FH2O	6425.604			
FN2		0.0446918	2.1898982	FN2	30.68047			
FC5	0.033227582			FC5	210.9008			
FC6	0.035852801			FC6	227.5635			
FC7	0.037610837			FC7	238.7221			
FC8	0.038649872			FC8	245.317			
FC9	0.039097023			FC9	248.1551			
FC10	0.039061084			FC10	247.927			
FC11	0.038634927			FC11	245.2221			
FC12	0.037897605			FC12	240.5422			
FC13	0.036916199			FC13	234.3131			
FC14	0.035747426			FC14	226.8947			
FC15	0.034439046			FC15	218.5902			
FC16	0.033031093			FC16	209.6537			
FC17	0.031556948			FC17	200.2971			
FC18	0.030044273			FC18	190.6959			
FC19	0.028515823			FC19	180.9945			
FC20	0.026990157			FC20	171.3109			
FC21	0.025482253			FC21	161.74			
FC22	0.024004039			FC22	152.3575			
FC23	0.022564857			FC23	143.2228			
FC24	0.021171865			FC24	134.3813			
FC25	0.019830378			FC25	125.8666			
FC26	0.018544172			FC26	117.7029			
FC27	0.017315736			FC27	109.9058			
FC28	0.016146498			FC28	102.4844			
FC29	0.015037008			FC29	95.44232			
FC30	0.013987104			FC30	88.77841			

Sensitivity Analysis for Tornado Chart

	Transportation	Revenue	Utility Cost
Worst NPV	4340888.263	4301423.444	3830297.08
Best NPV	4347010.557	4357711.62	4403804.644

Monte Carlo Variables

Probability	Frequency	NPV
0.086797011	1940	1086713
0.171938616	3843	1738621
0.25520111	5704	2390529
0.337971455	7554	3042437
0.420965505	9409	3694345
0.504541184	11277	4346253
0.58668516	13113	4998161
0.671513579	15009	5650068
0.754418147	16862	6301976
0.835577826	18676	6953884
0.920316764	20570	7605792
0.999910519	22349	8257700

Small Unit FTR Sizing

SMALL UNIT

FTR Sizing

A (ft^2)	L	D	n	A (m^2)
0.785398163	2.5	1	14	0.072965877

Utilites

Steam Flow (lbmole/hr)	Steam Flow lb/hr	Steam Flow klb/hr	Cost of Steam/hour	Cost of Steam/year	P/A .08, 20	PWC of Utilities	F/A .08, 20	FWC
	2505.94179	2.50594179	$10.02	$70,246.56	9.818147407	$689,691.08	45.7619643	$3,214,620.58

Catalyst Cost

Price of Catalyst ($/lb)	Amount (lb)	Cost ($)
10	2.505	25.05

Total Cost	$549.58

Purchasing Cost of the Small Unit

Purchase Cost						
C_p	192.3849314					
	Purchase Cost of Base Conditions					
	$C_p^{\ 0}$	109.2763864				
		K_1	2.7652	Range 10-100 m^2		
		K_2	0.7282	Assumed MultiPipe **		
		K_3	0.0783			
		A (m^2)	0.07296588	Area of Reboiler		
	Pressure Factor					
	F_p	0.978075145				
		P (barg)	33.6734694	Design Pressure		
		C_1	0.6072	Pressure Range:		
		C_2	-0.912	40 < P < 100 barg		
		C_3	0.3327			
	Material of Construction (MOC) Factor					
	F_m	1.8				
		Assumed SS-Shell/SS-tube, Table A.3/Figure A.18				
Installed Cost						
$C_{BM,\ Heat\ Exchanger}$	$524.53					
	B_1	1.74	Heat Exchanger Multiple Pipe, Table A.4			
	B_2	1.55				

Medium Unit FTR Sizing

MEDIUM UNIT

FTR Sizing

A (ft^2)	L	D	n	A (m^2)
1.227185	2.5	1.25	14	0.114009183

Utilites

Steam Flow (lbmole/hr)	Steam Flow lb/hr	Steam Flow klb/hr	Cost of Steam/hour	Cost of Steam/year	P/A .08, 20	PWC of Utilities	F/A .08, 20	FWC
	14336.66283	14.33666283	$57.35	$502,356.67	9.818147407	$4,932,211.79	45.7619643	$22,988,827.79

Catalyst Cost

Price of Catalyst ($/lb)	Amount (lb)	Cost ($)
10	3.6098	36.098

Total Cost	$711.16

Purchasing Cost of the Medium Unit

Purchase Cost						
C_p	247.9576053					
	Purchase Cost of Base Conditions					
	C_p^0	140.637348				
		K_1	2.7652	Range 10-100 m^2		
		K_2	0.7282	Assumed MultiPipe **		
		K_3	0.0783			
		A (m^2)	0.11400918	Area of Reboiler		
	Pressure Factor					
	F_p	0.979499593				
		P (barg)	34.137931	Design Pressure		
		C_1	0.6072	Pressure Range:		
		C_2	-0.912	40 < P < 100 barg		
		C_3	0.3327			
	Material of Construction (MOC) Factor					
	F_m	1.8				
		Assumed SS-Shell/SS-tube, Table A.3/Figure A.18				
Installed Cost						
$C_{BM,\ Heat\ Exchanger}$	$675.06					
	B_1	1.74	Heat Exchanger Multiple Pipe, Table A.4			
	B_2	1.55				

Large Unit FTR Sizing

LARGE UNIT

FTR Sizing

A (ft^2)	L	D	n	A (m^2)
1.767146	4.5	1.5	14	0.164173223

Utilites

Steam Flow (lbmole/hr)	Steam Flow lb/hr	Steam Flow klb/hr	Cost of Steam/hour	Cost of Steam/year	P/A .08, 20	PWC of Utilities	F/A .08, 20	FWC
	16813.52112	16.81352112	$ 67.25	$589,145.78	9.818147407	$ 5,784,320.11	45.7619643	$ 26,960,468.16

Total Cost	
$ 882.28	

Catalyst Cost

Price of Catalyst ($/lb)	Amount (lb)	Cost ($)
10	4.4293	$ 44.29

Purchasing Cost of the Large Unit

Purchase Cost					
C_p	307.8032334				
	Purchase Cost of Base Conditions				
	C_p^0	174.5807732			
		K_1	2.7652	Range 10-100 m^2	
		K_2	0.7282	Assumed MultiPipe **	
		K_3	0.0783		
		A (m^2)	0.164173223	Area of Reboiler	
	Pressure Factor				
	F_p	0.979499593			
		P (barg)	34.13793103	Design Pressure	
		C_1	0.6072	Pressure Range:	
		C_2	-0.912	$40 < P < 100$ barg	
		C_3	0.3327		
	Material of Construction (MOC) Factor				
	F_m	1.8			
		Assumed SS-Shell/SS-tube, Table A.3/Figure A.18			
Installed Cost					
M, Heat Excha	$837.99				
	B_1	1.74	Heat Exchanger Multiple Pipe, Table A.4		
	B_2	1.55			

Capital Cost for the Small Unit

Small Unit	
	Capital Cost
Syngas	$1,868,827.58
CO2 Recovery Syster	$375,232.73
FTR	$549.58
Separation	$102,977.17
Air Separation	
HIU	$10,061,097.19
Total	$12,408,684.26
CEPCI 2001	397
CEPCI 2022	776.9
Scaled Capital Cost	$13,308,161.69
Total Capital Invest.	$13,308,161.69

Capital Cost in 2024	
	Capital Cost
Syngas	$1,982,639.18
CO2 Recovery Syster	$398,084.40
HIU	$10,673,818.01

Capital Cost for the Medium Unit

Medium Unit	
	Capital Cost
Syngas	$5,552,529.06
CO2 Recovery Syste	$845,763.14
FTR	$711.16
Separation	$234,346.19
Air Sepration	
HIU	$27,905,278.66
Total	$34,538,628.21
CEPCI 2001	397
CEPCI 2022	776.9
Scaled Capital Cost	$36,968,502.40
Total Capital Invest.	$36,968,502.40

Capital Cost in 2024	
	Capital Cost
Syngas	$5,890,678.07
CO2 Recovery Syste	$897,270.12
HIU	$29,604,710.13

Capital Cost for the Large Unit

Large Unit	
	Capital Cost
Syngas	$8,757,127.37
CO2 Recovery Syste	$1,308,608.76
FTR	$882.28
Separation	$313,299.68
Air Separation Unit	
HIU	$36,802,218.56
Total	$47,182,136.65
CEPCI 2001	397
CEPCI 2022	776.9
Scaled Capital Cost	$50,491,896.88
Total Capital Invest.	$50,491,896.88

Capital Cost in 2024	
	Capital Cost
Syngas	9290436.428
CO2 Recovery Syste	1388303.033
HIU	39043473.67

HIU Costing

Costing			
Base HIU	8290000	$	
Base Capacity	1.13	kg/s	
scaling factor	0.55		
	kg/s	$	bbl/day of feed
Small Unit	1.606833	10061097	873.2175067
Medium Unit	10.26833	27905279	5580.223067
Large Unit	16.98333	36802219	9229.422667

Utilities for Air Separation Plant

Oxygen					
	O2 mole flowrate (lbr	lb/hour	short ton/hr	$/hour	$/yr
Small Unit	21	669.48	0.33474	33.474	234585.8
Medium Unit	125	3985	1.9925	199.25	1396344
Large Unit	219	6981.72	3.49086	349.086	2446395

Electricity Option			
	Short Ton O2/hr	kWh/hr	$/hr
Small Unit	0.33474	3347.4	100.422
Medium Unit	1.9925	19925	597.75
Large Unit	3.49086	34908.6	1047.258

Steam Option					
	Short Ton O2/hr	lb 600 psig stea	klb/hr	$/hr	$/yr
Small Unit	0.33474	3347.4	3.3474	16.737	117292.9
Medium Unit	1.9925	19925	19.925	99.625	698172
Large Unit	3.49086	34908.6	34.9086	174.543	1223197

Cooling Water Requirement					
	Short Ton O2/hr	Short Ton O2/d	Water gpm	$/hr	$/yr
Small Unit	0.33474	8.03376	3213.504	96.40512	675607.1
Medium Unit	1.9925	47.82	19128	573.84	4021471
Large Unit	3.49086	83.78064	33512.256	1005.36768	7045617

