AIChE Senior Design Project

Modular Distributed Gas-to-Liquids (GTL) Synthesis

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Group Number: 1890

Team Members: Garrison Allen, Natalie Evans, Jacob Lewellen, Dakota Plumlee

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Executive Summary

This project intends to evaluate the use of modular gas-to-liquid plants for well-site implementation based on energy efficiency and economic value. The design presented would allow the company to have the flexibility of the modular design and various reactor sizes, while reducing greenhouse gas emissions. In addition, the presented design incorporates inherent safety measures to control the hazards and risks that come with the unit.

Simulations of the design process were conducted and interpreted with the help of polymath code for reactor modeling and Aspen HYSYS for reactor and separation simulations. For the required equipment to design, the group recommends for the FTR operating a PBR and for the separations unit to install three 3-phase separators, one air fan, and two heat exchangers.

After overall design of the three units and entire process it was found that the energy efficiency of the hydrocarbon products over the methane in the initial feed was 93.9%.

An economic analysis of the project had an NPV of $5.99 million over the project evaluation life. Diesel was found to be the largest revenue contributor at $2112.60/day for the 500 MSCFD unit, $10,562.99/day for the 2.5 MMSCFD, and $21,125.99/day for the 5 MMSCFD. Naphtha was found to be the second largest revenue contributor at $5,558.45/day for the 5 MMSCFD unit. The fixed capital investment was found to be $180 million in year zero for construction of the equipment. The units will undergo turnaround every three years for routine maintenance and catalyst replacement. Trucking costs are dependent upon the deployment and redeployment schedule of the units. An optimization analysis was performed to determine the number of each size unit that should be fabricated. It was found that the optimal time to fabricate units was in year 0 before production begins. A grouping optimization was then used to group wells in order to optimize central plant capacity. Once these groups were complete, the best route and production schedule for each unit was determined. An NPV analysis was conducted and determined to be $5.99 million over the project life with a rate of return of 8.53% with a 95% confidence interval between 1.97% and 19.03%. Based on these analyses and the assumptions upon which the design is based, we recommend proceeding on an environmental and energy efficiency basis, but not on an economic basis.

In addition to the energy efficiency and economic aspects of the project, the safety and environmental aspects of the project were strongly considered in the design. The proposed design uses the best available technology available to control hazards associated with the process. As opposed to flaring, the design process generates transportation fuels to reduce CO_2 emissions.

Introduction

The company has discovered new gas deposits in remote areas that are difficult to profitably recover by traditional means. They are considering the option of modular deployment of gas-to-liquid (GTL) plants, based on the design of a Fischer-Tropsch reactor (FTR) reactor and separations system. These modular units were to be designed on the idea of three standard sizes or parallel implementation. Included with the plant is a predesigned Syngas Unit and Hydro-isomerization Unit. Along with the design of the modular units, the company requires an analysis of the supply chain and network optimization of deployment of the GTL modules over a project evaluation life of 20 years.

Summary

The process of the entire GTL plant is broken down into four units comprised of a syngas unit, CO2 recovery unit, a FTR unit, and a separations unit. At the syngas unit steam and CO2 are introduced to the feed and sent to a furnace to be heated. The feed is then mixed with oxygen in a steam methane reforming (SMR) and water-gas shift (WGS) reactor to allow partial oxidation, SMR, and WGS reactions to occur and produce carbon monoxide and hydrogen. This feed is then cooled and sent to the CO2 recovery unit.

The feed enters the CO2 recovery unit to be stripped of CO2, before being sent to the FTR. The CO2 recovery is made up of an absorber and a regenerator. The feed, sent to the absorber, is removed of CO2 with water as a solvent, moved along to a heater, bringing heat back to the feed. The removed CO2 is sent to a regenerator that separates the CO2 from the water, then recycled back into the syngas unit.

The heated feed from the CO2 recovery unit is run through a PBR, converting it into hydrocarbons, ranging from C2 to C45+. After the feed stream has converted, it is sent to the separations unit and split into different product, credit, and waste streams. The streams are either reintroduced to the process, sent to a waste treatment, or stored onsite, where they will later be transported to a central plant.

Discussion

Syngas Generation

The Syngas Unit was not included in actual equipment design and optimization for the project. The required objectives to meet were:

1. Have a Steam to CH4 ratio of a .5 mol steam/ 1 mol CH4 minimum in preheater feed
2. Meet a Hydrogen to Carbon Monoxide ratio of 2:1
3. Run the partial oxidation reaction to completion with oxygen as the limiting reagent
4. Run the SMR and WGS to equilibrium

The syngas unit consists of the feed from the well, a fired furnace, a SMR/WGS reactor, and a heat exchanger that cools the reactor product stream, before entering the carbon dioxide recovery system. The entire unit was modelled in Aspen HYSYS, simulating the SMR, partial oxidation, and WGS reactions. The feed from the well is assumed to be 100% methane, not requiring pre-treatment for sulfur compounds. The feed is mixed with low pressure steam at a ratio of .68 for coking prevention, maintaining the hydrogen to carbon monoxide ratio. Supplemental carbon dioxide is also mixed in at a ratio of .25 mol CO2/ 1 mol CH4, keeping the hydrogen to carbon monoxide ratio at 2.

The mixed feed is sent to a furnace, raising the temperature, before going into a SMR/WGS reactor. The furnace operates at 85% efficiency and can heat to a maximum limit of 1000°F, which is in place for metallurgical integrity purposes. The fired furnace was fueled with methane and produced tail gas, in order to lower cost in fuel utility. The excess air percentage was assumed to be 7.5%, based on a heuristic of natural gas fuels[1]. The furnace heated the feed to a temperature of 700°F, before entering the SMR/WGS reactor.

The SMR and WGS reactions were modelled in a Gibbs reactor, in Aspen HYSYS. A Gibbs reactor was chosen for the three reactions involved in syngas production. A Gibbs reactor in Aspen HYSYS works by minimizing the Gibbs free energy of a reaction. The SMR and WGS reactions are equilibrium reactions, in which the Gibbs free energy is minimized or equaled to zero, when these reactions reach equilibrium. This allows reliable simulation of SMR and WGS reactions, assuming the Gibbs reactor is adiabatic[2]. High purity oxygen is introduced to the reactor for partial oxidation to occur, raising the temperature for non-catalytic SMR reactions. A ratio of 1:2 oxygen to methane was used to create a partial oxidation reaction, raising the temperature to 1645°F. This temperature allows for SMR to occur, as well as WGS, resulting in the Gibbs reactor to yield a molar ratio of 2.006:1 of hydrogen to oxygen. This feed from the reactor enters a heat exchanger, is cooled down and run through the carbon dioxide recovery system to maintain pipe integrity.

CO_2 Recovery

A CO_2 recovery system was necessary to reduce the utility cost associated with carbon dioxide production, as well as minimize the quantities of inert components in the Fischer-Tropsch Reactor. An absorber column was selected to clean a 95 weight percentage of the CO_2 out of the FTR feed; the CO2 would then be stripped out by a stripper column, in addition to any residual H2S ubiquitous to reservoir product. Water was selected as the absorption medium over the more traditional amine system, due to concerns of product contamination and preclude the toxicity hazards associated with amine systems. The absorber was simulated using the Sour Soave-Redlich-Kwong (Sour SRK) property package of Aspen HYSYS. This was utilized over other properties, as it simulates the chemical absorption, neglected by other property packages, without the need to input additional reaction parameters. Results from the Sour SRK simulation had strong similarities to simulations run with the Acid Gas- Liquid Treating fluid property package. The absorber removes 95 weight percent of the CO2 and 99 weight percent of the H2O from the Fisher-Tropsch Reactor feed. The column operates at a temperature high of 470 °F and a pressure high of 445 psia, requiring 20.55 kgal/hr for 500 MSCFD units, 102.73 kgal/hr for 2.5 MMSCFD units and 205.45 kgal/hr for 5 MMSCFD units of process water. The reactor requires a heat exchanger utilizing 369 MBTU/hr for 500 MSCFD, 1.85 MMBTU/hr for 2.5 MMSCFD and 3.69 MMBTU/hr for 5 MMSCFD of high-pressure steam heating capacity in the CO2 recovery unit, bringing the reactor feed to the appropriate temperature.

Fischer-Tropsch Reactor

The Fischer-Tropsch Reactor (FTR) was designed and optimized to achieve the following objectives:

1. Minimize the inherent hazards associated with the system
2. Maximize energy efficiency
3. Maximize project net present value (NPV)
4. Create a safe, consistent, and reliable operation and control

The reactor system was designed by determining the optimal reactor operating temperature, then optimizing conversion via a separate reactor model. The selectivity of the hydrocarbon products of the FTR was determined via the Anderson-Schulz-Flory (ASF) distribution and the parameters described[3], which were ultimately a function of the operating temperature. Based on the relevant Fischer-Tropsch literature on cobalt catalyst[4], the hydrocarbons generated were assumed to be straight-chain alkanes, ranging from 1 to 100 carbon atoms. The conversion of the wax fraction (i.e. longer than n-icosane) to shorter hydrocarbons and the production of diesel in the Hydroisomerization Unit (HIU) was calculated via the parameters provided[3]. Since methane production is directly proportional to the production of ethane, propane, and butane, the components most likely to be used in a form other than a finished fuel product, the optimal energy efficiency was determined from the methane production. Revenue in terms of both mass and volume was approximated by assuming perfect separation of products and accounting for the utility cost of the HIU[3]. Other costs and utilities were assumed to have a negligible impact, compared to increases in the revenue generated. The optimal reactor temperature in terms of environmental impact, revenue by volume, and revenue by mass were 380.6 °F, 378.1 °F, and 380.7 °F, respectively; the differences between these values are too insufficiently small for common measurement and/or control systems to distinguish[5]. Minor changes in temperature have a minimal impact on the value of the hydrocarbon products near the optimal temperature.

Several reactor designs were considered as part of the FTR design. A number of continuous process FTR designs have previously been implemented, including packed-bed, circulating fluidized bed, fixed fluidized bed, and slurry bed reactors[6]. Recent research efforts have resulted in advancing the development of microchannel reactors for FTR applications[7,8]. Slurry bed reactors are the most implemented FTR, especially for recent applications[6]. However, these reactors do not effectively scale down to the flow rates encountered at a well-site[6,9], thus inappropriate for this application. Fluidized bed reactors are notoriously difficult to start up and control, especially at small scales[9]. Well-sites, typically, do not have the personnel needed to attend to these complexities. Microchannel reactors have not been implemented outside of pilot plants, and many of the parameters associated with their operation and design are not well known[8]. Because of the uncertainty associated with these reactor parameters and the immature nature of the technology, estimates are unlikely to accurately predict the cost and design requirements, and it is likely that the project timeline would be substantially delayed for the completion of R&D work, substantially decreasing NPV. Based on the comparison of these alternatives, the packed-bed reactor (PBR) design was considered the simplest design, most readily controlled, and most capable of handling the greatest variation in flow rates, as is expected with well site operation. Boiling water was selected to use on the shell side, due to the inherent safety improvement over alternatives, ability to make revenue from steam, and ease of control methodology implementation.

To determine the optimal reactor geometry, temperature profile, maximal conversion, and other reactor parameters, a single tube of the reactor was modeled in the Polymath differential equation solver software. The reactor mass balance was modeled in terms of conversion with respect to catalyst volume. The reaction kinetics and the catalyst parameters, relevant to the Ergun Equation, were provided[3]. The reaction stoichiometry of hydrogen and product hydrocarbon were based on the average chain length, as a function of temperature. The average chain length was determined via an empirical correlation of hydrocarbon products from the selectivity model. The energy balance was determined by the given heat of reaction and the overall heat transfer coefficient[3]; specific heat capacities, as a function of temperature for components that were not hydrocarbon products, were utilized from published empirical correlations[10]. Specific heat capacities for hydrocarbon products, modeled as a function of temperature, were determined by multiplying the phase fraction by the molar heat capacity of the phase[10]. The phases for each component at a specified temperature were then calculated. Heat capacities of pseudo components, utilized for hydrocarbons chains, longer than n-icosane, were calculated with Aspen HYSYS, using a Peng-Robinson property package. The specific heat capacity for the product hydrocarbons at a given temperature was calculated by multiplying the heat capacity for the component or pseudo component by the mole fraction, then summing the components and pseudo components of a given temperature. The heat capacities, as a function of temperature, were fit to an empirical correlation (See Figure in Appendix)

FTR geometry was designed and optimized with two goals in mind: optimize heat transfer and maximize conversion. Maximal energy efficiency, conversion, and revenue occurs when the reactor operates isothermally; however, operating under these conditions does not provide sufficient heat exchange to accommodate even slight upward deviations in the FTR feed temperature. To ensure safe operating conditions, the FTR was required to be able to sustain a 55 °F upward deviation in feed temperature, without the temperature in the reactor encountering a temperature of 620 °F; this represents a sufficiently high pressure differential such that a pressure relief valve could be sized such that a relief device would be practicable and provide sufficient control[11,12,13]. A slight, yet downward sloping temperature gradient was required to accomplish this. Optimal operating FTR feed temperature was selected as 386 °F, with the boiler producing shell side operating at 377 °F and 125 psig, thus producing a medium pressure steam utility. A differential between these two temperatures meets the upward deviation requisite that were adequately measured and controlled[5]. The design was specified so the 2,500 MSCFD FTR could be contained within a single module. Optimization was accomplished by selecting a tube diameter, maximizing the number of tubes, and varying length to maximize conversion subject to the modular units listed above, in addition to the maximum pressure drop of 50 psi specified[3]. Product specifications of the syncrude were determined by averaging the temperature gradient.

