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Dear AIChE,

Enclosed is the design proposal for the 2022 AIChE Design Competition.

This report details the engineering assumptions and techniques used to complete the project, as well as the results and recommendations.

Thank you for this opportunity!

Sincerely,

Andrew Coffey, Trace Francis, April Johnsen, Genna Orr

CHE 4224

Spring 2022

AIChE Design Project

April 11, 2022

Group Number

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Table of Contents

Executive Summary

The goal of this project was to design modular Gas-To-Liquids (GTL) processing plants to be deployed at a network of wellheads, which can convert stranded natural gas resources into liquid material. The reasoning for this project was to reduce greenhouse gas emissions at wellheads by avoiding venting/flaring methane to the atmosphere. The task of the project was to provide the preliminary design package for a Fischer-Tropsch Reaction unit (FTR) and separation facilities, as well as optimizing the deployment of the GTL plant modules into the field. The wellhead network was located around a central plant location where units could be initially deployed from as well as a location to transport product to for further refining. The modules in this design were sized based on feed capacities of 500 MSCF/day, 2,500 MSCF/day, and 5,000 MSCF/day respectively. The central plant had a maximum capacity of 30,000 MSCF/day of natural gas feed, requiring optimization of which wells could be in service at a given time. Permits will be requested at the beginning of 2023, and after approval at six months, the equipment can undergo construction. Production is planned to begin at the start of 2024.

The economic analysis of this project was evaluated over a 20-year period with a 7-year straight line depreciation. The company's hurdle rate was 8%, and an inflation of 3% per year was assumed. After the preliminary design for sixteen well sites, with a starting natural gas feed ranging from 2,870-14,700 MSCF/day, the net present value was determined to be $2.29B. The total capital investment is estimated to be $1.33B with a payback period of 3.5 years. This project requires two large units, seven medium units, and twelve small units totaling to 21 module units to be constructed. The project requires the employment of 154 operators. Due to the central plant's capacity and the varied service of wells in any given year, the combined production of these active wells processed on average 21,800 MSCF/day of natural gas each year, producing on average 1.78 MMbbl/year of naphtha and 5.22 MMbbl/year of diesel.

An inherently safer design was implemented after assessing the potential hazards. Since this design not only overcomes the hurdle rate and is designed to reduce major safety risks the final recommendation is to move forward with this project.

Introduction

AIChE has identified a potential area of growth in the chemical engineering industry. Natural gas is often flared or vented at isolated well sites which leads to higher carbon dioxide (CO_2) emissions, as opposed to being transported or further processed into usable product. The goal of this project was to investigate ways to convert stranded natural gas resources into liquid petroleum gas (LPG), naphtha, and diesel products through the use of modular GTL plants. This paper details the engineering and supply chain decisions that were made to consider this problem. *Figure 1: Block Flow Diagram* illustrates the generalized process.

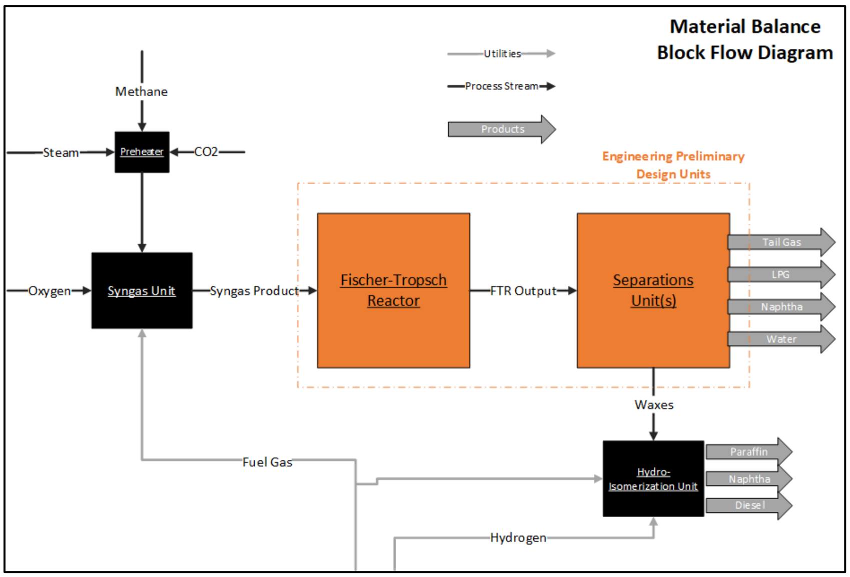

Figure 1: Block Flow Diagram

The GTL process will utilize steam methane reforming (SMR) to make syngas (CO/H_2 mixture), then the product will be fed to a Fischer-Tropsch Reaction unit (FTR). The reactor will utilize Fischer-Tropsch (FT) synthesis to create higher chain carbon molecules to be sold as LPG, naphtha, and diesel. A separations unit will follow the FTR to purify product streams. A hydro-isomerization unit (HIU) will collect the heavy bottoms product from the main fractionation column to form additional naphtha product, as well as diesel and waxes to further process in a downstream processing plant. The design statement calls for a detailed design of the FTR and separations facilities. Due to the nature of the units being modular and having various sizes, these units were considered at three different operating capacities based on the feed of methane: 5,000 MSCF/day, 2,500 MSCF/day, and 500 MSCF/day. This allowed for a variation of module sizes to be used at any given site based on need.

The GTL plant has four major feeds: methane, steam, carbon dioxide, and oxygen. The flow rates of oxygen, carbon dioxide, and steam depend on the modules maximum methane feed. The first stream is methane which comes directly from the wellhead at 500 psig and 100 ^oF. Carbon dioxide and high pressure (HP) steam, at those same conditions, are purchased for feeds to the syngas unit. The three feeds are mixed and must be heated before being sent into the syngas unit. The preheater is an electric furnace, heating the combined feed to 1,000 °F. The syngas unit is upstream of the FTR unit immediately following the feed preheater. This process converts methane feedstock into carbon monoxide and hydrogen in a ratio of two moles of hydrogen to one mole of carbon monoxide. This ratio will lead to the desired distribution of alkane products. Carbon monoxide and hydrogen will function as the feed to the Fischer-Tropsch reactor. The syngas unit can operate between a pressure of 300-500 psig and at a temperature of 1,600-1,950 ^oF. When simulating the syngas unit, the lower bound of the reactor temperature range was the target, as it reduced the amount of duty on the preheater. The two products are formed through three simultaneous reactions: steam reforming, partial oxidation, and the shift reaction. While the steam reforming and the shift reaction will end in equilibrium, the partial oxidation will go to completion. These reactions are listed in *Table 1: Syngas Unit*.

Table 1: Syngas Unit

Steam Reforming:	$CH_4 + H_2O \leftrightarrow CO + 3H_2$
Partial Oxidation:	$CH_4 + \frac{3}{2}O_2 \longrightarrow CO + 2H_2O$
Shift Reaction:	$CO + H_2O \leftrightarrow CO_2 + H_2$

The steam reforming reaction serves as the primary method for converting methane into carbon monoxide and hydrogen. This is an equilibrium reaction and will not completely consume the methane feed. The partial oxidation reaction helps to convert more of the methane. Finally, the shift reaction functions to balance out the products into the desired 2:1 ratio.

The first part of this project was the design and optimization of the FTR based on the syngas product. Following the FTR design, was the detailed design of the separation units. It was determined that a 3-phase separator would follow the FTR to reduce flow rates to the distillation column by removing water. The distillation column separates the C10 and lighter material from the C11+ material. C11+ material is sent to the HIU which does not need to be designed for this project. Products from the HIU were calculated given the equations from the design statement.

The second part of the project includes supply chain analysis and the optimized deployment of the modules based on plant capacities and well productions. Given that the well production rates decline over time, unit deployment was designed to provide the maximum product to the central plant.

Discussion

When beginning the preliminary stages of the FTR design, it was determined that Aspen HYSYS would be utilized to simulate the processes. The fluid package utilized throughout the simulations was Peng-Robinson. Using the specifications for the feed conditions, material streams were created and fed into an exchanger at a desired outlet temperature of 1,000 °F. The type of steam chosen for the feed was HP steam because of its high temperature allowing for less duty in the syngas preheater. The feed was then fed to an equilibrium reactor which is simulating the syngas reactor. The equations from *Table 1: Syngas Unit* were inserted into the reactor with the partial oxidation reaction having an equilibrium constant set to 1.0 MM to simulate a conversion reaction. An oxygen stream with 99% purity was fed to the reactor as a reactant for the partial oxidation reaction to occur. The reactor was then optimized by varying the inlet flow rates of steam, CO_2, and oxygen to help meet FTR reactant specifications. Varying the oxygen stream flow rate affected the temperature of the reactor since it was the exothermic reaction, so once a temperature was reached in the range of 1,600-1,950°F, the other two reactant stream (CO_2 and HP steam) flow rates were varied to help reach a product specification of 1.0 mole fraction CO to 2.0 mole fraction H_2.

Following the reactor was a 2-phase separator which acted to separate excess water from the product stream. The rational for implementing the drum was to reduce the overall flow rate downstream which will allow for a smaller-volume FTR and absorption column. After the separator, an absorption column was used with a water solvent to remove excess CO_2. Similar to the rational for the separator placement, the column was implemented to help reduce the volume of flow which will eventually lead to the FTR and to also purify the feed stream.

The Fischer-Tropsch reactor was designed and costed as a jacketed, non-agitated packed bed reactor. The goal of the reactor was to utilize the 2:1 H_2/CO feed to create the desired distribution of LPG's, naphtha, and diesel. This was done using the Anderson-Schulz-Flory distribution which is effective from 390$^oF-450^o$F. The higher end of the temperature distribution led to heavier carbon chain material (diesel), which is more valuable. The reactor was then designed to maintain the temperature at roughly 450^{o}F. The reaction is highly exothermic which made it necessary to have the reactor jacketed to maintain the temperature of the reactor. This led to the need for a large amount of steam as coolant to maintain that temperature. Pressure drop is also a large factor in the reaction since the vapor flow becomes a two-phase flow, which greatly decreases the conversion of the reaction. The overall pressure drop was reduced by increasing the number of tubes within the reactor while decreasing the tube diameter.

The reactor was modelled using differential equations paired with auxiliary equations. Polymath was considered for modelling the FTR, but it struggled with troubleshooting errors. It was then modelled within MATLAB due to its superior troubleshooting and computational abilities. This was done using a function file which contained all the reactor's differential equations and auxiliary equations. The function file was adjusted depending on the conditions of the reactor and the size of the modular unit's inlet flow. The script file was used to specify the initial conditions and to run the function file. In order to graph the resulting conversion, pressure drop, and temperature changes within the reactor, the function file was used. The script and function file can be seen in the appendix.

Following the FTR is a 3-phase separator which splits water from the hydrocarbon streams to help reduce the size of downstream units. Separators are typically much cheaper compared to other units, so the decision to implement these before larger facilities aims to reduce flow and therefore reducing the size of the more expensive units. The light liquid stream from the separator contains most of the C5-C30 material, which can be separated into the high-value products. This stream was then fed to a distillation column which was operating at atmospheric pressure. The rational for the column was to separate the C11+ material from the C10 and lighter material. The column has a partial condenser to separate tail gas from the naphtha stream. The bottoms products are then fed to the HIU to be separated into diesel and wax products. The amount of naphtha in the HIU feed was effectively negligible so it was included into the diesel stream.

Conclusions

A modular gas plant design is a viable option to produce diesel and naphtha from methane coming from hard-to-reach wellheads. Preliminary design of the units focused on the FTR and separation sections. Supply chain and network analysis were also a large portion of the design. Costing transportation of product trucks as well as module movement is a large expense, but maintenance and repair were the biggest factor to loss of revenue on a yearly basis. Once the preliminary design for all sixteen well sites was completed, it was determined that the project requires two large units, seven medium units, and twelve small units for a total of 21 modules to be built. Due to the central plant's capacity, the number of wells in service at any given year varies. The combined production of these active wells processed on average 21,800 MSCF/day of natural gas each year, producing on average 1.78 MMbbl/year of naphtha and 5.22 MMbbl/year of diesel. After conducting a net present value analysis, the NPV was determined to be $2.29B. The total capital investment was estimated to be $1.33B with a payback period of 3.5 years. In addition to being economically attractive, this project is also a relatively safe design as it does not require exotic materials, most of the components are compatible with one another and the overall design is simplified to not include unnecessary equipment. Finally, this design is better for the environment since it is turning methane into products rather than venting to the atmosphere or including continuous flaring.

Recommendations

Overall, it is recommended to move forward with this project. Given the positive environmental impact it has when compared to traditional methods, as well as the final NPV, the project is attractive in all aspects. When considering the volatility of the oil and gas market, the sensitivity analysis examined the worst-case scenario for oil prices and the project still yielded a positive NPV. When moving forwarded with the project, it is recommended to reexamine product specifications. For this project, the product specifications were left loose and often times unreasonable. In conclusion, verifying product specifications is recommended.

Another recommendation is to optimize the number of trucks being used to deploy the modules. It is recommended that the spacing for the modules within the containers is maximized to reduce the total number of containers being deployed.

Since the last three years of production are not profitable, another recommendation is to decommission the units at those wells and salvage material early.

Project Premise

The project was premised on the idea of investigating ways to reduce CO_2 emissions from well sites. When natural gas resources are stranded too far from conventional gathering pipeline systems, it cannot be collected along with crude oil. This results in the gas being burned using an on-site flare. The alternative to these methods of getting rid of natural gas is to implement a modular GTL processing plant among a network of producing wells. This allows for natural gas to be converted into longer-carbon-chain liquids to be transported and further processed at a refinery. Successful design and implementation of a network eliminates the need for methane to be vented or flared at stranded well sites and allows for more non-renewable resources to be utilized in other markets.

Heat and Material Balance

Table 2: Mass Balance for Medium Unit, Table 3: Mole Balance for Medium Unit, and Table 4: Energy Balance for Medium Unit present the mass, mole, and energy balances for the medium unit. The 1% difference in the mass balance is likely due to significant figure errors from entering values back and forth from several platforms (Aspen HYSY, Excel, MATLAB, and PowerPoint). There is a slight loss of heat from the system, but it is shown as 0% because it is negligible.

Table 2: Mass Balance for Medium Unit

Medium		
	Stream In (lb/hr)	Stream Out (lb/hr)
Methane	6594	
Carbon Dioxide	6601	
HP Steam	6950	
Oxygen	6392	
Solvent	3603000	
Process Water Return		3630000
Naphtha		6538
Diesel		18880
Wax		9253
Sums	3629537	3664671
Percent Difference %	1%	

Table 3: Mole Balance for Medium Unit

Medium		
	Stream In (lbmole/hr)	Stream Out (lbmole/hr)
Methane	411	
Carbon Dioxide	150	
HP Steam	385.8	
Oxygen	200	
Solvent	200000	
Process Water Return		201100
Naphtha		62.43
Diesel		94.66
Wax		26.59
Sums	201146.8	201283.68
Percent Difference %	0%	

Table 4: Energy Balance for Medium Unit

Medium		
	Heat In (Btu/hr)	Heat Out (Btu/hr)
Stream		
Methane	13260000	
Carbon Dioxide	25460000	
HP Steam	39120000	
Oxygen	31030	
Feed Preheater	8078000	
Condenser		28220000
Solvent	24590000000	
FTR Preheater	2671000	
Tower Condenser		1761000
Tower Reboiler	5530000	
Isom Reboiler	4883000	
Naphtha		5965000
Process Water Return		24720000000
Isom Condenser		4025000
Wax		3249000
Diesel		12450000
Sums	24689033030	24775670000
Percent Difference %	0%	

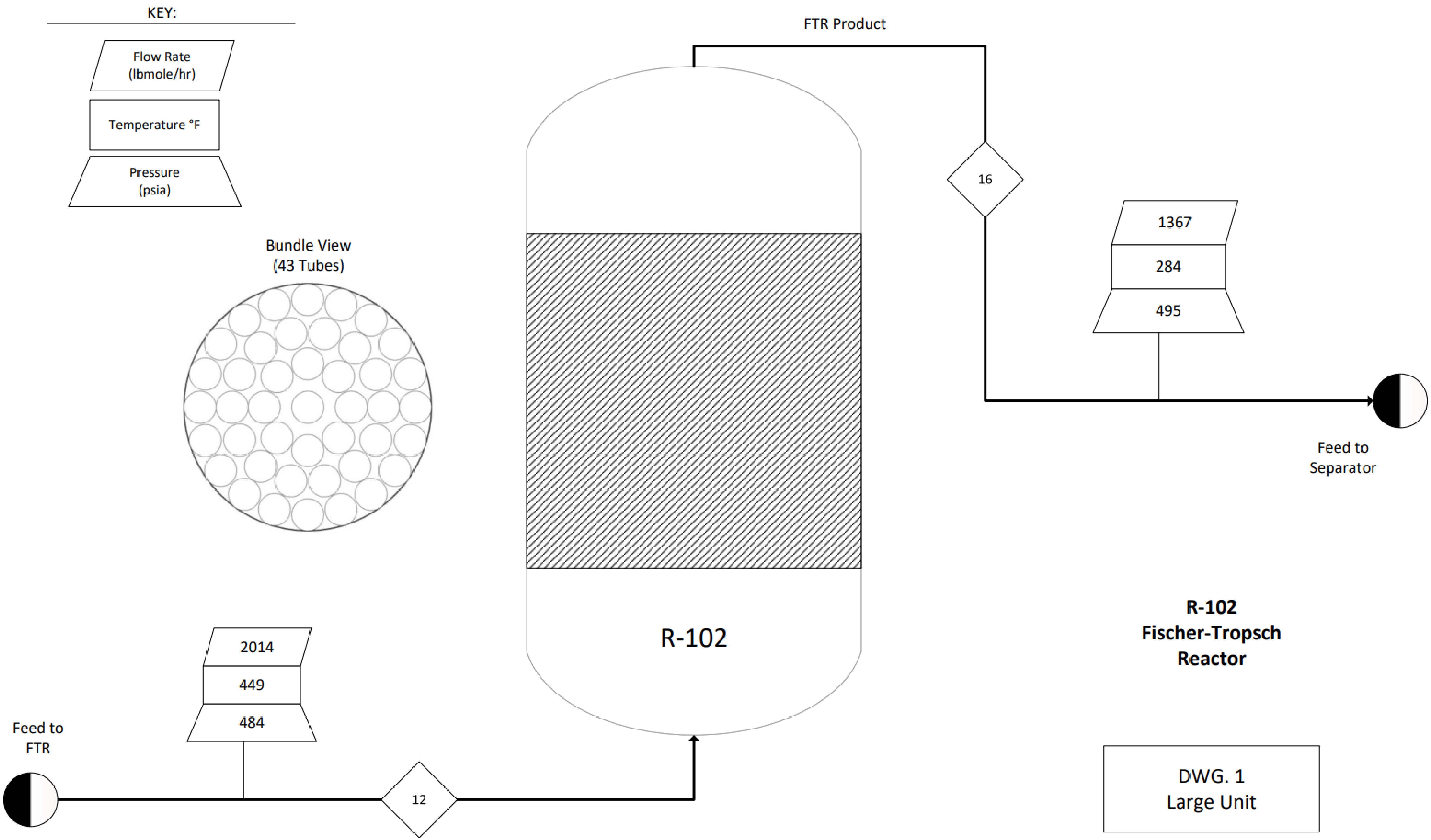

Figure 2: FTR Unit Process Flow Diagram

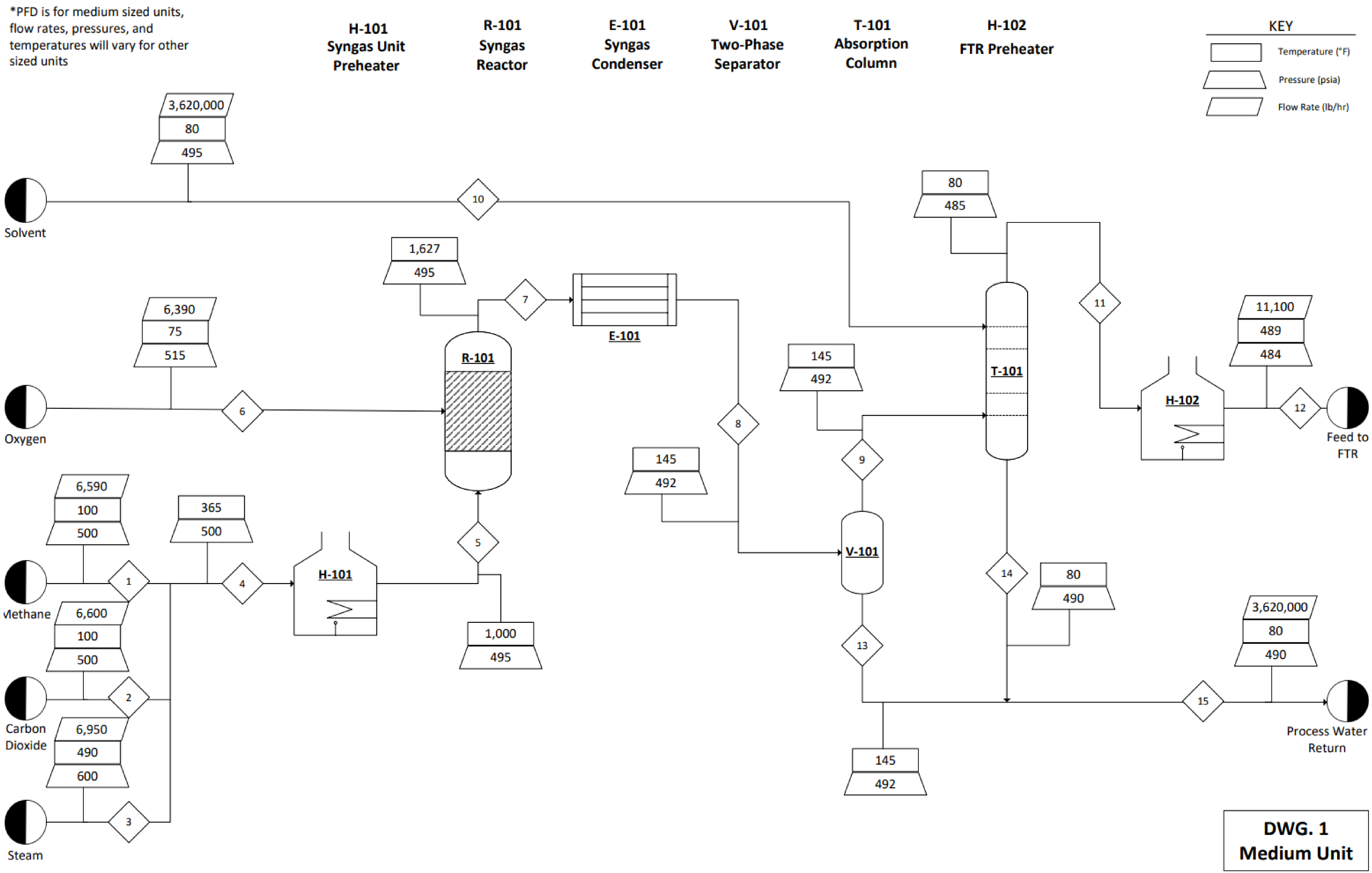

Figure 3: PFD for Medium GTL Plant Process (DWG. 1)

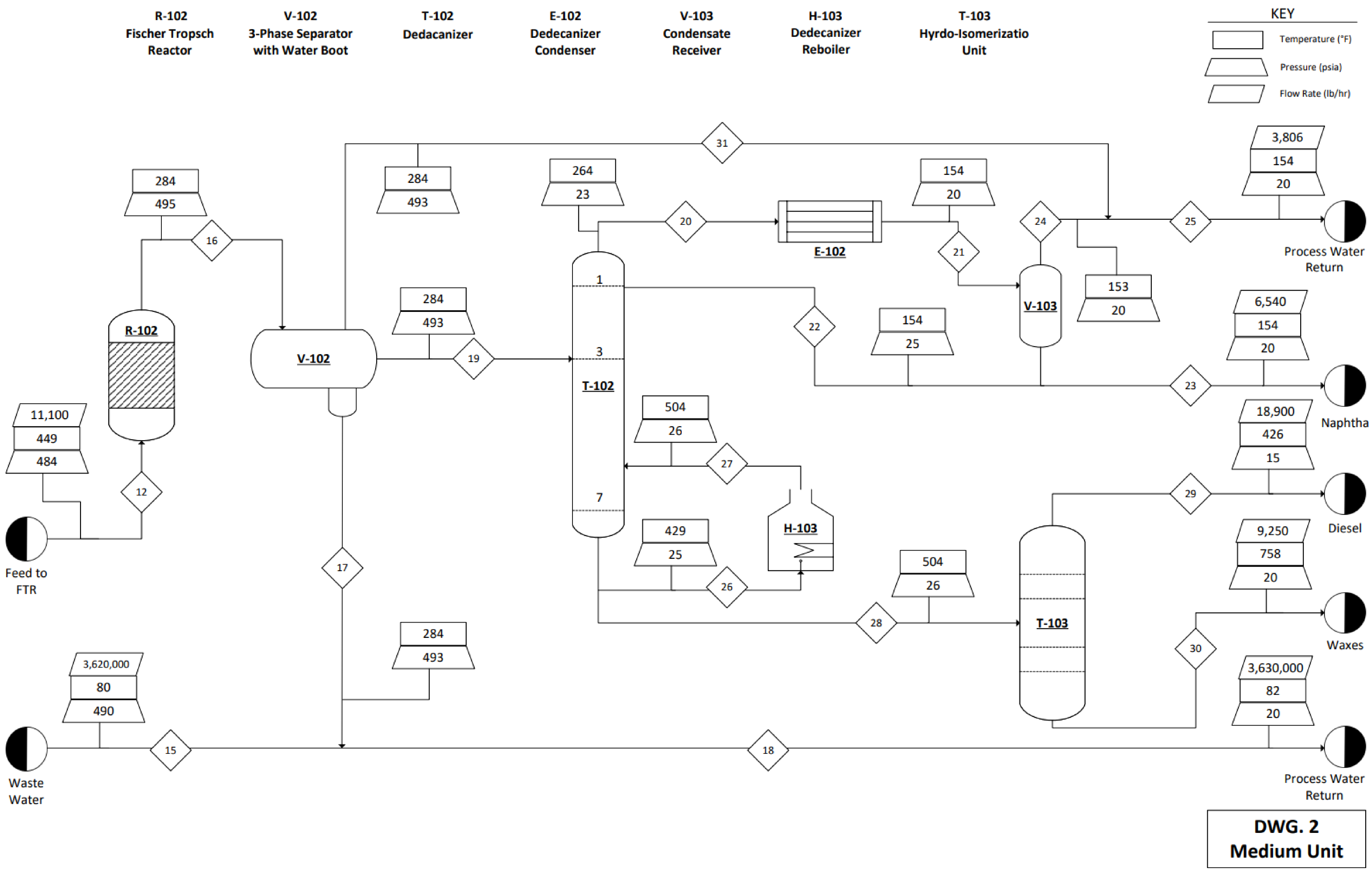

Figure 4: PFD for Medium GTL Plant Process (DWG. 2)