Product Formation of Small HIU

Product Formation of Small HIU				
Component	Inlet (lb/hr)		Product (lb/hr)	Outlet (lb/hr)
Methane	0.236		0.181165	0.417165
Ethane	0.0106		0.0905825	0.1011825
Propane	0.046		0.6340775	0.6800775
Butane	0.1622		0.6340775	0.7962775
Naphtha	68.7368		4.529125	73.265925
Diesel	120.737		12.0474725	132.7844725
C-11+	18.1165		0	0
Oxygen	0			0
Nitrogen	0.0007			0.0007
Hydrogen	0.0005			0.0005
Carbon Monoxide	0.0178			0.0178
Carbon Dioxide	0.0457			0.0457
Water	4.4395			4.4395

Product Formation of Medium HIU

Product Formation of Medium HIU				
Component	Inlet (lb/hr)		Product (lb/hr)	Outlet (lb/hr)
Methane	1.1336		1.07638	2.20998
Ethane	0.0861		0.53819	0.62429
Propane	0.3763		3.76733	4.14363
Butane	1.3347		3.76733	5.10203
Naphtha	488.163		26.9095	515.0725
Diesel	729.709		71.57927	801.28827
C-11+	107.638		0	0
Oxygen	0			0
Nitrogen	0.0057			0.0057
Hydrogen	0.0041			0.0041
Carbon Monoxide	0.1505			0.1505
Carbon Dioxide	0.2641			0.2641
Water	29.3745			29.3745

Product Formation of Large HIU

Product Formation of Large HIU				
Component	Inlet (lb/hr)		Product (lb/hr)	Outlet (lb/hr)
Methane	2.354		1.88189	4.23589
Ethane	0.1183		0.940945	1.059245
Propane	0.5128		6.586615	7.099415
Butane	1.8104		6.586615	8.397015
Naphtha	744.533		47.04725	791.58025
Diesel	1260.3		125.145685	1385.445685
C-11+	188.189		0	0
Oxygen	0			0
Nitrogen	0.008			0.008
Hydrogen	0.0056			0.0056
Carbon Monoxide	0.2087			0.2087
Carbon Dioxide	0.4714			0.4714
Water	47.2738			47.2738

Utilities for Each Unit of the HIU

Utilities of HIU for Small Unit						
Utilities	Cost					
Fuel Gas	0.08	MBTU/bbl	209.5722	$/day	8.732175	$/hr
Hydrogen	300	SCF/bbl	82.20	$/day	3.425196	$/hr
MP Steam	10	lb/bbl	34.9287	$/day	1.455363	$/hr
Electricity	2.5	kWh/bbl	87.32175	$/day	3.638406	$/hr
Cooling Wate	300	gal/bbl	130.9826	$/day	5.457609	$/hr

Utilities of HIU for Medium Unit						
Utilities	Cost					
Fuel Gas	0.08	MBTU/bbl	1339.254	$/day	55.80223	$/hr
Hydrogen	300	SCF/bbl	525.32	$/day	21.88842	$/hr
MP Steam	10	lb/bbl	223.2089	$/day	9.300372	$/hr
Electricity	2.5	kWh/bbl	558.0223	$/day	23.25093	$/hr
Cooling Wate	300	gal/bbl	837.0335	$/day	34.87639	$/hr

Utilities of HIU for Large Unit						
Utilities	Cost					
Fuel Gas	0.08	MBTU/bbl	2215.061	$/day	92.29423	$/hr
Hydrogen	300	SCF/bbl	868.86	$/day	36.20241	$/hr
MP Steam	10	lb/bbl	369.1769	$/day	15.38237	$/hr
Electricity	2.5	kWh/bbl	922.9423	$/day	38.45593	$/hr
Cooling Wate	300	gal/bbl	1384.413	$/day	57.68389	$/hr

Small Unit Separator Sizing

SMALL UNIT SEPARATOR SIZING											
Constants		V-102		Constants		V-103		Constants		V-104	
P (psia)	481.00	Q_V (ft^3/s)	0.17	ρ_v (lb/ft3)	1.28	Q_V (ft^3/s)	0.16	ρ_v	1.78	Q_V (ft^3/s)	0.01
ρ_v (lb/ft3)	1.25	Q_L (ft^3/s)	0.01	ρ_L (lb/ft3)	38.38	Q_L (ft^3/s)	0.00	ρ_L	33.44	Q_L (ft^3/s)	0.00
ρ_L (lb/ft3)	37.54	U_T (ft/s)	1.55	ρ_H	58.21	Q_{LL} (ft^3/s)	0.00	ρ_H	52.39	Q_{LL} (ft^3/s)	0.00
K	0.29	U_V (ft/s)	1.16	P (psia)	467.00	Q_{HL} (ft^3/s)	0.00	P (psia)	467.00	Q_{HL} (ft^3/s)	0.00
m_V (lb/hr)	752.05	T_H (s)	120.00	μ_L (cP)	0.25	U_T (ft/s)	1.55	μ_L (cP)	0.19	U_T (ft/s)	1.22
m_L (lb/hr)	783.41	T_S (s)	60.00	μ_H (cP)	0.23	U_V (ft/s)	1.17	μ_H (cP)	0.12	U_V (ft/s)	0.91
		V_H (ft^3)	0.70	K	0.29	T_H (min)	120.00	K	0.29	T_H (min)	120.00
H/D to A/At		V_S (ft^3)	0.35	ks	0.16	T_S (min)	60.00	ks	0.16	T_S (min)	60.00
a	4.75593E-05	L/D (est.)	2.00	m_V (lb/hr)	715.78	V_H (ft^3)	0.01	m_V (lb/hr)	41.94	V_H (ft^3)	0.21
b	3.924091	D (ft)	1.50	m_L (lb/hr)	10.45	V_S (ft^3)	0.00	m_L (lb/hr)	212.55	V_S (ft^3)	0.11
c	0.174875	A_T (ft^2)	1.77	m_{LL} (lb/hr)	10.45	L/D (est)	2.00	m_{LL} (lb/hr)	212.55	L/D (est)	2.00
d	-6.358805	H_{LLL} (ft)	0.75	m_{HL} (lb/hr)	25.83	D (ft)	0.50	m_{HL} (lb/hr)	528.93	D (ft)	1.50
e	5.668973	H_{LLL}/D	0.50			A_T (ft^2)	0.20			A_T (ft^2)	1.77
f	4.018448	A_{LLL}/A_T	0.50			H_V (ft)	0.10			H_V (ft)	0.30
g	-4.916411	A_{LLL} (ft^2)	0.88			H_V/D	0.20			H_V/D	0.20
h	-1.801705	H_V (ft)	0.30			A_V/A_T	0.14			A_V/A_T	0.14
i	-0.145348	H_V/D	0.20			A_V (ft^2)	0.03			A_V (ft^2)	0.25
		A_V/A_T	0.14			H_{LLL} (ft)	0.75			H_{LLL} (ft)	0.75
A/At to H/D		A_V (ft^2)	0.25			H_{LLL}/D	1.50			H_{LLL}/D	0.50
a	0.00153756	L_{min} (ft)	2.00			A_{LLL}/A_T	1.45			A_{LLL}/A_T	0.50
b	26.787101	r (ft)	0.75			A_{LLL} (ft^2)	0.28			A_{LLL} (ft^2)	0.88
c	3.299201	V (ft^3)	3.53			H_W (ft)	0.40			H_W (ft)	1.20
d	-22.923932					L_2 (ft)	0.00			L_2 (ft)	1.00
e	24.353518					H_W/2 (ft)	0.20			H_W/2 (ft)	0.60
f	-14.844824					H_{HL} (ft)	0.20			H_{HL} (ft)	0.60
g	-36.999376					H_{LL} (ft)	0.20			H_{LL} (ft)	0.60
h	10.529572					H_{HL}/D	0.40			H_{HL}/D	0.40
i	9.892851					A_{HL}/A_T	0.37			A_{HL}/A_T	0.37
						A_{HL} (ft^2)	0.07			A_{HL} (ft^2)	0.66
						A_{LL} (ft^2)	0.10			A_{LL} (ft^2)	0.86
						U_{HL} (ft/s)	12.78			U_{HL} (ft/s)	15.88
						U_{LH} (ft/s)	14.19			U_{LH} (ft/s)	24.93
						t_{HL} (s)	0.70			t_{HL} (s)	0.57
						t_{LH} (s)	0.17			t_{LH} (s)	0.29
						L_1 (ft)	0.50			L_1 (ft)	0.50
						L min (ft)	0.50			Lmin (ft)	1.50
						ϕ (s)	0.09			ϕ (s)	0.33
						U_{VA} (ft/s)	5.56			U_{VA} (ft/s)	0.03
						L_{min} (ft)	0.50			L_{min} (ft)	0.50
						H_{HLL} (ft)	0.40			H_{HLL} (ft)	1.20
						r (ft)	0.25			r (ft)	0.75
						V (ft^3)	0.10			V (ft^3)	2.65

Small Unit Separator Sizing

SMALL UNIT SEPARATOR COSTING							
V-102			V-103			V-104	
K1	3.5565		K1	3.5565		K1	3.5565
K2	0.3776		K2	0.3776		K2	0.3776
K3	0.0905		K3	0.0905		K3	0.0905
V (ft^3)	3.534291735		V (ft^3)	0.09817477		V (ft^3)	2.6507188
V (m^3)	0.100079052		V (m^3)	0.002779974		V (m^3)	0.07505929
Cp°	$1,859.81		Cp°	$1,522.65		Cp°	$1,763.23
P (psi)	481		P (psi)	467		P (psi)	467
P (bar)	33.16326531		P (bar)	32.19801434		P (bar)	32.1980143
D (ft)	1.5		D (ft)	0.5		D (ft)	1.5
D (m)	0.45717769		D (m)	0.152392563		D (m)	0.45717769
Fp	1.949573655		Fp	0.968800365		Fp	1.9064011
Fm	3.1		Fm	3.1		Fm	3.1
B1	1.49		B1	1.49		B1	1.49
B2	1.52		B2	1.52		B2	1.52
Cbm_0	$8,927.07		Cbm_0	$7,308.71		Cbm_0	$8,463.51
			Total	$24,699.29			