The optimized reactor geometry consists of tubes of diameter 1 ¼” and length 36’ 4” in a tube sheet arrangement, akin to a floating head shell and tube exchanger, selected for ease of maintenance during the regular 3-year turnaround. 1000 and 5000 tubes were utilized in the 500 MSCFD and 2,500 MSCFD FTR modules, respectively. Since the tubes for the 5,000 MSCFD unit cannot be contained within a cylinder of outer diameter 8', the 5000 MSCFD consisted of two 2500 FTR modules, running in parallel. Each tube contains 15.4 lb of catalyst. The reactor conversion of 0.972 made the addition of any other reactors unnecessary. Pressure dropped in the reactor from 29.8 atm to 27.6 atm. The converged differential equation model implemented in the differential equation solver PolyMath can be found in the Appendix.

Separations

Once the product and effluent from the Fischer Tropsch reactor had been produced, the following points were required to be met:

1. Separate into individual products/effluents (if profitability available) the following, along with their state of matter
    a. Tail Gas (TG) (Vapor)
    b. Liquid Petroleum Gases (LPG) (Liquid)
    c. Naphtha (Liquid)
    d. Distillate (Liquid)
    e. Produced Water/ Produced Steam Condensate (Liquid/Vapor)
    f. Waste Water (Liquid)
2. Keep distillate products of C20+ above 250 °F
3. Meet Water Solubility limit in distillate stream to HIU
4. Maintain a Naphtha RVP @ 100 °F between 8-14
5. Minimize the capital cost of the process unit
6. Minimize the required utilities needed for operation
7. Reduce complexity of process for remote operation/control

A computer simulation in Aspen HYSYS was selected as the prime source of modelling for the potential separation processes. The Peng-Robinson fluid package was selected since non-idealities from the elevated pressures encountered were thought to dominate over the impact of polar interactions, its successful use in simulating similar systems, and the presence of an RVP property within the Peng-Robinson package[14]; results obtained had strong agreement with results generated with the CPA package. Many potential separation processes were researched for initial design consideration by the group such as the traditional "stick built" refinery approach, separating the products via distillation columns. Due to the amount of water volume from the FTR, the constraint of maintaining streams with high distillate composition to it, and the potential for these units to be operated in remote locations, the group discovered that three phase separators, which are quite common in upstream oil and gas facilities[15], could be used as a main piece of equipment for the project's needed separation process.

An initial separation occurs in V-101/V201/V-301 of water, heavy hydrocarbon product and light hydrocarbon product. This allows for most of the water to be separated from potential product streams. A delta pressure of 159 psi was determined in V-101/V201/V-301, by limiting the amount of C20+ components to a negligible amount. By making the composition of C20+ components negligible in the vapor stream (Stream 103/203/303), better separation of Naphtha components from lighter hydrocarbons could be executed below the 250 °F constraint. C20+ was assumed for this cutoff, allowing untraceable to negligible amounts in streams below 250 °F, making it unlikely for the process streams under the constraint conditions to crystalize.

A fin fan air cooler was chosen to cool Stream 103/203/303 to 100°F, allowing for better separation of TG and LPG from the liquid naphtha product stream, rather than a larger pressure drop across V-102/202/302 that would inhibit the ability for any produced TG to be introduced resourcefully into a process as fuel gas. The fin fan air cooler was also selected because of its remote operation practicality in comparison to cooling water heat exchangers. V-102/202/302 separated out the condensed aqueous phase (Stream 107/207/307), the liquid naphtha (Stream 106/206/306) and the light end hydrocarbons and inert gases. Stream 106/206/306 of Naphtha product met the top end constraint of RVP of natural gasoline[5] of 14 psia. The top end of the RVP specification was used, as the group assumed it to mean that more LPG/TG components could be used and considered as the higher priced naphtha. Separation of the LPG from the TG was considered but ruled out based upon the assumption, even at complete and total separation of LPG from TG, it would not meet the necessary threshold to justify the lowest cost compressor system, the process would be inherently safer, due to less equipment especially of complexity and the inherent hazard of storing LPG under pressure, and the rich TG was best economically justified for utility needs. The hydrocarbon/inert vapor mixture of stream 105/205/305 is introduced as the majority of the fuel gas to the fire heater that heats the reactants prior to introduction to the syngas reactor.

V-103/203/303 was used to separate the water from the distillate stream, meeting the required temperature dependent solubility limit, as described in the GPSA Handbook[5]. A maximization of distillate product volume was conducted by increasing the pressure drop across V-103/203/303 to 47 psi, which allowed for the most hydrocarbon to drop out of stream 110/209/309. Produced and separated water of stream 108/207/307 was introduced into V-103/203/303, increasing the amount of steam in stream 110/209/309, condensed across the cooling water heat exchanger, E-102/E-202/E-302, to a vapor fraction assumed to be negligibly close to 0%. Stream 111/210/310 is then disposed of as wastewater, meeting the required 75% water purity on a volume basis. The produced water of V-102/202/302 for the 500 MSCFD modular unit size was of a volume that decreased the temperature of Stream 12 below the requested 250 °F. The group was able to solve this issue by introducing heat to the produced water stream of V-102/202/302 with the created medium pressure steam of the FTR at E-103. Produced water of V103/203/303 was combined with V-101/201/301 and used as a steam condensate credit, meeting the volume purity specification of 99.9%.

The post FTR separation streams' of met specification standard volumes are listed below per modular size.

Table 1: Post FTR Separation Streams Specifications

Stream	500 MSCFD Unit	2.5 MMSCFD Unit	5 MMSCFD Unit
TG (105/205/305)	138.4 MSCFD	690.0 MSCFD	1380 MSCFD

Naphtha (106/206/306)	5.73 STBD	28.7 STBD	57.3 STBD
Distillate (112/212/312)	37.9 STBD	189.5 STBD	379.0 STBD
Wastewater (111/210/310)	7.9 STBD	79.0 STBD	158 STBD
Steam Condensate (115/214/314)	57.4 STBD	114.8 STBD	229.6 STBD

The tail gas was determined to have a LHL of 3.222*10^5 Btu/lbmol; this value was found from Aspen HYSYS simulation and used to calculate the ability of tail gas to meet modular fuel gas needs throughout the process.

Sizing of separation equipment was approached by using a well-known and established upstream oil and gas manual by Richard Sivalls[16]. The manual distinguishes approaches for selecting liquid-gas three phase separators via correlations, established for liquid and vapor flowrate, in addition to, what the manual describes as, high pressure process vessels (200-2000 psig) and low-pressure process vessels (<125psig).

Optimization of the separation unit was approached by varying temperature and pressure variables over the three phase separators and heat exchangers in the process. Main dependent variables that were looked upon were volumetric flowrate of naphtha and distillate product streams, the Reid vapor pressure

Hydroisomerization Unit

The distillate stream was fed into the HIU; HIU products and utility requirements were calculated as described[3].

Conclusions

For this process, the design team was able to design and model a creative solution while still meeting given specifications. It can be concluded from the technical design and economic analysis that the project is feasible. This process was 93.9% energy efficient, making it an attractive for the company on an environmental basis.

In addition to the project being economically feasible, the design is inherently safer. Many efforts were made to design a process that used the best available technology to control the process and contribute to the efforts to reduce greenhouse gas emissions. Through this technology the design team was able to control any remaining hazards which makes this design inherently safer than alternatives.

The system design consists of the syngas feed stream into a packed-bed Fischer-Tropsch reactor which is then fed to a separations unit consisting of three three-phase separators before going to the hydroisomerization unit. Through further analyses, it was determined that the energy efficiency of the system was 93.9%.

In an economic analysis, NPV was found to be $5.99 million with an ROR of 8.53%. The project is profitable, and the ROR exceeds the minimum discount rate, making this project economically feasible with enhancements in the safety and environmental aspects.

Recommendations

For the project based on the results from preliminary design, it is recommended to remove some of the assumptions provided that simplify the process. The assumption of a 100% methane feed from the well is atypical compared to real well feeds where methane can range from 75% to 98% (mole basis) with an array of other materials from the well including ethane, propane, and butane as well as many impurities like hydrogen sulfide, water vapor and mercury[17]. Another assumption taken is that sulfur treatment is not necessarily from the well and as previously stated, wells typically have hydrogen sulfide which usually cost 300 to 600$/ton to remove excluding the cost of the equipment to treat sulfur[18]. Both these assumptions remove essential costs from the project that could impact the economics.

Due to the limitations associated with chemical reaction engineering and scale-up from laboratory tests, the FTR should be empirically modeled in both laboratory and pilot plant scales; experiments should be conducted to understand steady-state behavior, start-up and shut down, and runaway conditions and phenomena. The CO2 absorber column was designed under temperature and pressure conditions not previously validated; although two independent physical property models resulting in similar results, vapor-liquid equilibrium data should be collected prior to implementation.

After the sensitivity and quantitative analyses were run, it was determined that this project is heavily dependent on the price of oil. With the current price of oil, this project cannot reliably produce an economically attractive NPV. Due to this, we are recommending against the project on an economic basis, but suggesting the project be looked into further on an environmental aspect.

Project Premises

The objective of the project is to design a modular GTL plant that follows a safe design and mitigates hazards to people and the environment. This also forbids flaring of hydrocarbons except in emergency situations throughout the process. Find the best possible solution in lowering environmental impact and maximizing energy efficiency, based on the methane feed used and the hydrocarbons produced from that methane. With this approach the group will be able to find the optimum finished liquid fuel production, where a reasonable cost and benefit balance is found. When developing these modular units, follow the principles of MCPI (modular chemical process intensification) to develop quick and easy deployable units. Finally, establish an economic analysis that includes the equipment capital investment, expense cost of designed equipment, and expense costs of the syngas, air separation plant, and hydroisomerization units. In this analysis the following can be assumed:

- Assess project evaluation life of 20 years
- Account for a 7-year straight line depreciation starting first
- 3% yearly inflation and 20% tax rate

- Multiply equipment cost by 4.8 to account for total capital investments
- Estimate total yearly operating expense beyond utilities as 3% of total capital investment
- Execute every 3 years a 1-month turnaround, for non-catalyst replacement purposes
- Factor in 35% depletion in feedstock source every year after 2 years
- Analyze suitable and realistic process control

Heat and Material Balance

Table 2: Utility Streams

	500 MSCFD	2500 MSCFD	5000 MSCFD
Steam and Steam Condensate	lb/hr consumed	lb/hr consumed	lb/hr consumed
SEP MPS	103	0	0
HIU MPS	14	70	141
MPS	2158	11307	22613
HPS Feed	676	3378	6756
Air Sep HPS	4388	21941	43883
CO2 Recovery HPS	381	1903	3805
FTR SC	2275	11377	22754
Used Air Sep Steam (SC)	-4388	-21941	-43883
Condensed CO2 Recovery Steam (SC)	-381	-1903	-3805
HIU SC	-14	-70	-141
Cooling Water			
Sep CWS	11029	55145	110289
Sep CWR	-11029	-55145	-110289
HIU CWS	29294	146471	292941
HIU CWR	-29294	-146471	-292941
Fuel Gas Stream	MBTU/hr	MBTU/hr	MBTU/hr
Purchased Fuel Gas	-1.705	-8.525	-17.050
Tailgas	-4.907	-24.536	-49.071
Furnace Fuel Gas	0.470	2.350	4.700
Steam Plant Fuel Gas	6.603	33.016	66.033
HIU Fuel Gas	0.046	0.229	0.459
Paraffins	-0.508	-2.539	-5.077
Electrical Streams	kWh/hr	kWh/hr	kWh/hr
FTR Elec	41.8	208.8	417.6
HIU Elec	3.5	17.6	35.2