Table 5: Medium Unit PFD Stream Table

Medium Unit		1 Methane	2 Carbon Dioxide	3 HP Steam	4 Syngas Feed Mixture	5 Hot Syngas Feed	6 99% Oxygen	7 Syngas Product	8 Condensed Syngas	9 Syngas Product Water Removed	10 Absorber Solvent
Vapour Fraction		1.00	1.00	1.00	0.91	1.00	1.00	1.00	0.74	1.00	0.00
Temperature	F	100	100	490	365	1,000	75	1,627	145	145	80
Pressure	psia	500	500	600	500	495	515	495	492	492	495
Molar Flow	lbmole/hr	411	150	386	947	947	200	1,608	1,608	1,188	200,000
Mass Flow	lb/hr	6594	6601	6950	20145	20145	6392	26537	26537	18963	3603020
Liquid Volume Flow	barrel/day	1508	548	477	2533	2533	386	3418	3418	2898	247207
Mass Enthalpy	Btu/lb	-2010	-3857	-5629	-3864	-3463	-5	-2630	-3693	-2472	-6826
Component Fraction											
Hydrogen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	6.16E+02	6.16E+02	6.16E+02	0.00E+00
Water	lbmole/hr	0.00E+00	0.00E+00	3.86E+02	3.86E+02	3.86E+02	0.00E+00	4.29E+02	4.29E+02	9.76E+00	2.00E+05
Carbon Monoxide	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.07E+02	3.07E+02	3.07E+02	0.00E+00
Nitrogen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.00E+00	2.00E+00	2.00E+00	2.00E+00	0.00E+00
Oxygen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.98E+02	2.15E-01	2.15E-01	2.15E-01	0.00E+00
Carbon Dioxide	lbmole/hr	0.00E+00	1.50E+02	0.00E+00	1.50E+02	1.50E+02	0.00E+00	1.73E+02	1.73E+02	1.72E+02	0.00E+00
Methane	lbmole/hr	4.11E+02	0.00E+00	0.00E+00	4.11E+02	4.11E+02	0.00E+00	8.15E+01	8.15E+01	8.15E+01	0.00E+00
Ethane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Propane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Butane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Pentane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Hexane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Heptane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Octane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Nonane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Decane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C11	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C12	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C13	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C14	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C15	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C16	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C17	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C18	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C19	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C20	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C21	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C22	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C23	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C24	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C25	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C26	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C27	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C28	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C29	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C30	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Actual Volume Flow	barrel/day	19807	6307	23783	62616	127992	9244	312719	67376	66847	244975
Mass Density	lb/ft3	1.42	4.47	1.25	1.38	0.673	2.96	0.363	1.68	1.21	62.87
Molar Density	lbmole/ft3	0.0887	0.1017	0.0693	0.0646	0.0316	0.0925	0.0220	0.1020	0.0760	3.49

Medium Unit		11 Syngas Product CO2 Removed	12 FTR Feed	13 Water Removed	14 CO2 Removed	15 Combined Water	16 FTR Product	17 Separator Water Draw	18 Process Water Return	19 Column Feed	20 Column OVHD
Vapour Fraction		1.00	1.00	0.00	0.00	0.00	0.24	0.00	0.00	0.00	1
Temperature	F	80	449	145	80	80	284	284	82	284	264
Pressure	psia	485	484	492	490	490	495	493	20	493	23
Molar Flow	lbmole/hr	1,000	1,000	420	200,188	200,608	682	270	201,106	245	139
Mass Flow	lb/hr	11132	11132	7574	3610851	3618425	46607	4856	3630359	37944	11399
Liquid Volume Flow	barrel/day	2242	2242	520	247863	248383	4423	333	249681	3503	1181
Mass Enthalpy	Btu/lb	-1547	-1307	-6751	-6820	-6819	-1463	-6612	-6808	-785	-804
Component Fraction											
Hydrogen	lbmole/hr	6.11E+02	6.11E+02	2.21E-02	4.48E+00	4.50E+00	3.60E+01	2.60E-02	4.05E+01	2.32E+00	1.67E-02
Water	lbmole/hr	1.29E+00	1.29E+00	4.19E+02	2.00E+05	2.00E+05	2.95E+02	2.70E+02	2.01E+05	5.95E+00	4.28E-02
Carbon Monoxide	lbmole/hr	3.04E+02	3.04E+02	1.15E-02	2.91E+00	2.92E+00	1.80E+01	1.01E-02	2.09E+01	1.68E+00	1.21E-02
Nitrogen	lbmole/hr	1.85E+00	1.85E+00	2.00E-04	1.47E-01	1.47E-01	1.81E+00	8.92E-04	1.96E+00	1.60E-01	1.15E-03
Oxygen	lbmole/hr	2.11E-01	2.11E-01	1.67E-05	4.04E-03	4.05E-03	7.61E-04	8.53E-07	4.81E-03	9.49E-05	6.82E-07
Carbon Dioxide	lbmole/hr	4.94E-03	4.94E-03	4.58E-01	1.72E+02	1.73E+02	8.60E-03	4.91E-05	1.73E+02	1.87E-03	1.34E-05
Methane	lbmole/hr	8.14E+01	8.14E+01	3.90E-05	2.89E-03	2.93E-03	8.36E+01	3.25E-03	8.35E+01	1.24E+01	8.95E-02
Ethane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.81E+00	7.81E-06	1.80E+00	5.33E-01	3.85E-03
Propane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.81E+00	4.57E-07	1.76E+00	8.13E-01	5.93E-03
Butane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.60E+01	2.46E-07	2.33E+01	1.60E+01	1.20E-01
Pentane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.34E+01	5.14E-09	1.72E+01	1.76E+01	1.37E-01
Hexane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.10E+01	8.00E-11	1.05E+01	1.79E+01	1.47E-01
Heptane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.89E+01	9.16E-13	5.41E+00	1.72E+01	1.47E-01
Octane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.70E+01	9.33E-15	2.40E+00	1.62E+01	1.36E-01
Nonane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.53E+01	8.66E-17	9.02E-01	1.49E+01	9.41E-02
Decane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.38E+01	4.29E-19	2.97E-01	1.35E+01	3.45E-02
n-C11	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.24E+01	3.32E-21	1.23E-01	1.23E+01	1.00E-02
n-C12	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.11E+01	1.49E-23	6.15E-02	1.11E+01	2.87E-03
n-C13	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.00E+01	1.09E-25	2.98E-02	9.98E+00	5.46E-04
n-C14	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	9.00E+00	4.32E-28	1.40E-02	8.99E+00	7.24E-05
n-C15	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	8.09E+00	1.15E-30	8.02E-03	8.09E+00	1.57E-05
n-C16	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	7.28E+00	2.85E-33	4.26E-03	7.28E+00	2.62E-06
n-C17	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	6.55E+00	7.38E-37	2.40E-03	6.55E+00	5.17E-07
n-C18	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	5.89E+00	8.36E-41	1.35E-03	5.89E+00	1.00E-07
n-C19	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	5.30E+00	3.80E-45	7.98E-04	5.30E+00	2.27E-08
n-C20	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.76E+00	7.02E-44	3.37E-04	4.76E+00	1.80E-09
n-C21	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.28E+00	1.66E-46	1.86E-04	4.28E+00	3.31E-10
n-C22	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.85E+00	2.43E-49	1.08E-04	3.85E+00	7.09E-11
n-C23	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.47E+00	7.26E-52	5.35E-05	3.46E+00	9.20E-12
n-C24	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.12E+00	1.47E-54	2.84E-05	3.12E+00	1.50E-12
n-C25	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.80E+00	7.25E-57	1.60E-05	2.80E+00	2.94E-13
n-C26	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.52E+00	6.37E-60	8.33E-06	2.52E+00	4.47E-14
n-C27	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.27E+00	2.99E-62	3.99E-06	2.27E+00	5.18E-15
n-C28	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.04E+00	3.10E-65	2.50E-06	2.04E+00	1.45E-15
n-C29	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.83E+00	1.16E-67	1.52E-06	1.83E+00	3.63E-16
n-C30	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.83E+00	9.01E-66	4.34E-07	1.83E+00	6.40E-18
Actual Volume Flow	barrel/day	51321	87036	529	245464	245992	15464	364	668655	4036	193700
Mass Density	lb/ft3	0.927	0.547	61.16	62.88	62.88	12.88	57.09	23.21	40.19	0.252
Molar Density	lbmole/ft3	0.0833	0.0491	3.39	3.49	3.49	0.19	3.17	1.29	0.260	0.00307

Medium Unit		21 Condensed OVHD	22 Reflux	23 Naphtha	24 Condensate Drum Vapor	25 Light Gases	26 Reboiler Feed	27 Boilup	28 Hydro-Isom Unit Feed	29 Diesel Product	30 Wax Product	31 Separator Vapor
Vapour Fraction		0.4429	0	0.00	1.00	1.00	0	1	0.00	0.00	0.00	1.00
Temperature	F	154	20	153	153	284	429	504	504	426	758	217
Pressure	psia	20	153	20	20	493	25	26	26	15	20	20
Molar Flow	lbmole/hr	139	15	62	62	167	284	163	121	95	27	229
Mass Flow	lb/hr	11140	1589	6539	3272	3806	54243	26109	28134	18880	9253	7078
Liquid Volume Flow	barrel/day	1181	157	647	377	587	4878	2399	2479	1690	789	965
Mass Enthalpy	Btu/lb	-958	-912	-912	-1073	-1646	-659	-524	-589	-660	-351	-1381
Component Fraction												
Hydrogen	lbmole/hr	1.67E-02	5.03E-05	3.14E-03	2.32E+00	3.37E+01	7.74E-12	1.35E-11	6.73E-12	0.00E+00	0.00E+00	3.60E+01
Water	lbmole/hr	4.28E-02	1.33E-03	8.29E-02	5.87E+00	1.96E+01	3.90E-09	6.75E-09	8.74E-09	0.00E+00	0.00E+00	2.55E+01
Carbon Monoxide	lbmole/hr	1.21E-02	6.81E-05	4.25E-03	1.67E+00	1.63E+01	2.05E-11	3.57E-11	2.18E-11	0.00E+00	0.00E+00	1.80E+01
Nitrogen	lbmole/hr	1.15E-03	5.85E-06	3.65E-04	1.59E-01	1.65E+00	1.73E-12	3.01E-12	1.83E-12	0.00E+00	0.00E+00	1.81E+00
Oxygen	lbmole/hr	6.82E-07	6.41E-09	4.00E-07	9.45E-05	6.65E-04	3.88E-15	6.74E-15	5.21E-15	0.00E+00	0.00E+00	7.60E-04
Carbon Dioxide	lbmole/hr	1.34E-05	3.69E-07	2.30E-05	1.84E-03	6.68E-03	7.59E-13	1.32E-12	1.51E-12	0.00E+00	0.00E+00	8.53E-03
Methane	lbmole/hr	8.95E-02	1.20E-03	7.52E-02	1.24E+01	7.11E+01	1.10E-09	1.91E-09	1.71E-09	0.00E+00	0.00E+00	8.35E+01
Ethane	lbmole/hr	3.85E-03	2.04E-04	1.28E-02	5.21E-01	1.28E+00	1.02E-09	1.77E-09	2.74E-09	0.00E+00	0.00E+00	1.80E+00
Propane	lbmole/hr	5.93E-03	8.53E-04	5.33E-02	7.60E-01	9.95E-01	1.70E-08	2.93E-08	7.02E-08	0.00E+00	0.00E+00	1.76E+00
Butane	lbmole/hr	1.20E-01	4.27E-02	2.66E+00	1.33E+01	9.99E+00	3.71E-06	6.33E-06	2.36E-05	0.00E+00	0.00E+00	2.33E+01
Pentane	lbmole/hr	1.37E-01	9.89E-02	6.17E+00	1.14E+01	5.78E+00	3.99E-05	6.73E-05	3.80E-04	0.00E+00	0.00E+00	1.72E+01
Hexane	lbmole/hr	1.47E-01	1.68E-01	1.05E+01	7.40E+00	3.13E+00	3.81E-04	6.31E-04	5.36E-03	0.00E+00	0.00E+00	1.05E+01
Heptane	lbmole/hr	1.47E-01	2.15E-01	1.34E+01	3.75E+00	1.66E+00	3.13E-03	5.07E-03	6.37E-02	0.00E+00	0.00E+00	5.41E+00
Octane	lbmole/hr	1.36E-01	2.24E-01	1.40E+01	1.55E+00	8.46E-01	2.27E-02	3.55E-02	6.55E-01	0.00E+00	0.00E+00	2.40E+00
Nonane	lbmole/hr	9.41E-02	1.63E-01	1.02E+01	4.66E-01	4.36E-01	0.104635	0.156665	4.23E+00	0.00E+00	0.00E+00	9.02E-01
Decane	lbmole/hr	3.45E-02	6.10E-02	3.81E+00	7.35E-02	2.23E-01	0.174493	0.245255	9.65E+00	0.00E+00	0.00E+00	2.97E-01
n-C11	lbmole/hr	1.00E-02	1.79E-02	1.11E+00	8.91E-03	1.14E-01	0.151684	0.196314	1.11E+01	8.10E+00	0.00E+00	1.23E-01
n-C12	lbmole/hr	2.87E-03	5.13E-03	3.20E-01	1.18E-03	6.04E-02	0.113499	0.132047	1.07E+01	7.71E+00	0.00E+00	6.15E-02
n-C13	lbmole/hr	5.46E-04	9.79E-04	6.11E-02	8.72E-05	2.97E-02	8.30E-02	8.39E-02	9.92E+00	7.09E+00	0.00E+00	2.98E-02
n-C14	lbmole/hr	7.24E-05	1.30E-04	8.11E-03	4.25E-06	1.40E-02	6.07E-02	5.07E-02	8.98E+00	6.43E+00	0.00E+00	1.40E-02
n-C15	lbmole/hr	1.57E-05	2.82E-05	1.76E-03	4.79E-07	8.01E-03	4.77E-02	3.35E-02	8.08E+00	5.80E+00	0.00E+00	8.02E-03
n-C16	lbmole/hr	2.62E-06	4.70E-06	2.93E-04	3.53E-08	4.26E-03	3.82E-02	2.19E-02	7.28E+00	5.21E+00	0.00E+00	4.26E-03
n-C17	lbmole/hr	5.17E-07	9.28E-07	5.79E-05	3.81E-09	2.40E-03	3.09E-02	1.37E-02	6.55E+00	4.69E+00	0.00E+00	2.40E-03
n-C18	lbmole/hr	1.00E-07	1.80E-07	1.12E-05	3.86E-10	1.35E-03	2.58E-02	8.79E-03	5.89E+00	4.22E+01	0.00E+00	1.35E-03
n-C19	lbmole/hr	2.27E-08	4.07E-08	2.54E-06	5.04E-11	7.98E-04	2.20E-02	5.76E-03	5.30E+00	3.80E+00	0.00E+00	7.98E-04
n-C20	lbmole/hr	1.80E-09	3.23E-09	2.02E-07	1.10E-12	3.37E-04	1.88E-02	3.55E-03	4.76E+00	3.42E+00	0.00E+00	3.37E-04
n-C21	lbmole/hr	3.31E-10	5.94E-10	3.71E-08	9.89E-14	1.86E-04	1.64E-02	2.32E-03	4.28E+00	0.00E+00	4.06E+00	1.86E-04
n-C22	lbmole/hr	7.09E-11	1.27E-10	7.95E-09	1.13E-14	1.08E-04	1.45E-02	1.55E-03	3.85E+00	0.00E+00	3.66E+00	1.08E-04
n-C23	lbmole/hr	9.20E-12	1.65E-11	1.03E-09	5.84E-16	5.35E-05	1.28E-02	9.69E-04	3.46E+00	0.00E+00	3.29E+00	5.35E-05
n-C24	lbmole/hr	1.50E-12	2.69E-12	1.68E-10	4.33E-17	2.84E-05	1.13E-02	6.27E-04	3.12E+00	0.00E+00	2.96E+00	2.84E-05
n-C25	lbmole/hr	2.94E-13	5.27E-13	3.29E-11	4.22E-18	1.60E-05	1.01E-02	4.17E-04	2.80E+00	0.00E+00	2.66E+00	1.60E-05
n-C26	lbmole/hr	4.47E-14	8.02E-14	5.01E-12	2.82E-19	8.33E-06	9.03E-03	2.67E-04	2.52E+00	0.00E+00	2.39E+00	8.33E-06
n-C27	lbmole/hr	5.18E-15	9.30E-15	5.81E-13	1.21E-20	3.99E-06	8.08E-03	1.66E-04	2.27E+00	0.00E+00	2.15E+00	3.99E-06
n-C28	lbmole/hr	1.45E-15	2.60E-15	1.62E-13	2.07E-21	2.50E-06	7.25E-03	1.16E-04	2.04E+00	0.00E+00	1.93E+00	2.50E-06
n-C29	lbmole/hr	3.63E-16	6.51E-16	4.07E-14	2.94E-22	1.52E-06	6.51E-03	8.04E-05	1.83E+00	0.00E+00	1.74E+00	1.52E-06
n-C30	lbmole/hr	6.40E-18	1.15E-17	7.17E-16	7.24E-25	4.34E-07	6.48E-03	3.80E-05	1.83E+00	0.00E+00	1.74E+00	4.34E-07
Actual Volume Flow	barrel/day	85770	167	688	84905	11124	6206	257030	3300	2141	1234	353282
Mass Density	lb/ft3	0.57	40.61	0.16	1.46	0.09	37.36	0.434215	40.61	37.69	32.05	36.45
Molar Density	lbmole/ft3	0.01	0.39	0.00311	0.0642	0.00277	0.195	0.00	0.388	0.19	0.09	0.16

Safety and Environmental Summary

Inherently safer consideration should always be addressed in the preliminary design stage since it is often incredibly expensive to incorporate these ideas into an existing plant. The main focus of inherent safety is to avoid hazards rather than attempting to control them by adding additional equipment.

This project includes several inherently safer design considerations. One of the concerns was having a large amount of hazardous material in storage or process equipment that could potentially be released. To minimize this risk, the equipment was sized accordingly based on each modular unit size to ensure that process vessels and piping did not contain excessive amounts of process material that could pose problems in the case of a fire or explosion. Similarly, there is no long-term product storage built at the well sites. Frac tanks on trailers are used to store products until they reach half capacity and are then taken back to the central plant so that products are not being stored onsite for long periods of time. The benefits of these changes are reduced capital cost from smaller equipment and reduced operating cost of buying the mobile storage tanks rather than renting them or building permanent ones on site. Unwanted material was also removed from the process stream. This leads to a lower overall amount of material. It can also lead to a reduction in the size of the equipment used in the process. In this project, an absorption column is used to remove the excess CO_2 while a flash drum removes the water from the system. This leads to an increase in the concentration of the reactants and increased the overall conversion of the reactor. It also reduced the volumetric flow rate of the stream and reduced the size of the reactor and all following equipment.

Another inherently safe factor utilized was substitution. The solvent used in the reactor was water rather than a more exotic solvent. Using water is a much safer than using a potentially toxic or highly flammable solvent. Other solvents are not as readily available, and many times must be regenerated after a certain point which is a significant added expense. For the FTR, medium pressure steam was used as the cooling medium rather than high pressure steam. This is because high pressure steam will cause the reactor to overheat and result in an undesired Anderson-Schulz-Flory probability distribution and the desired products will not be able to be obtained at a temperature over 449 degrees Fahrenheit. Cooling water was also considered but due to the possibility that the water could vaporize, it was determined that the vapor state was the preferred cooling medium. Medium pressure steam is also at a lower temperature which will be less likely to cause an explosion, fire, or material release. Using the medium pressure steam for the reactor also reduces the cost of operation since it is cheaper than the high-pressure steam.

Project safety management is a key management system that needs to be in place at every plant to achieve safe operation. This is done by defining safe operating limits and putting procedures in place to make sure that the process is operating within those limits. In addition to providing general safety information, standard operating procedures for this GTL plant will be clearly defined and written with step-by-step instruction that requires annual certification.

One of the best ways to help make safe practices a habit is to have continuous training programs and yearly, companywide training conferences. People will often pay closer attention when they know they will be quizzed on the material so tests or quizzes after training sessions should be held to confirm that everyone has retained the information provided.

All equipment in the plant needs to be inspected and undergo frequent testing to verify that all processes are operating properly. This should also include a pre-startup safety review and process hazard analysis as each module is redeployed.

Management of change is a significant part of having good safety culture. This system needs to be implemented to require proper documentation for all maintenance, shift changes, contracted work, etc. A mandatory authorization permit for any hazardous activities within the plant should be required as well as the proper training.

To prevent incidents a full-scale emergency planning and response team will be put into effect before start-up. For any incidents that occur, an investigation needs to be held within 48 hours of the occurrence and a written report should be issued and recorded.

After a conducting a chemical reactivity analysis, it was determined that there are potentially hazardous conditions throughout the process since it contains oxygen. Hydrocarbons, especially, those considered to be naphtha, are highly flammable. Oxygen generates heat that creates an exothermic reaction at ambient temperatures. This reaction could be intense, violent, or explosive as shown in the deflagration analysis below.

These substances can also create gaseous products that may cause increases in pressure. An increase in pressure may lead to an overpressure event which will be handled with pressure relief devices and rupture disks to prevent any damage to equipment. Reactions from these substances can produce toxic gases such as carbon monoxide, chlorine, hydrogen halide, nitrogen oxides, and acid fumes.

This process also uses a lot of water as the solvent for the reactor. Water when mixed with carbon dioxide and oxygen can be corrosive. The process water return line will need to be monitored and possibly treated to prevent corrosion of the pipes and equipment.

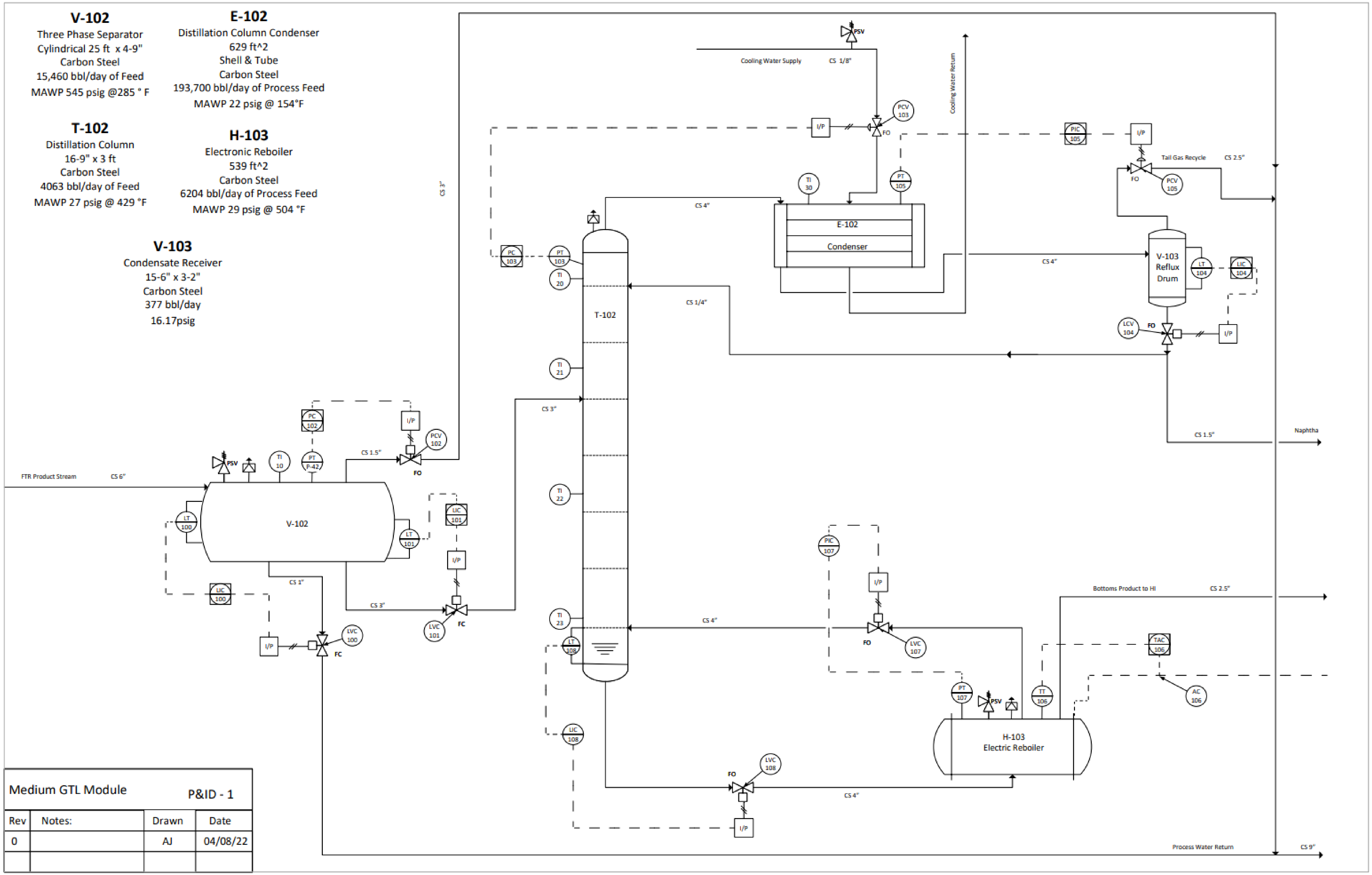

Figure 5: P&ID of Medium Unit Major Fractionator

The modular gas plant is a refinery that deals with volatile carbon chains. If the system were to experience a catastrophic failure, it could lead to flammable liquid or vapors escaping the system and creating health and environmental hazards. Plume dispersion, puff dispersion, and a deflagration were modeled to determine the potential worst-case effects. The Pasquill-Gifford coefficients as well as atmospheric stability classes were used to determine how the plume was spread. The best and worst cases, class A and F, were modelled, and the wells were in a rural setting. It was determined that the worst-case scenario occurs within the large unit. A rupture within the FTR or FTR product line could release a large amount of vapor or flammable liquids. The line is also under pressure at 495 psia and 449 ^oF. The condition of each model assumes the ambient conditions of the area to assist with specifying variables in the continuity equation.