Small Unit Heat Exchanger Costing

SMALL UNIT HEAT EXCHANGER COSTING										
E-103			E-104			E-105			E-106	
K1	3.8062		K1	3.8062		K1	3.8062		K1	3.8062
K2	0.8924		K2	0.8924		K2	0.8924		K2	0.8924
K3	-0.1671		K3	-0.1671		K3	-0.1671		K3	-0.1671
A (ft^2)	19.484554		A (ft^2)	1.53853862		A (ft^2)	7.28466728		A (ft^2)	0.805141006
A (m^2)	1.8101761		A (m^2)	0.14293505		A (m^2)	0.67676839		A (m^2)	0.074800119
Cp°	$10,594.76		Cp°	$856.98		Cp°	$4,467.66		Cp°	$ 388.48
P (psi)	485		P (psi)	471		P (psi)	471		P (psi)	487
P (bar)	33.439512		P (bar)	32.4742481		P (bar)	32.4742481		P (bar)	33.57925946
C1	0		C1	0		C1	0		C1	0
C2	0		C2	0		C2	0		C2	0
C3	0		C3	0		C3	0		C3	0
Fp	1		Fp	1		Fp	1		Fp	1
Fm	2.7		Fm	2.7		Fm	2.7		Fm	2.7
B1	1.63		B1	1.63		B1	1.63		B1	1.63
B2	1.66		B2	1.66		B2	1.66		B2	1.66
Cbm_0	$50,854.87		Cbm_0	$4,113.51		Cbm_0	$21,444.79		Cbm_0	$ 1,864.72
			Total	$78,277.88						

Small Unit Heat Exchanger Utility Costing

SMALL UNIT HEAT EXCHANGER UTILITY COSTING												
	E-103			E-104			E-105			E-106		
	lb/hr	22011.991		lb/hr	1292.264		lb/hr	168.0899		lb/hr	700.7348	
	klb/hr	22.011991		klb/hr	1.292264		klb/hr	0.16809		klb/hr	0.700735	
PCW	$/hr	11.005996	PCW	$/hr	0.646132	HP Steam	$/hr	0.840449	PCW	$/hr	0.350367	
	$/yr	77130.018		$/yr	4528.093		$/yr	5889.869		$/yr	2455.375	
						Total Utili	12.84294442					

Small Unit Heat Exchanger Sizing

SMALL UNIT HEAT EXCHANGER SIZING										
E-103			E-104			E-105			E-106	
Duty (Btu/hr)	659039.02		Duty (Btu/hr)	34891.13		Duty (Btu/hr)	122369.43		Duty (Btu/hr)	20980.00
Thi (F)	450.00		Thi (F)	275.00		Thi (F)	490.00		Thi (F)	420.00
Tci (F)	90.00		Tci (F)	90.00		Tci (F)	275.00		Tci (F)	90.00
Tho (F)	275.00		Tho (F)	250.00		Tho (F)	490.00		Tho (F)	250.00
Tco (F)	120.00		Tco (F)	120.00		Tco (F)	420.00		Tco (F)	120.00
ΔTlm (F)	250.55		ΔTlm (F)	157.49		ΔTlm (F)	129.22		ΔTlm (F)	222.71
U (Btu/hr*ft^2*°F	150.00		U (Btu/hr*ft^2*°F)	160.00		U (Btu/hr*ft^2*°F)	130.00		U (Btu/hr*ft^2*°F)	130.00
A (ft^2)	19.48		A (ft^2)	1.54		A (ft^2)	7.28		A (ft^2)	0.81
Cv (Btu/lbm °F)	1.00		Cv (Btu/lbm °F)	1.00		λ (Btu/lb)	728.00		Cv (Btu/lbm °F)	1.00
m (lb/hr)	22011.99		m (lb/hr)	1292.26		m (lb/hr)	168.09		m (lb/hr)	700.73
F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90
cooler			cooler			heater			cooler	

Medium Unit Separator Sizing

MEDIUM UNIT SEPARATOR SIZING											
Constants		V-102		Constants		V-103		Constants		V-104	
P (psia)	481.00	Q_V (ft^3/s)	0.71	ρ_v (lb/ft3)	1.33	Q_V (ft^3/s)	0.66	ρ_v	1.88	Q_V (ft^3/s)	0.05
ρ_V (lb/ft3)	1.31	Q_L (ft^3/s)	0.04	ρ_L (lb/ft3)	38.13	Q_L (ft^3/s)	0.00	ρ_L	33.20	Q_L (ft^3/s)	0.01
ρ_L (lb/ft3)	37.42	U_T (ft/s)	1.51	ρ_H	58.21	Q_{LL} (ft^3/s)	0.00	ρ_H	52.40	Q_{LL} (ft^3/s)	0.01
K	0.29	U_V (ft/s)	1.13	P (psia)	467.00	Q_{HL} (ft^3/s)	0.00	P (psia)	467.00	Q_{HL} (ft^3/s)	0.02
m_V (lb/hr)	3344.53	T_H (s)	120.00	μ_L (cP)	0.23	U_T (ft/s)	1.52	μ_L (cP)	0.19	U_T (ft/s)	1.18
m_L (lb/hr)	4899.21	T_S (s)	60.00	μ_H (cP)	0.23	U_v (ft/s)	1.14	μ_H (cP)	0.12	U_v (ft/s)	0.88
		V_H (ft^3)	4.36	K	0.29	T_H (min)	120.00	K	0.29	T_H (min)	120.00
H/D to A/At		V_S (ft^3)	2.18	ks	0.16	T_S (min)	60.00	ks	0.16	T_S (min)	60.00
a	-4.75593E-05	L/D (est.)	3.00	m_V (lb/hr)	3178.66	V_H (ft^3)	0.05	m_V (lb/hr)	366.85	V_H (ft^3)	1.36
b	3.924091	D (ft)	2.00	m_L (lb/hr)	55.84	V_S (ft^3)	0.02	m_L (lb/hr)	1358.24	V_S (ft^3)	0.68
c	0.174875	A_T (ft^2)	3.14	m_{LL} (lb/hr)	55.84	L/D (est)	3.00	m_{LL} (lb/hr)	1358.24	L/D (est)	3.00
d	-6.358805	H_{LLL} (ft)	0.75	m_{HL} (lb/hr)	110.03	D (ft)	1.00	m_{HL} (lb/hr)	3174.13	D (ft)	2.00
e	5.668973	H_{LLL}/D	0.38			A_T (ft^2)	0.79			A_T (ft^2)	3.14
f	4.018448	A_{LLL}/A_T	0.34			H_V (ft)	0.20			H_V (ft)	0.40
g	-4.916411	A_{LLL} (ft^2)	1.08			H_V/D	0.20			H_V/D	0.20
h	-1.801705	H_V (ft)	0.40			A_V/A_T	0.14			A_V/A_T	0.14
i	-0.145348	H_V/D	0.20			A_V (ft^2)	0.11			A_V (ft^2)	0.45
		A_V/A_T	0.14			H_{LLL} (ft)	0.07			H_{LLL} (ft)	0.75
A/At to H/D		A_V (ft^2)	0.45			H_{LLL}/D	0.07			H_{LLL}/D	0.38
a	0.00153756	L_{min} (ft)	4.50			A_{LLL}/A_T	0.03			A_{LLL}/A_T	0.34
b	26.787101	r (ft)	1.00			A_{LLL} (ft^2)	0.03			A_{LLL} (ft^2)	1.08
c	3.299201	V (ft^3)	14.14			H_W (ft)	0.80			H_W (ft)	1.60
d	-22.923932					L_2 (ft)	0.50			L_2 (ft)	1.50
e	24.353518					$H_W/2$ (ft)	0.40			$H_W/2$ (ft)	0.80
f	-14.844824					H_{HL} (ft)	0.40			H_{HL} (ft)	0.80
g	-36.999376					H_{LL} (ft)	0.40			H_{LL} (ft)	0.80
h	10.529572					H_{HL}/D	0.40			H_{HL}/D	0.40
i	9.892851					A_{HL}/A_T	0.37			A_{HL}/A_T	0.37
						A_{HL} (ft^2)	0.29			A_{HL} (ft^2)	1.17
						A_{LL} (ft^2)	0.38			A_{LL} (ft^2)	1.52
						U_{HL} (ft/s)	14.01			U_{HL} (ft/s)	16.82
						U_{LH} (ft/s)	14.38			U_{LH} (ft/s)	25.23
						t_{HL} (s)	0.06			t_{HL} (s)	0.54
						t_{LH} (s)	0.33			t_{LH} (s)	0.38
						L_1 (ft)	0.50			L_1 (ft)	0.50
						L min (ft)	1.00			Lmin (ft)	2.00
						ϕ (s)	0.18			ϕ (s)	0.45
						U_{VA} (ft/s)	5.92			U_{VA} (ft/s)	0.12
						L_{min} (ft)	1.50			L_{min} (ft)	0.50
						H_{HLL} (ft)	0.80			H_{HLL} (ft)	1.60
						r (ft)	0.50			r (ft)	1.00
						V (ft^3)	0.79			V (ft^3)	6.28

Medium Unit Separator Costing

MEDIUM UNIT SEPARATOR COSTING							
V-102			V-103			V-104	
K1	3.5565		K1	3.5565		K1	3.5565
K2	0.3776		K2	0.3776		K2	0.3776
K3	0.0905		K3	0.0905		K3	0.0905
V (ft^3)	14.13716694		V (ft^3)	0.785398163		V (ft^3)	6.28318531
V (m^3)	0.40031621		V (m^3)	0.022239789		V (m^3)	0.17791832
Cp°	$2,634.35		Cp°	$1,512.24		Cp°	$2,109.89
P (psi)	481		P (psi)	467		P (psi)	467
P (bar)	33.16326531		P (bar)	32.19801434		P (bar)	32.1980143
D (ft)	2		D (ft)	1		D (ft)	2
D (m)	0.609570253		D (m)	0.304785126		D (m)	0.60957025
Fp	2.432764873		Fp	1.43760073		Fp	2.37520146
Fm	3.1		Fm	3.1		Fm	3.1
B1	1.49		B1	1.49		B1	1.49
B2	1.52		B2	1.52		B2	1.52
Cbm_0	$12,644.89		Cbm_0	$7,258.75		Cbm_0	$10,127.47
			Total	$30,031.11			