Table 3: Process Streams

500 MSCFD						Component Flows (lb/hr)												
	Vapor Fraction	Temperature (°F)	Pressure (psia)	Mass Flowrate (lb/hr)	Enthalpy (10^6BTU/hr)	Methane	Ethane	Propane	n-Butane	Water	Carbon Dioxide	Carbon Monoxide	Hydrogen	Nitrogen	Naphtha	Diesel	Waxes	Oxygen
Methane	1.00	100	515	881	-1.77	881.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	1.00	100	515	604	-2.33	0.0	0.0	0.0	0.0	0.0	604.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Oxygen	1.00	75	515	869	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	86.9	0.0	0.0	0.0	860.1
Cleaned Syngas	1.00	386	438	1482	-2.10	194.5	0.0	0.0	0.0	2.7	36.2	1084.2	156.8	7.5	0.0	0.0	0.0	0.0
CO2 Waste	0.00	88	445	311	-1176.00	0.0	0.0	0.0	0.0	0.0	311.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0
FTR Steam*	1.00	386	125	2275	-12.83	0.0	0.0	0.0	0.0	2275.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Syncrude	0.56	377	405	1443	-5.10	202.5	0.6	0.9	1.2	680.7	36.2	29.9	0.3	6.3	62.4	124.2	298.3	0.0
Tailgas*	1.00	93	88	298	-0.62	201.8	0.6	0.9	1.1	2.4	35.8	29.9	0.3	6.3	19.2	0.1	0.0	0.0
Naphtha	0.00	93	88	60	-0.05	0.2	0.0	0.0	0.1	0.0	0.1	0.0	0.0	0.0	34.7	24.9	0.2	0.0
Waste Water	0.01	100	29	83	-0.52	0.5	0.0	0.0	0.0	74.3	0.2	0.1	0.0	0.0	4.6	3.2	0.0	0.0
Distillate	0.00	258	35	398	-0.32	0.0	0.0	0.0	0.0	0.4	0.0	0.0	0.0	0.0	3.9	96.0	298.1	0.0
SEP SC*	0.06	259	35	604	-3.97	0.0	0.0	0.0	0.0	603.5	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Paraffins*	-	-	-	25	-	3.0	1.5	10.4	10.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HUI Naphtha	-	-	-	78	-	-	-	-	-	-	-	-	-	-	78.0	-	-	-
Diesel	-	-	-	291	-	-	-	-	-	-	-	-	-	-	-	291.0	-	-
H2	-	-	-	2	-	-	-	-	-	-	-	-	2.2	-	-	-	-	-

2500 MSCFD						Component Flows (lb/hr)												
	Vapor Fraction	Temperature (°F)	Pressure (psia)	Mass Flowrate (lb/hr)	Enthalpy (10^6BTU/hr)	Methane	Ethane	Propane	n-Butane	Water	Carbon Dioxide	Carbon Monoxide	Hydrogen	Nitrogen	Naphtha	Diesel	Waxes	Oxygen
Methane	1.00	100	515	4406	-8.9	4405.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	1.00	100	515	3022	-11.7	0.0	0.0	0.0	0.0	0.0	3021.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Oxygen	1.00	75	515	4344	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	434.4	0.0	0.0	0.0	4300.7
Cleaned Syngas	1.00	386	438	7409	-10.5	972.3	0.0	0.0	0.0	13.4	180.8	5421.1	783.8	37.3	0.0	0.0	0.0	0.0
CO2 Waste	0.00	88	445	1557	-5.9	0.0	0.0	0.0	0.0	0.0	1557.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
FTR Steam*	1.00	386	125	11375	-64.2	0.0	0.0	0.0	0.0	11375.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Syncrude	0.56	377	405	7217	-25.5	1012.3	3.0	4.4	5.8	3403.5	180.9	149.6	1.3	31.7	312.0	621.1	1491.6	0.0
Tailgas*	1.00	93	88	1491	-3.1	1008.8	3.0	4.3	5.4	12.2	179.1	149.3	1.3	31.6	95.9	0.5	0.0	0.0
Naphtha	0.00	93	88	301	-0.3	0.9	0.0	0.1	0.3	0.0	0.4	0.1	0.0	0.0	173.3	124.6	1.2	0.0
Waste Water	0.01	100	29	415	-2.6	2.6	0.0	0.0	0.1	371.7	1.1	0.3	0.0	0.1	23.1	16.2	0.0	0.0
Distillate	0.00	258	35	1992	-1.6	0.0	0.0	0.0	0.0	1.8	0.0	0.0	0.0	0.0	19.7	479.9	1490.3	0.0

SEP SC*	0.06	259	35	3018	-19.9	0.0	0.0	0.0	0.0	3017.7	0.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Paraffins*	-	-	-	127	-	14.9	7.4	52.1	52.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HUI Naphtha	-	-	-	390	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Diesel	-	-	-	1455	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2	-	-	-	11	-	-	-	-	-	-	-	-	11	-	-	-	-	-

5000 MSCFD						Component Flows (lb/hr)												
	Vapor Fraction	Temperature (°F)	Pressure (psia)	Mass Flowrate (lb/hr)	Enthalpy (10^6BTU/hr)	Methane	Ethane	Propane	n-Butane	Water	Carbon Dioxide	Carbon Monoxide	Hydrogen	Nitrogen	Naphtha	Diesel	Waxes	Oxygen
Methane	1.00	100	515	8812	-18	8812	0	0	0	0	0	0	0	0	0	0	0	0
CO2	1.00	100	515	6043	-23	0	0	0	0	0	6043	0	0	0	0	0	0	0
Oxygen	1.00	75	515	8688	0	0	0	0	0	0	0	0	0	869	0	0	0	8601
Cleaned Syngas	1.00	386	438	14818	-21	1945	0	0	0	27	362	10842	1568	75	0	0	0	0
CO2 Waste	0.00	88	445	3114	-12	0	0	0	0	0	3114	0	0	0	0	0	0	0
FTR Steam*	1.00	386	125	22750	-128	0	0	0	0	22750	0	0	0	0	0	0	0	0
Syncrude	0.56	377	405	14434	-51	2025	6	9	12	6807	362	299	3	63	624	1242	2983	0
Tailgas*	1.00	93	88	2983	-6	2018	6	9	11	24	358	299	3	63	192	1	0	0
Naphtha	0.00	93	88	602	-1	2	0	0	1	0	1	0	0	0	347	249	2	0
Waste Water	0.01	100	29	830	-5	5	0	0	0	743	2	1	0	0	46	32	0	0
Distillate	0.00	258	35	3983	-3	0	0	0	0	4	0	0	0	0	39	960	2981	0
SEP SC*	0.06	259	35	6036	-40	0	0	0	0	6035	1	0	0	0	0	0	0	0
Paraffins*	-	-	-	253	-	30	15	104	104	0	0	0	0	0	0	0	0	0
HUI Naphtha	-	-	-	780	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Diesel	-	-	-	2910	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2	-	-	-	22	-	-	-	-	-	-	-	-	22	-	-	-	-	-

"*" Denotes product flow may also be considered utility stream.

500 MSCFD FTR Unit PFD

P-101A/B	R-101	V-101	E-101	V-102	E-102	V-103	E-103
Boiler Feed Water Pump	Packed Bed Fischer-Tropsch Reactor	Syncrude Three-Phase Separator	Light End Condenser	Light-Ends Three-Phase Separator	Light-Ends Water Condenser	Heavy-Ends Three-Phase Separator	Cooling Water Condenser