The plume is modelled using point source dispersion modelling. A modified form of the continuity equation was used to determine the concentration of volatile carbon chains at various distances. *Figure 6: Plume Dispersion Effect* shows how the concentration of the carbon chains that vaporize spread over a distance from 5 meters to 640 meters when escaping at 12 kg/s. A continuous plume is unlikely due to the control systems in place causing a shutdown of the system thus stopping the 12kg/s flow.

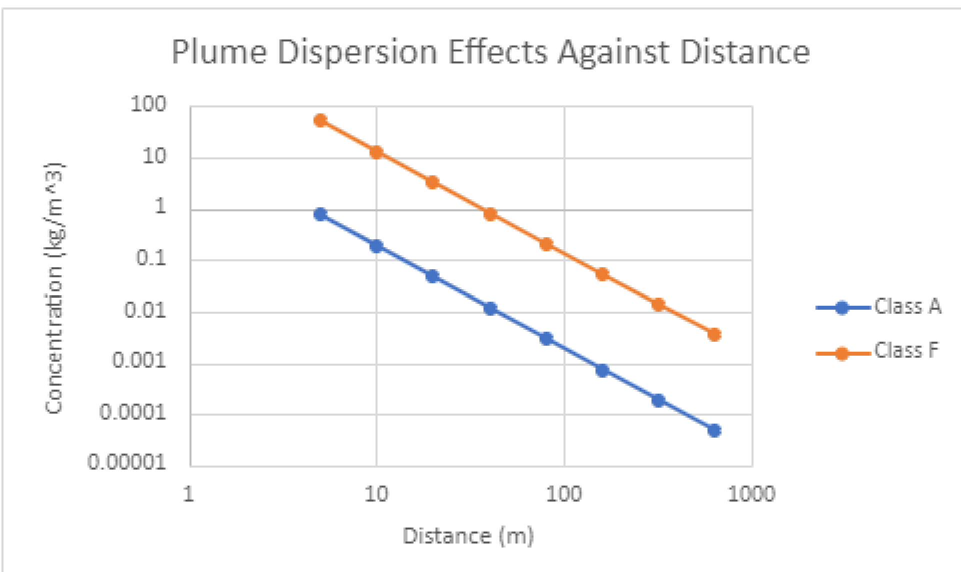

Figure 6: Plume Dispersion Effect

In the event the FTR experiences a rupture or of the FTR product line rupturing, a puff model was developed again using the continuity equation. This event assumes that the control system would quickly stop flow upon a sudden pressure drop. The puff model then assumes that all the material within the reactor would be released over a 18 seconds, the amount of time for the reactor volume to move through the reactor. This leads to 216 kg of hydrocarbons being released with 90 kg being vaporized. This 90 kg is composed of the lighter carbons from C1-C12 which are flammable. *Figure 7: Puff Dispersion Case Comparison* shows the dispersion from 5 m to 1,460 m. The average lower flammability limit is 1.63% by volume while the average upper flammability limit is 10.4% by volume. The mixture will only be flammable between these bounds. Since the vapor is a hydrocarbon, it is likely for the mixture to combust at some point upon release due to the many ignition sources located within the GTL plant.

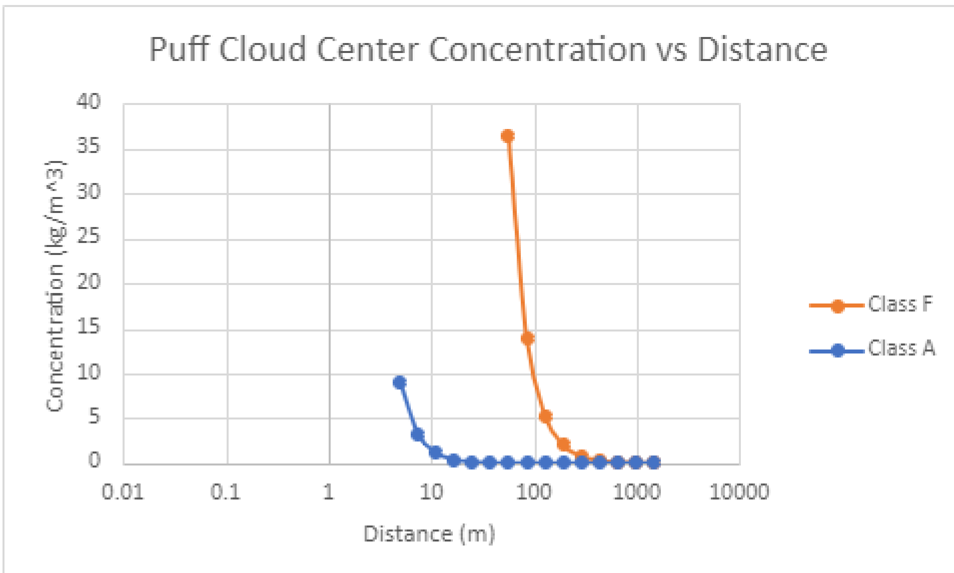

Figure 7: Puff Dispersion Case Comparison

An uncongested vapor cloud deflagration was conducted to determine the potential risks associated with a vapor leak of flammable hydrocarbon gases. The puff model's 90 kg mass was used to determine the equivalent mass of TNT and the resulting damage. The resulting mass of TNT was determined to be 18.6 kg of TNT using an efficiency of 30% for hydrocarbons. *Figure 8: Deflagration Pressures vs Distance* **Error! Reference source not found.**shows the resulting pressure wave at varying distances. From 0-10 meters, the resulting damage is a total loss of equipment. Beyond that, severe damage is seen all the way to 65 meters, at this point the damage lowers to minor damage to structures and less.

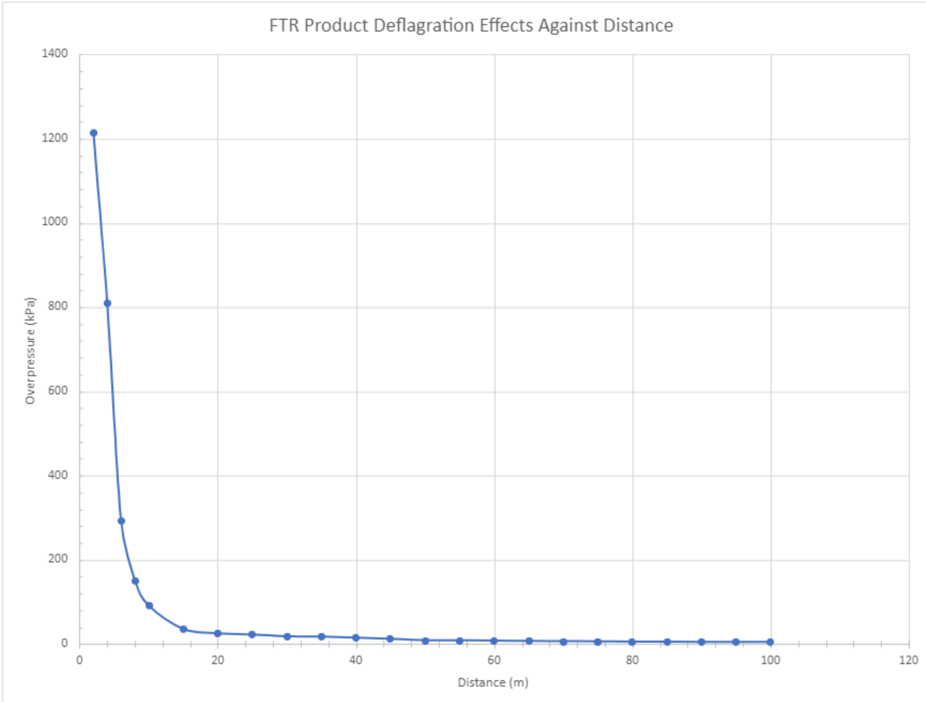

Figure 8: Deflagration Pressures vs Distance

From a safety perspective, there is always risk associated with chemical processes, but there is nothing in this process that is exceptionally unsafe. At this point in the design stage, the only potential for terminating the project is purely economic.

One of the major concerns requiring significant attention in the detailed design stage would be start-up processes. More specifically, since all the units are modular, each unit will have to be taken off the truck and assembled at the well site. The assembly will include a structure that will hold the condenser and condensate receiver up in the air next to the fractionation column in the separation unit. Such a structure may need safety railing and or harnesses for routine maintenance.

Preventing a runaway reaction or loss of containment from the Fischer-Tropsch reactor is a major concern that will require considerable attention and should be adequately addressed with process safety management practices. The most important would be the aforementioned methods like standard operating procedures, management of change procedures, and proper training.

A risk management plan will need to be developed to minimize the frequency and severity of accidents within the plant. A focus on communication with the nearest communities and adjacent landowners will be necessary to implement a plan to handle emergencies properly. This project also focuses on being environmentally conscious. In many similar processes, methane is vented to the atmosphere, or a flare is used to burn off excess hydrocarbons. In this project the goal was to convert as much methane into products as possible.

Overall, this design is inherently more safe, all of the hazards have been identified and consequences of risk have been addressed and mitigated. A strong safety culture will be realized once the process safety management procedures are put into place and a risk management plan is established.

Equipment Information Summary

All equipment information can be found in *Table 6: Medium GTL Plant Process Equipment Summary.* This table only summarizes equipment specifications for the medium-sized units, the small and large units can be found in the appendix. Equipment design specifications were determined based on safety of the process and maximizing the NPV. Any specifications marked in *Table 6: Medium GTL Plant Process Equipment Summary* as "N/A" implies that detailed design was not necessary for the scope of the project. The material of construction for all of the equipment and piping is carbon steel since it is the cheapest option that can safely operate given all operating conditions and materials.

All heaters in this design were determined to be electric heaters. After a cost analysis for both feed preheaters (H-101 & H-102) and the column reboiler (H-103), it was determined that fuel gas would cost more than electricity when supplying the duty to run at the desired outlet conditions.

The condensers were all determined to be U-Tube heat exchangers since it is the cheapest option that provided the largest heat transfer area. The syngas condenser (E-101) was not sized since it was within the scope of the syngas unit which did not need to be detailed; however, specifications regarding expected inlet/outlet conditions were provided along with the duties to achieve a calculated area if needed for future reference.

The main fractionation vessel consisted of 7, single-pass sieve trays with a material of construction being carbon steel. The tower has a condenser and reboiler to maintain product specifications and aid in efficient separation and product temperatures. The column was optimized to maximize the bottoms production in order to produce more C11+ material. The temperature and pressure of the column were determined by examining the relative volatilities of C10 and lighter material compared to C11+ material. Due to the nature of these two product streams, it was a simple separation which allowed for the tower being able to operate at atmospheric pressure, which resulted in lower condenser and reboiler duties. Diesel was a more profitable product compared to naphtha and LPG, so the goal of the design was to produce as much of the heavier products as possible to maximize the project NPV.

A three-phase separator was implemented to provide liquid holdup to remove water and vapor from the main hydrocarbon stream. The purpose of the placement of the vessel was to remove volume from the feed stream to the main fractionation vessel (T-102). This resulted in an overall smaller volume for the distillation column. The orientation of the 3-phase separator was determined to be horizontal as they are cheaper since they don't require as much structuring, and the length and diameter calculated fit within the required dimensions. The condensate receiver following the distillation column is vertical since it contains a higher vapor volume.

A packed bed reactor (PBR) was designed to serve as the FTR (R-102). A PBR was chosen due to the requirement of continuous flow and the use of a catalyst. PBR's are generally versatile and help to increase contact between multiple phases. This reaction leads to the creation of a two-phase liquid/gas flow in contact with a cobalt catalyst. A tubular design was selected to help increase catalyst contact and conversion of the feed into the desired products. Tube numbers ranging from 3-70 were tested, ultimately leading to 42 tubes being the optimum number to increase conversion while balancing catalyst weight and cooling water cost. The reactor was kept at constant temperature to lead to the desired ratio of products. This led to the reactor further being designed as a jacketed reactor to regulate temperature. The reactor was designed to be horizontal to reduce the cost as it requires less structure. This was chosen after a compatibility matrix of the reactants and products was conducted. The conditions of the reactor are also reasonable as the maximum temperature and pressure of carbon steel are 750 ^{o}F and 60,000 psi respectively.

Table 6: Medium GTL Plant Process Equipment Summary

Heat Exchangers	**E-101**	**E-102**	**Heaters**	**H-101**	**H-102**	**H-103**		
Type	U-Tube	U-Tube	Type	Electric Heater	Electric Heater	Electric Reboiler		
Area (m2)	N/A	11.9	MOC	CS	CS	CS		
Duty (BTU/hr)	28,200,000	1,780,000	Duty (Btu/hr)	8,080,000	2,670,000	5,610,000		
Shell								
Temperature (°C)	886	141						
Pressure (bar)	34.1	1.59						
Phase	Vapor → Liquid	Vapor → Liquid						
MOC	CS	CS						
Tube								
Temperature (°C)	32.2	32.2						
Pressure (bar)	3.45	3.45						
Phase	Liquid	Liquid						
MOC	CS	CS						
Vessels/Towers/ Reactors	**V-101**	**V-102**	**V-103**	**T-101**	**T-102**	**T-103**	**R-101**	**R-102**
Temperature (°C)	62.8	140	67.2	26.7	221	403	888	254
Pressure (bar)	33.9	34.1	1.59	33.8	5.37	1.38	34.1	33.4
Orientation	N/A	Horizontal	Vertical	Vertical	Vertical	Vertical	Horizontal	Horizontal
MOC	CS	CS	CS	CS	CS	CS	CS	CS
Size								
Height/Length (m)	N/A	7.50	4.74	N/A	5.11	N/A	N/A	3.45
Diameter (m)	N/A	3.09	0.96	N/A	0.914	N/A	N/A	1.00
Internals	N/A	Heavy Liquid Boot	Mist Eliminator	N/A	Sieve Trays	N/A	N/A	42 Tubes

Unit Control and Instrumentation Description

Controls is an integral part of designing a safer, more reliable design. A full control system was developed for the major fractionator and a preliminary plan was considered for the FTR. The systems were designed considering safety, and access to the operators to monitor the system in the field and in a central control location.

The control systems in place in the major fractionator primarily maintain temperature, pressure, liquid levels, and flow rates. Level controllers are used in the separators and condensate receivers in order to maintain proper separation of phases and components. This is done by using a control valve on a liquid outlet which will adjust the flow of liquid. This will serve to maintain the liquid seal. Temperature controllers are used throughout the system. The temperature controller uses a transmitter and controller to adjust the flow rates of the cooling or heating fluids to the necessary flow rate. This allows the process fluid to maintain the proper temperature. Pressure controllers are used to regulate the flow rate of gasses throughout the system. A pressure indicator records the pressure and sends a signal to a controller which in turn regulates the gas flow line. This system also helps prevent the pressure of the vessels from going above operating conditions.

The control system is depicted in *Figure 5: P&ID of Medium Unit Major Fractionator*. The control system is the same across the sizes of modules, but this description will follow the medium unit's P&ID.

Table 7: Unit Control and Instrumentation Description describes the instruments that are not connected to the central control system, and their purpose.

The FTR requires temperature controls to regulate and monitor the temperature due to the exothermic nature of the reaction. The temperature needs to be monitored not only for the safety of the equipment but also for the risk of causing a runaway reaction or creating undesirable products. A backup to the steam cooling system will need to be implemented, as well as a control system. This system should also include pressure relief devices to prevent over pressuring the equipment.

Table 7: Unit Control and Instrumentation Description

Type of Device	ID	Locations	Description
Pressure Relief Valve	PSV	(V-102) (E-102) (H-103)	If overpressure occurs, PSV will allow relief to protect the process vessel until overpressure event desists.
Rupture Disk	RD	(V-102) (T-102) (H-103)	Disk will rupture due to overpressure to prevent equipment breakage but may lead to backpressure.
Temperature Indicator	TI	(V-102) (T-102) (E-102)	Monitors temperature at different points throughout the distillation column, and the three-phase separator, and overhead condenser. These devices are local allowing operators to monitor the tower in person.

Economics

This design was evaluated over a 20-year project life with a hurdle rate of 8%. Based on timing for permits, construction is planned to begin in the latter half of 2023, so that the first units can be deployed in the first part of 2024. Units were calculated with a first of a kind (FOAK) cost initially and then a learning rate was applied to the subsequent units. The heat exchangers, FTR, vessels, distillation column, and trays were all calculated using the Turton Appendix A equations for capital costs. The capital cost for the syngas unit, HIU, steam plant, and CO2 recovery unit were cost based on their calculated capacities and scaled for each unit size. The FOAK cost was applied to each size unit since the dimensions will change between units. In *Table 8:Cost of Module by Name* (Total is Sum x 4.8) the total capital cost was multiplied by the supplied multiplier of 4.8 and applied to the total equipment sum. This table also shows the salvage value calculated at the end of the project life.

Based on the unit's positions/usage throughout the well systems over the project life the utilities were calculated based on the capacity of the units deployed (for example a medium unit may be deployed because smalls were unavailable, so the medium runs at a reduced capacity). The utilities reported in *Table 9:Utility Cost per Unit & Usage per Year* are the services used and the associated usages per year of each unit at full capacity. Some utilities where credited for returning them, those credits were accounted for in the utilities total.

The other operating costs include labor, maintenance (including catalyst recharge), module transportation and product transportation. The number of operators needed for the units employed was calculated to be 154 and the salary estimated at $80,000 per year. This salary was inflated 3% every year over the project life. Maintenance was calculated to be 10% of the capital cost each year plus the catalyst recharge for all the units every three years. The module transportation was calculated based on the unit capacity and the distance traveled. For example, a unit named Hedy, moved from well 2C to 2B in year ten, and this transportation was approximately 83 miles and cost $1.4M. Further information of total transportation cost and specific distances traveled can be found in the Appendix. Finally, once the amount of product produced per unit was finalized, the number of product trucks needed per day was determined and priced. Each product truck was priced based on the number of miles driven. The cost for product trucks was $1.25 per mile, for total product truck costs per well see the Appendix.

Table 8:Cost of Module by Name (Total is Sum x 4.8)

Number ID	Name	# OAK	Unit Cost	Salvage
L_1	April	1	($40,248,000)	$4,049,800
L_2	Genna	2	($32,199,000)	$3,244,900
M_1	Hedy	1	($25,691,000)	$2,594,100
M_2	Irene	2	($20,553,000)	$2,080,300
M_3	Katherine	3	($18,038,000)	$1,828,800
M_4	Margaret	4	($16,443,000)	$1,669,300
M_5	Mary	5	($15,303,000)	$1,555,300
M_6	Rachel	6	($14,431,000)	$1,468,100
M_7	Shirley	7	($13,732,000)	$1,398,200
S_1	Alexa	1	($10,536,000)	$1,078,600
S_2	Annie	2	($8,429,000)	$867,900
S_3	Bessie	3	($7,398,000)	$764,800
S_4	Beth	4	($6,744,000)	$699,400
S_5	Caroline	5	($6,276,000)	$652,600
S_6	Dijanna	6	($5,918,000)	$616,800
S_7	Dorothy	7	($5,632,000)	$588,200
S_8	Ella	8	($5,395,000)	$564,500
S_9	Evelyn	9	($5,194,000)	$544,400
S_10	France	10	($5,021,000)	$527,100
S_11	Gertrude	11	($4,869,000)	$511,900
S_12	Jane	12	($4,735,000)	$498,500
S_13	June	13	($4,614,000)	$486,400
		Total:	**($1,331,515,200)**	**$28,289,900**

Table 9:Utility Cost per Unit & Usage per Year

Utility	Price/Unit	Large Usage	Medium Usage	Small Usage
Process Water Return (kgal)	- $0.35/kgal	16680208	8200236	8959728
Electricity (kWh)	$0.04/kWh	81466782	34854368	7123323
Hydrogen (lb)	$0.06//lb	3345225	1500150	327177
Cooling Water (kgal)	$0.15/kgal	26631407	13263693	2650551
Oxygen (Short Ton)	$100/short ton	44781	22398	4468
MP Steam (klb)	$2/klb	21672	9747	2152
HP Steam (klb)	$3/klb	96710	48706	9741
Fuel Gas (MBTU)	$3/MBTU	171276	76808	16751
CO2 (MSCF)	$400/MSCF	25845	12923	2579
Direct credits are accounted for in the price of the utility, credits alone are noted as negative.				

The income for this project was solely from the products of naphtha and diesel. The goal was to optimize the amount of diesel as the sale price is $90/bbl and naphtha was $75/bbl. The design did not create a viable LPG stream, and since LPG is only $0.30/lb the focus was not placed on excessive equipment to create and LPG product. Since the design was able to produce a substantial amount of diesel and naphtha, this project was ultimately profitable.

Table 10:Production of LPG, Naphtha, and Diesel by Unit Size

Large		Medium		Small	
LPG (lb)	0	LPG (lb)	0	LPG (lb)	0
Naptha (bbl)	541076	Naptha (bbl)	200954	Naptha (bbl)	18460
Diesel (bbl)	1114856	Diesel (bbl)	625172	Diesel (bbl)	120012

The bottom line of the preliminary design was calculating a rate of return that was greater than 8% and completing an NPV analysis. That final ROR exceeded the hurdle rate substantially coming in at 23%. The project NPV was determined to be $2.29B with a capital cost of $1.33B. The average cashflow per year is about $181M, with the first years being the most profitable and the final years being the least. The cash flow below provides more details, note that in the final three years the maintenance estimates are the same as early on, which is not a completely fair assumption since most units are not in service. The negative values were kept maintaining a conservative estimate of the project.

Table 11: Cash Flow Table for Modular GTL Plants

Cash Flow Table for Modular GTL Plants																						
End of Year		0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
Net Revenue																						
Income		-	796,000	796,000	814,000	627,000	702,000	803,000	809,000	784,000	826,000	783,000	765,000	753,000	760,000	722,000	526,000	316,000	236,000	127,000	76,000	49,000
Salvage		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	28,000
Utilities		-	(79,000)	(79,000)	(77,000)	(51,000)	(72,000)	(78,000)	(74,000)	(36,000)	(40,000)	(42,000)	(62,000)	(68,000)	(73,000)	(69,000)	(58,000)	(27,000)	(16,000)	(2,000)	7,000	5,000
Module Transportation		-	(5,000)	-	(2,000)	(2,000)	(3,000)	(4,000)	(3,000)	(4,000)	(2,000)	(4,000)	(3,000)	(3,000)	(2,000)	(4,000)	(1,000)	(2,000)	(2,000)	(1,000)	(1,000)	-
Product Transportation		-	(1,000)	(1,000)	(1,000)	(1,000)	(1,000)	(1,000)	(1,000)	(1,000)	(1,000)	(1,000)	(2,000)	(2,000)	(2,000)	(2,000)	(2,000)	(2,000)	(2,000)	(1,000)	(1,000)	(1,000)
Maintenence and Repair		-	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)	(134,000)
Labor		-	(53,000)	(53,000)	(54,000)	(54,000)	(54,000)	(55,000)	(55,000)	(56,000)	(56,000)	(57,000)	(57,000)	(57,000)	(58,000)	(59,000)	(59,000)	(60,000)	(60,000)	(61,000)	(61,000)	(62,000)
Depreciation Straight Line - Seven Years																						
Assets		-	(191,000)	(191,000)	(191,000)	(191,000)	(191,000)	(191,000)	(191,000)	-	-	-	-	-	-	-	-	-	-	-	-	-
Taxable Income		-	336,000	340,000	359,000	198,000	250,000	343,000	355,000	556,000	595,000	548,000	510,000	492,000	493,000	458,000	275,000	95,000	25,000	(70,000)	(113,000)	(115,000)
- Tax	0.2	-	(68,000)	(68,000)	(72,000)	(40,000)	(50,000)	(69,000)	(71,000)	(112,000)	(119,000)	(110,000)	(102,000)	(99,000)	(99,000)	(92,000)	(55,000)	(19,000)	(5,000)	14,000	23,000	23,000
Net Income		-	269,000	272,000	287,000	159,000	200,000	274,000	284,000	445,000	476,000	439,000	408,000	393,000	395,000	367,000	220,000	76,000	20,000	(56,000)	(91,000)	(92,000)
Depreciation																						-
Assets		-	191,000	191,000	191,000	191,000	191,000	191,000	191,000	-	-	-	-	-	-	-	-	-	-	-	-	-
Capital Cost																						
Assets		(1,332,000)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cash Flows																						
Cash Flow		(1,332,000)	459,000	463,000	478,000	349,000	390,000	465,000	474,000	445,000	476,000	439,000	408,000	393,000	395,000	367,000	220,000	76,000	20,000	(56,000)	(91,000)	(92,000)
Discount factor (P/F)		1.0000	1.0800	1.1664	1.2597	1.3605	1.4693	1.5869	1.7138	1.8509	1.9990	2.1589	2.3316	2.5182	2.7196	2.9372	3.1722	3.4259	3.7000	3.9960	4.3157	4.6610
Discount Cash Flow		(1,332,000)	425,000	397,000	380,000	257,000	266,000	294,000	277,000	241,000	239,000	204,000	175,000	157,000	146,000	125,000	70,000	23,000	6,000	(15,000)	(22,000)	(20,000)
NPV at 8% Hurdle Rate		**2,293,000**	Hurdle Rate:	8%	DCFROR:	23%	Numbers in cashflow need to be multiplied by 1000 to retrieve the appropriate values.															