Medium Unit Heat Exchanger Costing

MEDIUM UNIT HEAT EXCHANGER COSTING										
E-103			E-104			E-105			E-106	
K1	4.188		K1	3.8062		K1	3.8062		K1	3.8062
K2	-0.2503		K2	0.8924		K2	0.8924		K2	0.8924
K3	0.1974		K3	-0.1671		K3	-0.1671		K3	-0.1671
A (ft^2)	114.68447		A (ft^2)	6.645263725		A (ft^2)	47.341428		A (ft^2)	6.869410869
A (m^2)	10.654546		A (m^2)	0.617365799		A (m^2)	4.39816683		A (m^2)	0.63818977
Cp°	$13,779.34		Cp°	$4,092.13		Cp°	$20,469.63		Cp°	4224.545471
P (psi)	485		P (psi)	471		P (psi)	471		P (psi)	487
P (bar)	33.439512		P (bar)	32.47424813		P (bar)	32.4742481		P (bar)	33.5774073
C1	0.03881		C1	0		C1	0		C1	0
C2	-0.11272		C2	0		C2	0		C2	0
C3	0.08183		C3	0		C3	0		C3	0
Fp	1.1405561		Fp	1		Fp	1		Fp	1
Fm	2.7		Fm	2.7		Fm	2.7		Fm	2.7
B1	1.63		B1	1.63		B1	1.63		B1	1.63
B2	1.66		B2	1.66		B2	1.66		B2	1.66
Cbm_0	$66,140.84		Cbm_0	$19,642.21		Cbm_0	$98,254.21		Cbm_0	$20,277.82
			Total	$204,315.08						

Medium Unit Heat Exchanger Utility Costing

MEDIUM UNIT HEAT EXCHANGER UTILITY COSTING											
	E-103			E-104			E-105			E-106	
	lb/hr	129560.8		lb/hr	5033.464		lb/hr	1092.379		lb/hr	5978.624
	klb/hr	129.5608		klb/hr	5.033464		klb/hr	1.092379		klb/hr	5.978624
PCW	$/hr	64.78038	PCW	$/hr	2.516732	HP Steam	$/hr	5.461893	PWC	$/hr	2.989312
	$/yr	453980.9		$/yr	17637.26		$/yr	38276.95		$/yr	20949.1
					Total Utili	75.74831					

Medium Unit Heat Exchanger Sizing

MEDIUM UNIT HEAT EXCHANGER SIZING										
E-103			E-104			E-105			E-106	
Duty (Btu/hr)	3879048.97		Duty (Btu/hr)	150701.92		Duty (Btu/hr)	795251.66		Duty (Btu/hr)	179000.00
Thi (F)	450.00		Thi (F)	275.00		Thi (F)	490.00		Thi (F)	420.00
Tci (F)	90.00		Tci (F)	90.00		Tci (F)	275.00		Tci (F)	90.00
Tho (F)	275.00		Tho (F)	250.00		Tho (F)	490.00		Tho (F)	250.00
Tco (F)	120.00		Tco (F)	120.00		Tco (F)	420.00		Tco (F)	120.00
ΔTlm (F)	250.55		ΔTlm (F)	157.49		ΔTlm (F)	129.22		ΔTlm (F)	222.71
U (Btu/hr*ft^2*°F	150.00		U (Btu/hr*ft^2*°F)	160.00		U (Btu/hr*ft^2*°F)	130.00		U (Btu/hr*ft^2*°F)	130.00
A (ft^2)	114.68		A (ft^2)	6.65		A (ft^2)	47.34		A (ft^2)	6.87
Cv (Btu/lbm °F)	1.00		Cv (Btu/lbm °F)	1.00		λ (Btu/lb)	728.00		Cv (Btu/lbm °F)	1.00
m (lb/hr)	129560.75		m (lb/hr)	5033.46		m (lb/hr)	1092.38		m (lb/hr)	5978.62
F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90
cooler			cooler			heater			cooler	

Large Unit Separator Sizing

LARGE UNIT SEPARATORS											
Constants		V-102		Constants		V-103		Constants		V-104	
P (psia)	481.00	Q_V (ft^3/s)	1.63	ρ_v (lb/ft3)	1.29	Q_V (ft^3/s)	1.52	ρ_v	1.80	Q_V (ft^3/s)	0.07
ρ_V (lb/ft3)	1.26	Q_L (ft^3/s)	0.06	ρ_L (lb/ft3)	38.34	Q_L (ft^3/s)	0.00	ρ_L	33.39	Q_L (ft^3/s)	0.02
ρ_L (lb/ft3)	37.52	U_T (ft/s)	1.54	ρ_H	58.21	Q_{LL} (ft^3/s)	0.00	ρ_H	52.40	Q_{LL} (ft^3/s)	0.02
K	0.29	U_V (ft/s)	1.16	P (psia)	467.00	Q_{HL} (ft^3/s)	0.00	P (psia)	467.00	Q_{HL} (ft^3/s)	0.03
m_V (lb/hr)	7380.21	T_H (s)	120.00	μ_L (cP)	0.25	U_T (ft/s)	1.55	μ_L (cP)	0.19	U_T (ft/s)	1.21
m_L (lb/hr)	8224.19	T_S (s)	60.00	μ_H (cP)	0.23	U_v (ft/s)	1.16	μ_H (cP)	0.12	U_v (ft/s)	0.91
		V_H (ft^3)	7.31	K	0.29	T_H (min)	120.00	K	0.29	T_H (min)	120.00
H/D to A/At		V_S (ft^3)	3.65	ks	0.16	T_S (min)	60.00	ks	0.16	T_S (min)	60.00
a	-4.75593E-05	L/D (est.)	3.00	m_V (lb/hr)	7022.01	V_H (ft^3)	0.09	m_V (lb/hr)	467.64	V_H (ft^3)	2.24
b	3.924091	D (ft)	2.00	m_L (lb/hr)	106.46	V_S (ft^3)	0.05	m_L (lb/hr)	2245.79	V_S (ft^3)	1.12
c	0.174875	A_T (ft^2)	3.14	m_{LL} (lb/hr)	106.46	L/D (est)	3.00	m_{LL} (lb/hr)	2245.79	L/D (est)	4.00
d	-6.358805	H_{LLL} (ft)	0.75	m_{HL} (lb/hr)	251.75	D (ft)	1.00	m_{HL} (lb/hr)	5510.76	D (ft)	2.00
e	5.668973	H_{LLL}/D	0.38			A_T (ft^2)	0.79			A_T (ft^2)	3.14
f	4.018448	A_{LLL}/A_T	0.34			H_V (ft)	0.20			H_V (ft)	0.40
g	-4.916411	A_{LLL} (ft^2)	1.08			H_V/D	0.20			H_V/D	0.20
h	-1.801705	H_V (ft)	0.40			A_V/A_T	0.14			A_V/A_T	0.14
i	-0.145348	H_V/D	0.20			A_V (ft^2)	0.11			A_V (ft^2)	0.45
		A_V/A_T	0.14			H_{LLL} (ft)	0.75			H_{LLL} (ft)	0.75
A/At to H/D		A_V (ft^2)	0.45			H_{LLL}/D	0.75			H_{LLL}/D	0.38
a	0.00153756	L_{min} (ft)	7.00			A_{LLL}/A_T	0.80			A_{LLL}/A_T	0.34
b	26.787101	r (ft)	1.00			A_{LLL} (ft^2)	0.63			A_{LLL} (ft^2)	1.08
c	3.299201	V (ft^3)	21.99			H_W (ft)	0.80			H_W (ft)	1.60
d	-22.923932					L_2 (ft)	3.50			L_2 (ft)	2.50
e	24.353518					H_W/2 (ft)	0.40			H_W/2 (ft)	0.80
f	-14.844824					H_{HL} (ft)	0.40			H_{HL} (ft)	0.80
g	-36.999376					H_{LL} (ft)	0.40			H_{LL} (ft)	0.80
h	10.529572					H_{HL}/D	0.40			H_{HL}/D	0.40
i	9.892851					A_{HL}/A_T	0.37			A_{HL}/A_T	0.37
						A_{HL} (ft^2)	0.29			A_{HL} (ft^2)	1.17
						A_{LL} (ft^2)	0.38			A_{LL} (ft^2)	1.52
						U_{HL} (ft/s)	13.01			U_{HL} (ft/s)	16.08
						U_{LH} (ft/s)	14.22			U_{LH} (ft/s)	24.99
						t_{HL} (s)	0.69			t_{HL} (s)	0.56
						t_{LH} (s)	0.34			t_{LH} (s)	0.38
						L_1 (ft)	0.50			L_1 (ft)	0.50
						L min (ft)	4.00			Lmin (ft)	3.00
						ϕ (s)	0.17			ϕ (s)	0.44
						U_{VA} (ft/s)	13.55			U_{VA} (ft/s)	0.16
						L_{min} (ft)	2.50			L_{min} (ft)	0.50
						H_{HLL} (ft)	0.80			H_{HLL} (ft)	1.60
						r (ft)	0.50			r (ft)	1.00
						V (ft^3)	3.14			V (ft^3)	9.42

Large Unit Separator Costing

LARGE UNIT COSTING							
V-102			V-103			V-104	
K1	3.5565		K1	3.5565		K1	3.5565
K2	0.3776		K2	0.3776		K2	0.3776
K3	0.0905		K3	0.0905		K3	0.0905
V (ft^3)	21.99114858		V (ft^3)	3.141592654		V (ft^3)	9.42477796
V (m^3)	0.622714104		V (m^3)	0.088959158		V (m^3)	0.26687747
Cp°	$3,038.46		Cp°	$1,818.21		Cp°	$2,342.40
P (psi)	481		P (psi)	467		P (psi)	467
P (bar)	33.16326531		P (bar)	32.19801434		P (bar)	32.1980143
D (ft)	2		D (ft)	1		D (ft)	2
D (m)	0.609570253		D (m)	0.304785126		D (m)	0.60957025
Fp	2.432764873		Fp	1.43760073		Fp	2.37520146
Fm	3.1		Fm	3.1		Fm	3.1
B1	1.49		B1	1.49		B1	1.49
B2	1.52		B2	1.52		B2	1.52
Cbm_0	14,584.59		Cbm_0	8,727.42		Cbm_0	11,243.51
			Total	$34,555.52			