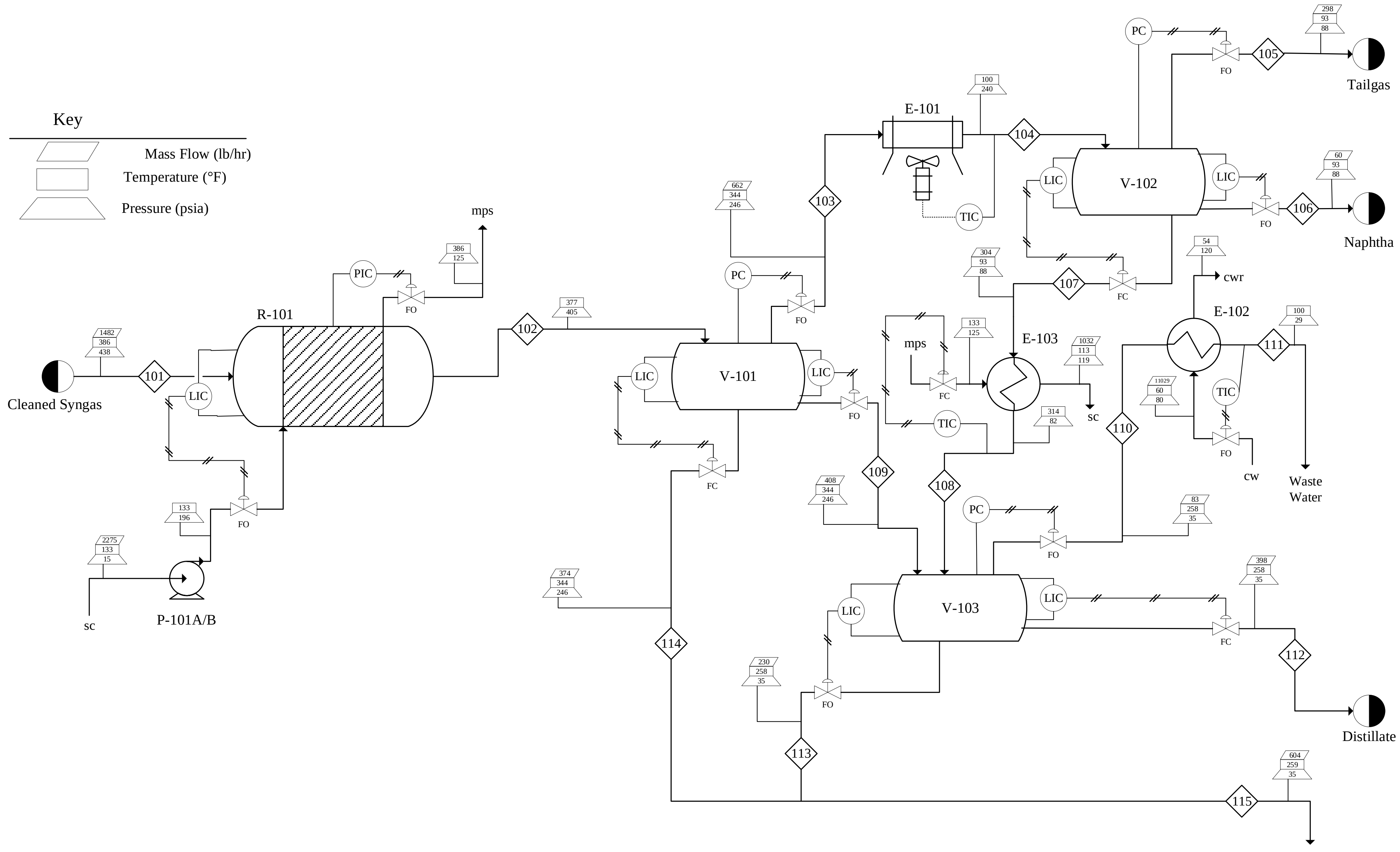

Figure 1: 500 MSCFD FTR Process Flow Diagram

Table 4: 500 MSCFD FTR Unit Stream Table

Stream Number	101	102	103	104	105	106	107	108	109	110	111	112	113	114	115
Vapor Fraction	1.00	0.56	1.00	0.46	1.00	0.00	0.00	0.11	0.00	1.00	0.01	0.00	0.00	0.00	0.06
Temperature (^oF)	386	377	344	100	93	93	93	314	344	258	100	258	258	344	259
Pressure (psia)	438	405	246	240	88	88	88	82	246	35	29	35	35	246	35
Mole Flowrate (lbmole/hr)	129.82	54.50	32.53	32.53	15.23	0.46	16.85	16.85	1.23	4.23	4.23	1.08	12.77	20.74	33.50
Mass Flowrate (lb/hr)	1482	1443	662	662	298	60	304	304	408	83	83	398	230	374	604
Enthalpy (-10^6BTU/hr)	-2.10	-5.10	-2.35	-2.75	-0.63	-0.06	-2.07	-1.97	-0.31	-0.43	-0.52	-0.32	-1.53	-2.45	-3.97
Density (lb/ft^3)	0.49	2.19	0.60	1.78	0.29	44.50	62.48	1.68	43.17	0.09	8.26	45.65	57.81	55.06	1.35
Vol. Flow (Barrel/day)	11649	2820	4702	1587	4332	6	21	774	40	3904	43	37	17	29	1918
Std. Id. Vol. Flow (Barrel/day)	294.8	141.9	81.5	81.5	55.0	5.7	20.8	20.8	34.7	6.0	6.0	33.8	15.8	25.6	41.4
Reid Vapor Pressure (psia)		1640.4				13.9			6.5	268.9	268.9	0.0			
Component Mass Flows (lb/hr)															
Methane	194.5	202.5	201.9	201.9	201.8	0.2	0.0	0.0	0.5	0.5	0.5	0.0	0.0	0.0	0.0
Ethane	0.0	0.6	0.6	0.6	0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	0.0	0.9	0.9	0.9	0.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Butane	0.0	1.2	1.1	1.1	1.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Pentane	0.0	8.4	8.2	8.2	7.1	1.2	0.0	0.0	0.2	0.2	0.2	0.0	0.0	0.0	0.0
n-Hexane	0.0	9.4	9.0	9.0	5.9	3.1	0.0	0.0	0.4	0.4	0.4	0.0	0.0	0.0	0.0
n-Heptane	0.0	10.3	9.5	9.5	3.7	5.8	0.0	0.0	0.8	0.6	0.6	0.1	0.0	0.0	0.0
n-Octane	0.0	10.9	9.5	9.5	1.7	7.9	0.0	0.0	1.4	0.9	0.9	0.4	0.0	0.0	0.0
n-Nonane	0.0	11.5	9.2	9.2	0.6	8.6	0.0	0.0	2.2	1.2	1.2	1.1	0.0	0.0	0.0
n-Decane	0.0	11.9	8.4	8.4	0.2	8.2	0.0	0.0	3.5	1.2	1.2	2.2	0.0	0.0	0.0
n-Undecane	0.0	12.2	7.2	7.2	0.1	7.1	0.0	0.0	5.0	1.1	1.1	3.9	0.0	0.0	0.0
n-Dodecane	0.0	12.4	5.8	5.8	0.0	5.8	0.0	0.0	6.6	0.9	0.9	5.8	0.0	0.0	0.0
n-Tridecane	0.0	12.6	4.2	4.2	0.0	4.2	0.0	0.0	8.3	0.6	0.6	7.8	0.0	0.0	0.0
n-Tetradecane	0.0	12.6	2.8	2.8	0.0	2.8	0.0	0.0	9.9	0.3	0.3	9.5	0.0	0.0	0.0
n-Pentadecane	0.0	12.7	1.9	1.9	0.0	1.9	0.0	0.0	10.7	0.2	0.2	10.5	0.0	0.0	0.0
n-Hexadecane	0.0	12.6	1.3	1.3	0.0	1.3	0.0	0.0	11.3	0.1	0.1	11.2	0.0	0.0	0.0
n-Heptadecane	0.0	12.5	0.8	0.8	0.0	0.8	0.0	0.0	11.7	0.1	0.1	11.7	0.0	0.0	0.0
n-Octadecane	0.0	12.4	0.5	0.5	0.0	0.5	0.0	0.0	11.9	0.0	0.0	11.9	0.0	0.0	0.0
n-Nonadecane	0.0	12.2	0.3	0.3	0.0	0.3	0.0	0.0	11.9	0.0	0.0	11.9	0.0	0.0	0.0
n-Icosane	0.0	12.0	0.2	0.2	0.0	0.2	0.0	0.0	11.8	0.0	0.0	11.8	0.0	0.0	0.0
C21-C25	0.0	56.2	0.2	0.2	0.0	0.2	0.0	0.0	56.0	0.0	0.0	56.0	0.0	0.0	0.0
C26-C29	0.0	39.7	0.0	0.0	0.0	0.0	0.0	0.0	39.7	0.0	0.0	39.7	0.0	0.0	0.0
C30-C35	0.0	50.2	0.0	0.0	0.0	0.0	0.0	0.0	50.2	0.0	0.0	50.2	0.0	0.0	0.0
C36-C47	0.0	70.1	0.0	0.0	0.0	0.0	0.0	0.0	70.1	0.0	0.0	70.1	0.0	0.0	0.0
C48+	0.0	82.1	0.0	0.0	0.0	0.0	0.0	0.0	82.1	0.0	0.0	82.1	0.0	0.0	0.0
Water	2.7	680.7	305.9	305.9	2.4	0.0	303.5	303.5	1.2	74.3	74.3	0.4	230.0	373.6	603.5
Carbon Dioxide	36.2	36.2	36.0	36.0	35.8	0.1	0.1	0.1	0.1	0.2	0.2	0.0	0.0	0.1	0.1
Carbon Monoxide	1084.2	29.9	29.9	29.9	29.9	0.0	0.0	0.0	0.1	0.1	0.1	0.0	0.0	0.0	0.0
Hydrogen	156.8	0.3	0.3	0.3	0.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	7.5	6.3	6.3	6.3	6.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

2500 MSCFD FTR Unit PFD

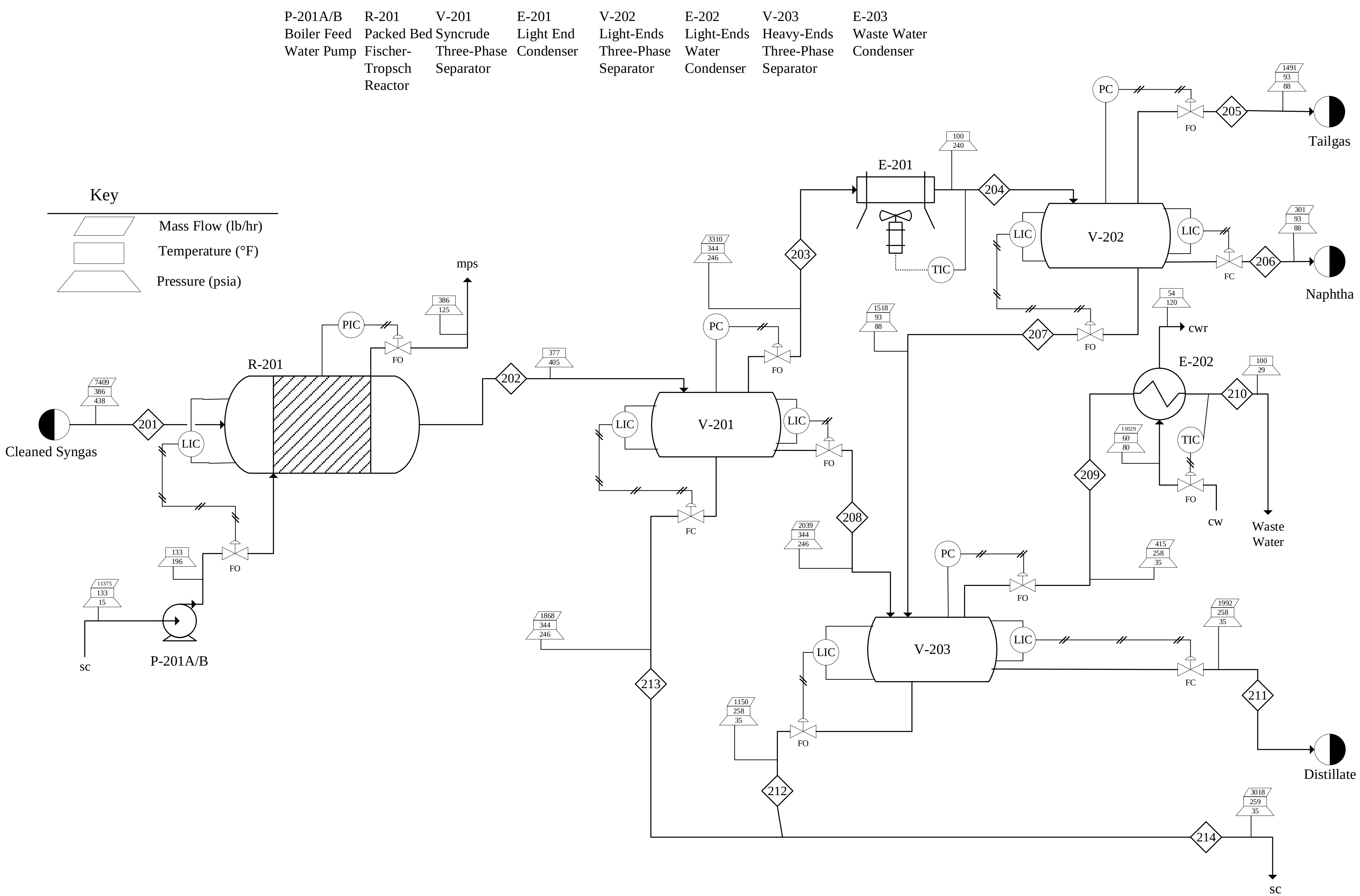

Figure 2: 2500 MSCFD FTR Unit Process Flow Diagram

Table 5: 2500 MSCFD FTR Unit Stream Table

Stream Number	201	202	203	204	205	206	207	208	209	210	211	212	213	214
Vapor Fraction	1.00	0.56	1.00	0.46	1.00	0.00	0.00	0.00	1.00	0.01	0.00	0.00	0.00	0.06
Temperature (^oF)	386	377	344	100	93	93	93	344	258	100	258	258	344	259
Pressure (psia)	438	405	246	240	88	88	88	246	35	29	35	35	246	35
Mole Flowrate (lbmole/hr)	649.1	272.5	162.7	162.7	76.2	2.3	84.2	6.1	21.1	21.1	5.4	63.8	103.7	167.5
Mass Flowrate (lb/hr)	7409	7217	3310	3310	1491	301	1518	2039	415	415	1992	1150	1868	3018
Enthalpy (10^6BTU/hr)	-10.49	-25.51	-11.73	-13.74	-3.12	-0.27	-10.34	-1.55	-2.15	-2.58	-1.60	-7.64	-12.23	-19.87
Density (lb/ft^3)	0.49	2.19	0.60	1.78	0.29	44.50	62.48	43.17	0.09	8.26	45.65	57.81	55.06	1.34
Volumetric Flow Rate (Barrel/day)	58247	14099	23508	7935	21659	29	104	202	19522	215	186	85	145	9594
Id. Volumetric Flow Rate (Barrel/day)	1474.0	709.3	407.6	407.6	274.9	28.5	104.1	173.5	29.9	29.9	168.8	78.9	128.2	207.1
Reid Vapor Pressure (psia)		1640.4				13.9		6.5	268.9	268.9	0.0			
Component Mass Flows (lb/hr)														
Methane	972.3	1012.3	1009.7	1009.7	1008.8	0.9	0.0	2.6	2.6	2.6	0.0	0.0	0.0	0.0
Ethane	0.0	3.0	3.0	3.0	3.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	0.0	4.4	4.4	4.4	4.3	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Butane	0.0	5.8	5.7	5.7	5.4	0.3	0.0	0.1	0.1	0.1	0.0	0.0	0.0	0.0
n-Pentane	0.0	42.2	41.1	41.1	35.3	5.8	0.0	1.2	1.1	1.1	0.1	0.0	0.0	0.0
n-Hexane	0.0	47.2	44.9	44.9	29.6	15.3	0.0	2.2	2.0	2.0	0.2	0.0	0.0	0.0
n-Heptane	0.0	51.3	47.3	47.3	18.4	28.8	0.0	4.0	3.2	3.2	0.7	0.0	0.0	0.0
n-Octane	0.0	54.6	47.7	47.7	8.4	39.4	0.0	6.9	4.7	4.7	2.2	0.0	0.0	0.0
n-Nonane	0.0	57.3	46.1	46.1	3.2	42.9	0.0	11.2	5.9	5.9	5.4	0.0	0.0	0.0
n-Decane	0.0	59.4	42.1	42.1	1.1	41.0	0.0	17.3	6.2	6.2	11.2	0.0	0.0	0.0
n-Undecane	0.0	61.0	36.0	36.0	0.3	35.7	0.0	25.0	5.5	5.5	19.5	0.0	0.0	0.0
n-Dodecane	0.0	62.1	29.0	29.0	0.1	28.9	0.0	33.1	4.3	4.3	28.8	0.0	0.0	0.0
n-Tridecane	0.0	62.9	21.2	21.2	0.0	21.1	0.0	41.7	2.8	2.8	38.9	0.0	0.0	0.0
n-Tetradecane	0.0	63.2	13.9	13.9	0.0	13.9	0.0	49.3	1.6	1.6	47.7	0.0	0.0	0.0
n-Pentadecane	0.0	63.3	9.7	9.7	0.0	9.7	0.0	53.5	1.0	1.0	52.5	0.0	0.0	0.0
n-Hexadecane	0.0	63.0	6.3	6.3	0.0	6.3	0.0	56.7	0.5	0.5	56.2	0.0	0.0	0.0
n-Heptadecane	0.0	62.6	4.0	4.0	0.0	4.0	0.0	58.6	0.3	0.3	58.3	0.0	0.0	0.0
n-Octadecane	0.0	61.9	2.4	2.4	0.0	2.4	0.0	59.4	0.2	0.2	59.3	0.0	0.0	0.0
n-Nonadecane	0.0	61.0	1.5	1.5	0.0	1.5	0.0	59.5	0.1	0.1	59.4	0.0	0.0	0.0
n-Icosane	0.0	60.0	0.9	0.9	0.0	0.9	0.0	59.2	0.0	0.0	59.1	0.0	0.0	0.0
C21-C25	0.0	281.1	1.1	1.1	0.0	1.1	0.0	280.0	0.0	0.0	280.0	0.0	0.0	0.0
C26-C29	0.0	198.4	0.1	0.1	0.0	0.1	0.0	198.3	0.0	0.0	198.3	0.0	0.0	0.0
C30-C35	0.0	250.8	0.0	0.0	0.0	0.0	0.0	250.8	0.0	0.0	250.8	0.0	0.0	0.0
C36-C47	0.0	350.6	0.0	0.0	0.0	0.0	0.0	350.6	0.0	0.0	350.6	0.0	0.0	0.0
C48+	0.0	410.6	0.0	0.0	0.0	0.0	0.0	410.6	0.0	0.0	410.6	0.0	0.0	0.0
Water	13.4	3403.5	1529.6	1529.6	12.2	0.0	1517.4	6.1	371.7	371.7	1.8	1150.0	1867.8	3017.7
Carbon Dioxide	180.8	180.9	180.0	180.0	179.1	0.4	0.5	0.7	1.1	1.1	0.0	0.0	0.3	0.3
Carbon Monoxide	5421.1	149.6	149.3	149.3	149.3	0.1	0.0	0.3	0.3	0.3	0.0	0.0	0.0	0.0
Hydrogen	783.8	1.3	1.3	1.3	1.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	37.3	31.7	31.7	31.7	31.6	0.0	0.0	0.1	0.1	0.1	0.0	0.0	0.0	0.0