Wellhead Site System Plot

The system for the wells in based around a central plant. The wells in the system all vary in distance from the central plant, these distances are depicted in *Table 12: Well Distances from the Central Plant*. The wells were assumed to be approximately 45° starting from 1A moving toward 2A then to 1B, 2B, 1C and so on around a full 360°. This assumption led to the approximation of each site from each other site as shown in *Table 13*. To better visualize the entire well system, *Figure 9: Wellhead Map* shows an approximation of the wells from the central plant. This map is not exactly to scale, however this visualization helped with optimization for module deployment. The images on each well site are in varying sizes so that the largest producing wells have the largest icons and the least producing have the smallest. The actual starting feed value of methane of each well is depicted and organized from the least productive well to the most as well as the rounded value used throughout the development of this plan in *Table 14*.

Table 12: Well Distances from the Central Plant

Well Name	1A	2A	1B	2B	1C	2C	1D	2D	1E	2E	1F	2F	1G	2G	1H	2H
Miles to Central Plant	62	85.8	21	84.7	17	2.4	73.4	42	6.2	98.8	91	13	59	87	89.3	36

Table 13:Well Distances from Other Wells in the System

Well Name	1A	1B	1C	1D	1E	1F	1G	1H	2A	2B	2C	2D	2E	2F	2G	2H
1A		49.4	64.3	125.2	68.2	141.8	85.6	63.2	37.1	83.7	63.0	102.1	157.9	68.0	85.3	31.9
1B			15.0	76.3	25.8	112.0	75.3	91.7	66.9	65.8	20.2	53.7	118.5	33.4	97.0	34.0
1C				62.5	18.1	103.7	76.0	102.0	80.8	69.3	14.8	38.8	106.5	29.4	102.9	45.3
1D					69.2	116.9	122.4	162.7	132.5	88.3	71.2	38.1	98.0	79.3	157.3	107.5
1E						86.7	59.3	93.8	91.6	87.3	5.7	36.3	93.1	12.1	89.6	41.8
1F							64.6	127.5	173.4	172.3	91.9	84.4	37.8	79.1	98.9	109.9
1G								63.3	121.3	141.0	61.2	84.5	93.7	47.3	39.6	56.1
1H									97.3	144.7	91.5	129.1	156.5	85.2	34.5	57.7
2A										65.3	85.8	119.3	184.6	95.4	122.2	65.5
2B											83.0	94.5	169.6	97.7	158.6	92.0
2C												40.3	98.8	14.8	89.4	37.7
2D													75.2	44.0	120.4	78.0
2E														90.1	131.6	126.8
2F															78.3	38.3
2G																66.6
2H																

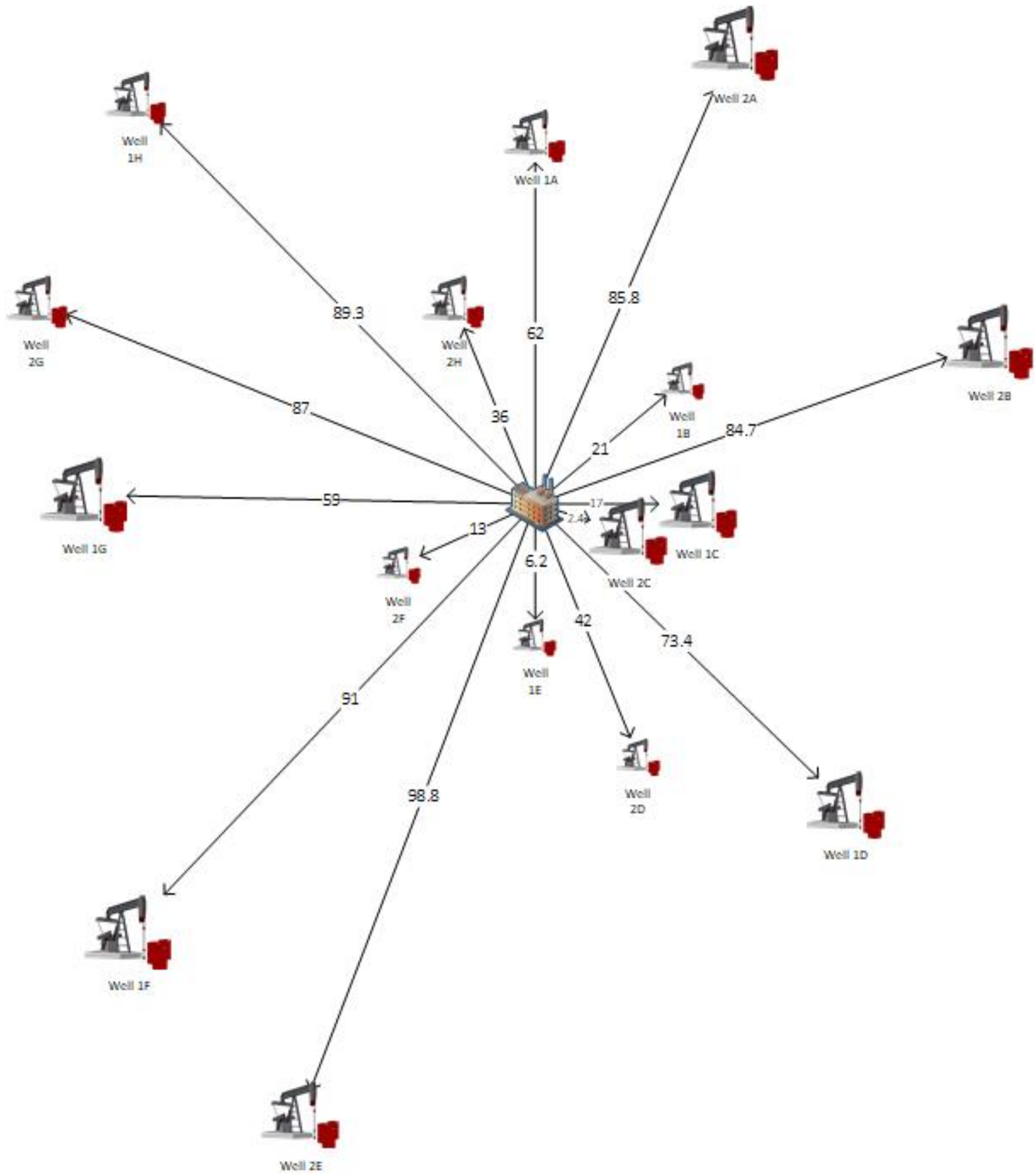

Figure 9: Wellhead Map

Table 14: Well Starting Methane Feed Values

Well Name	Original Production	Rounded Estimated Feed
2F	2874	2900
1B	2989	3000
1E	3290	3300
2D	3840	3800
1A	4494	4500
2G	6076	6100
2H	6468	6500
1H	7081	7100
1D	7447	7400
2E	7939	7900
1C	8365	8400
2C	8742	8700
2A	9182	9200
1G	11155	11200
2B	12258	12300
1F	14737	14700

Unit Deployment and Redeployment Logistics

The central plant can only process the product equivalent of 30,000 MSCF of natural gas feed each day. Each well starts with the feed depicted in *Table 14: Well Starting Methane Feed Values* and produces that same amount for 2 years then steadily decreases by 35% each year. The graph in *Figure 10: Well 1A Methane Supply* depicts the decline of Well 1A to represent the decrease. In order to optimize the wells in service at any given time, the startup of the wells was varied over the project life.

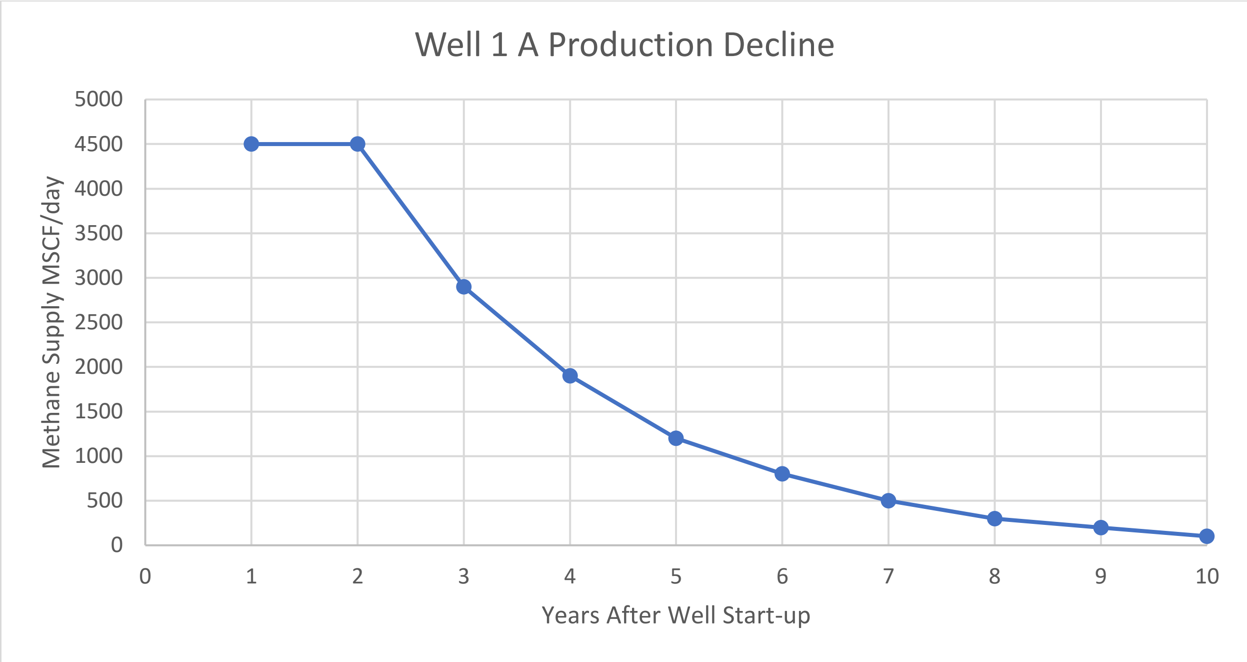

Figure 10: Well 1A Methane Supply

Since the production of each well decreases over time, there was less of a need for larger modular units later in the project. There were occasions where a well may be taken out of service in the middle of its useful production life, to optimize the use of methane while still meeting the central plant's capacity. *Figure 13: Well Feed Value (MSCF/day) and Years in Service* shows what wells were determined to be in service each year. *Figure 11: Optimized Service Years of Wells in System over the Project Life* shows further information such as the feed for each year, showing the total of the methane as well as the number of wells in service each year. This table represents the method that was used to optimize the utilization of each well, each array of the feed and its decline was shuffled though out the years to find the best time to bring the well into service. It was important to try to get as close to the central plant's 30,000 MSCF/day capacity while also minimizing the number of wells in service at a time, as that translates to total minimum number of modular units.

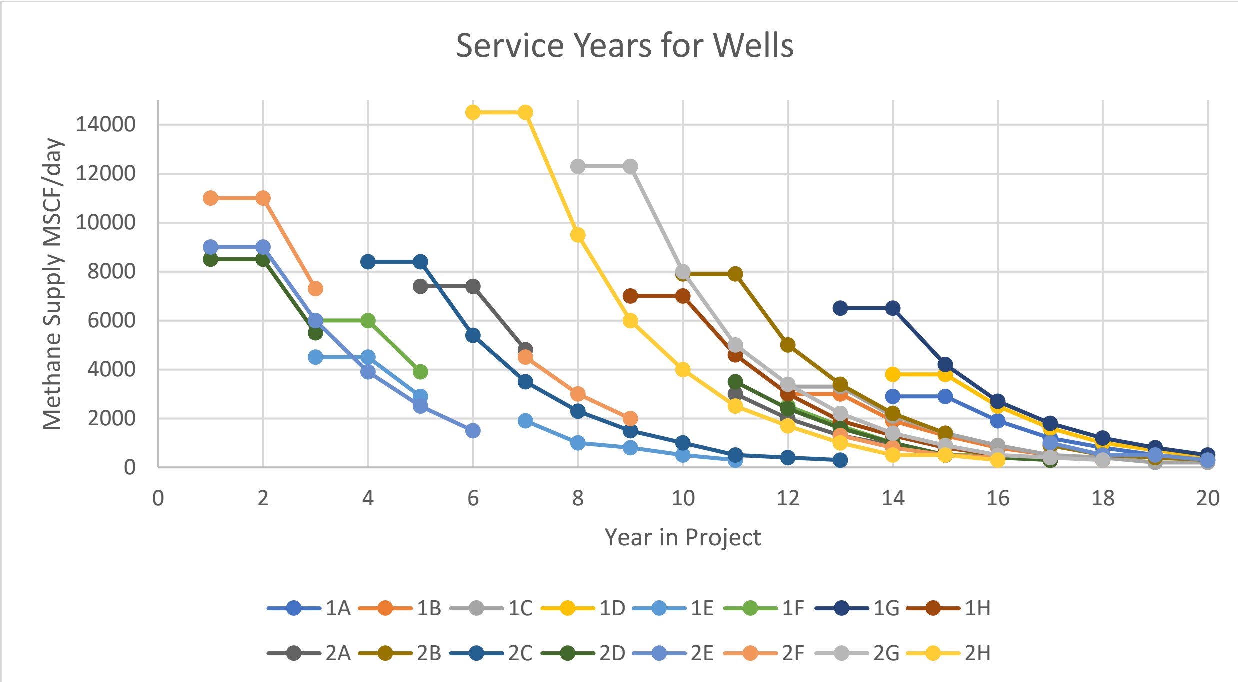

Figure 11: Optimized Service Years of Wells in System over the Project Life

The unit sizes were based on feed of methane with three sizes, 500 MSCF/day, 2,500 MSCF/day, and 5,000 MSCF/day. Based on the optimized service of the wells, modules had to be assigned to active wells each year of the project. This was accomplished by displaying the map of the wells next to its total production and the calculated number of units needed to meet the feed of the well. An example of two years of the well sites with their modules is displayed in *Figure 14: Example of Year 1 and Year 10 Module Distribution*. When refining the total number of modules, the turndown of the unit was considered and it was determined that the units could not turn down more than approximately 60%, so when a well only had, for example 100 MSCF/day remaining, that production was lost.

The final number of units needed of each size was determined to be twelve small, seven medium, and two larges. In *Table 15:Module Fleet Inventory Deployment Plan* the full fleet of modules is displayed as well as their location. Due to the cost of moving the modules, unused units would remain at the site they were last used while they waited to be deployed to the next site.

Well Name	Year																			
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
1A														2900	2900	1900	1200	800	500	300
1B												3000	3000	1900	1300	800	500	300		
1C												3300	3300	2100	1400	900	500	400	200	200
1D														3800	3800	2500	1600	1000	700	400
1E			4500	4500	2900		1900	1000	800	500	300									
1F			6000	6000	3900							2500	1700	1000	500	500	300			
1G													6500	6500	4200	2700	1800	1200	800	500
1H									7000	7000	4600	3000	1900	1300	800	500	300			
2A					7400	7400	4800				3000	2000	1300	900	500	400				
2B										7900	7900	5000	3400	2200	1400		900	500	400	300
2C				8400	8400	5400	3500	2300	1500	1000	500	400	300							
2D	8500	8500	5500								3500	2400	1600	1000	500	400	300			
2E	9000	9000	6000	3900	2500	1500											1000	500	500	300
2F	11000	11000	7300				4500	3000	2000				1300	800	500	400				
2G								12300	12300	8000	5000	3400	2200	1400	900	500	400	300		
2H						14500	14500	9500	6000	4000	2500	1700	1000	500	500	300				
# Wells in Service	3	3	5	4	5	4	5	5	6	6	8	10	12	13	13	12	11	8	6	6
Total Methane Supply (MSCF/day)	28500	28500	29300	22800	25100	28800	29200	28100	29600	28400	27300	26700	27500	26300	19200	11800	8800	5000	3100	2000

Figure 13: Well Feed Value (MSCF/day) and Years in Service

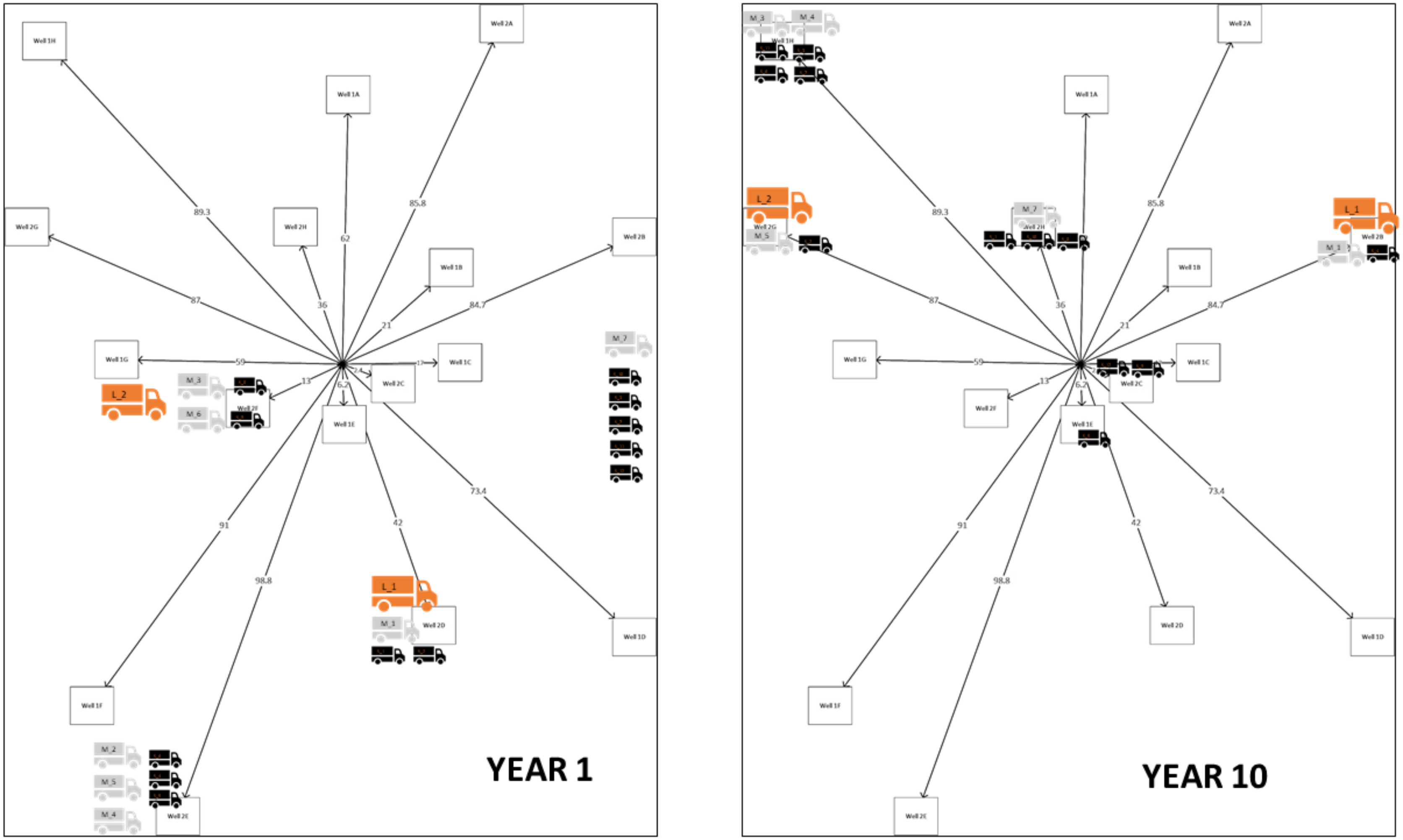

Figure 14: Example of Year 1 and Year 10 Module Distribution

Table 15:Module Fleet Inventory Deployment Plan

Number ID	Name	Year 0	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Year 14	Year 15	Year 16	Year 17	Year 18	Year 19	Year 20
L_1	April	Constuction	2D	2D	2D	2C	2C	2H	2H	2H	2H	2B	2B	2B	2B	1D	1D	1D	1D	1D	1D	1D
L_2	Genna	Constuction	2F	2F	2F	2F	2A	2A	2H	2G	2G	2G	2G	2G	1G	1G	1G	1G	1G	1G	1G	1G
M_1	Hedy	Constuction	2D	2D	1E	1E	1E	2C	2C	2C	2C	2B	2B	1B	1B	1B	1B	1B	1D	1D	1D	1D
M_2	Irene	Constuction	2E	2E	1F	1F	1F	2H	2A	2G	2G	2G	2A	2A	2A	1A	1A	1A	1A	1A	1A	1A
M_3	Katherine	Constuction	2F	2F	1F	1F	1F	2H	2F	2F	1H	1H	1H	1H	1H	1H	1H	1H	1H	1H	1H	1H
M_4	Margaret	Constuction	2E	2E	2E	2E	2E	2H	2H	2H	1H	1H	1H	1F	1F	1G	1G	2E	2E	2E	2E	2E
M_5	Mary	Constuction	2E	2E	2E	2C	2C	2C	1E	2G	2G	2G	2D	2D	2D	2B	2B	2B	2B	2B	2B	2B
M_6	Rachel	Constuction	2F	2F	2F	2F	2A	2A	2A	2G	2G	1H	1H	1C	1C	1C	1C	1C	1C	1C	1C	1C
M_7	Shirley	Constuction	CP	CP	CP	CP	CP	CP	CP	CP	2F	2H	2H	2H	2G	2G	2G	2G	1G	1G	1G	1G
S_1	Alexa	Constuction	2D	2D	2D	2C	2C	2C	2C	2C	2H	2H	2H	1C	1C	1A	1A	1B	1C	1C	1C	1C
S_2	Annie	Constuction	2D	2D	2D	2C	2C	2H	2H	2H	1H	2G	2G	2G	1G	1F	1H	1H	1H	1H	1H	1H
S_3	Bessie	Constuction	CP	CP	1E	1E	1E	1E	2H	2H	1H	1H	2A	2A	2H	2H	1H	1H	1H	1H	1G	1G
S_4	Beth	Constuction	2E	2E	2E	2E	2E	2E	2F	1E	1E	2B	2B	2B	2H	2H	2H	2H	2B	2B	2B	2B
S_5	Caroline	Constuction	2E	2E	2E	2E	2E	2E	2F	1E	1E	1E	1E	1E	2F	2F	1F	1C	2E	2E	2E	2E
S_6	Dijanna	Constuction	2E	2E	1E	1E	1E	2H	2H	2H	1H	1H	1H	1B	1B	2A	2A	1B	1B	1B	1B	1B
S_7	Dorothy	Constuction	CP	CP	1E	1E	1E	2H	2H	2H	2H	1H	1H	1H	2F	2F	2F	2F	2B	1A	1A	1A
S_8	Ella	Constuction	2F	2F	1F	1F	1F	2E	2C	2C	2H	2H	2D	2D	2F	2D	2D	2D	2D	2D	1D	1D
S_9	Evelyn	Constuction	2F	2F	1F	1F	1F	1F	2F	2F	2F	2C	2C	1C	1C	2A	2A	2A	2A	1A	1A	1A
S_10	France	Constuction	CP	CP	CP	CP	1F	1F	1F	1F	1F	2H	2D	2D	1G	1E	2G	2G	2G	2G	1A	1A
S_11	Gertrude	Constuction	CP	CP	1E	1E	1E	2H	2H	2H	1H	1H	1H	1H	1G	2D	2G	1F	1F	1D	1G	1G
S_12	Jane	Constuction	CP	CP	2D	2E	2E	2E	2F	2F	2F	2C	2C	2C	2C	2C	2C	1C	2E	1D	1D	1D
Units were named to make it easier for identification, all units are named after female scientiest of great renown.																		CP: Central Plant			Not In Service	

Based on the deployed modules the product truck development plan was created. This plan was designed by finding the maximum number of product trucks needed every five years, as they are contracted in five-year increments. Using the total production per day, the number of trucks was determined and then priced based on the requirement to return to the central plant to off load product. It was estimated that any given driver could work a maximum of 11 hours in a day and that the average speed of travel was 65 mph. It was also assumed to take about 45 minutes to load and unload a truck. This information helped model the number of trips required per day to any given site. The number of trips were then applied to how many trips a specific could truck make in a day based on the distance to the central plant and the allowable number of work hours. The total number of contracted product trucks are displayed in *Table 16: Total Product Trucks Contracted.*

Table 16: Total Product Trucks Contracted

Years:	**1-5**	**6-10**	**11-15**	**16-20**
Trucks Contracted	16	20	30	27

Summary of NPV and Sensitivity Analysis

The costs for the project have the potential to suddenly change as it is hard to predict future trends in inflation and demand. Variable analysis was conducted to determine the potential effects in changes to costs. Tornado charts were assembled to compare the effects of each new cost. *Figure 15: Single Variable Analysis Tornado Chart* shows the relative impact of varying costs for capital, labor, total feed, and total utilities as well as varying income from naphtha and diesel. Income was found to have the most profound effects on the profitability of the project. The price of oil can vary greatly over time and can be affected by many factors. The price low point was chosen to be a 45% reduction from original prices while the high point was a 45% increase in price. This is based upon trends in the last 20 years. Ultimately, income is the most important factor for the success of the project in a field that experience upturns and downturns frequently. The next major factor is the capital cost of the system. All the equipment is modular. This leads to an increase in the cost as it is specialized for being capable of being disassembled and moved relatively quickly. The cost of this equipment is difficult to estimate. A 4.8 multiplier was used to initially estimate the cost of equipment. A lower bound of 1.7 was selected as it is the Lang factor for a prefabricated modular plant. The upward bound was chosen to be 7.9 as it was the opposite direction of the Lang factor. This multiplier greatly affects the profit of the project as a multiplier of 7.9 would reduce the NPV by $1B.