Large Unit Heat Exchanger Costing

LARGE UNIT HEAT EXCHANGER COSTING										
E-103			E-104			E-105			E-106	
K1	4.188		K1	3.8062		K1	3.8062		K1	3.8062
K2	-0.2503		K2	0.8924		K2	0.8924		K2	0.8924
K3	0.1974		K3	-0.1671		K3	-0.1671		K3	-0.1671
A (ft^2)	181.731811		A (ft^2)	13.52147142		A (ft^2)	85.47153497		A (ft^2)	8.93023413
A (m^2)	16.88345404		A (m^2)	1.256187016		A (m^2)	7.940573117		A (m^2)	0.829646701
Cp°	$15,072.36		Cp°	$7,815.50		Cp°	$29,779.77		Cp°	5404.071011
P (psi)	485		P (psi)	471		P (psi)	471		P (psi)	487
P (bar)	33.4395124		P (bar)	32.47424813		P (bar)	32.47424813		P (bar)	33.5774073
C1	0.03881		C1	0		C1	0		C1	0
C2	-0.11272		C2	0		C2	0		C2	0
C3	0.08183		C3	0		C3	0		C3	0
Fp	1.140556102		Fp	1		Fp	1		Fp	1
Fm	2.7		Fm	2.7		Fm	2.7		Fm	2.7
B1	1.63		B1	1.63		B1	1.63		B1	1.63
B2	1.66		B2	1.66		B2	1.66		B2	1.66
Cbm_0	$72,347.35		Cbm_0	$37,514.38		Cbm_0	$142,942.90		Cbm_0	$ 25,939.54
			Total	$278,744.16						

Large Unit Heat Exchanger Utility Costing

LARGE UNIT HEAT EXCHANGER UTILITY COSTING											
	E-103			E-104			E-105			E-106	
	lb/hr	228116.81		lb/hr	11379.84		lb/hr	1774.99		lb/hr	7772.211
	klb/hr	228.11681		klb/hr	11.37984		klb/hr	1.77499		klb/hr	7.772211
PCW	$/hr	114.05841	PCW	$/hr	5.689921	HP Steam	$/hr	8.874949	PCW	$/hr	3.886106
	$/yr	799321.32		$/yr	39874.97		$/yr	62195.64		$/yr	27233.83
				Total Utility	132.5094						

Large Unit Heat Exchanger Sizing

LARGE UNIT HEAT EXCHANGER SIZING										
E-103			E-104			E-105			E-106	
Duty (Btu/hr)	6829817.44		Duty (Btu/hr)	340712.46		Duty (Btu/hr)	1292192.59		Duty (Btu/hr)	232700.00
Thi (F)	450.00		Thi (F)	275.00		Thi (F)	490.00		Thi (F)	420.00
Tci (F)	90.00		Tci (F)	90.00		Tci (F)	275.00		Tci (F)	90.00
Tho (F)	275.00		Tho (F)	250.00		Tho (F)	490.00		Tho (F)	250.00
Tco (F)	120.00		Tco (F)	120.00		Tco (F)	420.00		Tco (F)	120.00
ΔTlm (F)	250.55		ΔTlm (F)	157.49		ΔTlm (F)	129.22		ΔTlm (F)	222.71
U (Btu/hr*ft^2*°F	150.00		U (Btu/hr*ft^2*°F)	160.00		U (Btu/hr*ft^2*°F)	130.00		U (Btu/hr*ft^2*°F)	130.00
A (ft^2)	181.73		A (ft^2)	13.52		A (ft^2)	85.47		A (ft^2)	8.93
Cv (Btu/lbm °F)	1.00		Cv (Btu/lbm °F)	1.00		λ (Btu/lb)	728.00		Cv (Btu/lbm °F)	1.00
m (lb/hr)	228116.81		m (lb/hr)	11379.84		m (lb/hr)	1774.99		m (lb/hr)	7772.21
F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90		F (h-ft2-°F/Btu)	0.90
cooler			cooler			heater			cooler	

Small Unit Pipe Diameter Sizing

Small Unit							
Stream	vapor frac	mass Flowrate (lb/hr)	density (lb/ft^3)	volumetric (ft^3/hr)		D (in)	D (in)
15	0.9989	1535	1.109	1384.129847	172.5548542	2	2
16	0.5173	1535	2.492	615.9711075	76.79106474	0.4006499	0.5
17	1	752.1	1.252	600.7188498	74.88961661	1.5	1.5
18	0.9559	752.1	1.342	560.4321908	69.86721311	1.5	1.5
19	1	715.8	1.279	559.6559812	69.77044566	1.5	1.5
20	0	10.45	38.38	0.272277228	0.033943894	0.05	0.25
21	0	25.83	58.21	0.443738189	0.055319361	0.06	0.25
22	0	783.4	49.62	15.78798871	1.968235926	0.35	0.5
23	0.0399	783.4	19.59	39.98979071	4.985393908	0.56	0.75
24	1	41.93	1.781	23.5429534	2.935021523	0.375	0.5
25	0.1053	41.93	13.32	3.147897898	0.392437938	0.16	0.25
26	0	212.5	33.44	6.354665072	0.792214912	0.22	0.25
27	0	528.9	52.39	10.09543806	1.258564612	0.28	0.5
28	0	554.8	52.7	10.52751423	1.312430108	0.29	0.5
29	0	1038	56.03	18.52578976	2.309548456	0.38	0.5
11	0	483.6	59.37	8.145528044	1.01547583	0.25	0.5

Medium Unit Pipe Diameter Sizing

Medium Unit							
Stream	vapor frac	mass Flowrate (lb/hr)	density (lb/ft^3)	volumetric (ft^3/hr	gal/min	D (in)	D (in)
15	1	8224	1.174	7005.110733	875.6388416	4	4
16	0.4267	8244	3.103	2656.783758	332.0979697	0.7188015	0.75
17	1	3345	1.308	2557.33945	319.6674312	2.5	2.5
18	0.955	3345	1.402	2385.877318	298.2346648	2.5	2.5
19	1	3179	1.335	2381.273408	297.659176	2.5	2.5
20	0	55.84	38.13	1.464463677	0.18305796	0.1069632	0.25
21	0	110	58.21	1.889709672	0.236213709	0.1215046	0.25
22	0	4899	49.1	99.77596741	12.47199593	0.8828928	1
23	0.0544	4899	16.54	296.191052	37.0238815	1.5211813	1.75
24	1	366.8	1.883	194.795539	24.34944238	1	1
25	0.0785	366.8	16.61	22.08308248	2.76038531	0.4153602	0.5
26	0	1358	33.2	40.90361446	5.112951807	0.5652959	0.75
27	0	3174	52.4	60.57251908	7.571564885	0.6879119	0.75
28	0	3284	51.11	64.25357073	8.031696341	0.7085062	0.75
29	0	5729	55.1	103.9745917	12.99682396	0.9012777	1
11	0	2445	58.29	41.94544519	5.243180648	0.5724498	0.75

Large Unit Pipe Diameter Sizing

Large Unit							
Stream	vapor frac	mass Flowrate (lb/hr)	density (lb/ft^3)	volumetric (ft^3/hr	gal/min	D (in)	D (in)
15	0.9998	15600	1.121	13916.1463	1739.518287	4	4
16	0.4887	15600	2.592	6018.518519	752.3148148	0.9939834	1
17	1	7380	1.261	5852.498017	731.5622522	3	3
18	0.9557	7380	1.351	5462.620281	682.8275352	3	3
19	1	7022	1.287	5456.099456	682.012432	3	3
20	0	106.5	38.24	2.785041841	0.34813023	0.1475064	0.25
21	0	251.7	58.21	4.323999313	0.540499914	0.1837967	0.25
22	0	8224	49.52	166.0743134	20.75928918	1.1390591	1.25
23	0.0422	8224	19.04	431.9327731	53.99159664	1.8369744	2
24	1	467.6	1.803	259.3455352	32.4181919	1	1
25	0.0998	467.6	13.91	33.61610352	4.20201294	0.5124703	0.75
26	0	2246	33.39	67.2656484	8.40820605	0.7249227	0.75
27	0	5511	52.4	105.1717557	13.14646947	0.9064515	1
28	0	5762	50.95	113.0912659	14.13640824	0.9399604	1
29	0	10610	55.26	192.0014477	24.00018096	1.2247495	1.25
12	0	4844	59.26	81.74147823	10.21768478	0.7991278	1

FTR Material Compatibility Chart

	METHANE	WATER	CARBON MONOXIDE	HYDROGEN	ETHANE	PROPANE	BUTANE	N-PENTANE	N-HEXANE	N-HEPTANE	N-OCTANE	N-NONANE	N-DECANE	UNDECANE	N-DODECANE	TRIDECANE	N-TETRADECANE	N-PENTADECANE	N-HEXADECANE	N-HEPTADECANE	N-OCTADECANE	N-NONADECANE
WATER	Compatible																					
CARBON MONOXIDE	Compatible	Compatible																				
HYDROGEN	Compatible	Compatible	Compatible																			
ETHANE	Compatible	Compatible	Compatible	Compatible																		
PROPANE	Compatible	Compatible	Compatible	Compatible	Compatible																	
BUTANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible																
N-PENTANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible															
N-HEXANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible														
N-HEPTANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible													
N-OCTANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible												
N-NONANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible											
N-DECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible										
UNDECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible									
N-DODECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible								
TRIDECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible							
N-TETRADECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible						
N-PENTADECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible					
N-HEXADECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible				
N-HEPTADECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible			
N-OCTADECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible		
N-NONADECANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	
N-EICOSANE	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible	Compatible