5000 MSCFD FTR Unit PFD

P-301A/B	R-301	V-301	E-301	V-302	E-302	V-303	E-303
Boiler Feed Water Pump	Packed Bed Fischer-Tropsch Reactor	Syncrude Three-Phase Separator	Light End Condenser	Light-Ends Three-Phase Separator	Light-Ends Water Condenser	Heavy-Ends Three-Phase Separator	Waste Water Condenser

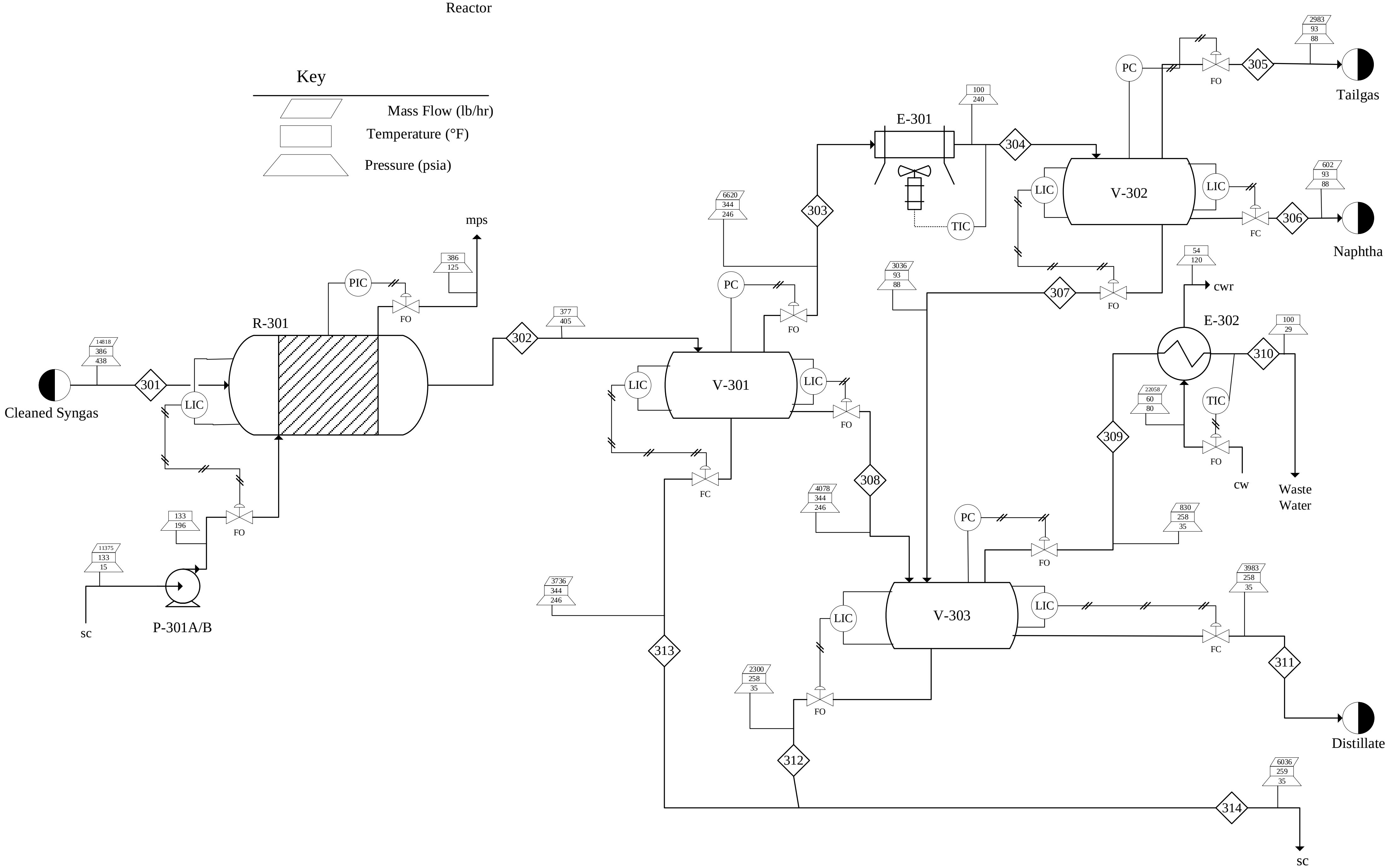

Figure 3: 5000 MSCFD FTR Unit Process Flow Diagram

Table 6: 5000 MSCFD FTR Unit Stream Table

Stream Number	301	302	303	304	305	306	307	308	309	310	311	312	313	314
Vapor Fraction	1.00	0.56	1.00	0.46	1.00	0.00	0.00	0.00	1.00	0.01	0.00	0.00	0.00	0.06
Temperature (^oF)	470	377	344	100	93	93	93	344	258	100	258	258	344	259
Pressure (psia)	438	405	246	240	88	88	88	246	35	29	35	35	246	35
Mole Flowrate (lbmole/hr)	1298.2	545.0	325.3	325.3	152.3	4.6	168.5	12.3	42.3	42.3	10.8	127.7	207.4	335.0
Mass Flowrate (lb/hr)	14818	14434	6620	6620	2983	602	3036	4078	830	830	3983	2300	3736	6036
Enthalpy (BTU/hr)	-20.97	-51.02	-23.47	-27.48	-6.25	-0.55	-20.68	-3.10	-4.31	-5.15	-3.19	-15.28	-24.46	-39.74
Density (lb/ft^3)	0.49	2.19	0.60	1.78	0.29	44.50	62.48	43.17	0.09	8.26	45.65	57.81	55.06	1.34
Volumetric Flow Rate (Barrel/day)	116493	28197	47016	15871	43318	58	208	404	39044	430	373	170	290	19187
Id. Volumetric Flow Rate (Barrel/day)	2948.1	1418.5	815.2	815.2	549.9	57.0	208.3	347.0	59.9	59.9	337.6	157.8	256.4	414.2
Reid Vapor Pressure (psia)		1640.4				13.9		6.5	268.9	268.9	0.0			
Component Mass Flows (lb/hr)														
Methane	1944.6	2024.5	2019.3	2019.3	2017.6	1.7	0.0	5.1	5.1	5.1	0.0	0.0	0.1	0.1
Ethane	0.0	6.0	6.0	6.0	6.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	0.0	8.8	8.7	8.7	8.6	0.1	0.0	0.1	0.1	0.1	0.0	0.0	0.0	0.0
n-Butane	0.0	11.6	11.4	11.4	10.9	0.6	0.0	0.2	0.2	0.2	0.0	0.0	0.0	0.0
n-Pentane	0.0	84.5	82.2	82.2	70.5	11.6	0.0	2.3	2.2	2.2	0.1	0.0	0.0	0.0
n-Hexane	0.0	94.3	89.9	89.9	59.2	30.6	0.0	4.4	4.0	4.0	0.5	0.0	0.0	0.0
n-Heptane	0.0	102.5	94.5	94.5	36.9	57.6	0.0	8.0	6.5	6.5	1.5	0.0	0.0	0.0
n-Octane	0.0	109.2	95.5	95.5	16.7	78.7	0.0	13.7	9.4	9.4	4.4	0.0	0.0	0.0
n-Nonane	0.0	114.6	92.2	92.2	6.3	85.9	0.0	22.4	11.7	11.7	10.7	0.0	0.0	0.0
n-Decane	0.0	118.8	84.2	84.2	2.1	82.0	0.0	34.7	12.4	12.4	22.3	0.0	0.0	0.0
n-Undecane	0.0	122.0	72.1	72.1	0.6	71.4	0.0	50.0	10.9	10.9	39.0	0.0	0.0	0.0
n-Dodecane	0.0	124.3	58.1	58.1	0.2	57.9	0.0	66.2	8.5	8.5	57.7	0.0	0.0	0.0
n-Tridecane	0.0	125.7	42.3	42.3	0.0	42.3	0.0	83.4	5.6	5.6	77.9	0.0	0.0	0.0
n-Tetradecane	0.0	126.5	27.9	27.9	0.0	27.9	0.0	98.6	3.2	3.2	95.4	0.0	0.0	0.0
n-Pentadecane	0.0	126.6	19.5	19.5	0.0	19.5	0.0	107.1	2.0	2.0	105.1	0.0	0.0	0.0
n-Hexadecane	0.0	126.1	12.6	12.6	0.0	12.6	0.0	113.5	1.1	1.1	112.4	0.0	0.0	0.0
n-Heptadecane	0.0	125.2	7.9	7.9	0.0	7.9	0.0	117.2	0.6	0.6	116.6	0.0	0.0	0.0
n-Octadecane	0.0	123.8	4.9	4.9	0.0	4.9	0.0	118.9	0.3	0.3	118.6	0.0	0.0	0.0
n-Nonadecane	0.0	122.1	3.1	3.1	0.0	3.1	0.0	119.0	0.2	0.2	118.8	0.0	0.0	0.0
n-Icosane	0.0	120.1	1.7	1.7	0.0	1.7	0.0	118.3	0.1	0.1	118.3	0.0	0.0	0.0
C21-C25	0.0	562.2	2.2	2.2	0.0	2.2	0.0	560.0	0.1	0.1	560.0	0.0	0.0	0.0
C26-C29	0.0	396.9	0.2	0.2	0.0	0.2	0.0	396.7	0.0	0.0	396.7	0.0	0.0	0.0
C30-C35	0.0	501.7	0.0	0.0	0.0	0.0	0.0	501.7	0.0	0.0	501.7	0.0	0.0	0.0
C36-C47	0.0	701.2	0.0	0.0	0.0	0.0	0.0	701.2	0.0	0.0	701.2	0.0	0.0	0.0
C48+	0.0	821.2	0.0	0.0	0.0	0.0	0.0	821.2	0.0	0.0	821.2	0.0	0.0	0.0
Water	26.9	6806.9	3059.2	3059.2	24.5	0.1	3034.7	12.2	743.4	743.4	3.6	2299.9	3735.5	6035.5
Carbon Dioxide	361.7	361.9	359.9	359.9	358.2	0.8	0.9	1.4	2.3	2.3	0.0	0.0	0.6	0.6
Carbon Monoxide	10842.2	299.3	298.7	298.7	298.6	0.1	0.0	0.5	0.5	0.5	0.0	0.0	0.1	0.1
Hydrogen	1567.5	2.6	2.6	2.6	2.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	74.6	63.4	63.3	63.3	63.3	0.0	0.0	0.1	0.1	0.1	0.0	0.0	0.0	0.0