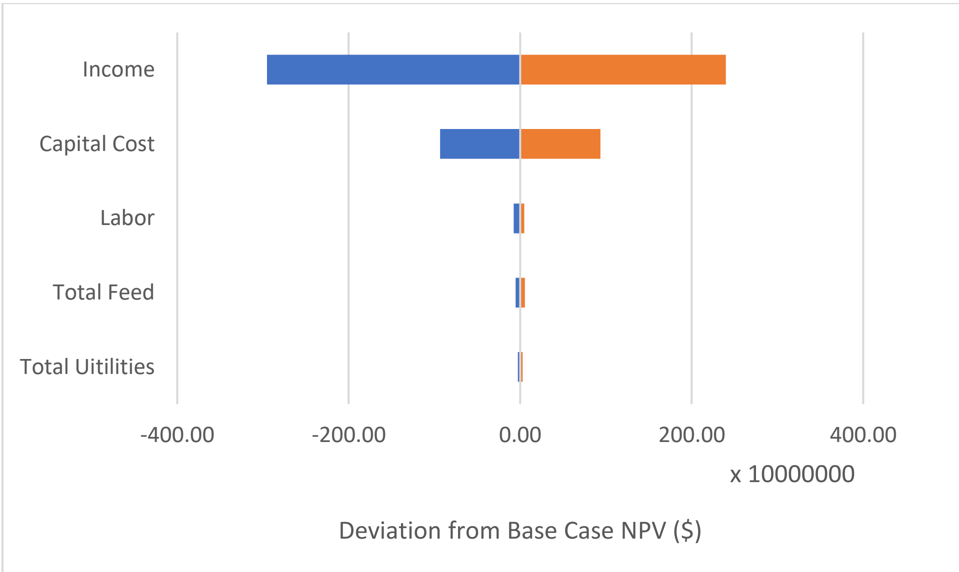

Figure 15: Single Variable Analysis Tornado Chart

Assuming the capital cost and income are fixed, labor was then compared to the cost of the feeds and utilities. *Figure 16: Labor Cost Comparison with Feed and Utilities* show the relative weight of varying the combined utilities and feed by ± 10%. This number was chosen to estimate the potential increase or decrease in utilities on top of inflation. The labor cost was varied by adjusting the average operator pay from $50k/year to $130k/year. The salary depends heavily on the experience of the operators and the location of the site. Ultimately, it was found that the increase in utilities and feed would cause less of an effect on the total NPV. Labor was found to have a similar effect. The increase in cost of utilities and feed will not lead to the project being economically unattractive even with the large of amount of water, steam, and other auxiliary feeds and utilities.

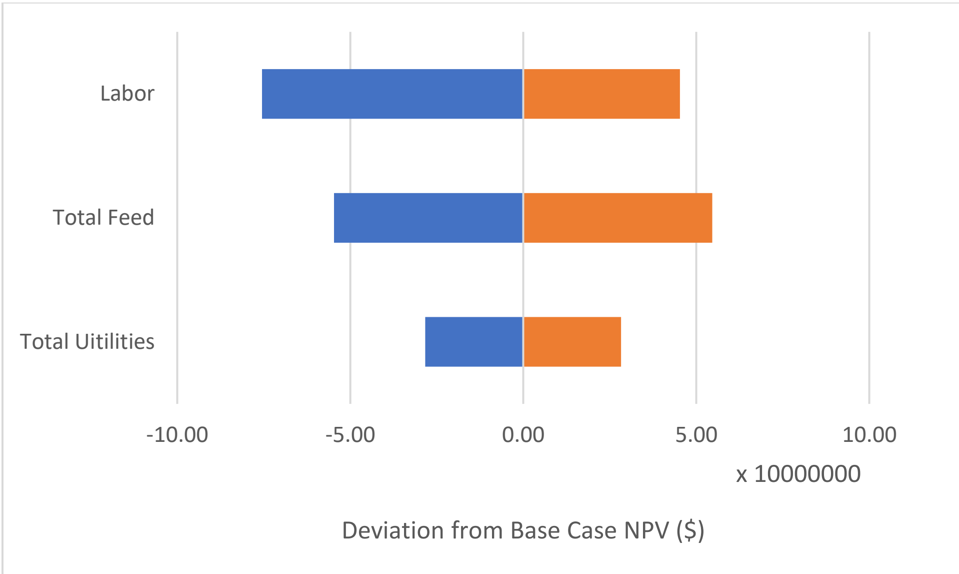

Figure 16: Labor Cost Comparison with Feed and Utilities

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Appendix

Large Unit		1	2	3	4	5	6	7	8	9	10
		Methane	Carbon Dioxide	HP Steam	Syngas Feed Mixture	Hot Syngas Feed	99% Oxygen	Syngas Product	Condensed Syngas	Syngas Product Water Removed	Absorber Solvent
Vapour Fraction		1.000	1.000	1.000	0.913	1.000	1.000	1.000	0.740	1.000	0.000
Temperature	F	100.0	100.0	490.0	364.1	1000.0	75.0	1628.0	145.0	145.0	80.0
Pressure	psia	500.0	500.0	600.0	500.0	495.0	514.7	495.0	489.0	489.0	495.0
Molar Flow	lbmole/hr	822	300	765	1887	1887	400	3210	3210	2375	200000
Mass Flow	lb/hr	13187	13203	13782	40172	40172	12784	52956	52956	37904	3603020
Liquid Volume Flow	barrel/day	3016	1095	946	5057	5057	772	6825	6825	5792	247207
Mass Enthalpy	Btu/lb	-2010.3	-3857.0	-5628.7	-3858.6	-3457.3	-4.9	-2623.8	-3686.5	-2469.5	-6826.1
Component Fraction											
Hydrogen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.23E+03	1.23E+03	1.23E+03	0.00E+00
Water	lbmole/hr	0.00E+00	0.00E+00	7.65E+02	7.65E+02	7.65E+02	0.00E+00	8.53E+02	8.53E+02	1.96E+01	2.00E+05
Carbon Monoxide	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	6.15E+02	6.15E+02	6.15E+02	0.00E+00
Nitrogen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.00E+00	4.00E+00	4.00E+00	4.00E+00	0.00E+00
Oxygen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.96E+02	4.27E-01	4.27E-01	4.27E-01	0.00E+00
Carbon Dioxide	lbmole/hr	0.00E+00	3.00E+02	0.00E+00	3.00E+02	3.00E+02	0.00E+00	3.44E+02	3.44E+02	3.43E+02	0.00E+00
Methane	lbmole/hr	8.22E+02	0.00E+00	0.00E+00	8.22E+02	8.22E+02	0.00E+00	1.63E+02	1.63E+02	1.63E+02	0.00E+00
Ethane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Propane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Butane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Pentane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Hexane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Heptane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Octane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Nonane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Decane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C11	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C12	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C13	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C14	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C15	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C16	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C17	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C18	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C19	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C20	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C21	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C22	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C23	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C24	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C25	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C26	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C27	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C28	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C29	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C30	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Actual Volume Flow	barrel/day	39614	12615	47160	124723	255108	18489	624384	135492	134440	244975
Mass Density	lb/ft3	1.42	4.47	1.25	1.38	0.67	2.96	0.36	1.67	1.21	62.87
Molar Density	lbmole/ft3	0.089	0.102	0.069	0.065	0.032	0.092	0.022	0.101	0.076	3.490

Large Unit		11	12	13	14	15	16	17	18	19	20
		Syngas Product CO2 Removed	FTR Feed	Water Removed	CO2 Removed	Combined Water	FTR Product	Water Mix	Process Water Return	Column Feed	Column OVHD
Vapour Fraction		1.000	1.000	0.000	0.000	0.000	0.209	0.000	0.003	0.000	1.000
Temperature	F	80.0	449.0	145.0	80.6	80.8	284.0	284.0	82.9	284.0	284.9
Pressure	psia	485.0	484.0	489.0	490.0	489.0	495.0	493.0	20.0	493.0	23.0
Molar Flow	lbmole/hr	2014	2014	834	200361	201195	1367	553	202127	528	293
Mass Flow	lb/hr	22684	22684	15052	3618239	3633291	100276	9953	3653350	84358	25318
Liquid Volume Flow	barrel/day	4526	4526	1033	248473	249506	9404	683	251659	7750	2561
Mass Enthalpy	Btu/lb	-1575.7	-1338.2	-6751.1	-6813.4	-6813.1	-1420.2	-6612.5	-6798.0	-781.9	-782.0
Component Fraction											
Hydrogen	lbmole/hr	1.23E+03	1.23E+03	4.36E-02	4.26E+00	4.30E+00	6.10E+01	5.15E-02	6.53E+01	4.77E+00	1.63E-02
Water	lbmole/hr	2.60E+00	2.60E+00	8.33E+02	2.00E+05	2.01E+05	5.99E+02	5.52E+02	2.01E+05	1.27E+01	4.36E-02
Carbon Monoxide	lbmole/hr	6.12E+02	6.12E+02	2.28E-02	2.77E+00	2.79E+00	3.25E+01	2.13E-02	3.53E+01	3.67E+00	0.00E+00
Nitrogen	lbmole/hr	3.86E+00	3.86E+00	3.95E-04	1.38E-01	1.38E-01	0.00E+00	0.00E+00	1.38E-01	0.00E+00	0.00E+00
Oxygen	lbmole/hr	4.23E-01	4.23E-01	3.28E-05	3.80E-03	3.84E-03	0.00E+00	0.00E+00	3.84E-03	0.00E+00	0.00E+00
Carbon Dioxide	lbmole/hr	6.48E+00	6.48E+00	9.01E-01	3.37E+02	3.38E+02	0.00E+00	0.00E+00	3.38E+02	0.00E+00	0.00E+00
Methane	lbmole/hr	1.63E+02	1.63E+02	7.71E-05	2.78E-03	2.85E-03	1.69E+02	7.59E-03	1.69E+02	3.03E+01	1.03E-01
Ethane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.67E+00	1.77E-05	3.61E+00	1.26E+00	4.33E-03
Propane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.37E+00	1.20E-06	4.11E+00	2.22E+00	7.74E-03
Butane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.37E+00	4.32E-08	3.54E+00	2.93E+00	1.05E-02
Pentane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	5.27E+01	1.17E-08	3.08E+01	4.19E+01	1.57E-01
Hexane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.74E+01	1.79E-10	1.62E+01	4.17E+01	1.62E-01
Heptane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.26E+01	2.02E-12	7.49E+00	3.97E+01	1.57E-01
Octane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.84E+01	2.03E-14	3.17E+00	3.69E+01	1.41E-01
Nonane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.45E+01	1.88E-16	1.27E+00	3.37E+01	1.06E-01
Decane	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.17E+01	9.48E-19	4.98E-01	3.13E+01	5.19E-02
n-C11	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.79E+01	7.15E-21	2.08E-01	2.77E+01	1.73E-02
n-C12	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.58E+01	3.31E-23	1.07E-01	2.57E+01	6.13E-03
n-C13	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.26E+01	2.33E-25	5.01E-02	2.25E+01	1.85E-03
n-C14	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.03E+01	9.28E-28	2.33E-02	2.03E+01	5.12E-04
n-C15	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.90E+01	2.58E-30	1.38E-02	1.90E+01	2.00E-04
n-C16	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.64E+01	6.11E-33	7.01E-03	1.64E+01	6.16E-05
n-C17	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.48E+01	1.58E-36	3.92E-03	1.48E+01	2.11E-05
n-C18	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.33E+01	1.79E-40	2.19E-03	1.33E+01	7.13E-06
n-C19	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.19E+01	8.15E-45	1.29E-03	1.19E+01	2.64E-06
n-C20	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.07E+01	1.50E-43	5.40E-04	1.07E+01	5.28E-07
n-C21	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	9.66E+00	3.54E-46	2.96E-04	9.66E+00	1.72E-07
n-C22	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	8.69E+00	5.18E-49	1.70E-04	8.69E+00	6.16E-08
n-C23	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	7.81E+00	1.55E-51	8.39E-05	7.81E+00	1.62E-08
n-C24	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	7.03E+00	3.14E-54	4.44E-05	7.03E+00	4.89E-09
n-C25	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	6.32E+00	1.54E-56	2.48E-05	6.32E+00	1.66E-09
n-C26	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	5.68E+00	1.36E-59	1.28E-05	5.68E+00	4.80E-10
n-C27	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	5.11E+00	6.35E-62	6.09E-06	5.11E+00	1.18E-10
n-C28	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.60E+00	6.59E-65	3.80E-06	4.60E+00	4.97E-11
n-C29	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.14E+00	2.46E-67	2.29E-06	4.14E+00	1.96E-11
n-C30	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.72E+00	1.71E-65	5.84E-07	3.72E+00	1.38E-12
Actual Volume Flow	barrel/day	103349	175314	1052	245915	246966	28764	745	1036112	8909	419288
Mass Density	lb/ft3	0.94	0.55	61.16	62.89	62.89	14.90	57.09	15.07	40.48	0.26
Molar Density	lbmole/ft3	0.083	0.049	3.390	3.483	3.482	0.203	3.169	0.834	0.253	0.003

Large Unit		21	22	23	24	25	26	27	28	29	30	31
		Condensed OVHD	Reflux	Naphtha	Condensate Drum Vapor	Light Gases	Reboiler Feed	Boilup	Hydro-Isom Unit Feed	Diesel Product	Wax Product	Separator Vapor
Vapour Fraction		0.318	0.000	0.000	1.000	1.000	0.000	1.000	0.000	0.000	0.000	1.000
Temperature	F	144.2	144.2	144.2	144.2	220.9	437.0	516.9	516.9	417.90	712.40	283.96
Pressure	psia	20.0	20.0	20.0	20.0	20.0	26.5	28.0	28.0	14.70	20.00	493.00
Molar Flow	lbmole/hr	293	32	168	93	380	642	375	267	178	89	286
Mass Flow	lb/hr	25320	3384	17792	4141	10106	123397	60972	62425	33790	20640	5965
Liquid Volume Flow	barrel/day	2588	332	1755	499	1470	11087	5579	5496	3042	2455	971
Mass Enthalpy	Btu/lb	-969.7	-915.9	-915.9	-1244.8	-1562.6	-653.4	-514.2	-578.3	-670.6	-398.3	-1783.2
Component Fraction												
Hydrogen	lbmole/hr	1.63E-02	6.58E-05	1.11E-02	4.76E+00	6.09E+01	1.73E-09	2.95E-09	3.63E-09	0.00E+00	0.00E+00	5.61E+01
Water	lbmole/hr	4.36E-02	1.92E-03	3.23E-01	1.24E+01	4.59E+01	2.21E-07	3.76E-07	1.16E-06	0.00E+00	0.00E+00	3.35E+01
Carbon Monoxide	lbmole/hr	0.00E+00	9.80E-05	1.65E-02	3.66E+00	3.25E+01	3.42E-09	5.84E-09	8.71E-09	0.00E+00	0.00E+00	2.88E+01
Nitrogen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Oxygen	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Carbon Dioxide	lbmole/hr	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Methane	lbmole/hr	1.03E-01	1.94E-03	3.27E-01	3.00E+01	1.69E+02	1.21E-07	2.06E-07	4.44E-07	0.00E+00	0.00E+00	1.39E+02
Ethane	lbmole/hr	4.33E-03	3.28E-04	5.51E-02	1.21E+00	3.61E+00	4.80E-08	8.14E-08	3.00E-07	0.00E+00	0.00E+00	2.41E+00
Propane	lbmole/hr	7.74E-03	1.57E-03	2.63E-01	1.96E+00	4.11E+00	4.88E-07	8.23E-07	4.62E-06	0.00E+00	0.00E+00	2.15E+00
Butane	lbmole/hr	1.05E-02	4.94E-03	8.30E-01	2.10E+00	3.54E+00	3.76E-06	6.30E-06	5.42E-05	0.00E+00	0.00E+00	1.44E+00
Pentane	lbmole/hr	1.57E-01	1.31E-01	2.20E+01	2.00E+01	3.08E+01	2.87E-04	4.75E-04	6.11E-03	0.00E+00	0.00E+00	1.08E+01
Hexane	lbmole/hr	1.62E-01	1.86E-01	3.12E+01	1.05E+01	1.62E+01	1.48E-03	2.41E-03	4.60E-02	0.00E+00	0.00E+00	5.69E+00
Heptane	lbmole/hr	1.57E-01	2.08E-01	3.49E+01	4.53E+00	7.49E+00	6.75E-03	1.08E-02	3.00E-01	0.00E+00	0.00E+00	2.96E+00
Octane	lbmole/hr	1.41E-01	1.99E-01	3.34E+01	1.69E+00	3.17E+00	2.79E-02	4.32E-02	1.75E+00	0.00E+00	0.00E+00	1.48E+00
Nonane	lbmole/hr	1.06E-01	1.52E-01	2.56E+01	5.19E-01	1.27E+00	8.82E-02	1.31E-01	7.64E+00	0.00E+00	0.00E+00	7.55E-01
Decane	lbmole/hr	5.19E-02	7.57E-02	1.27E+01	1.06E-01	4.98E-01	1.57E-01	2.20E-01	1.85E+01	0.00E+00	0.00E+00	3.92E-01
n-C11	lbmole/hr	1.73E-02	2.53E-02	4.25E+00	1.43E-02	2.08E-01	1.51E-01	1.96E-01	2.34E+01	2.42E+01	0.00E+00	1.94E-01
n-C12	lbmole/hr	6.13E-03	8.98E-03	1.51E+00	2.28E-03	1.07E-01	1.21E-01	1.42E-01	2.42E+01	2.47E+01	0.00E+00	1.05E-01
n-C13	lbmole/hr	1.85E-03	2.72E-03	4.57E-01	2.60E-04	5.01E-02	8.72E-02	9.05E-02	2.21E+01	2.24E+01	0.00E+00	4.98E-02
n-C14	lbmole/hr	5.12E-04	7.51E-04	1.26E-01	2.56E-05	2.33E-02	6.41E-02	5.59E-02	2.01E+01	2.03E+01	0.00E+00	2.33E-02
n-C15	lbmole/hr	2.00E-04	2.94E-04	4.93E-02	5.10E-06	1.38E-02	5.23E-02	3.90E-02	1.89E+01	1.91E+01	0.00E+00	1.38E-02
n-C16	lbmole/hr	6.16E-05	9.04E-05	1.52E-02	6.75E-07	7.01E-03	4.00E-02	2.48E-02	1.64E+01	1.65E+01	0.00E+00	7.01E-03
n-C17	lbmole/hr	2.11E-05	3.10E-05	5.20E-03	1.25E-07	3.92E-03	3.21E-02	1.57E-02	1.48E+01	1.49E+01	0.00E+00	3.92E-03
n-C18	lbmole/hr	7.13E-06	1.05E-05	1.76E-03	2.17E-08	2.19E-03	2.66E-02	1.01E-02	1.33E+01	1.34E+01	0.00E+00	2.19E-03
n-C19	lbmole/hr	2.64E-06	3.87E-06	6.50E-04	4.55E-09	1.29E-03	2.25E-02	6.71E-03	1.19E+01	1.20E+01	0.00E+00	1.29E-03
n-C20	lbmole/hr	5.28E-07	7.74E-07	1.30E-04	2.39E-10	5.40E-04	1.92E-02	4.21E-03	1.07E+01	1.08E+01	0.00E+00	5.40E-04
n-C21	lbmole/hr	1.72E-07	2.52E-07	4.24E-05	3.72E-11	2.96E-04	1.67E-02	2.78E-03	9.66E+00	0.00E+00	1.37E+01	2.96E-04
n-C22	lbmole/hr	6.16E-08	9.03E-08	1.52E-05	6.97E-12	1.70E-04	1.46E-02	1.87E-03	8.69E+00	0.00E+00	1.23E+01	1.70E-04
n-C23	lbmole/hr	1.62E-08	2.37E-08	3.99E-06	7.09E-13	8.39E-05	1.29E-02	1.19E-03	7.81E+00	0.00E+00	1.11E+01	8.39E-05
n-C24	lbmole/hr	4.89E-09	7.18E-09	1.21E-06	9.49E-14	4.44E-05	1.14E-02	7.77E-04	7.03E+00	0.00E+00	9.94E+00	4.44E-05
n-C25	lbmole/hr	1.66E-09	2.43E-09	4.09E-07	1.57E-14	2.48E-05	1.02E-02	5.22E-04	6.32E+00	0.00E+00	8.95E+00	2.48E-05
n-C26	lbmole/hr	4.80E-10	7.04E-10	1.18E-07	1.94E-15	1.28E-05	9.06E-03	3.38E-04	5.68E+00	0.00E+00	8.05E+00	1.28E-05
n-C27	lbmole/hr	1.18E-10	1.73E-10	2.90E-08	1.71E-16	6.09E-06	8.09E-03	2.13E-04	5.11E+00	0.00E+00	7.22E+00	6.09E-06
n-C28	lbmole/hr	4.97E-11	7.29E-11	1.22E-08	4.33E-17	3.80E-06	7.25E-03	1.50E-04	4.60E+00	0.00E+00	6.50E+00	3.80E-06
n-C29	lbmole/hr	1.96E-11	2.88E-11	4.84E-09	9.55E-18	2.29E-06	6.51E-03	1.05E-04	4.14E+00	0.00E+00	5.86E+00	2.29E-06
n-C30	lbmole/hr	1.38E-12	2.03E-12	3.40E-10	8.82E-20	5.84E-07	5.82E-03	4.59E-05	3.72E+00	0.00E+00	5.25E+00	5.84E-07
Actual Volume Flow	barrel/day	12960	353	1854	127407	590527	14190	555142	7432	3818	3724	19214
Mass Density	lb/ft3	0.84	41.02	41.02	0.14	0.07	37.17	0.47	35.90	37.83	32.88	1.33
Molar Density	lbmole/ft3	0.010	0.387	0.387	0.003	0.003	0.193	0.003	0.154	0.200	0.102	0.064

Table 1: Large Unit PFD Stream Table

Small Unit		1	2	3	4	5	6	7	8	9	10
		Methane	Carbon Dioxide	HP Steam	Syngas Feed Mixture	Hot Syngas Feed	99% Oxygen	Syngas Product	Condensed Syngas	Syngas Product Water Removed	Absorber Solvent
Vapour Fraction		1.00	1.00	1.00	0.91	1.00	1.00	1.00	0.74	1.00	0.00
Temperature	*F*	500.0	500.0	600.0	500.0	495.0	514.7	495.0	489.0	489.0	495.0
Pressure	*psia*	100.0	100.0	490.0	364.7	1000.0	75.0	1627.2	145.0	145.0	80.0
Molar Flow	*lbmole/hr*	82.0	29.9	77.0	188.9	188.9	39.9	320.8	320.8	237.1	200000.0
Mass Flow	*lb/hr*	1315.5	1317.2	1386.6	4019.4	4019.4	1275.2	5294.5	5294.5	3783.7	3603020.1
Liquid Volume Flow	*barrel/day*	300.9	109.3	95.1	505.3	505.3	77.0	682.0	682.0	578.2	247206.8
Mass Enthalpy	*Btu/lb*	-2010.3	-3857.0	-5628.7	-3863.8	-3462.8	-4.9	-2630.0	-3693.3	-2472.3	-6826.1
Component Fraction											
Hydrogen	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.23E+02	1.23E+02	1.23E+02	0.00E+00
Water	*lbmole/hr*	0.00E+00	0.00E+00	7.70E+01	7.70E+01	7.70E+01	0.00E+00	8.56E+01	8.56E+01	1.96E+00	2.00E+05
Carbon Monoxide	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	6.12E+01	6.12E+01	6.12E+01	0.00E+00
Nitrogen	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.99E-01	3.99E-01	3.99E-01	3.99E-01	0.00E+00
Oxygen	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.95E+01	4.28E-02	4.28E-02	4.28E-02	0.00E+00
Carbon Dioxide	*lbmole/hr*	0.00E+00	2.99E+01	0.00E+00	2.99E+01	2.99E+01	0.00E+00	3.45E+01	3.45E+01	3.44E+01	0.00E+00
Methane	*lbmole/hr*	8.20E+01	0.00E+00	0.00E+00	8.20E+01	8.20E+01	0.00E+00	1.63E+01	1.63E+01	1.63E+01	0.00E+00
Ethane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Propane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Butane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Pentane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Hexane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Heptane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Octane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Nonane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Decane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C11	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C12	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C13	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C14	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C15	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C16	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C17	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C18	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C19	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C20	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C21	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C22	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C23	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C24	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C25	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C26	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C27	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C28	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C29	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
n-C30	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Actual Volume Flow	*barrel/day*	3952	1259	4745	12493	25536	1844	62389	13525	13419	244975
Mass Density	*lb/ft3*	1.42	4.47	1.25	1.38	0.67	2.96	0.36	1.67	1.21	62.87
Molar Density	*lbmole/ft3*	0.089	0.102	0.069	0.065	0.032	0.092	0.022	0.101	0.076	3.490