Small Unit Vapor Deflagration Analysis

Small Unit										
R-102			V-102			V-103			V-104	
Pressure (psia)	445.00		Pressure (psia)	441.00		Pressure (psia)	437.00		Pressure (psia)	437.00
Volume (ft3)	2.36		Volume (ft3)	3.53		Volume (ft3)	0.10		Volume (ft3)	2.65
R (ft^3*psi/R*lbmol)	10.73		R (ft3*psi/R*lbmol)	10.73		R (ft3*psi/R*lbmol)	10.73		R (ft3*psi/R*lbmol)	10.73
Temp (F)	450.00		Temp (F)	275.00		Temp (F)	250.00		Temp (F)	420.00
Temp (°R)	909.67		Temp (°R)	734.67		Temp (°R)	709.67		Temp (°R)	879.67
n (lbmol)	0.11		n (lbmol)	0.20		n (lbmol)	0.01		n (lbmol)	0.12
MW (lb/lbmol)	23.04		MW (lb/lbmol)	23.04		MW (lb/lbmol)	21.82		MW (lb/lbmol)	24.35
m (lb)	2.47		m (lb)	4.56		m (lb)	0.12		m (lb)	2.99
Molar Enthalpy of Vapor (Btu/lbmo	-69128.24		Molar Enthalpy of Vapor (Btu/lbmol	-69127.45		Molar Enthalpy of Vapor (Btu/lbmol	-42933.84		Molar Enthalpy of Vapor (Btu/lbmole	-99961.72
Molar Enthalpy of Liquid (Btu/lbmo	-90204.45		Molar Enthalpy of Liquid (Btu/lbmol	-90208.19		Molar Enthalpy of Liquid (Btu/lbmol	-56009.49		Molar Enthalpy of Liquid (Btu/lbmole	-119104.05
λ	21076.21		λ	21080.74		λ	13075.65		λ	19142.33
Cp (BTU/lbmol*F)	12.71		Cp (BTU/lbmol*F)	11.54		Cp (BTU/lbmol*F)	11.35		Cp (BTU/lbmol*F)	12.25
T1 (F)	450.00		T1 (F)	275.00		T1 (F)	250.00		T1 (F)	420.00
Tb (F)	382.67		Tb (F)	382.67		Tb (F)	186.14		Tb (F)	406.29
mv (lb)	0.10		mv (lb)	-0.27		mv (lb)	0.01		mv (lb)	0.03
η	0.02		η	0.02		η	0.02		η	0.02
ΔH_c (Btu/lb)	10315.89		ΔH_c (Btu/lb)	10315.89		ΔH_c (Btu/lb)	16237.14		ΔH_c (Btu/lb)	4631.63
E_{TNT} (Btu/lb)	2014.62		E_{TNT} (Btu/lb)	2014.62		E_{TNT} (Btu/lb)	2014.62		E_{TNT} (Btu/lb)	2014.62
m_{TNT} (kg)	0.00		m_{TNT} (lb)	-0.012482		m_{TNT} (lb)	0.000499		m_{TNT} (lb)	0.0005472
m_{TNT} (g)	4.67									
r_1 (m)	10.00		r_1 (m)	10.00		r_1 (m)	10.00		r_1 (m)	10.00
z_e ($m/kg^{1/3}$)	713.59		z_e ($m/kg^{1/3}$)	-267.0536		z_e ($m/kg^{1/3}$)	6682.771		z_e ($m/kg^{1/3}$)	6091.9951
p_s	0.00		p_s	0.01		p_s	0.00		p_s	0.00
p_O (kPa)	0.24		p_O (kPa)	0.63		p_O (kPa)	0.03		p_O (kPa)	0.03

Medium Unit Vapor Deflagration Analysis

Medium Unit										
R-102			V-102			V-103			V-104	
Pressure (psia)	445.00		Pressure (psia)	441.00		Pressure (psia)	437.00		Pressure (psia)	437.00
Volume (ft3)	4.60		Volume (ft3)	14.14		Volume (ft3)	0.79		Volume (ft3)	6.28
R (ft^3*psi/R*lbmol)	10.73		R (ft3*psi/R*lbmol)	10.73		R (ft3*psi/R*lbmol)	10.73		R (ft3*psi/R*lbmol)	10.73
Temp (F)	450.00		Temp (F)	275.00		Temp (F)	250.00		Temp (F)	420.00
Temp (°R)	909.67		Temp (°R)	734.67		Temp (°R)	709.67		Temp (°R)	879.67
n (lbmol)	0.21		n (lbmol)	0.79		n (lbmol)	0.05		n (lbmol)	0.29
MW (lb/lbmol)	24.00		MW (lb/lbmol)	24.00		MW (lb/lbmol)	22.82		MW (lb/lbmol)	24.88
m (lb)	5.03		m (lb)	18.98		m (lb)	1.03		m (lb)	7.24
Molar Enthalpy of Vapor (Btu/lbmo	-74323.89		Molar Enthalpy of Vapor (Btu/lbmol	-74322.27		Molar Enthalpy of Vapor (Btu/lbmol	-42974.84		Molar Enthalpy of Vapor (Btu/lbmole	-99681.57
Molar Enthalpy of Liquid (Btu/lbmo	-97033.83		Molar Enthalpy of Liquid (Btu/lbmol	-97037.20		Molar Enthalpy of Liquid (Btu/lbmol	-56282.41		Molar Enthalpy of Liquid (Btu/lbmole	-118848.41
λ	22709.94		λ	22714.93		λ	13307.57		λ	19166.83
Cp (BTU/lbmol*F)	12.93		Cp (BTU/lbmol*F)	11.79		Cp (BTU/lbmol*F)	11.58		Cp (BTU/lbmol*F)	12.57
T1 (F)	450.00		T1 (F)	275.00		T1 (F)	250.00		T1 (F)	420.00
Tb (F)	387.14		Tb (F)	387.14		Tb (F)	183.83		Tb (F)	405.31
mv (lb)	0.18		mv (lb)	-1.10		mv (lb)	0.06		mv (lb)	0.07
η	0.02		η	0.02		η	0.02		η	0.02
ΔH_c (Btu/lb)	9276.43		ΔH_c (Btu/lb)	9276.4328		ΔH_c (Btu/lb)	15466.01		ΔH_c (Btu/lb)	5051.01
E_{TNT} (Btu/lb)	2014.62		E_{TNT} (Btu/lb)	2014.62		E_{TNT} (Btu/lb)	2014.62		E_{TNT} (Btu/lb)	2014.62
m_{TNT} (kg)	0.01		m_{TNT} (lb)	-0.046172		m_{TNT} (lb)	0.004126		m_{TNT} (lb)	0.0015863
m_{TNT} (g)	7.53									
r_1 (m)	10.00		r_1 (m)	10.00		r_1 (m)	10.00		r_1 (m)	10.00
z_e ($m/kg^{1/3}$)	442.42		z_e ($m/kg^{1/3}$)	-72.19423		z_e ($m/kg^{1/3}$)	807.9643		z_e ($m/kg^{1/3}$)	2101.2652
p_s	0.00		p_s	0.02		p_s	0.00		p_s	0.00
p_O (kPa)	0.38		p_O (kPa)	2.34		p_O (kPa)	0.21		p_O (kPa)	0.08

Large Unit Vapor Deflagration Analysis

Large Unit							
R-102		V-102		V-103		V-104	
Pressure (psia)	445.00	Pressure (psia)	441.00	Pressure (psia)	437.00	Pressure (psia)	437.00
Volume (ft3)	7.95	Volume (ft3)	21.99	Volume (ft3)	3.14	Volume (ft3)	9.42
R (ft^3*psi/R*lbmol)	10.73	R (ft3*psi/R*lbmol)	10.73	R (ft3*psi/R*lbmol)	10.73	R (ft3*psi/R*lbmol)	10.73
Temp (F)	450.00	Temp (F)	275.00	Temp (F)	250.00	Temp (F)	420.00
Temp (°R)	909.67	Temp (°R)	734.67	Temp (°R)	709.67	Temp (°R)	879.67
n (lbmol)	0.36	n (lbmol)	1.23	n (lbmol)	0.18	n (lbmol)	0.44
MW (lb/lbmol)	23.22	MW (lb/lbmol)	23.22	MW (lb/lbmol)	21.98	MW (lb/lbmol)	24.46
m (lb)	8.42	m (lb)	28.56	m (lb)	3.96	m (lb)	10.67
Molar Enthalpy of Vapor (Btu/lbmo	-70089.92	Molar Enthalpy of Vapor (Btu/lbmol	-70089.05	Molar Enthalpy of Vapor (Btu/lbmol	-42831.40	Molar Enthalpy of Vapor (Btu/lbmole	-99911.10
Molar Enthalpy of Liquid (Btu/lbmo	-91518.06	Molar Enthalpy of Liquid (Btu/lbmol	-91521.66	Molar Enthalpy of Liquid (Btu/lbmol	-55916.30	Molar Enthalpy of Liquid (Btu/lbmole	-119059.82
λ	21428.15	λ	21432.61	λ	13084.90	λ	19148.72
Cp (BTU/lbmol*F)	12.73	Cp (BTU/lbmol*F)	11.57	Cp (BTU/lbmol*F)	11.37	Cp (BTU/lbmol*F)	12.31
T1 (F)	450.00	T1 (F)	275.00	T1 (F)	250.00	T1 (F)	420.00
Tb (F)	383.59	Tb (F)	383.60	Tb (F)	185.57	Tb (F)	406.08
mv (lb)	0.33	mv (lb)	-1.67	mv (lb)	0.22	mv (lb)	0.10
η	0.02	η	0.02	η	0.02	η	0.02
ΔH_c (Btu/lb)	10066.33	ΔH_c (Btu/lb)	10066.332	ΔH_c (Btu/lb)	16030.14	ΔH_c (Btu/lb)	4714.55
E_{TNT} (Btu/lb)	2014.62	E_{TNT} (Btu/lb)	2014.62	E_{TNT} (Btu/lb)	2014.62	E_{TNT} (Btu/lb)	2014.62
m_{TNT} (kg)	0.02	m_{TNT} (lb)	-0.075988	m_{TNT} (kg)	0.016029	m_{TNT} (lb)	0.0020284
m_{TNT} (g)				m_{TNT} (g)	16.02935		
r_1 (m)	10.00	r_1 (m)	10.00	r_1 (m)	10.00	r_1 (m)	10.00
z_e (m/kg$^{1/3}$)	221.22	z_e (m/kg$^{1/3}$)	-43.86662	z_e (m/kg$^{1/3}$)	207.9519	z_e (m/kg$^{1/3}$)	1643.2973
p_s	0.01	p_s	0.04	p_s	0.01	p_s	0.00
p_O (kPa)	0.76	p_O (kPa)	3.88	p_O (kPa)	0.81	p_O (kPa)	0.10

Small Unit Revenue

Small Unit		
Separations		
LPG	41.94	lb/hr
Naphtha	10.45	lb/hr
HIU		
Naphtha	73.27	lb/hr
Diesel	132.78	lb/hr

Revenue of Small Unit						
	Flow (lb/hr)	Density (lb/ft^3)	Vol. Flow (ft^3/hr)	Vol. Flow (bbl/hr)	Vol. Flow (gal/hr)	Revenue ($/hr)
LPG	0.21	13.3	0.016	0.0028	0.1183	12.58
Naphtha	83.72	45.0	1.860	0.3313	13.9164	24.85
Diesel	132.78	53.0	2.505	0.4462	18.7415	40.16
Total Revenue						77.59

Medium Unit Revenue

Medium Unit		
Separations		
LPG	366.8	lb/hr
Naphtha	55.84	lb/hr
HIU		
Naphtha	515.07	lb/hr
Diesel	801.29	lb/hr