Simplified GTL Plant Process Flow Diagram

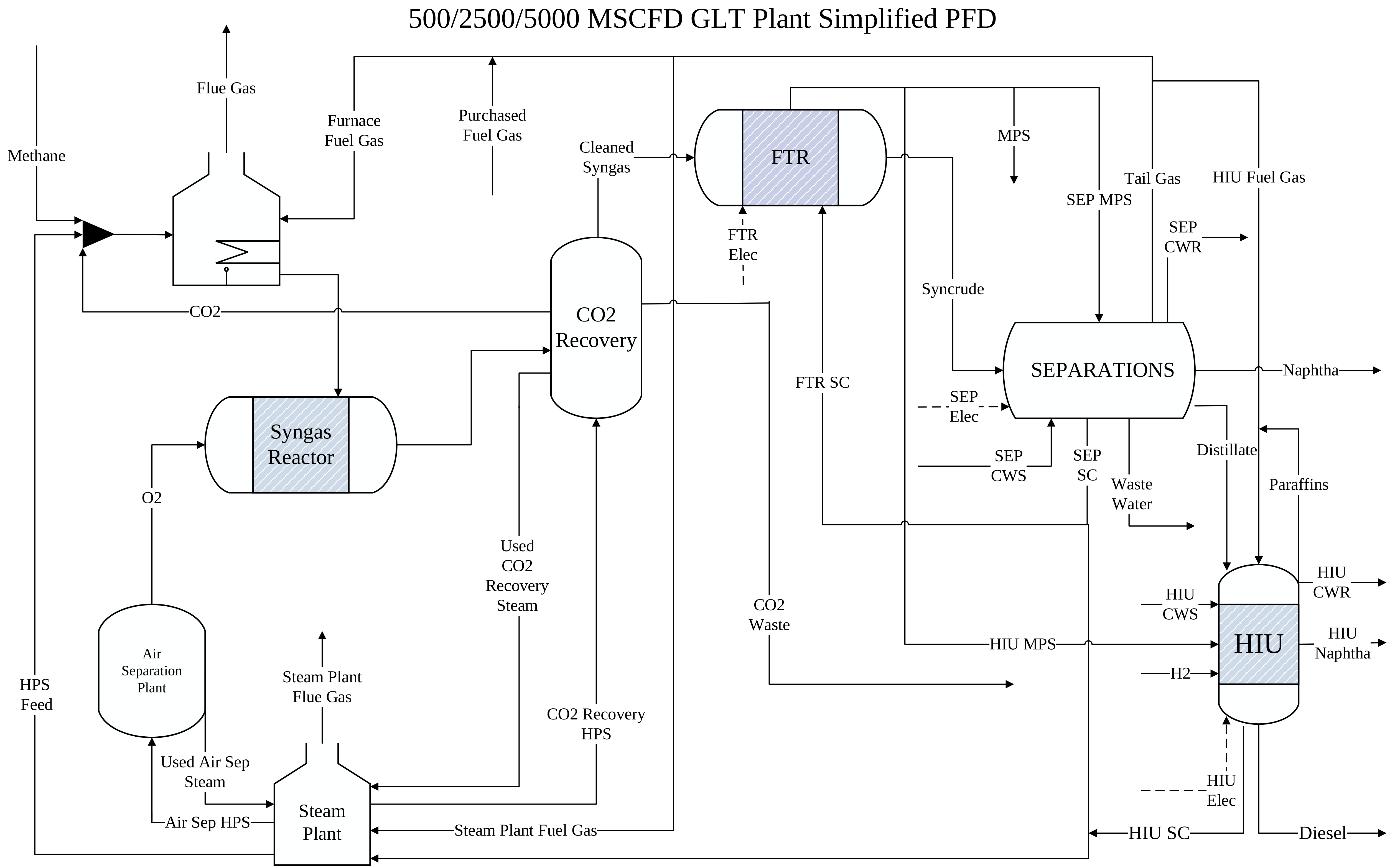

Figure 4: Simplified GTL Plant Process Flow Diagram

Safety and Environmental Summary

The primary consideration for all design decisions is the safety and environmental impact. A concerted effort was placed on identifying process hazards, evaluating design alternatives with an inherent safety approach, designing passive, active, and procedural controls, to prevent and mitigate the impact of remaining process hazards. Energy efficiency, the second highest weighted decision factor, is 93.9%.

Four waste streams are generated as part of this process: carbon dioxide purge from CO_2 recovery, wastewater from separations, and flue gases from the boiler and fired-heater exhaust. Each can be controlled to the Best-Available Control Technology (BACT); the laws and regulations of the Texas Commission on Environmental Quality were utilized, as a model law for this purpose[19]. Of the chemical species potentially present in the process, nitrogen oxides produced through combustion (NOx), carbon monoxide (CO), and volatile organic carbons (VOCs) are subject to BACT requirements. The CO2 purge stream contains a small, albeit substantive, quantity of CO that cannot be readily controlled by changes to operating conditions, which would not generate a much greater environmental impact. The BACT, an oxidative catalyst module, was installed on the CO2 recovery unit, to oxidize up to 98% of the remaining CO to CO2, as well as any trace VOCs[20]. VOCs containing wastewater will be disposed of by a properly vetted contractor, as the hydrocarbons cannot be adequately removed without prohibitively high capital investment. Low NOx burners, meeting CO omissions requirements, represent the BACT for both boilers and fired heaters[21],and will be installed on the respective equipment.

Inherent Safety Evaluation

Inherently safer design served as an integral guiding consideration in all design decisions, especially, when evaluating amongst design alternatives. Each of the inherently safer design principles, defined by the Centers for Chemical Process safety, were applied to this project22; examples are as follows:

<u>Substitution</u>

- Water was selected as the absorber solvent in the CO2 recovery unit, opposed to the more toxic and flammable amine solutions.
- Ambient air was used to cool lines 103/203/303, reducing the hazards associated with fouling and corrosion-causing components, which would have resulted from using cooling water.

<u>Minimization</u>

- A plug-flow FTR reactor was used in place of a larger batch, or fluidized bed reactors, to reduce the quantity of reactants and the consequences of a reactor runaway.
- Consuming the LPG fraction and tail gas as fuel, gas on-site minimizes the stored quantity of compressed, flammable gases.
- Oxygen is completely consumed in the syngas reactor, reducing the oxidizer concentration in other parts of the process.

Moderation

- Consuming the LPG fraction precludes the need for higher pressure processes, required to condense the stream.
- Three-phase separators allow for the naphtha-tail gas and distillate-water fractionations to occur at lower pressures than if a single, large column were used.
- Cooling the CO2 recovery feed, between the syngas reactor and absorber, reduces the temperature for the bulk of the process.

Simplification

- Using a single fired heater reduces the equipment required to heat the syngas reactor.
- Using a packed bed, rather than a fluidized bed, greatly simplifies the control and operational parameters.
- Utilizing three-phase separators, rather than distillation columns, reduces the complexities, in terms of additional heat exchange equipment, start-up, and process control, associated with distillation columns.
- Implementing three-phase separators, rather than two-phase separators, allows for a greater degree of separation to occur with less equipment.
- Ambient air was used to cool lines 103/203/303, rather than cooling water, reducing the equipment associated with the latter.

Process Safety Management

Process Hazards

There are few hazards in the GTL plant that exceed the baseline risks, associated with a wellsite. Hydrogen and all hydrocarbons produced are flammable, many being able to form explosive mixtures with air. Proper facility citing, pressure relief, and fireproofing are vital to safe operation. Facility citing should consider the hazards associated with oil and gas extraction and endeavor, in order to minimize the hazards of extraction operations. Additionally, this process produces carbon monoxide, a toxic, colorless, odorless gas; proper monitoring for carbon monoxide and combustible gases is essential to create a safe work environment. All relief devices and transient hydrocarbon waste streams should be vented to a flare. More information on the inherent hazards of the chemical species involved can be found in the Appendix.

Special consideration should be taken toward the prevention and mitigation of an FTR runaway and an uncongested vapor cloud explosion.

FTR Runaway Hazard

The FTR represents the single most inherently hazardous unit in the GTL Plant, predominantly due to the runaway hazard of the exothermic Fischer-Tropsch Reaction. In exothermic reactors, insufficient cooling capacity will cause an uncontrollable increase in temperature, with severe consequences. However, there is no known increase in stoichiometric pressure in the FTR, as opposed to the circumstances behind more infamous runaways[23]. Reactor parameters at operating conditions, minimum runaway conditions, and minimum runaway conditions at 0 psig are shown in Table 7. To anticipate the hazards associated with a runaway

FTR, a worst-case scenario runaway, assuming feed temperature 458°F, average hydrocarbon length of 1.50 (the average length at 1664°F), and hydrocarbon heat capacity of methane was modeled; the temperature profile is shown in Figure 5. However, laboratory and pilot plant studies should be pursued to confirm this runaway model and verify the lack of secondary reactions.

Table 7: Operating and Runaway Conditions

Parameter	Operating Condition	Runway	Runaway at 0 psig Jacket
Feed temperature (°F)	386	449	548
Feed flow (mol/hr)	58.9	16.8	2.0
Feed Pressure (psia)	438	617	>15,000
Jacket Pressure (psia)	189.5	215.5	N/A

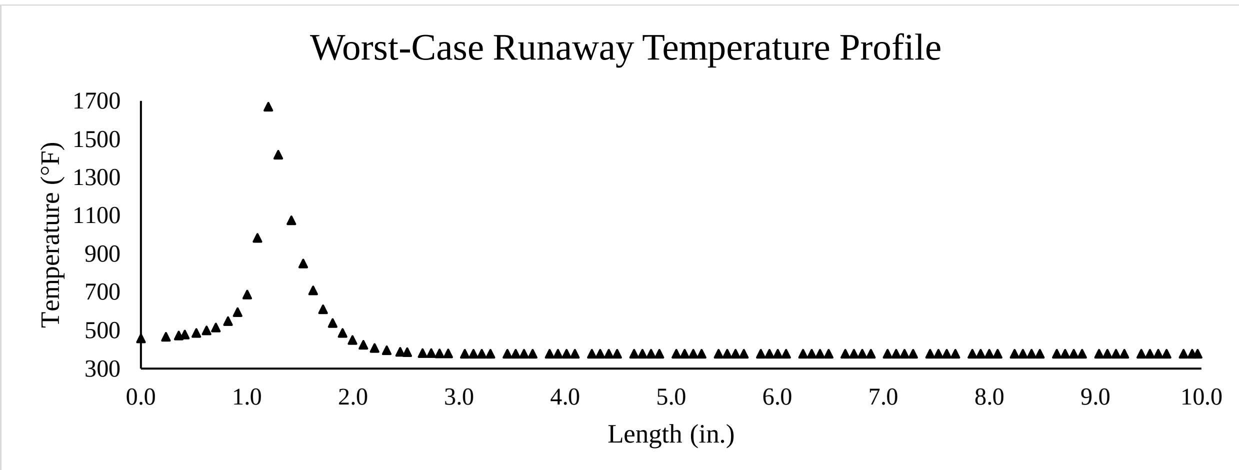

Figure 5: Worst-Case Runaway Temperature Profile

From Table 7 and Figure 5, it is apparent that the focus of FTR safety efforts should be upon prevention, as opposed to mitigation of effects. Notably, no pressure readings above the feed pressure were observed in the reactor model, and outlet temperature was an unreliable means of determining the presence of a runaway. At the temperatures concentrated over the length predicted, a loss of metallurgical integrity in most materials, including carbon steel, would occur. Adding heat resistant coatings to the FTR or using heat resistant materials of construction should be considered to limit the impact of a runaway event, although it is unlikely the reactor could sustain such an event without substantive damage, requiring major repair.

To prevent a runaway event, several active safety systems should be put into effect. First, the control system regulates the pressure of the two-phase region on the shell side of the reactor, as well as the level in the reactor. It is apparent from the runaway conditions at 0 psig, simply lowering the pressure on the shell side of the FTR can be sufficient to prevent reactor runaway. There are systems designed to detect and alarm when the process upsets, including when feed temperature exceeds its bounds, there are signs of reaction runaway, the flow rate or outlet pressure increase. Appropriate safety interlocks should be designed to increase reaction cooling, quenching the reactor with steam, when a runaway is detected. Provisions should be made to supply cooling media in the event of P-101A/B, P-201A/B, or P-301A/B failure, either through a redundant pump with an independent power supply or the utilization of cooling water. The shell side of the reactor is protected by both a pressure safety valve and a rupture disk; both devices should be set so that the overpressure experienced is less than the minimum runaway pressures.