Small Unit		11	12	13	14	15	16	17	18	19	20
		Syngas Product CO2 Removed	FTR Feed	Water Removed	CO2 Removed	Combined Water	FTR Product	Water Mix	Process Water Return	Column Feed	Column OVHD
Vapour Fraction		1.00	1.00	0.00	0.00	0.00	0.23	0.00	0.00	0.00	1.00
Temperature	*F*	485.0	484.0	489.0	490.0	489.0	495.0	4.5	4.5	4.5	300.9
Pressure	*psia*	80.0	449.0	145.0	80.0	80.1	284.0	131.0	81.6	131.0	23.0
Molar Flow	*lbmole/hr*	193.2	193.2	83.7	200043.9	200127.7	133.0	11.9	200230.0	30.7	6.0
Mass Flow	*lb/hr*	2140.8	2140.8	1510.9	3604662.9	3606173.8	9748.3	214.6	3609144.6	6781.2	675.4
Liquid Volume Flow	*barrel/day*	433.7	433.7	103.7	247351.4	247455.1	916.3	14.7	247771.2	600.6	65.4
Mass Enthalpy	*Btu/lb*	-1549.5	-1308.2	-6751.1	-6824.7	-6824.6	-1411.0	-6774.9	-6821.3	-831.2	-692.7
Component Fraction											
Hydrogen	*lbmole/hr*	1.18E+02	1.18E+02	4.38E-03	4.65E+00	4.65E+00	6.01E+00	6.07E-07	1.07E+01	5.97E-04	1.00E-04
Water	*lbmole/hr*	2.49E-01	2.49E-01	8.36E+01	2.00E+05	2.00E+05	5.76E+01	1.19E+01	2.00E+05	5.37E-02	9.50E-03
Carbon Monoxide	*lbmole/hr*	5.82E+01	5.82E+01	2.28E-03	3.03E+00	3.04E+00	3.00E+00	3.41E-07	6.04E+00	6.70E-04	1.13E-04
Nitrogen	*lbmole/hr*	2.45E-01	2.45E-01	3.96E-05	1.54E-01	1.54E-01	3.53E-01	1.30E-07	5.07E-01	6.79E-05	1.15E-05
Oxygen	*lbmole/hr*	3.86E-02	3.86E-02	3.31E-06	4.21E-03	4.21E-03	7.61E-04	1.79E-10	4.97E-03	2.64E-07	4.50E-08
Carbon Dioxide	*lbmole/hr*	1.26E-10	1.26E-10	9.08E-02	3.44E+01	3.45E+01	8.60E-03	9.56E-08	3.45E+01	9.78E-06	1.71E-06
Methane	*lbmole/hr*	1.63E+01	1.63E+01	7.74E-06	2.99E-03	3.00E-03	1.62E+01	1.89E-08	1.62E+01	7.38E-03	1.27E-03
Ethane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.53E-01	1.60E-11	3.52E-01	6.58E-04	1.17E-04
Propane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.53E-01	2.45E-13	3.51E-01	1.96E-03	3.57E-04
Butane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	5.07E+00	3.19E-14	4.99E+00	8.39E-02	1.55E-02
Pentane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.56E+00	1.39E-16	4.36E+00	2.12E-01	3.94E-02
Hexane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.10E+00	3.99E-19	3.61E+00	4.99E-01	9.29E-02
Heptane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.69E+00	6.70E-22	2.67E+00	1.02E+00	1.89E-01
Octane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.32E+00	7.98E-25	1.61E+00	1.71E+00	3.00E-01
Nonane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.98E+00	6.53E-28	7.79E-01	2.21E+00	2.55E-01
Decane	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.68E+00	1.66E-31	3.15E-01	2.37E+00	7.35E-02
n-C11	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.41E+00	7.89E-35	1.13E-01	2.30E+00	1.59E-02
n-C12	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	2.17E+00	1.49E-38	4.31E-02	2.13E+00	4.71E-03
n-C13	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.95E+00	6.13E-42	1.33E-02	1.94E+00	1.28E-03
n-C14	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.76E+00	9.56E-46	3.88E-03	1.75E+00	2.60E-04
n-C15	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.58E+00	7.80E-50	1.61E-03	1.58E+00	7.09E-05
n-C16	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.42E+00	5.83E-54	5.53E-04	1.42E+00	1.50E-05
n-C17	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.28E+00	1.11E-59	2.33E-04	1.28E+00	3.39E-06
n-C18	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.15E+00	5.89E-66	9.25E-05	1.15E+00	7.52E-07
n-C19	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.03E+00	7.32E-73	3.99E-05	1.03E+00	1.90E-07
n-C20	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	9.29E-01	8.08E-71	9.08E-06	9.29E-01	1.99E-08
n-C21	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	8.36E-01	5.46E-75	3.45E-06	8.36E-01	4.21E-09
n-C22	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	7.52E-01	1.72E-79	1.42E-06	7.52E-01	1.02E-09
n-C23	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	6.76E-01	1.73E-83	4.35E-07	6.76E-01	1.59E-10
n-C24	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	6.08E-01	9.27E-88	1.51E-07	6.08E-01	3.04E-11
n-C25	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	5.47E-01	2.03E-91	5.87E-08	5.47E-01	6.81E-12
n-C26	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.92E-01	2.90E-96	1.94E-08	4.92E-01	1.22E-12
n-C27	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	4.42E-01	6.33E-100	5.46E-09	4.42E-01	1.74E-13
n-C28	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.98E-01	1.48E-101	2.57E-09	3.98E-01	5.35E-14
n-C29	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	3.58E-01	1.33E-101	1.14E-09	3.58E-01	1.50E-14
n-C30	*lbmole/hr*	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.95E+00	7.26E-101	7.96E-10	1.95E+00	2.05E-15
Actual Volume Flow	*barrel/day*	9913	16811	106	245089	245195	2921	15	730585	624	9960
Mass Density	*lb/ft3*	0.92	0.54	61.16	62.87	62.87	14.27	61.45	21.12	46.46	0.29
Molar Density	*lbmole/ft3*	0.083	0.049	3.390	3.489	3.489	0.195	3.411	1.172	0.211	0.003

Small Unit		21	22	23	24	25	26	27	28	29	30	31
		Condensed OVHD	Reflux	Naphtha	Condensate Drum Vapor	Light Gases	Reboiler Feed	Boilup	Hydro-Isom Unit Feed	Diesel Product	Wax Product	Separator Vapor
Vapour Fraction		0.01	0.00	1.00	1.00	1.00	0.00	1.00	0.00	0.00	0.00	[illegible]
Temperature	*F*	183.9	183.9	20.0	4.5	4.5	430.7	510.1	20.0	425.7	774.0	[illegible]
Pressure	*psia*	20.0	20.0	183.9	131.0	131.0	26.0	26.0	183.9	14.7	20.0	[illegible]
Molar Flow	*lbmole/hr*	6.0	0.7	0.1	90.4	90.4	76.5	51.1	5.3	18.36	6.992	[illegible]
Mass Flow	*lb/hr*	676.1	74.0	3.6	2752.6	2756.2	14418.7	8238.9	597.8	3640	2539	[illegible]
Liquid Volume Flow	*barrel/day*	65.8	7.2	0.4	301.0	301.4	1297.7	754.6	58.2	326	216	[illegible]
Mass Enthalpy	*Btu/lb*	-888.5	-885.3	-1499.6	-2421.3	-2420.1	-659.6	-519.2	-885.3	-660.7	-335.5	[illegible]
Component Fraction												
Hydrogen	*lbmole/hr*	1.00E-04	1.10E-05	5.39E-04	6.01E+00	6.01E+00	2.24E-14	3.34E-14	5.81E-05	0.00E+00	0.00E+00	3.47E-15
Water	*lbmole/hr*	9.50E-03	4.91E-03	2.77E-02	4.56E+01	4.56E+01	3.60E-10	5.36E-10	2.60E-02	0.00E+00	0.00E+00	1.43E-10
Carbon Monoxide	*lbmole/hr*	1.13E-04	1.99E-05	5.64E-04	3.00E+00	3.00E+00	8.72E-14	5.97E-14	1.06E-04	0.00E+00	0.00E+00	1.65E-14
Nitrogen	*lbmole/hr*	1.15E-05	1.91E-06	5.78E-05	3.53E-01	3.53E-01	7.99E-15	1.19E-14	1.01E-05	0.00E+00	0.00E+00	1.50E-15
Oxygen	*lbmole/hr*	4.50E-08	1.17E-08	2.02E-07	7.61E-04	7.61E-04	1.15E-16	1.72E-16	6.21E-08	0.00E+00	0.00E+00	2.74E-17
Carbon Dioxide	*lbmole/hr*	1.71E-06	8.37E-07	5.35E-06	8.59E-03	8.60E-03	4.01E-14	1.30E-13	4.44E-06	0.00E+00	0.00E+00	1.41E-14
Methane	*lbmole/hr*	1.27E-03	4.18E-04	5.16E-03	1.62E+01	1.62E+01	7.00E-12	1.04E-11	2.22E-03	0.00E+00	0.00E+00	1.92E-12
Ethane	*lbmole/hr*	1.17E-04	7.55E-05	2.58E-04	3.52E-01	3.52E-01	1.29E-11	1.91E-11	4.00E-04	0.00E+00	0.00E+00	6.07E-12
Propane	*lbmole/hr*	3.57E-04	2.97E-04	3.85E-04	3.51E-01	3.51E-01	4.00E-10	5.92E-10	1.57E-03	0.00E+00	0.00E+00	2.89E-10
Butane	*lbmole/hr*	1.55E-02	1.45E-02	7.13E-03	4.99E+00	4.99E+00	1.82E-07	2.68E-07	7.68E-02	0.00E+00	0.00E+00	2.02E-07
Pentane	*lbmole/hr*	3.94E-02	3.86E-02	7.59E-03	4.35E+00	4.36E+00	4.29E-06	6.29E-06	2.05E-01	0.00E+00	0.00E+00	7.15E-06
Hexane	*lbmole/hr*	9.29E-02	9.27E-02	7.47E-03	3.60E+00	3.61E+00	9.06E-05	1.31E-04	4.92E-01	0.00E+00	0.00E+00	2.24E-04
Heptane	*lbmole/hr*	1.89E-01	1.91E-01	6.51E-03	2.67E+00	2.67E+00	1.52E-03	2.16E-03	1.01E+00	0.00E+00	0.00E+00	5.44E-03
Octane	*lbmole/hr*	3.00E-01	3.03E-01	4.42E-03	1.61E+00	1.61E+00	1.89E-02	2.64E-02	1.61E+00	0.00E+00	0.00E+00	9.69E-02
Nonane	*lbmole/hr*	2.55E-01	2.58E-01	1.66E-03	7.77E-01	7.79E-01	1.16E-01	1.57E-01	1.37E+00	0.00E+00	0.00E+00	8.37E-01
Decane	*lbmole/hr*	7.35E-02	7.43E-02	2.15E-04	3.15E-01	3.15E-01	1.96E-01	2.54E-01	3.94E-01	0.00E+00	0.00E+00	1.97E+00
n-C11	*lbmole/hr*	1.59E-02	1.61E-02	2.06E-05	1.13E-01	1.13E-01	1.62E-01	1.99E-01	8.54E-02	2.51E+00	0.00E+00	2.21E+00
n-C12	*lbmole/hr*	4.71E-03	4.76E-03	2.95E-06	4.31E-02	4.31E-02	1.16E-01	1.32E-01	2.52E-02	2.39E+00	0.00E+00	2.10E+00
n-C13	*lbmole/hr*	1.28E-03	1.30E-03	3.37E-07	1.33E-02	1.33E-02	8.13E-02	8.38E-02	6.87E-03	2.19E+00	0.00E+00	1.93E+00
n-C14	*lbmole/hr*	2.60E-04	2.63E-04	2.72E-08	3.88E-03	3.88E-03	5.69E-02	5.09E-02	1.39E-03	1.99E+00	0.00E+00	1.75E+00
n-C15	*lbmole/hr*	7.09E-05	7.18E-05	4.03E-09	1.61E-03	1.61E-03	4.32E-02	3.38E-02	3.80E-04	1.79E+00	0.00E+00	1.58E+00
n-C16	*lbmole/hr*	1.50E-05	1.52E-05	4.02E-10	5.53E-04	5.53E-04	3.34E-02	2.22E-02	8.04E-05	1.61E+00	0.00E+00	1.42E+00
n-C17	*lbmole/hr*	3.39E-06	3.42E-06	5.09E-11	2.33E-04	2.33E-04	2.60E-02	1.39E-02	1.82E-05	1.45E+00	0.00E+00	1.28E+00
n-C18	*lbmole/hr*	7.52E-07	7.61E-07	6.15E-12	9.25E-05	9.25E-05	2.10E-02	8.95E-03	4.03E-06	1.30E+00	0.00E+00	1.15E+00
n-C19	*lbmole/hr*	1.90E-07	1.93E-07	9.19E-13	3.99E-05	3.99E-05	1.75E-02	5.89E-03	1.02E-06	1.17E+00	0.00E+00	1.03E+00
n-C20	*lbmole/hr*	1.99E-08	2.02E-08	3.01E-14	9.08E-06	9.08E-06	1.46E-02	3.64E-03	1.07E-07	1.05E+00	0.00E+00	9.29E-01
n-C21	*lbmole/hr*	4.21E-09	4.26E-09	3.28E-15	3.45E-06	3.45E-06	1.25E-02	2.39E-03	2.26E-08	0.00E+00	2.37E-01	8.36E-01
n-C22	*lbmole/hr*	1.02E-09	1.03E-09	4.40E-16	1.42E-06	1.42E-06	1.09E-02	1.60E-03	5.47E-09	0.00E+00	2.13E-01	7.52E-01
n-C23	*lbmole/hr*	1.59E-10	1.61E-10	2.95E-17	4.35E-07	4.35E-07	9.51E-03	1.00E-03	8.52E-10	0.00E+00	1.92E-01	6.76E-01
n-C24	*lbmole/hr*	3.04E-11	3.07E-11	2.72E-18	1.51E-07	1.51E-07	8.39E-03	6.53E-04	1.63E-10	0.00E+00	1.73E-01	6.08E-01
n-C25	*lbmole/hr*	6.81E-12	6.88E-12	3.20E-19	5.87E-08	5.87E-08	7.44E-03	4.35E-04	3.65E-11	0.00E+00	1.55E-01	5.47E-01
n-C26	*lbmole/hr*	1.22E-12	1.24E-12	2.70E-20	1.94E-08	1.94E-08	6.62E-03	2.80E-04	6.56E-12	0.00E+00	1.40E-01	4.92E-01
n-C27	*lbmole/hr*	1.74E-13	1.76E-13	1.55E-21	5.46E-09	5.46E-09	5.90E-03	1.75E-04	9.32E-13	0.00E+00	1.25E-01	4.42E-01
n-C28	*lbmole/hr*	5.35E-14	5.41E-14	2.99E-22	2.57E-09	2.57E-09	5.28E-03	1.22E-04	2.87E-13	0.00E+00	1.13E-01	3.98E-01
n-C29	*lbmole/hr*	1.50E-14	1.52E-14	4.95E-23	1.14E-09	1.14E-09	4.74E-03	8.51E-05	8.04E-14	0.00E+00	1.01E-01	3.58E-01
n-C30	*lbmole/hr*	2.05E-15	2.08E-15	1.12E-24	7.96E-10	7.96E-10	2.57E-02	2.20E-04	1.10E-14	0.00E+00	5.54E+00	1.95E+00
Actual Volume Flow	*barrel/day*	172	8	101	542756	543235	1656	81398	63	411	339	[illegible]
Mass Density	*lb/ft3*	16.82	40.46	0.15	0.02	0.02	37.22	0.43	40.44	37.86	32.05	[illegible]
Molar Density	*lbmole/ft3*	0.150	0.359	0.003	0.001	0.001	0.197	0.003	0.359	0.191	0.088	0.151

Table 2: Small Unit PFD Stream Tables

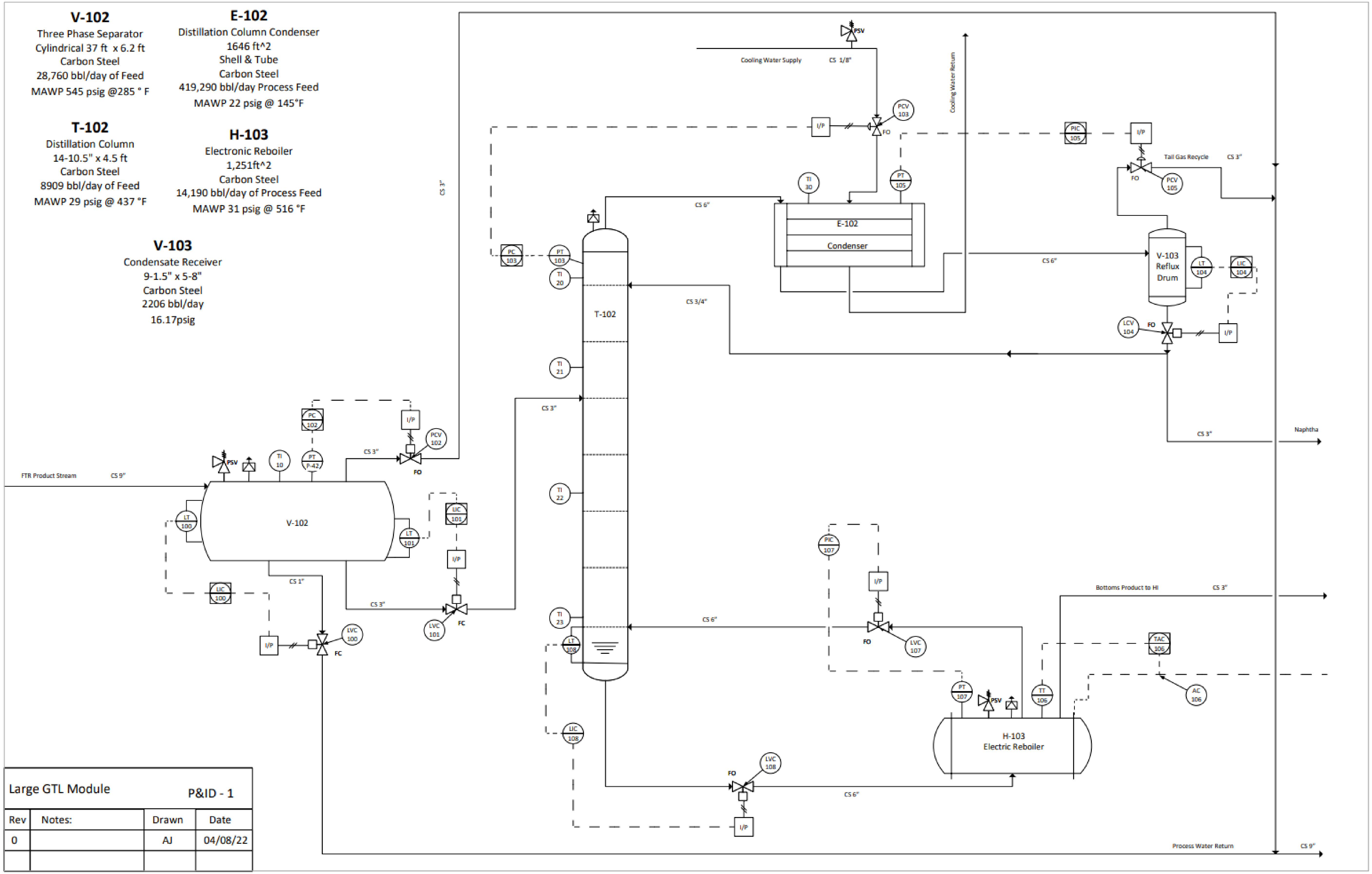

Figure 1: Large GTL Module P&ID

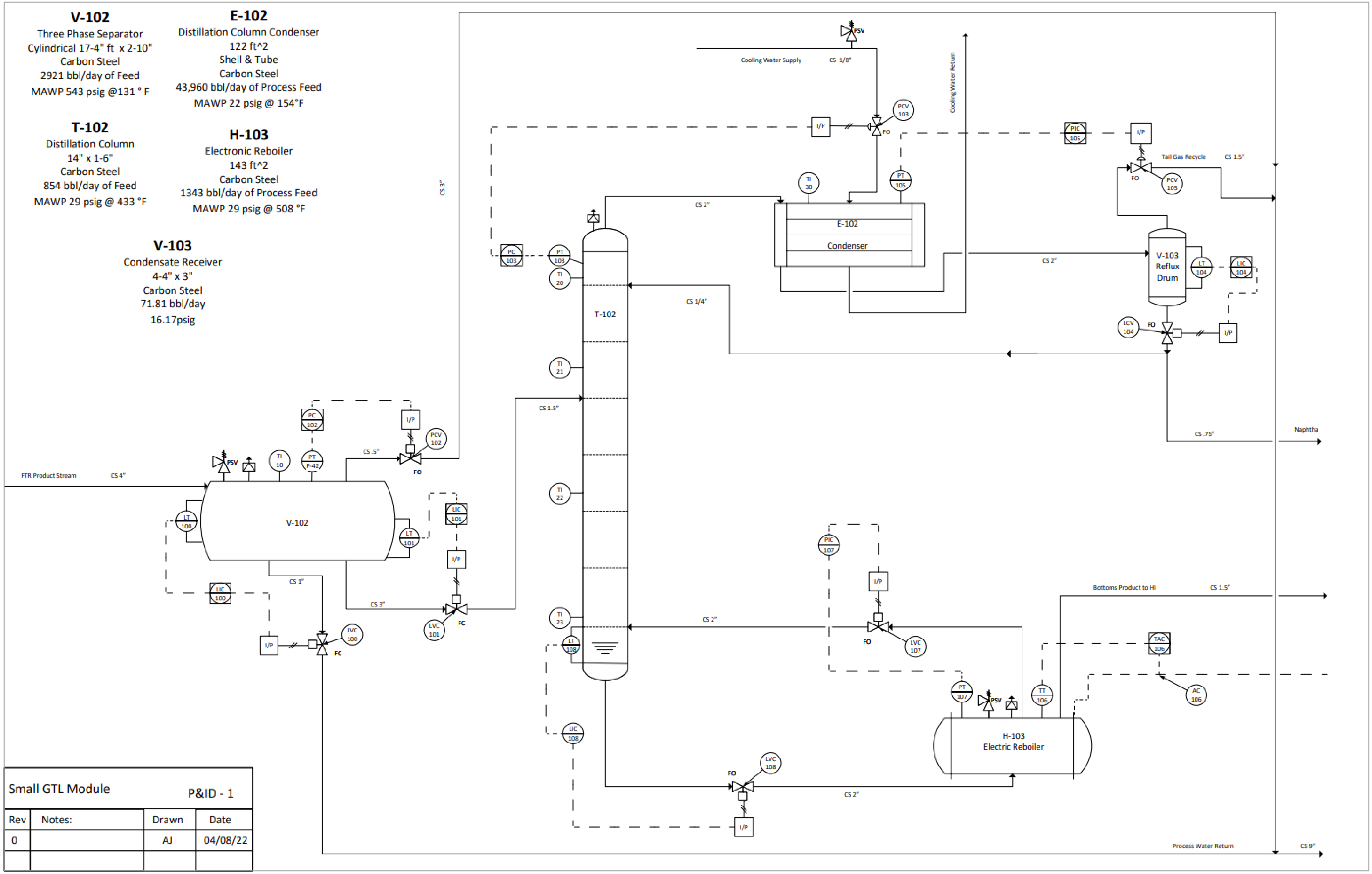

Figure 2: Small GTL Module P&ID

Transportation of the Modules Total Per Year																					
Year		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
Total		$ 4,200,000	$ -	$ 1,700,000	$ 1,600,000	$ 2,300,000	$ 3,600,000	$ 2,600,000	$ 3,600,000	$ 1,700,000	$ 3,100,000	$ 2,200,000	$ 2,500,000	$ 1,900,000	$ 3,500,000	$ 800,000	$ 1,300,000	$ 1,500,000	$ 600,000	$ 500,000	$ -