Revenue of Medium Unit						
	Flow (lb/hr)	Density (lb/ft^3)	Vol. Flow (ft^3/hr)	Vol. Flow (bbl/hr)	Vol. Flow (gal/hr)	Revenue ($/hr)
LPG	2.53	16.6	0.152	0.0272	1.1406	110.04
Naphtha	570.91	45.0	12.687	2.2596	94.9049	169.47
Diesel	801.29	53.0	15.119	2.6927	113.0953	242.35
Total Revenue						521.86

Large Unit Revenue

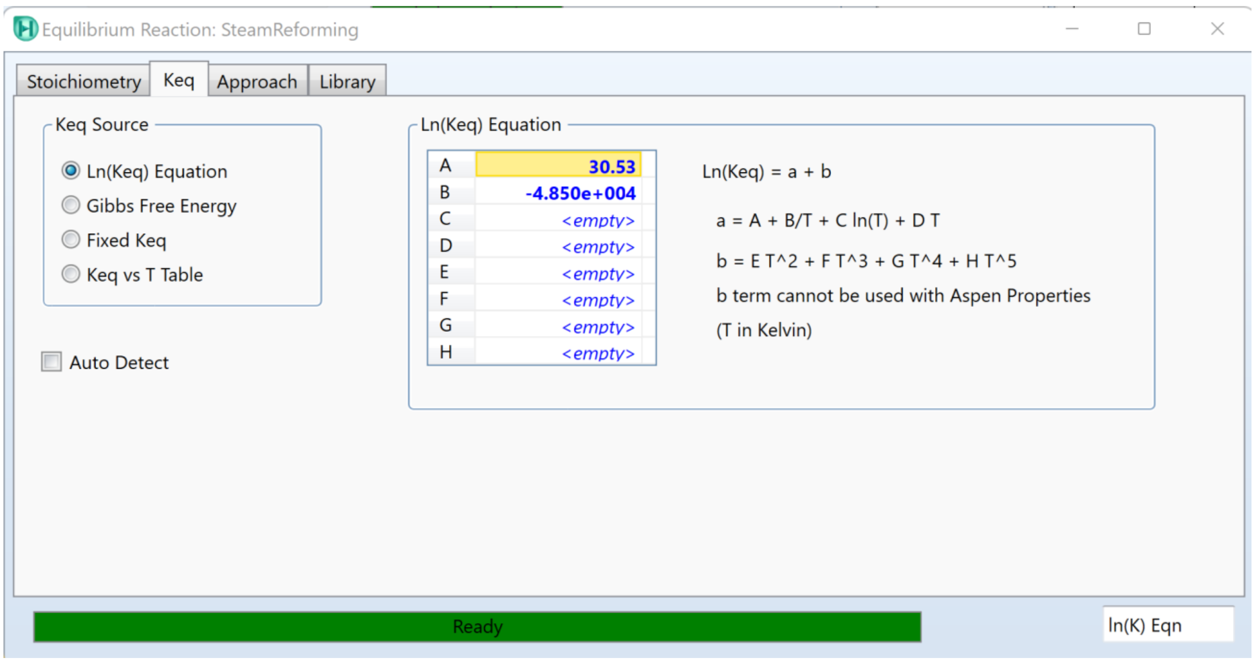
Large Unit		
Separations		
LPG	467.6	lb/hr
Naphtha	106.5	lb/hr
HIU		
Naphtha	791.58	lb/hr
Diesel	1385.45	lb/hr

Revenue of Large Unit						
	Flow (lb/hr)	Density (lb/ft^3)	Vol. Flow (ft^3/hr)	Vol. Flow (bbl/hr)	Vol. Flow (gal/hr)	Revenue ($/hr)
LPG	2.52	13.9	0.181	0.0322	1.3537	140.28
Naphtha	898.08	45.0	19.957	3.5546	149.2913	266.59
Diesel	1385.45	53.0	26.140	4.6558	195.5444	419.02
Total Revenue						685.62

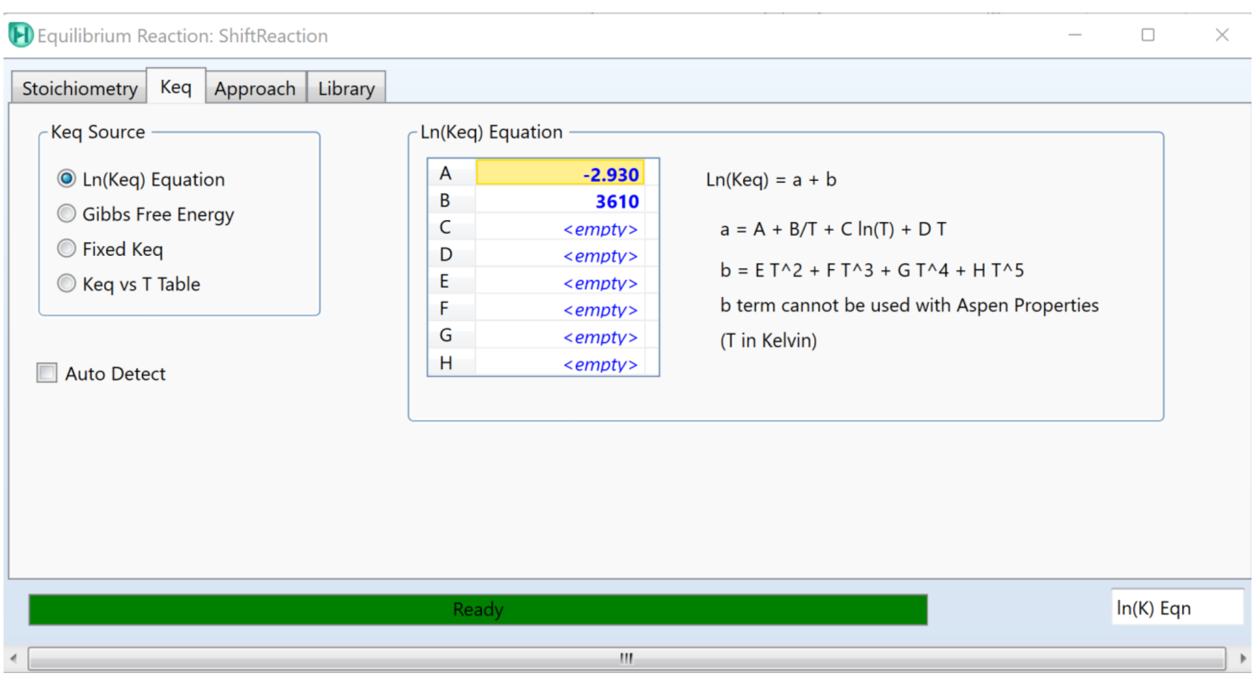
Wastewater Treatment Cost

Waste Water Treatment Cost					
	(lb/hr)	(gal/hr)	(kgal/hr)	$/hr	
small	1038.40	124.5088	0.124509	0.747053	5235.346
medium	5729.20	686.9542	0.686954	4.121725	28885.05
large	10606.03	1271.707	1.271707	7.63024	53472.73

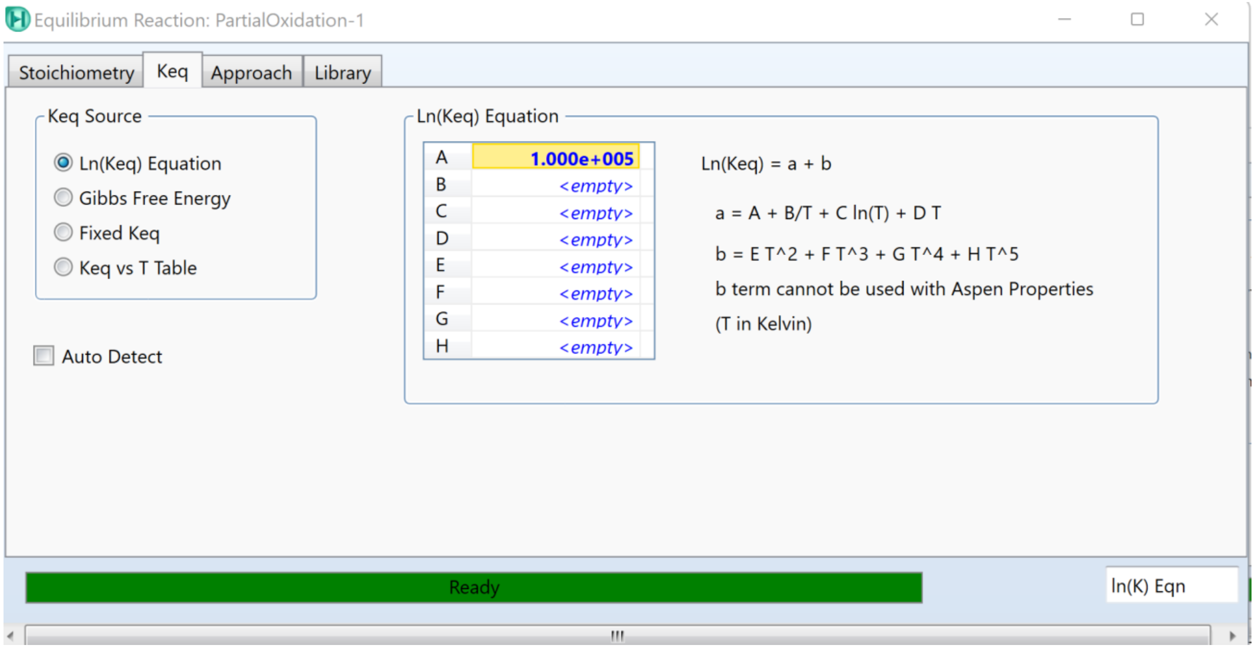
Syngas Steam Reforming Equilibrium Equation