P&IDs of the Major Fractionator

500 MSCFD FTR Unit P&ID

P-101A/B	R-101	V-101	E-101	V-102	E-102	V-103	E-103
Boiler Feed Water Pump	Packed Bed Fischer-Tropsch Reactor	Syncrude Three-Phase Separator	Light End Condenser	Light-Ends Three-Phase Separator	Light-Ends Water Condenser	Heavy-Ends Three-Phase Separator	Waste Water Condenser

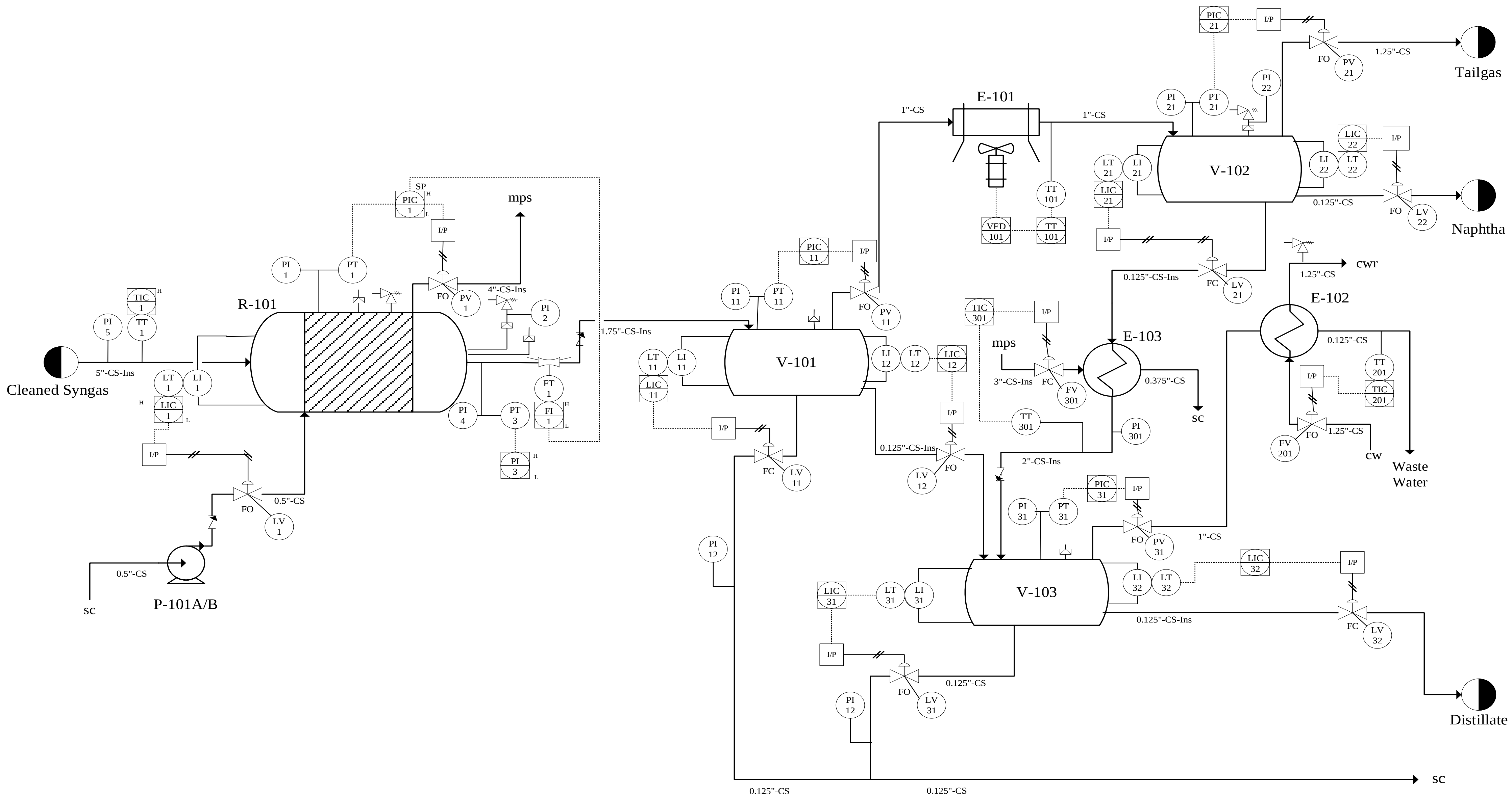

Figure 6: 500 MSCFD FTR Unit P&ID

2500 MSCFD FTR Unit P&ID

P-201A/B	R-201	V-201	E-201	V-202	E-202	V-203	E-203
Boiler Feed Water Pump	Packed Bed Fischer-Tropsch Reactor	Syncrude Three-Phase Separator	Light End Condenser	Light-Ends Three-Phase Separator	Light-Ends Water Condenser	Heavy-Ends Three-Phase Separator	Waste Water Condenser

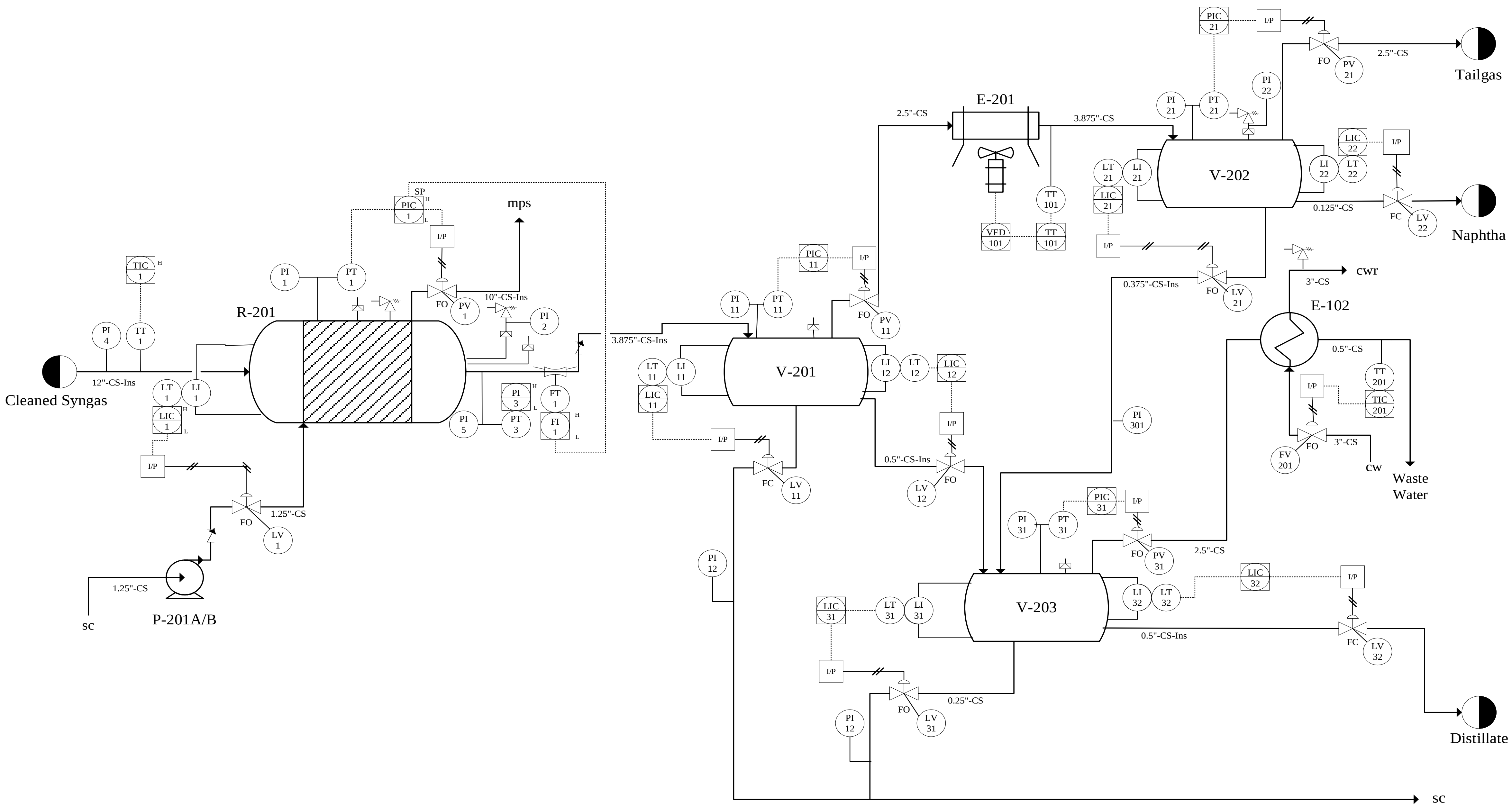

Figure 7: 2500 MSCFD FTR Unit P&ID

5000 MSCFD FTR Unit P&ID

P-301A/B	R-301	V-301	E-301	V-302	E-302	V-303	E-303
Boiler Feed Water Pump	Packed Bed Fischer-Tropsch Reactor	Syncrude Three-Phase Separator	Light End Condenser	Light-Ends Three-Phase Separator	Light-Ends Water Condenser	Heavy-Ends Three-Phase Separator	Waste Water Condenser

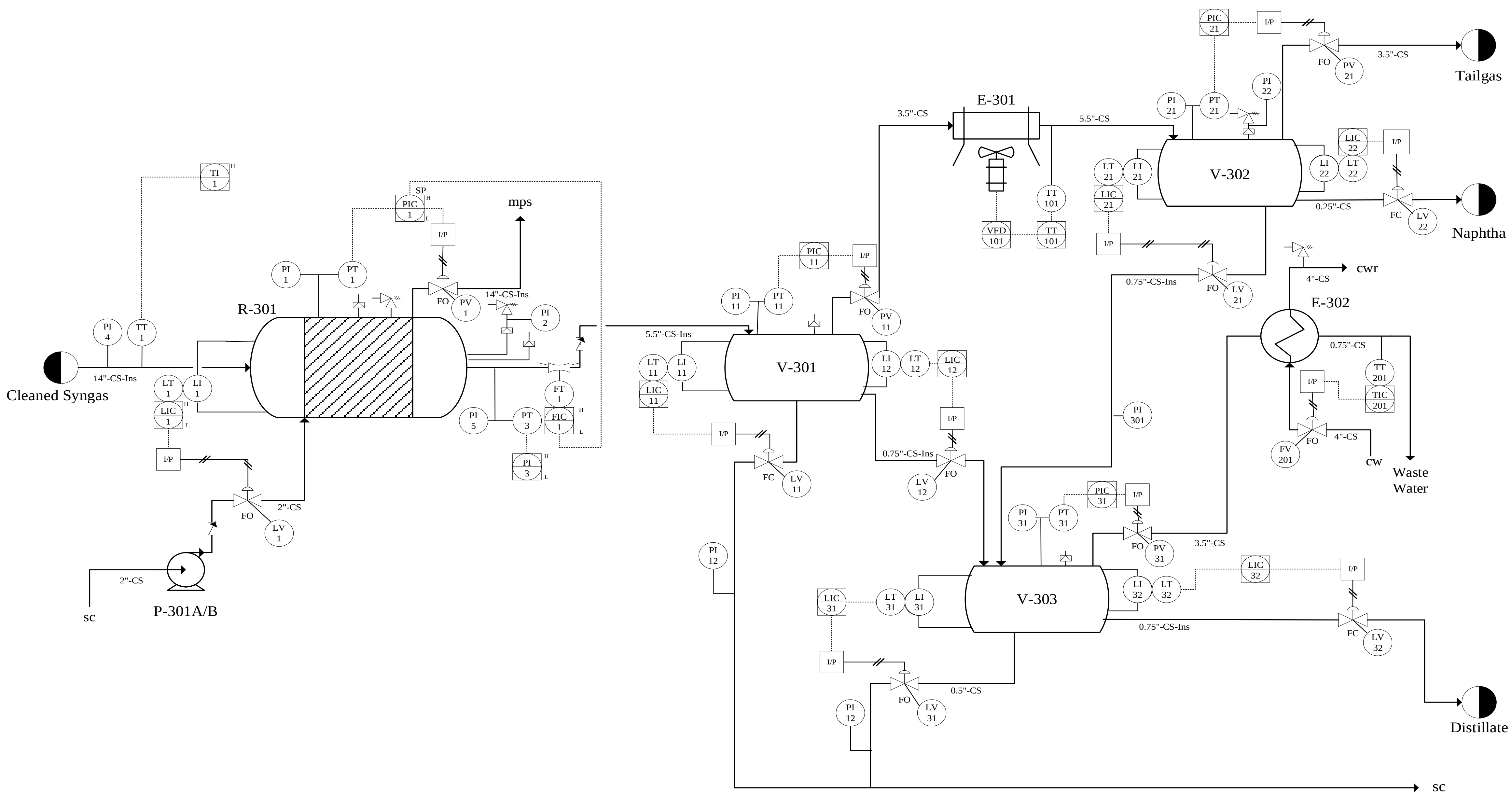

Figure 8: 5000 MSCFD FTR Unit P&ID

Uncongested Vapor Cloud Deflagration

The deflagration of an uncongested vapor cloud is able to occur at any point, where a potential vapor leak can be found in the process. This is most likely to occur in the Syngas unit and the FTR. These units were determined to be the most possible areas for leaks to occur, due to the high pressures, temperatures, as well as having the largest capacity for vapors to reside. The most probable ignition source locations for deflagration in the process are at the syngas unit furnace and the FTR feed heat exchanger. The best possible actions for prevention and mitigation of deflagration are having daily leak checks by operators, emergency shut offs, in case of leaks, an installation of automated block valves on fuel and feed lines, and an increasing distance of ignition sources from flammable and explosive materials.

Safety Summary

There are several substantive hazards associated with the GTL process that require a dedicated process safety plan to manage; however, few hazards are greater or different to those typically encountered on a wellsite. Inherently safer design principles were applied throughout the design to minimize the safety risks of the project. For the hazards that could not be avoided through design alternatives, passive and active measures to prevent or mitigate the hazards were explored, with several recommended for implementation in the latter stages of design. Special effort was taken to determine the conditions under which the FTR would runaway, with the bulk of safety measures focused on mitigating this aspect of the design. Safety was central to the design of the control system as well as line sizing. Finally, the impacts of and mitigation for an uncongested vapor cloud deflagration was considered. The result of these efforts is a plant mitigated to an acceptable level of risk.