Transportation of the Modules Total Per Year																					
Year		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
Total		$ 4,200,000	$ -	$ 1,700,000	$ 1,600,000	$ 2,300,000	$ 3,600,000	$ 2,600,000	$ 3,600,000	$ 1,700,000	$ 3,100,000	$ 2,200,000	$ 2,500,000	$ 1,900,000	$ 3,500,000	$ 800,000	$ 1,300,000	$ 1,500,000	$ 600,000	$ 500,000	$ -
M_4	Margaret	99	0	0	0	0	127	0	0	58	0	0	127	0	65	0	94	0	0	0	0
M_5	Mary	99	0	0	99	0	0	6	90	0	0	120	0	0	95	0	0	0	0	0	0
M_6	Rachel	13	0	0	0	95	0	0	122	0	34	0	102	0	0	0	0	0	0	0	0
M_7	Shirley	0	0	0	0	0	0	0	0	13	38	0	0	67	0	0	0	40	0	0	0
S_1	Alexa	42	0	0	40	0	0	0	0	38	0	0	45	0	64	0	49	15	0	0	0
S_3	Bessie	0	0	6	0	0	0	42	0	58	0	97	0	65	0	58	0	0	0	63	0
S_4	Beth	99	0	0	0	0	0	90	12	0	87	0	0	92	0	0	0	92	0	0	0
S_5	Caroline	99	0	0	0	0	0	90	12	0	0	0	0	12	0	79	104	106	0	0	0
S_6	Dijanna	99	0	93	0	0	42	0	0	58	0	0	92	0	67	0	67	0	0	0	0
S_7	Dorothy	0	0	6	0	0	42	0	0	0	58	0	0	85	0	0	0	98	84	0	0
S_8	Ella	13	0	79	0	0	38	99	0	38	0	78	0	44	44	0	0	0	0	38	0
S_9	Evelyn	13	0	79	0	0	0	79	0	0	15	0	15	0	81	0	0	0	37	0	0
S_10	France	0	0	0	0	91	0	0	0	0	110	78	0	85	59	90	0	0	0	85	0
S_11	Gertrude	0	0	6	0	0	42	0	0	58	0	0	0	63	85	120	99	0	117	122	0
S_12	Jane	0	0	42	75	0	0	90	0	0	15	0	0	0	0	0	15	106	98	0	0

Travel Cost																					
Number ID	Name	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Year 14	Year 15	Year 16	Year 17	Year 18	Year 19	Year 20
L_1	April	$ 630,000	$ -	$ -	$ 605,080	$ -	$ 566,029	$ -	$ -	$ -	$ 1,380,496	$ -	$ -	$ -	$ 1,325,051	$ -	$ -	$ -	$ -	$ -	$ -
L_2	Genna	$ 195,000	$ -	$ -	$ -	$ 1,431,542	$ -	$ 982,405	$ 999,014	$ -	$ -	$ -	$ -	$ 593,487	$ -	$ -	$ -	$ -	$ -	$ -	$ -
M_1	Hedy	$ 315,000	$ -	$ 272,621	$ -	$ -	$ 42,961	$ -	$ -	$ -	$ 622,652	$ -	$ 493,434	$ -	$ -	$ -	$ -	$ 572,588	$ -	$ -	$ -
M_2	Irene	$ 741,000	$ -	$ 283,576	$ -	$ -	$ 824,466	$ 491,202	$ 916,432	$ -	$ -	$ 916,432	$ -	$ -	$ 278,239	$ -	$ -	$ -	$ -	$ -	$ -
M_3	Katherine	$ 97,500	$ -	$ 593,596	$ -	$ -	$ 824,466	$ 287,065	$ -	$ 638,821	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
M_4	Margaret	$ 741,000	$ -	$ -	$ -	$ -	$ 951,274	$ -	$ -	$ 432,817	$ -	$ -	$ 956,228	$ -	$ 484,264	$ -	$ 702,801	$ -	$ -	$ -	$ -
M_5	Mary	$ 741,000	$ -	$ -	$ 741,219	$ -	$ -	$ 42,961	$ 671,670	$ -	$ -	$ 903,136	$ -	$ -	$ 709,061	$ -	$ -	$ -	$ -	$ -	$ -
M_6	Rachel	$ 97,500	$ -	$ -	$ -	$ 715,771	$ -	$ -	$ 916,432	$ -	$ 258,512	$ -	$ 765,236	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
M_7	Shirley	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ 97,500	$ 287,065	$ -	$ -	$ 499,507	$ -	$ -	$ -	$ 296,744	$ -	$ -	$ -
S_1	Alexa	$ 63,000	$ -	$ -	$ 60,508	$ -	$ -	$ -	$ -	$ 56,603	$ -	$ -	$ 67,972	$ -	$ 96,433	$ -	$ 74,151	$ 22,506	$ -	$ -	$ -
S_2	Annie	$ 63,000	$ -	$ -	$ 60,508	$ -	$ 56,603	$ -	$ -	$ 86,563	$ 51,702	$ -	$ -	$ 59,349	$ 96,853	$ 191,246	$ -	$ -	$ -	$ -	$ -
S_3	Bessie	$ -	$ -	$ 9,300	$ -	$ -	$ -	$ 62,693	$ -	$ 86,563	$ -	$ 145,986	$ -	$ 98,240	$ -	$ 86,563	$ -	$ -	$ -	$ 94,921	$ -
S_4	Beth	$ 148,200	$ -	$ -	$ -	$ -	$ -	$ 135,117	$ 18,109	$ -	$ 130,891	$ -	$ -	$ 138,050	$ -	$ -	$ -	$ 138,050	$ -	$ -	$ -
S_5	Caroline	$ 148,200	$ -	$ -	$ -	$ -	$ -	$ 135,117	$ 18,109	$ -	$ -	$ -	$ -	$ 18,109	$ -	$ 118,719	$ 155,580	$ 159,706	$ -	$ -	$ -
S_6	Dijanna	$ 148,200	$ -	$ 139,653	$ -	$ -	$ 62,693	$ -	$ -	$ 86,563	$ -	$ -	$ 137,604	$ -	$ 100,325	$ -	$ 100,325	$ -	$ -	$ -	$ -
S_7	Dorothy	$ -	$ -	$ 9,300	$ -	$ -	$ 62,693	$ -	$ -	$ -	$ 86,563	$ -	$ -	$ 127,764	$ -	$ -	$ -	$ 146,550	$ 125,489	$ -	$ -
S_8	Ella	$ 19,500	$ -	$ 118,719	$ -	$ -	$ 56,715	$ 148,244	$ -	$ 56,603	$ -	$ 117,000	$ -	$ 65,949	$ 65,949	$ -	$ -	$ -	$ -	$ 57,222	$ -
S_9	Evelyn	$ 19,500	$ -	$ 118,719	$ -	$ -	$ -	$ 118,719	$ -	$ -	$ 22,192	$ -	$ 22,217	$ -	$ 121,252	$ -	$ -	$ -	$ 55,648	$ -	$ -
S_10	France	$ -	$ -	$ -	$ -	$ 136,500	$ -	$ -	$ -	$ -	$ 164,893	$ 117,000	$ -	$ 126,762	$ 88,987	$ 134,334	$ -	$ -	$ -	$ 128,025	$ -
S_11	Gertrude	$ -	$ -	$ 9,300	$ -	$ -	$ 62,693	$ -	$ -	$ 86,563	$ -	$ -	$ -	$ 94,921	$ 126,762	$ 180,627	$ 148,421	$ -	$ 175,369	$ 183,669	$ -
S_12	Jane	$ -	$ -	$ 63,000	$ 112,820	$ -	$ -	$ 135,117	$ -	$ -	$ 22,192	$ -	$ -	$ -	$ -	$ -	$ 22,217	$ 159,706	$ 146,959	$ -	$ -
S_13	June																				

Table 3: Transportation Cost per Year

Column Sizing

Dedecanizer

Number of Stages	Feed Entering Stage	OVHD Pressure	Reboiler Pressure	Reflux Ratio	Condenser Duty (BTU/hr)	Reboiler Duty (BTU/hr)	N*RR	Number of Trays	Feed Entering Tray	Tray Diameter (ft)	Tray Spacing (ft)	Tower Height (ft)	Tower Volume (ft^3)	Tray Area (ft^2)	Tray Diameter (m)	Tray Spacing (m)	Tower Height (m)	Tower Volume (m^3)	Tray Area (m^2)
4	2	20	27.9	0.122	4.651E+06	1.300E+07	0.49	7	3	4.50	2.25	14.86	236.31	15.90	1.3716	0.686	4.53	6.69	1.48
4	2	20	27.9	0.122	1.779E+06	5.607E+06	0.49	7	3	3.00	2.50	15.10	106.75	7.07	0.9144	0.762	5.11	3.36	0.66
4	2	20	27.9	0.122	3.430E+05	1.189E+06	0.49	7	3	1.50	2.00	15.31	27.05	1.77	0.4572	0.610	4.15	0.68	0.16

Table 4: Distillation Column Sizing for Each Unit Size

Costing

Dedecanizer

Cp0,Tower	Cp0,Trays	Fp,Tower	Fp,Trays	Fq,Trays	CBM,Trays	CBM,Tower	CBM,Total (2001)	CBM,Total (2022)
8672.49	1207.78	1.19	1.00	1.00	$7,976	$38,303	$46,279	$85,098
5760.09	844.43	0.96		1.00	$5,668	$23,027	$28,696	$52,765
2647.28	773.45	0.73		1.00	$5,261	$9,474	$14,736	$27,096

Table 5: Distillation Column Costing for Each Unit Size

Heat Exchangers	E-101	E-102	Heaters	H-101	H-102	H-103		
Type	U-Tube	U-Tube	Type	Electric Heater	Electric Heater	Electric Reboiler		
Area (m2)	N/A	31.1	MOC	CS	CS	CS		
Duty (BTU/hr)	56,300,000	4,650,000	Duty (Btu/hr)	16,100,000	6,710,000	13,000,000		
Shell								
Temperature (°C)	886	141						
Pressure (bar)	34.1	1.59						
Phase	Vapor → Liquid	Vapor → Liquid						
MOC	CS	CS						
Tube								
Temperature (°C)	32.2	32.2						
Pressure (bar)	3.45	3.45						
Phase	Liquid	Liquid						
MOC	CS	CS						
Vessels/Towers/ Reactors	**V-101**	**V-102**	**V-103**	**T-101**	**T-102**	**T-103**	**R-101**	**R-102**
Temperature (°C)	62.8	140	67.2	26.7	221	403	888	254
Pressure (bar)	33.9	34.1	1.59	33.8	5.37	1.38	34.1	33.4
Orientation	N/A	Horizontal	Vertical	Vertical	Vertical	Vertical	Horizontal	Horizontal
MOC	CS	CS	CS	CS	CS	CS	CS	CS
Size								
Height/Length (m)	N/A	11.5	2.79	N/A	4.53	N/A	N/A	2.87
Diameter (m)	N/A	1.89	1.73	N/A	1.37	N/A	N/A	1.524
Internals	N/A	Heavy Liquid Boot	Mist Eliminator	N/A	Sieve Trays	N/A	N/A	42 Tubes

Table 6: Large GTL Plant Process Equipment Summary

Heat Exchangers	E-101	E-102	Heaters	H-101	H-102	H-103		
Type	U-Tube	U-Tube	Type	Electric Heater	Electric Heater	Electric Reboiler		
Area (m2)	N/A	2.29	MOC	CS	CS	CS		
Duty (BTU/hr)	5,630,000	343,000	Duty (Btu/hr)	1,610,000	545,000	1,190,000		
Shell								
Temperature (°C)	886	141						
Pressure (bar)	34.1	1.59						
Phase	Vapor → Liquid	Vapor → Liquid						
MOC	CS	CS						
Tube								
Temperature (°C)	32.2	32.2						
Pressure (bar)	3.45	3.45						
Phase	Liquid	Liquid						
MOC	CS	CS						
Vessels/Towers/ Reactors	**V-101**	**V-102**	**V-103**	**T-101**	**T-102**	**T-103**	**R-101**	**R-102**
Temperature (°C)	62.8	140	67.2	26.7	221	403	888	254
Pressure (bar)	33.9	34.1	1.59	33.8	5.37	1.38	34.1	33.4
Orientation	N/A	Horizontal	Vertical	Vertical	Vertical	Vertical	Horizontal	Horizontal
MOC	CS	CS	CS	CS	CS	CS	CS	CS
Size								
Height/Length (m)	N/A	5.27	1.30	N/A	4.15	N/A	N/A	1.78
Diameter (m)	N/A	0.86	0.91	N/A	0.457	N/A	N/A	0.75
Internals	N/A	Heavy Liquid Boot	Mist Eliminator	N/A	Sieve Trays	N/A	N/A	42 Tubes

Table 7: Small GTL Plant Process Equipment Summary

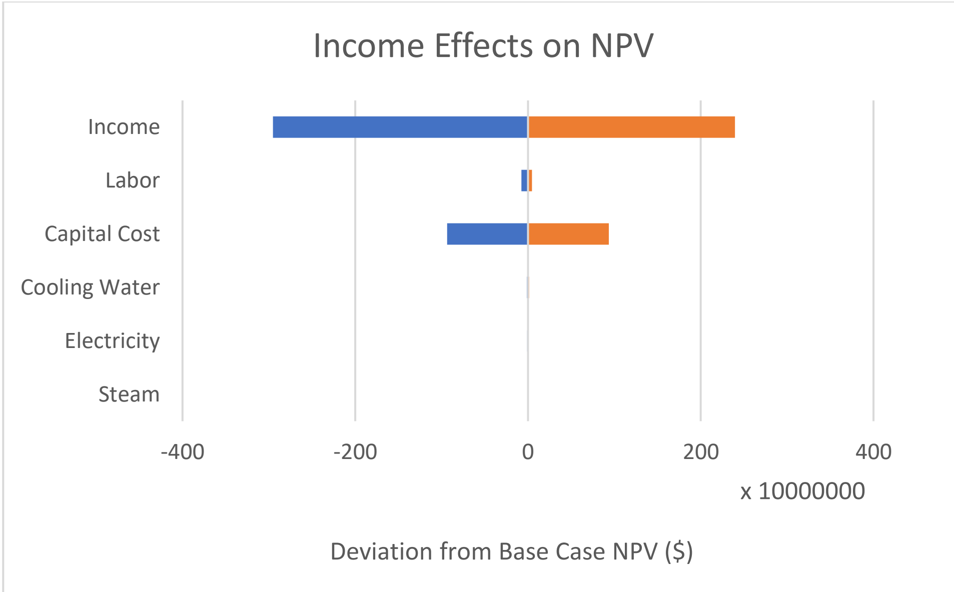

Figure 3: Income Effects on NPV

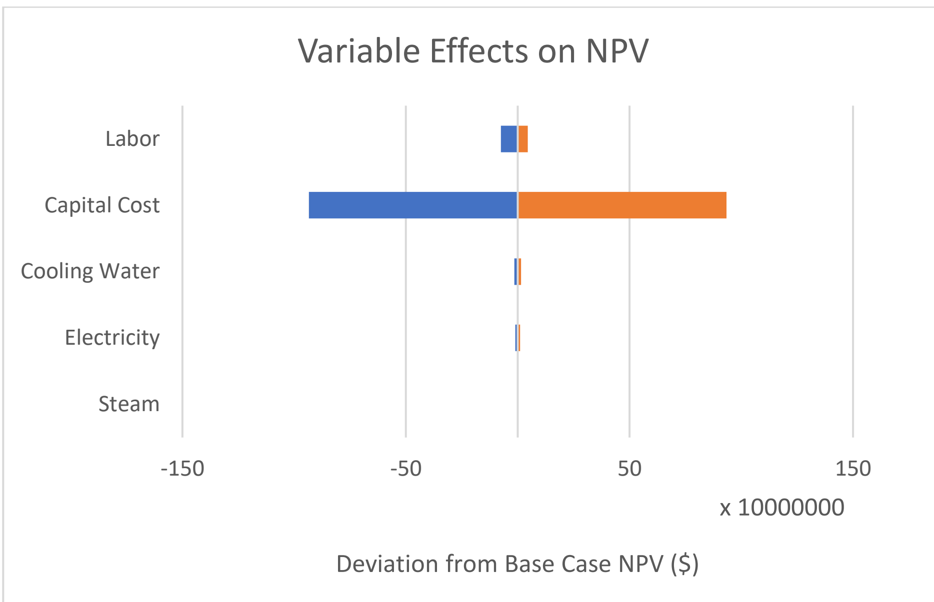

Figure 4: Variable Effects on NPV

FOAK Plant Cost LARGE									
Costing Heuristics from Turton									
Equipment Type	**Equipment Description**	**Name/Tag**	**MOC**	**Equipment Cost Attribute**	**Units**	**Pressure (barg)**	**Diameter (m)**		**Final Cost**
HEX	Shell & Tube (U-Tube)	Condenser	CS & CS					$	134,000
HEX	Shell & Tube (U-Tube)	Reboiler	CS & CS					$	347,000
FTR	Jacketed Nonagitated (similar to price of "process units"		CS	5.20	m^3	35.90	1.52	$	226,000
Seperator	Horizontial	Three Phase Seperator	CS	32.57	m^3	33.12	1.89	$	452,000
Seperator	Vertical		CS	6.69	m^3	5.37	1.37	$	71,000
Receiver			CS			2001 Cost:	21770.05	$	41,000
Trays	Seive		CS			2001 Cost:	7976	$	15,000
							Turton Total	$	1,286,000

Project Prompt Costing								
	Base Cost	**Base Capacity**	**Units**	**Scaling Factor**	**Operating Capacity**	**Units**		**Cost**
Syngas Unit	21,620,000	12.2	kg/s	0.67	5.1	kg/s	$	11,991,000
HI Unit	8,290,000	1.13	kg/s	0.55	7.9	kg/s	$	24,116,000
CO2 Recovery	5,300,000	5.54	kg/s	0.55	1.7	kg/s	$	2,735,000
Steam Plant	**Unit Cost:**	300	**Usage:**	400	Hp		$	120,000
						Packet Total	$	38,962,000
						Total Cost for Large Unit	**$**	**40,248,000**

Table 7: FOAK Plant Cost for Large Unit

FOAK Plant Cost MEDIUM									
Costing Heuristics from Turton									
Equipment Type	**Equipment Description**	**Name/Tag**	**MOC**	**Equipment Cost Attribute**	**Units**	**Pressure (barg)**	**Diameter (m)**		**Final Cost**
HEX	Shell & Tube (U-Tube)	Condenser	CS & CS					$	105,000
HEX	Shell & Tube (U-Tube)	Reboiler	CS & CS					$	257,000
FTR	Jacketed Nonagitated (similar to price of "process units"		CS	2.80	m^3	35.90	1.00	$	163,000
Seperator	Horizontial	Three Phase Seperator	CS	12.35	m^3			$	50,000
Seperator	Vertical		CS	3.36	m^3	5.37	0.91	$	43,000
Condensate Receiver			CS			2001 Cost:	14847.186	$	28,000
Trays	Seive		CS			2001 Cost:	5668	$	11,000
							Turton Total	$	657,000

Project Prompt Costing								
	Base Cost	**Base Capacity**	**Units**	**Scaling Factor**	**Operating Capacity**	**Units**		**Cost**
Syngas Unit	21,620,000	12.2	kg/s	0.67	2.5	kg/s	$	7,552,000
HI Unit	8,290,000	1.13	kg/s	0.55	3.5	kg/s	$	15,553,000
CO2 Recovery	5,300,000	5.54	kg/s	0.55	0.83	kg/s	$	1,868,000
Steam Plant	**Unit Cost:**	300	**Usage:**	202	Hp		$	61,000
						Packet Total	$	25,034,000
						Total Cost for Medium Unit	**$**	**25,691,000**

Table 8: FOAK Plant Cost for Medium Unit

FOAK Plant SMALL									
Costing Heuristics from Turton									
Equipment Type	**Equipment Description**	**Name/Tag**	**MOC**	**Equipment Cost Attribute**	**Units**	**Pressure (barg)**	**Diameter (m)**		**Final Cost**
HEX	Shell & Tube (U-Tube)	Condenser	CS & CS					$	95,000
HEX	Shell & Tube (U-Tube)	Reboiler	CS & CS					$	193,000
FTR	Jacketed Nonagitated (similar to price of "process units"		CS	0.80	m^3	35.90	0.75	$	84,000
Seperator	Horizontial	Three Phase Seperator	CS	3.09	m^3			$	24,000
Seperator	Vertical		CS	0.68	m^3	5.37	0.46	$	18,000
Condensate Receiver			CS			2001 Cost:	8247.97	$	16,000
Trays	Seive		CS			2001 Cost:	5261	$	10,000
						Turton Total		$	440,000

Project Prompt Costing								
	Base Cost	**Base Capacity**	**Units**	**Scaling Factor**	**Operating Capacity**	**Units**		**Cost**
Syngas Unit	21,620,000	12.2	kg/s	0.67	0.51	kg/s	$	2,565,000
HI Unit	8,290,000	1.13	kg/s	0.55	0.78	kg/s	$	6,755,000
CO2 Recovery	5,300,000	5.54	kg/s	0.55	0.17	kg/s	$	770,000
Steam Plant	**Unit Cost:**	300	**Usage:**	20	Hp		$	6,000
					Packet Total		$	10,096,000
					Total Cost for Small Unit		**$**	**10,536,000**

Table 9: FOAK Plant Cost for Small Unit

Heat Exchanger Design				Heat Exchanger Utilities				
Dedecanizer								
Condenser								
Condenser Duty (BTU/hr)	Ao (m^2)	Cp,0	Cbm,2022	Mass Flow Rate (lb/hr)	kGal/yr	S.F.	Adj. kgal/yr	Costs ($/yr)
4.651E+06	31.1	1.797E+04	$134,479	2.323E+05	2.440E+05	80%	1.952E+05	**$97,600**
1.779E+06	11.9	1.404E+04	$105,104	8.886E+04	9.333E+04		7.466E+04	**$37,332**
3.430E+05	2.3	1.276E+04	$95,498	1.713E+04	1.799E+04		1.440E+04	**$7,198**
Reboiler								
Reboiler Duty (BTU/hr)	Ao (m^2)	$Cp_{,0}$	Cbm,2022	kWh/h	kWh/yr	S.F.	Adj. kWh/yr	Costs ($/yr)
1.300E+07	157.0	2.800E+04	$346,608	3811.4	3.339E+07	80%	2.671E+07	**$1,068,402**
5.607E+06	67.7	2.078E+04	$257,145	1643.3	1.440E+07		1.152E+07	**$460,658**
1.189E+06	14.3	1.559E+04	$192,964	348.4	3.052E+06		2.442E+06	**$97,671**

Table 10: Condenser/Reboiler Sizing and Pricing for Each Unit

Syngas Preheater (electricity)					
Preheater Duty (BTU/hr)	kWh/h	kWh/yr	S.F.	Adj. kWh/yr	Costs ($/yr)
1.612E+07	4724.3	4.138E+07	80%	3.311E+07	**$1,324,317**
8.078E+06	2367.4	2.074E+07		1.659E+07	**$663,637**
1.612E+06	472.4	4.138E+06		3.311E+06	**$132,432**
Preheater (fuel gas)					
Preheater Duty (BTU/hr)	MBtu/hr	MBtu/yr	S.F.	Adj. MBtu/yr	Costs ($/yr)
1.612E+07	16120.0	1.412E+08	80%	1.130E+08	$338,906,880
8.078E+06	8078.0	7.076E+07		5.661E+07	$169,831,872
1.612E+06	1612.0	1.412E+07		1.130E+07	$33,890,688
Syngas Condenser					
Syngas Condenser Duty (BTU/hr)	Mass Flow Rate (lb/hr)	kGal/yr	S.F.	Adj. kgal/yr	Costs ($/yr)
5.627E+07	2.811E+06	2.952E+06	80%	2.362E+06	**$1,180,894**
2.822E+07	1.410E+06	1.481E+06		1.184E+06	**$592,231**
5.630E+06	2.812E+05	2.954E+05		2.363E+05	**$118,152**
FTR Preheater					
Preheater Duty (BTU/hr)	kWh/h	kWh/yr	S.F.	Adj. kWh/yr	Costs ($/yr)
6.710E+06	1966.5	1.723E+07	80%	1.378E+07	**$551,251**
2.671E+06	782.8	6.857E+06		5.486E+06	**$219,432**
5.449E+05	159.7	1.399E+06		1.119E+06	**$44,766**

Table 11: Heater Costing for each Unit

Dedecanizer HEX Details				
	Condenser (U-Tube)		**Reboiler (Thermosyphon)**	
	Temp. (°F)	Press. (psi)	Temp. (°F)	Press. (psi)
$T_{c,in}$	90	50	437	26.5
$T_{c,out}$	110	47	516.9	28
$T_{h,in}$	284.9	23	490	600
$T_{h,out}$	144.2	20	490	598.5
Cp/hfg	1.001	BTU/lbm*°F	1203.85	Btu/lb
U_0	150	Btu/(F*ft^2*hr)	200	Btu/(F*ft^2*hr)
F	0.9		1	
ΔT,lm	103.03		38.49	
Area (ft^2)	25	ft2	154	ft2

Table 12: Condenser/Reboiler Specifications (Steam on reboiler can be negated)

HEX Specifications			
U-Tube		Thermosyphon (Fixed Tube)	
K1	4.1884	K1	4.3247
K2	-0.25	K2	-0.303
K3	0.1974	K3	0.1634
C1	0.0388	C1	0.03881
C2	-0.113	C2	-0.11272
C3	0.0818	C3	0.08183
Fm	1.35	Fm	1.35
B1	1.63	B1	1.63
B2	1.66	B2	1.66
Teflon Tube			
K1	3.8062		
K2	0.8924		
K3	-0.167		
C1	0		
C2	0		
C3	0		
Fm	1		
B1	1.63		
B2	1.66		