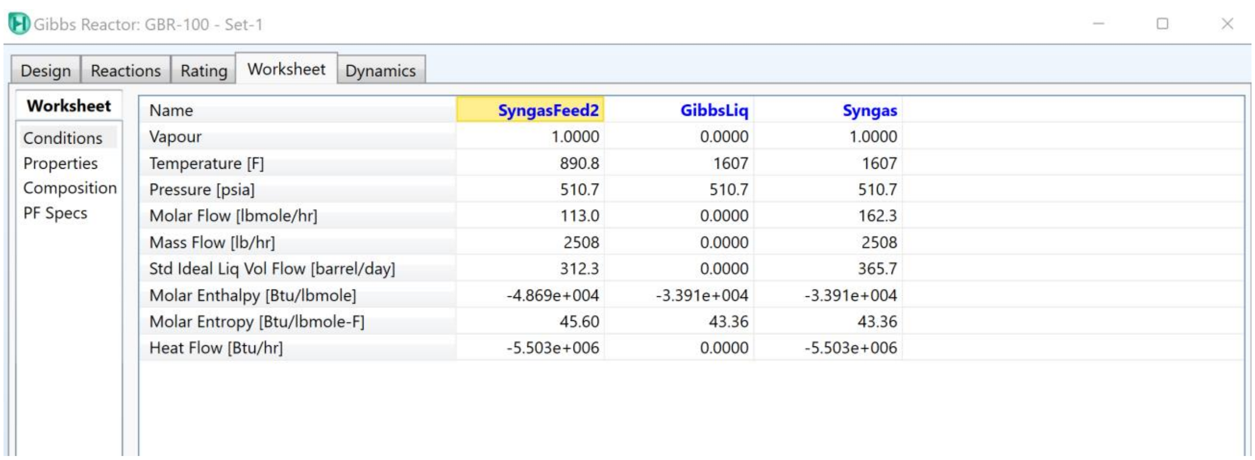
Syngas Water Gas Shift Equilibrium Equation



Syngas Partial Oxidation Equilibrium Equation



Small Unit Syngas Gibbs Reactor



Name	SyngasFeed2	GibbsLiq	Syngas
Vapour	1.0000	0.0000	1.0000
Temperature [F]	890.8	1607	1607
Pressure [psia]	510.7	510.7	510.7
Molar Flow [lbmole/hr]	113.0	0.0000	162.3
Mass Flow [lb/hr]	2508	0.0000	2508
Std Ideal Liq Vol Flow [barrel/day]	312.3	0.0000	365.7
Molar Enthalpy [Btu/lbmole]	-4.869e+004	-3.391e+004	-3.391e+004
Molar Entropy [Btu/lbmole-F]	45.60	43.36	43.36
Heat Flow [Btu/hr]	-5.503e+006	0.0000	-5.503e+006

Medium Unit Syngas Gibbs Reactor

Gibbs Reactor: GBR-100 - Set-1

Design | Reactions | Rating | Worksheet | Dynamics

Worksheet: Conditions, Properties, Composition, PF Specs

Name	SyngasFeed2	GibbsLiq	Syngas
Vapour	1.0000	0.0000	1.0000
Temperature [F]	870.4	1699	1699
Pressure [psia]	510.7	510.7	510.7
Molar Flow [lbmole/hr]	575.2	0.0000	870.3
Mass Flow [lb/hr]	1.274e+004	0.0000	1.274e+004
Std Ideal Liq Vol Flow [barrel/day]	1564	0.0000	1884
Molar Enthalpy [Btu/lbmole]	-4.507e+004	-2.979e+004	-2.979e+004
Molar Entropy [Btu/lbmole-F]	45.09	42.37	42.37
Heat Flow [Btu/hr]	-2.592e+007	0.0000	-2.592e+007

Large Unit Syngas Gibbs Reactor

Gibbs Reactor: GBR-100 - Set-1

Design | Reactions | Rating | Worksheet | Dynamics

Worksheet: Conditions, Properties, Composition, PF Specs

Name	SyngasFeed2	GibbsLiq	Syngas
Vapour	1.0000	0.0000	1.0000
Temperature [F]	886.2	1627	1627
Pressure [psia]	510.7	510.7	510.7
Molar Flow [lbmole/hr]	1134	0.0000	1649
Mass Flow [lb/hr]	2.515e+004	0.0000	2.515e+004
Std Ideal Liq Vol Flow [barrel/day]	3122	0.0000	3681
Molar Enthalpy [Btu/lbmole]	-4.781e+004	-3.288e+004	-3.288e+004
Molar Entropy [Btu/lbmole-F]	45.48	43.12	43.12
Heat Flow [Btu/hr]	-5.423e+007	0.0000	-5.423e+007

Small Unit Syngas Separator

Separator: V-101

Design | Reactions | Rating | Worksheet | Dynamics

Worksheet: Conditions, Properties, Composition, PF Specs

Name	CooledSyngas	Water	Syngaswowater
Vapour	0.8347	0.0000	1.0000
Temperature [F]	**212.0**	212.0	212.0
Pressure [psia]	506.7	506.7	506.7
Molar Flow [lbmole/hr]	162.3	26.83	135.5
Mass Flow [lb/hr]	2508	483.6	2025
Std Ideal Liq Vol Flow [barrel/day]	365.7	33.20	332.5
Molar Enthalpy [Btu/lbmole]	-4.919e+004	-1.205e+005	-3.506e+004
Molar Entropy [Btu/lbmole-F]	29.72	17.02	32.24
Heat Flow [Btu/hr]	-7.983e+006	-3.233e+006	-4.750e+006

Medium Unit Syngas Separator

Separator: V-101

Design | Reactions | Rating | Worksheet | Dynamics

Worksheet

Conditions
Properties
Composition
PF Specs

Name	CooledSyngas	Water	Syngaswowater
Vapour	0.8441	0.0000	1.0000
Temperature [F]	212.0	212.0	212.0
Pressure [psia]	506.7	506.7	506.7
Molar Flow [lbmole/hr]	870.3	135.6	734.7
Mass Flow [lb/hr]	1.274e+004	2445	1.030e+004
Std Ideal Liq Vol Flow [barrel/day]	1884	167.8	1716
Molar Enthalpy [Btu/lbmole]	-4.517e+004	-1.205e+005	-3.125e+004
Molar Entropy [Btu/lbmole-F]	28.97	17.02	31.17
Heat Flow [Btu/hr]	-3.931e+007	-1.635e+007	-2.296e+007

Large Unit Syngas Separator

Separator: V-101

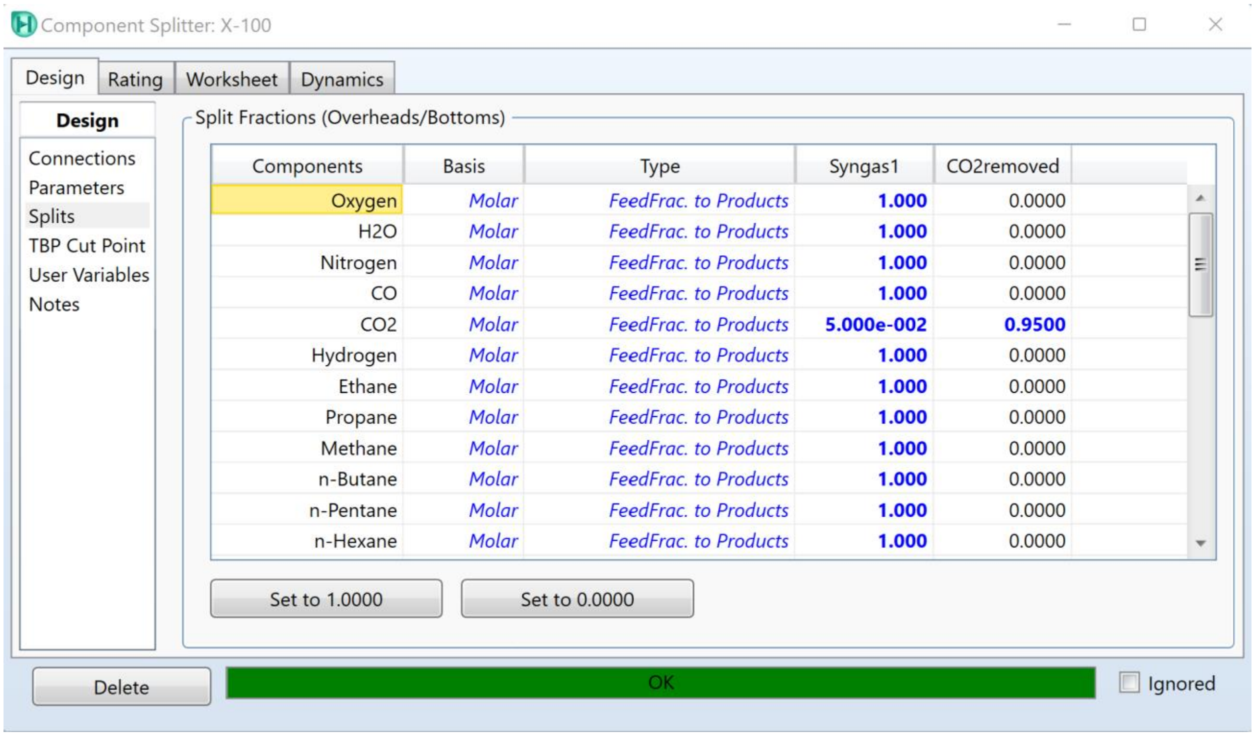
Design | Reactions | Rating | Worksheet | Dynamics

Worksheet

Conditions
Properties
Composition
PF Specs

Name	CooledSyngas	Water	Syngaswowater
Vapour	0.8371	0.0000	1.0000
Temperature [F]	212.0	212.0	212.0
Pressure [psia]	506.7	506.7	506.7
Molar Flow [lbmole/hr]	1649	268.7	1381
Mass Flow [lb/hr]	2.515e+004	4844	2.031e+004
Std Ideal Liq Vol Flow [barrel/day]	3681	332.5	3348
Molar Enthalpy [Btu/lbmole]	-4.817e+004	-1.205e+005	-3.409e+004
Molar Entropy [Btu/lbmole-F]	29.55	17.02	31.98
Heat Flow [Btu/hr]	-7.945e+007	-3.238e+007	-4.707e+007

Syngas CO_2 Splitter



Components	Basis	Type	Syngas1	CO2removed
Oxygen	Molar	FeedFrac. to Products	1.000	0.0000
H2O	Molar	FeedFrac. to Products	1.000	0.0000
Nitrogen	Molar	FeedFrac. to Products	1.000	0.0000
CO	Molar	FeedFrac. to Products	1.000	0.0000
CO2	Molar	FeedFrac. to Products	5.000e-002	0.9500
Hydrogen	Molar	FeedFrac. to Products	1.000	0.0000
Ethane	Molar	FeedFrac. to Products	1.000	0.0000
Propane	Molar	FeedFrac. to Products	1.000	0.0000
Methane	Molar	FeedFrac. to Products	1.000	0.0000
n-Butane	Molar	FeedFrac. to Products	1.000	0.0000
n-Pentane	Molar	FeedFrac. to Products	1.000	0.0000
n-Hexane	Molar	FeedFrac. to Products	1.000	0.0000