Table 13: Additional Heat Exchanger Sizing Information

Tower/Tray Material Factors	
Fm, CS	1
FBM, CS	1
η_{Large}	60.57%
η_{Medium}	59.59%
η_{Small}	58.80%

Column Costing Factors		
	Tower	Trays (Sieve)
k1	3.4947	2.9949
k2	0.4485	0.4465
k3	0.1074	0.3961
B1	2.25	
B2	1.82	

Additional Tower Spacing	
Top Spacing (m)	Bottom Spacing (m)
1.2192	1.8288

CEPCI Factors	
CEPCI, 2001	397
CEPCI, 2021	730

Design Pressures		
Dedecanizer		
P, reb (psi)	P, reb adj.	P,reb (bar)
27.9	77.9	**5.37**

Pressure Drops		
ΔP Condens	6	psi
ΔP Reboiler	1.5	psi
ΔP Stage	0.1	psi

Pressure Factors			
Condenser		Reboiler	
Fp,debut	1.09	Fp,debut	2.28

Conversions	
psi→bar	0.0689476
Gal→lbs	8.34
ft→m	0.3048
ft^2→m^2	0.092903
psig→psia	14.3
BTU → kWh	0.00029307
°F → °C	

Table 14: Additional Distillation Column Sizing Information

Feed Conditions		
Variable	Value	Units
P_{Gauge}	8.3	psig
$P_{Absolute}$	23.0	psia
Mist Eliminator?		Yes
ρ_V	0.1389	$\frac{lb}{ft^3}$
ρ_L	41.02	
μ_V	0.01035	Cp
μ_L	0.3151	
W_V	4141	$\frac{lb}{hr}$
W_L	21174	

Atmospheric Pressure with Respect to Elevation		
Elev (ft)	0	ft
Elev (m)	0	m
$P_{atmospheric}$	101325	Pa
$P_{atmospheric}$	14.7	psi

Table 15: Large Condensate Receiver Feed Conditions

2-Phase Separator Sizing			
Variable	Equation	Value	Units
Q_V	$Q_V = \frac{W_V}{(3{,}600)(\rho_V)}$	8	$\frac{ft^3}{s}$
Q_L	$Q_L = \frac{W_L}{(60)(\rho_L)}$	8.60	$\frac{ft^3}{min}$
U_T	$U_T = K\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{1/2}$	6.16	$\frac{ft}{s}$
U_V	$U_T = 0.75U_T$	4.62	ft/s
V_H	$V_H = (T_H)(Q_L)$	86.031204	ft^3
V_S	$V_S = (T_S)(Q_L)$	43.015602	ft^3
L/D	From Table 4	1.5	
D	$D = \left(\frac{4(V_H + V_S)}{(\pi)(0.6)(L/D)}\right)^{1/3}$	**5.67**	ft
A_T	$A_T = \frac{\pi}{4}D^2$	25.3	ft^2
H_{LLL}	$H_{LLL} = 0.5D + 7$	9.83	in
A_{LLL}	From Table 5	2.2	ft^2
H_V		2	ft
A_V	From Table 5	8.91	ft^2
L	$L = \frac{V_H + V_S}{A_T - A_V - A_{LLL}}$	9.14	ft
φ	$\varphi = \frac{H_V}{U_V}$	0.433	s
U_{VA}	$U_{VA} = \frac{Q_V}{A_V}$	0.9	$\frac{ft}{s}$
L_{Min}	$L_{Min} = U_{VA}\varphi$	0.4	ft
L_{Actual}	**9.1**		ft
Volume	$V = (L_{Actual})(A_T)$	231	ft^3
		6.54	$\mathbf{m^3}$

Table 16: Large Condensate Receiver Sizing

Feed Conditions		
Variable	Value	Units
P_{Gauge}	8.3	psig
$P_{Absolute}$	23.0	psia
Mist Eliminator?		Yes
ρ_V	0.1539	$\frac{lb}{ft^3}$
ρ_L	40.47	
μ_V	0.009254	Cp
μ_L	0.2864	
W_V	15.94	$\frac{lb}{hr}$
W_L	3570.6	

Atmospheric Pressure with Respect to Elevation		
Elev (ft)	0	ft
Elev (m)	0	m
$P_{atmospheric}$	101325	Pa
$P_{atmospheric}$	14.7	psi

Table 17: Medium Condensate Receiver Feed Conditions

2-Phase Separator Sizing			
Variable	Equation	Value	Units
Q_V	$Q_V = \frac{W_V}{(3{,}600)(\rho_V)}$	0	$\frac{ft^3}{s}$
Q_L	$Q_L = \frac{W_L}{(60)(\rho_L)}$	1.47	$\frac{ft^3}{min}$
U_T	$U_T = K\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{1/2}$	5.81	$\frac{ft}{s}$
U_V	$U_T = 0.75U_T$	4.36	ft/s
V_H	$V_H = (T_H)(Q_L)$	14.70472	ft^3
V_S	$V_S = (T_S)(Q_L)$	7.3523598	ft^3
L/D	From Table 4	1.5	
D	$D = \left(\frac{4(V_H + V_S)}{(\pi)(0.6)(L/D)}\right)^{1/3}$	**3.15**	ft
A_T	$A_T = \frac{\pi}{4}D^2$	7.8	ft^2
H_{LLL}	$H_{LLL} = 0.5D + 7$	9.00	in
A_{LLL}	From Table 5	1.4	ft^2
H_V		2	ft
A_V	From Table 5	4.95	ft^2
L	$L = \frac{V_H + V_S}{A_T - A_V - A_{LLL}}$	15.56	ft
φ	$\varphi = \frac{H_V}{U_V}$	0.459	s
U_{VA}	$U_{VA} = \frac{Q_V}{A_V}$	0.0	$\frac{ft}{s}$
L_{Min}	$L_{Min} = U_{VA}\varphi$	0.0	ft
L_{Actual}	**15.6**		ft
Volume	$V = (L_{Actual})(A_T)$	121	ft^3
		3.4309	m^3

Table 18: Medium Condensate Receiver Sizing

Feed Conditions		
Variable	Value	Units
P_{Gauge}	8.3	psig
$P_{Absolute}$	23.0	psia
Mist Eliminator?		Yes
ρ_V	0.1542	$\frac{lb}{ft^3}$
ρ_L	40.46	
μ_V	0.009189	Cp
μ_L	0.2855	
W_V	3.636	$\frac{lb}{hr}$
W_L	671.82	
Atmospheric Pressure with Respect to Elevation		
Elev (ft)	0	ft
Elev (m)	0	m
$P_{atmospheric}$	101325	Pa
$P_{atmospheric}$	14.7	psi

Table 19: Small Condensate Receiver Feed Conditions

2-Phase Separator Sizing			
Variable	Equation	Value	Units
Q_V	$Q_V = \frac{W_V}{(3{,}600)(\rho_V)}$	0.0065	$\frac{ft^3}{s}$
Q_L	$Q_L = \frac{W_L}{(60)(\rho_L)}$	0.28	$\frac{ft^3}{min}$
U_T	$U_T = K\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{1/2}$	5.81	$\frac{ft}{s}$
U_V	$U_T = 0.75U_T$	4.36	ft/s
V_H	$V_H = (T_H)(Q_L)$	2.7674246	ft^3
V_S	$V_S = (T_S)(Q_L)$	1.3837123	ft^3
L/D	From Table 4	1.5	
D	$D = \left(\frac{4(V_H + V_S)}{(\pi)(0.6)(L/D)}\right)^{1/3}$	**3.00**	ft
A_T	$A_T = \frac{\pi}{4}D^2$	7.1	ft^2
H_{LLL}	$H_{LLL} = 0.5D + 7$	9.00	in
A_{LLL}	From Table 5	1.4	ft^2
H_V		2	ft
A_V	From Table 5	4.71	ft^2
L	$L = \frac{V_H + V_S}{A_T - A_V - A_{LLL}}$	4.26	ft
φ	$\varphi = \frac{H_V}{U_V}$	0.459	s
U_{VA}	$U_{VA} = \frac{Q_V}{A_V}$	0.0	$\frac{ft}{s}$
L_{Min}	$L_{Min} = U_{VA}\varphi$	0.0	ft
L_{Actual}	**4.3**		ft
Volume	$V = (L_{Actual})(A_T)$	30	ft^3
		0.8532	$\boldsymbol{m^3}$

Table 20: Small Condensate Receiver Sizing

$$\eta_{section} = 0.503(\mu_L \alpha)^{-0.226}$$

Large Unit

Dedecanizer			
	μ	K (light)	K (heavy)
Tray 1	0.2533	0.2713	0.1455
Tray 2	0.3548	0.4530	0.2539
Tray 3	0.2544	1.0445	0.6450
Tray 4	0.2285	1.6734	1.0921
Avg.	0.2728	0.8605	0.5341

α	1.6111
$\eta_{,Large}$	**60.57%**

Medium Unit

Dedecanizer			
	μ	K (light)	K (heavy)
Tray 1	0.2739	0.2205	0.1142
Tray 2	0.3926	0.3790	0.2069
Tray 3	0.2642	0.9271	0.5650
Tray 4	0.2296	1.4945567	0.9693002
Avg.	0.2901	0.7552933	0.4638618

α	1.6283
$\eta_{,Medium}$	**59.59%**

Small Unit

Dedecanizer			
	μ	K (light)	K (heavy)
Tray 1	0.2740	0.2104	0.1084
Tray 2	0.4256	0.3771	0.2047
Tray 3	0.2803	0.9202	0.5563
Tray 4	0.2422	1.5059	0.9682
Avg.	0.3055	0.7534	0.4594

α	1.6400
η_{Small}	**58.80%**

Table 21: O’Connell Correlation for Tray Efficiency

Feed Conditions		
Variable	Value	Units
P_{gauge}	480.3	psig
$P_{absolute}$	495.0	psia
Mist Eliminator?		Yes
ρ_V	0.02122	$\frac{lb}{ft^3}$
ρ_L	46.11	
ρ_H	61.42	
$\dot{m}_V$	26710	$\frac{lb}{hr}$
$\dot{m}_{LL}$	71200	
$\dot{m}_{HL}$	2366	
μ_V	1.01E-02	$\frac{lb}{ft*s}$
μ_L	1.417	
μ_H	0.4953	

Table 22: Large 3-Phase Separator Feed Conditions

3-Phase Separator Sizing

Variable	Equation	Value	Units
Q_V	$Q_V = \frac{W_V}{3{,}600\rho_V}$	349.64	$\frac{ft^3}{s}$
Q_{LL}	$Q_{LL} = \frac{W_{LL}}{60\rho_L}$	25.74	$\frac{ft^3}{min}$
Q_{HL}	$Q_{HL} = \frac{W_{HL}}{60\rho_H}$	0.64	$\frac{ft^3}{min}$
U_T	$U_T = K\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{1/2}$	14.54	$\frac{ft}{s}$
U_V	$U_V = 0.75U_T$	10.90	ft/s
V_H	$V_H = T_H Q_L$	263.8	ft^3
V_S	$V_S = T_S Q_L$	131.9	ft^3
L/D		3.5	
D	$D = \left(\frac{4(V_H + V_S)}{0.6\pi(L/D)}\right)^{1/3}$	**6.21**	ft
A_T	$A_T = \frac{\pi D^2}{4}$	30.32	ft^2
H_V	Larger of 0.2D or 2.0	2	ft
A_V	$A_V = y^* A_T$	8.43	ft^2
H_{LLV}	Set as minimum from figure	1	ft
H_{LLB}	Set as minimum from figure	0.5	ft
A_{LLV}	$A_{LLV} = y * A_T$	3.16	ft^2
L	$L = \frac{V_H + V_S}{A_T - A_V - A_{LLV}}$	21.12	ft
Φ	$\Phi = H_V / U_V$	0.183	s
U_{VA}	$U_{AV} = Q_V / A_V$	41.5	ft/s
L_{min}	$L_{min} = U_{VA}\Phi$	7.6	ft
L_{actual}		21.1	ft
U_{HL}	$U_{HL} = \frac{k_S(\rho_H - \rho_L)}{\mu_L}$	1.76	$\frac{ft}{s}$
t_{HL}	$t_{HL} = \frac{12(H_{LLB} + D - H_V)}{U_{HL}}$	32.12	s
θ_{LL}	$\theta_{LL} = \frac{(A_T - A_V)L}{Q_{LL}}$	17.97	s
$L_{adjusted}$	$L = \frac{t_{HL} Q_{LL}}{(A_T - A_V)}$	**37.76**	ft
L/D		6.08	
H_{HLL}	$H_{HLL} = D - H_V$	4.21	ft
A_{NLL}	$A_{NLL} = A_{LLV} + V_H / L$	10.14	ft^2
H_{NLL}	From Table 3	21.89	ft
H_{HL}	Guess	2	ft
U_{LH}	$U_{LH} = \frac{k_S(\rho_H - \rho_L)}{\mu_H}$	5.04	$\frac{ft}{s}$
U_P	$U_P = 0.75U_{LH}$	3.78	ft/s
D_B	$D_B = \sqrt{\frac{4 * 12Q_{HL}}{\pi U_P}}$	**1.61**	ft
t_{LH}	$t_{LH} = \frac{12H_H}{U_{LH}}$	2.38	s
θ_{HL}	$\theta_{HL} = \frac{\pi D_p^2 H_{HL}}{4Q_{HL}}$	5.59266	s
V_{Vessel}		1145	ft^3
V_{Boot}		5.10	ft^3
V_{Total}		1150	ft^3
		32.6	m^3

Table 23: Large 3-Phase Separator Sizing

Feed Conditions		
Variable	Value	Units
P_{gauge}	480.3	psig
$P_{absolute}$	495.0	psia
Mist Eliminator?		Yes
ρ_V	0.0217	$\frac{lb}{ft^3}$
ρ_L	46.22	
ρ_H	61.5	
$\dot{m}_V$	13930	$\frac{lb}{hr}$
$\dot{m}_{LL}$	31340	
$\dot{m}_{HL}$	1332	
μ_V	1.01E-02	$\frac{lb}{ft * s}$
μ_L	1.472	
μ_H	0.5088	

Table 24: Medium 3-Phase Separator Feed Conditions

3-Phase Separator Sizing

Variable	Equation	Value	Units
Q_V	$Q_V = \frac{W_V}{3,600\rho_V}$	178.32	$\frac{ft^3}{s}$
Q_{LL}	$Q_{LL} = \frac{W_{LL}}{60\rho_L}$	11.30	$\frac{ft^3}{min}$
Q_{HL}	$Q_{HL} = \frac{W_{HL}}{60\rho_H}$	0.36	$\frac{ft^3}{min}$
U_T	$U_T = K\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{1/2}$	14.39	$\frac{ft}{s}$
U_V	$U_V = 0.75U_T$	10.80	ft/s
V_H	$V_H = T_H Q_L$	116.6	ft^3
V_S	$V_S = T_S Q_L$	58.3	ft^3
L/D			3.5
D	$D = \left(\frac{4(V_H + V_S)}{0.6\pi(L/D)}\right)^{1/3}$	**4.73**	ft
A_T	$A_T = \frac{\pi D^2}{4}$	17.60	ft^2
H_V	Larger of 0.2D or 2.0	2	ft
A_V	$A_V = y^*A_T$	7.07	ft^2
H_{LLV}	Set as minimum from figure	1	ft
H_{LLB}	Set as minimum from figure	0.5	ft
A_{LLV}	$A_{LLV} = y * A_T$	2.71	ft^2
L	$L = \frac{V_H + V_S}{A_T - A_V - A_{LLV}}$	22.37	ft
Φ	$\Phi = H_V/U_V$	0.185	s
U_{VA}	$U_{AV} = Q_V/A_V$	25.2	ft/s
L_{min}	$L_{min} = U_{VA}\Phi$	4.7	ft
L_{actual}		22.4	ft
U_{HL}	$U_{HL} = \frac{k_S(\rho_H - \rho_L)}{\mu_L}$	1.69	$\frac{ft}{s}$
t_{HL}	$t_{HL} = \frac{12(H_{LLB} + D - H_V)}{U_{HL}}$	22.93	s
θ_{LL}	$\theta_{LL} = \frac{(A_T - A_V)L}{Q_{LL}}$	20.84	s
$L_{adjusted}$	$L = \frac{t_{HL}Q_{LL}}{(A_T - A_V)}$	**24.61**	ft
L/D			5.20
H_{HLL}	$H_{HLL} = D - H_V$	2.73	ft
A_{NLL}	$A_{NLL} = A_{LLV} + V_H/L$	7.45	ft^2
H_{NLL}	From Table 3	10.53	ft
H_{HL}	Guess	2	ft
U_{LH}	$U_{LH} = \frac{k_S(\rho_H - \rho_L)}{\mu_H}$	4.90	$\frac{ft}{s}$
U_P	$U_P = 0.75U_{LH}$	3.67	ft/s
D_B	$D_B = \sqrt{\frac{4 * 12Q_{HL}}{\pi U_P}}$	**1.23**	ft
t_{LH}	$t_{LH} = \frac{12H_H}{U_{LH}}$	2.45	s
θ_{HL}	$\theta_{HL} = \frac{\pi D_P^2 H_{HL}}{4Q_{HL}}$	5.53043	s
V_{Vessel}		433	ft^3
V_{Boot}		2.95	ft^3
V_{Total}		436	ft^3
		12.3	m^3

Table 25: Medium 3-Phase Separator Sizing

Feed Conditions		
Variable	Value	Units
P_{gauge}	480.3	psig
$P_{absolute}$	495.0	psia
Mist Eliminator?		Yes
ρ_V	0.02168	$\frac{lb}{ft^3}$
ρ_L	46.46	
ρ_H	61.45	
$\dot{m}_V$	2753	$\frac{lb}{hr}$
$\dot{m}_{LL}$	6781	
$\dot{m}_{HL}$	214.6	
μ_V	1.00E-02	$\frac{lb}{ft*s}$
μ_L	1.673	
μ_H	0.5011	

Table 26: Small 3-Phase Separator Feed Conditions

3-Phase Separator Sizing			
Variable	Equation	Value	Units
Q_V	$Q_V = \frac{W_V}{3{,}600\rho_V}$	35.27	$\frac{ft^3}{s}$
Q_{LL}	$Q_{LL} = \frac{W_{LL}}{60\rho_L}$	2.43	$\frac{ft^3}{min}$
Q_{HL}	$Q_{HL} = \frac{W_{HL}}{60\rho_H}$	0.06	$\frac{ft^3}{min}$
U_T	$U_T = K\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{1/2}$	14.44	$\frac{ft}{s}$
U_V	$U_V = 0.75U_T$	10.83	ft/s
V_H	$V_H = T_H Q_L$	24.9	ft^3
V_S	$V_S = T_S Q_L$	12.5	ft^3
L/D			3.5
D	$D = \left(\frac{4(V_H + V_S)}{0.6\pi(L/D)}\right)^{1/3}$	**2.83**	ft
A_T	$A_T = \frac{\pi D^2}{4}$	6.29	ft^2
H_V	Larger of 0.2D or 2.0	2	ft
A_V	A_V = y*A_T	4.75	ft^2
H_{LLV}	Set as minimum from figure	1	ft
H_{LLB}	Set as minimum from figure	0.5	ft
A_{LLV}	$A_{LLV} = y * A_T$	1.99	ft^2
L	$L = \frac{V_H + V_S}{A_T - A_V - A_{LLV}}$	-82.86	ft
Φ	$\Phi = H_V/U_V$	0.185	s
U_{VA}	$U_{AV} = Q_V/A_V$	7.4	ft/s
L_{min}	$L_{min} = U_{VA}\Phi$	1.4	ft
L_{actual}		1.4	ft
U_{HL}	$U_{HL} = \frac{k_S(\rho_H - \rho_L)}{\mu_L}$	1.46	$\frac{ft}{s}$
t_{HL}	$t_{HL} = \frac{12(H_{LLB} + D - H_V)}{U_{HL}}$	10.92	s
θ_{LL}	$\theta_{LL} = \frac{(A_T - A_V)L}{Q_{LL}}$	0.87	s
$L_{adjusted}$	$L = \frac{t_{HL}Q_{LL}}{(A_T - A_V)}$	**17.29**	ft
L/D			6.11
H_{HLL}	$H_{HLL} = D - H_V$	0.83	ft
A_{NLL}	$A_{NLL} = A_{LLV} + V_H/L$	3.43	ft^2
H_{NLL}	From Table 3	1.54	ft
H_{HL}	Guess	2	ft
U_{LH}	$U_{LH} = \frac{k_S(\rho_H - \rho_L)}{\mu_H}$	4.88	$\frac{ft}{s}$
U_P	$U_P = 0.75U_{LH}$	3.66	ft/s
D_B	$D_B = \sqrt{\frac{4 * 12Q_{HL}}{\pi U_P}}$	**0.49**	ft
t_{LH}	$t_{LH} = \frac{12H_H}{U_{LH}}$	2.46	s
θ_{HL}	$\theta_{HL} = \frac{\pi D_P^2 H_{HL}}{4Q_{HL}}$	4.79235	s
V_{Vessel}		109	ft^3
V_{Boot}		0.48	ft^3
V_{Total}		109	ft^3
		3.1	m^3

Table 27: Small 3-Phase Separator Sizing

MATLAB Code

Function File Code

```
function f = FTRFunctionFinal(W,Y)

X = Y(1);

z = Y(2);

T = Y(3);

Ta = Y(4);


Dt = 20; %cm

Tube = 42;


db = 0.8; %gm/cm^3


mc = 395000; %g/hr


Vt = W/db;

Vtft = Vt*3.53*10^-5;

Vtotal=Vtft*Tube %ft^3


F1o = 60922/Tube; %CO

F2o = 122434/Tube; %H2

F3o = 259/Tube; %H2O

F4o = 0; %C Chain

F5o = 16254/Tube; %CH4

F6o = 42/Tube; %O2

F7o = 636/Tube; %CO2
```

```
Fi = F5o+F6o+F7o;

Fto = F1o+F2o+F3o+F4o+Fi;


S1 = 1;

S2 = 2;

S3 = 1;

S4 = 1;


O2 = F2o/F1o;

O3 = F3o/F1o;

O4 = F4o/F1o;


F1 = F1o*(1-X);

F2 = F1o*(O2-(S2/S1)*X);

F3 = F1o*(O3+(S3/S1)*X);

F4 = F1o*(O4+(S4/S3)*X);

Ft = F1+F2+F3+F4+Fi;


To = 505;


T1 = exp((-4492)*((1/T)-(1/473)));

T2 = exp(8237*((1/T)-(1/473)));


Pto = 32;

P1o = Pto*0.331;

P2o = Pto*0.6251;

P3o = Pto*0.0013;
```

```
P4o = Pto*0;

Pio = (Fi/Fto)*Pto;


P1 = Pto*(F1/Ft)*(T/To)*z;

P2 = Pto*(F2/Ft)*(T/To)*z;

P3 = Pto*(F3/Ft)*(T/To)*z;

P4 = Pto*(F4/Ft)*(T/To)*z;

Pi = Pto*(Fi/Ft)*(T/To)*z;

Pt = P1+P2+P3+P4+Pi;


k = 0.173;

k2 = 4.512;


Cp1 = 0.0282; %BTU/g*K

Cp2 = 0.0275;

Cp3 = 0.0549;

Cp4 = 0.0438;

Cp5 = 0.0438;

Cp6 = 0.0295;

Cp7 = 0.0423;

Cpc = 0.0459;


Hrx = -154.444; %BTU/gmole CO


Do = 0.00885; %gm/cm^3

D = Do*(Pt/Pto)*(T/To)*(Ft/Fto);
```

```
por = 0.4;

dc = db/(1-por);


u = 57804; %cm/hr


G = u*D;


Ua = abs((0.385*((G)^(0.8)))/(Dt^(0.2)));


a = 0.00000305; %differ from excel


r1 = (-k*T1*P2*P1)/((1+k2+T2*P1)^2);


dXdW = -r1/F1o;

dzdW = ((-a)/(2*z))*(T/To)*(Ft/Fto)*1.5;

dTdW = (r1*Hrx-(Ua*(T-Ta)/dc))/(F1*Cp1+F2*Cp2+F3*Cp3+F4*Cp4+F5o*Cp5+F6o*Cp6+F7o*Cp7);

dTadW = ((Ua/dc)*(T-Ta))/(mc*Cpc);

f = [dXdW;dzdW;dTdW;dTadW];

end
```

Script Code

```
clc

Wspan = [0,100000]; %Range for independent variable or catalyst weight

% list values for dependent variables in order 1Conversion 2Pressure Drop

%Guy-Lussac for temp and pressure

y0 = [0,1,505,451]; % Initial Values for 0, 1, 473, 298

%FTRNoP sees only conversion, Reactor Temp and Coolant Temp
```

```
[W, y] = ode78(@FTRFunctionFinal,Wspan,y0);


%Plotting this stuff

%Conversion

plot(W,y(:,1));

%legend('Catalyst Weight','Conversion')

ylabel('Conversion');

xlabel('Catalyst Weight');

figure;


%Pressure Drop

%plot(W,y(:,2));

%legend('Catalyst Weight','Conversion')

%ylabel('Pressure Drop');

%xlabel('Catalyst Weight');

%figure;


%CHANGED THE ARRAY POSTIONS


%Reactor Temperature

plot(W,y(:,3));

%legend('Catalyst Weight','Conversion')

ylabel('Reactor Temperature');

xlabel('Catalyst Weight');

%figure;


%Coolant Temperature
```

```
%plot(W,y(:,4));

%legend('Catalyst Weight','Conversion')

%ylabel('Coolant Temperature');

%xlabel('Catalyst Weight');
```