Heat Exchangers	**E-101**	**E-102**	**E-103**	**Heat Exchangers**	**E-201**	**E-202**		**Heat Exchangers**	**E-301**	**E-302**	
Type	Fin Fan Air	Floating Head	Double Pipe	Type	Floating Head	Floating Head		Type	Floating Head	Floating Head	
Area (ft^2)	14.2	38.7	17.7	Area (ft^2)	936	7082		Area (ft^2)	936	7082	
Duty (10^3 BTU/hr)	93	441	110	Duty (10^6 BTU/hr)	49.2	19.4		Duty (10^6 BTU/hr)	49.2	19.4	
Shell				Shell				Shell			
Temp. (°F)	75	343.9	94.1	Temp. (°F)	500	131		Temp. (°F)	500	131	
Press. (PSIA)	175.0	132.0	175.0	Press. (PSIA)	175.0	132.0		Press. (PSIA)	175.0	132.0	
Phase	Condensing	Condensing	2-Phase	Phase	Condensing	Condensing		Phase	Condensing	Condensing	
MOC	CS	CS	CS	MOC	CS	CS		MOC	CS	CS	
Stream	Air	process	Process	Stream	Mps	process		Stream	Mps	process	
Tube				Tube				Tube			
Temp. (°F)	343.9	353	80	Temp. (°F)	253	120		Temp. (°F)	253	120	
Press. (PSIA)	246.0	139.7	60	Press. (PSIA)	130.0	64.7		Press. (PSIA)	130.0	64.7	
Phase	Liquid	Vapor	Liquid	Phase	Liquid	Liquid		Phase	Liquid	Liquid	
MOC	CS	CS	CS	MOC	CS	CS		MOC	CS	CS	
Vessels	**V-101**	**V-102**	**V-103**	**Vessels**	**V-201**	**V-202**	**V-203**	**Vessels**	**V-301**	**V-302**	**V-303**
Orientation	Horizontal	Horizonal	Horizonal	Orientation	Horizontal	Horizonal	Horizonal	Orientation	Horizontal	Horizonal	Horizonal
Temp. (°F)	343.9	93.9	257.8	Temp. (°F)	120	109	105	Temp. (°F)	120	109	105
Press. (PSIA)	246	98	35	Press. (PSIA)	131.1	130	220.9	Press. (PSIA)	131.1	130	220.9
MOC	CS	CS	CS	MOC	CS	CS	CS	MOC	CS	CS	CS
Height/length (ft)	17.6	12.9	13.1	Height/length (ft)	17.6	12.9	13.1	Height/length (ft)	17.6	12.9	13.1
Diameter	14	10	8	Diameter	14	10	8	Diameter	14	10	8
Reactor	**R-101**			**Reactor**	**R-201**			**Reactor**	**R-301**	**R-302**	
Type	Floating Head			Tubesheet Type	Floating Head			Tubesheet Type	Floating Head	Floating Head	
Duty (10^6 BTU/hr)				Duty (10^6 BTU/hr)				Duty (10^6 BTU/hr)			
Shell				Shell				Shell			

Temp. (°F)	386	Temp. (°F)	386			Temp. (°F)	386	386	
Press. (PSIA)	196	Press. (PSIA)	196			Press. (PSIA)	196	196	
Phase	Evaporating	Phase	Evaporating			Phase	Evaporating	Evaporating	
MOC	CS	MOC	CS			MOC	CS	CS	
Stream	SC	Stream	SC			Stream	SC	SC	
Tube		Tube				Tube			
Number	1000	Number	2500			Number	2500	2500	
Diameter (in)	1.25	Diameter (in)	1.25			Diameter (in)	1.25	1.25	
Length (ft)	36.33	Length (ft)	36.33			Length (ft)	36.33	36.33	
Temp. (°F)	253	Temp. (°F)	253			Temp. (°F)	253	253	
Press. (PSIA)	130.0	Press. (PSIA)	130.0			Press. (PSIA)	130.0	130.0	
MOC	CS	MOC	CS			MOC	CS	CS	
Pumps	**P-101A/B**	**Pumps**	**P-201A/B**			**Pumps**	**P-301A/B**	**P-302A/B**	
Flow (gpm)	20	Flow (gpm)	20			Flow (gpm)	20	20	
Fluid Density (lb/ft^3)	62.4	Fluid Density (lb/ft^3)	62.4			Fluid Density (lb/ft^3)	62.4	62.4	
Brake Power (hp)	10.6	Brake Power (hp)	10.6			Brake Power (hp)	10.6	10.6	
ΔP (PSI)	235	ΔP (PSI)	235			ΔP (PSI)	235	235	
Discharge (PSIA)	250	Discharge (PSIA)	250			Discharge (PSIA)	250	250	
MOC	CS	MOC	CS			MOC	CS	CS	

Figure 9: Equipment Information Summary

Unit Control and Instrumentation Description

There are two handles on the FTR, both on the cooling jacket side: LV1 and PV1. LV1 controls the flow of steam condensate from P101A/B, P201A/B, or P301A/B into the FTR through LIC1, based off the liquid level in the FTR jacket. The pressure in the jacket is controlled through a cascade loop, so any runaways can be quickly mitigated. The pressure is primarily controlled by PV1 through PIC1, as measured by PT1. The secondary controller, FI1, uses the flow rate of the effluent hydrocarbons to establish an external set point for LIC1. The FTR is supported by a wide swath of additional instrumentation, primarily to prevent or detect potential runaway conditions. As the cleaned syngas enters the FTR, the temperature is measured by TT1 and transmitted by TI1 to the control room. Leaving the FTR, the pressure is measured by PT3, before the outlet flow is measured by FI1. To prevent runaways, TIC1, LIC1, PIC1, PI3, and F1 were outfitted with appropriate alarms.

Each of the three-phase separators (V-101/V-102/V-103) implemented the same control methodology. The flow of the aqueous phase out of LV11/LV21/LV31 were controlled by LIC11/LIC21/LIC31, based on the level of the aqueous/organic liquid phase interface. It is likely the total liquid level, as transmitted by LT12/LT22/LT32, controlled the flow of organic phase through LV12/LV22/LV32. Pressure in the vessel was controlled by PIC11/PIC21/PIC31 through PV-11/PV-21/PV31.

The controlled variable for all heat exchangers was the temperature of the effluent process fluid, while the manipulated variable was the flow of the cooling or heating media. FV201 and FV301 were controlled by TIC201 and TIC301 respectively, while the fan of E-101/E-201/E-301 was controlled via TT101, through a variable frequency drive.

All piping in the FTR and separation unit were sized, based on either 3-phase, 2-phase, liquid, and gas flows. The basis for sizing 3-phase flow was varying the diameter to adjust the line velocity and move liquid and gas volumetric flows into a stratified flow regime, based on a flow regime chart in GPSA[5]. Calculation of 2-phase flow was accomplished by using a diameter calculation specified in the PDH course on optimum pipe sizing[24]. For liquid flow, the inner diameter of the pipe was assumed, then plugged into the continuity equation to calculate the line velocity of the fluid. Based on heuristic data, found from Norsok Standard[25], a velocity range of 2.6 ft/s and 19.7 ft/s was used to figure the optimum diameter of the pipe. For gas flow, diameter was varied and used in the technique, found from “Pipe Line Rules of Thumb Handbook,” to find the optimum diameter. Heuristic ranges of 15 ft/s to 60-80 ft/s, found from PetroWiki[26], were used for determining an appropriate line diameter for gas flow. Welds for pipe connections were selected to lower the chance of mass flow leaks between units and equipment.

Economics

Capital Costs

The capital costs for each unit are shown in Tables 8. The cost of the syngas unit, hydroisomerization unit, CO_2 unit, and steam plant were created from the base capital expenditures, multiplied by a capacity factor, and the sixth-tenths rule, as prescribed by AIChE3. The cost of the FTR was based on the layout of the reactor, pressure, and size. The separators' cost was based on the load and the maximum allowable working pressure (MAWP). Finally, the cost of the heat exchangers was based on a purchased equipment cost scaling factor. All equipment was costed, using a CEPCI scaling factor of 1.67. This was found based on historical CEPCI data[27] and an analyzing a trendline. Equipment was also multiplied by a factor of 4.8, to account for additional costs and working capital, and a Lang factor of 1.7, per the project statement.

Table 8: Capital Cost of 500 MSCFD Unit

Equipment	Cost
SynGas	$2,126,000
Hydroisomerization	$1,492,000
CO_2 Recovery	$961,000
Steam Plant	$69,000
FTR	$2,020,000
Separators	$101,000
Heat Exchangers	$692,000
TOTAL	**$7,461,000**

Table 9: Capital Cost of 2.5 MMSCFD Unit

Equipment	Cost
SynGas	$6,251,000
Hydroisomerization	$3,264,000
CO_2 Recovery	$2,329,000
Steam Plant	$345,000
FTR	$3,640,000
Separators	$112,000
Heat Exchangers	$1,059,000
TOTAL	**$17,360,000**

Table 10: Capital Cost of 5 MMSCFD Unit

Equipment	Cost
SynGas	$9,949,000
Hydroisomerization	$5,305,000
CO_2 Recovery	$3,410,000
Steam Plant	$690,000
FTR	$8,990,000
Separators	$125,000
Heat Exchangers	$1,588,000
TOTAL	**$30,062,000**

Operating Costs

The operating costs were found using the specifications given in the problem statement. The utility specifications can be found in Table 11 and the total utility costs for each unit can be found in Table 12.

Table 11: Utility Specifications

Utility	Cost	Credit
HP Steam	$5/klb consumed	$4/klb produced
MP Steam	$4/klb consumed	$3/klb produced
LP Steam	$3.5/klb consumed	$2.5/klb produced
Electricity	$0.04/kWh consumed	$0.03/kWh produced
Fuel Gas	$3/MBTU consumed	$2/MBTU produced
Hydrogen	$0.06/lb consumed	
Carbon Dioxide	$400/MSCF consumed	
Steam Condensate		$2/klb produced
Process/Cooling Tower Water	$0.50/kgal consumed	$0.35/klb produced
Waste Water Treatment	$6/kgal produced	

Table 12: Utility Costs

Unit	Total Utility Cost
500 MSCFD	$1,257,055
2.5 MMSCFD	$6,285,300
5 MMSCFD	$12,570,550

In addition to utilities, operational labor was accounted for, assuming each operator would work 40 hours/week and 50 weeks/year. The hourly wage of a plant and system operator in the oil and gas extraction industry was found through the Bureau of Labor Statistics[28]. Operational labor costs are found in Table 13.

Table 13: Operator Costs

Number of Operators	4
Hourly Wage	$39.50
Total Operator Cost per Year	$316,000

Operating costs other than utilities and labor were accounted for by taking 3% of the total capital investment.

Logistics Considerations

Groupings

For the purpose of optimizing the capacity of the central plant, an approach of grouping wells together was taken. An optimization analysis, similar to the transportation problem, used to decide optimal transportation locations and resource allocations[29], was run in Excel to optimize the combinations, while trying to create groups with total production as close to 30,000 MSCFD as possible. The analysis resulted in groups shown below in Table 14. These groupings were the basis of optimization and used during the redeployment scheduling process discussed.

Table 14: Groupings

Group 1	Group 2	Group 3	Group 4
1B	2A	1C	1G
1E	1D	1H	1A
1F	2D	2E	1B
2C	2F	2G	
	2H		

Optimization Process

In order to optimize the unit capacity on each well, all possible unit combinations for production of each well were analyzed. Production was based on year 1 production, due to it being the highest well head flow rate. Once all possible combinations were identified, a present worth cost analysis was performed for each combination on each well. The analysis was conducted through the first 11 years, including annual production revenue, annual utility costs, annual trucking costs, and annual equipment costs.

With combinations analyzed, the cases were ranked from best to worst. The top 3 cases for each well were organized in a table and considered for implementation. A theoretical Net Present Value (NPV) was calculated, as if all required modules were built in a well's 11-year operation life. The cost of all potential combinations of reactors was also considered. The 3 best cases were analyzed to optimize the number of each size unit that would be fabricated. This included trying to keep the number of each size unit equal throughout all groups while still maximizing NPV. The best cases, based on NPV across a 11-year life only considering utilities, revenue and number of units without accounting for fixed capital costs, are shown below in

Table 15. Capital costs were not accounted for under the assumption that all units would be fabricated during year 0 before production begins. The best and worst case NPVs of the top 3 cases for each well were analyzed and totaled by group. The totals of the top 3 cases were analyzed and used to determine which case would be used for which group.

Table 15: Best and Worst Case

Well ID	Best Case NPV	Worst Case NPV	# of 500 MSCFD	# of 2.5 MMSCFD	# of 5 MMSCFD
1A	$17.04 MM	($13.89) MM	1	1	0
1B	$11.46 MM	($14.45) MM	0	0	1
1C	$32.10 MM	($29.75) MM	0	0	2
1D	$28.15 MM	($20.82) MM	0	1	1
1E	$12.66 MM	($23.43) MM	0	2	0
1F	$55.44 MM	($37.34) MM	0	0	3
1G	$43.84 MM	($36.05) MM	0	1	2
1H	$26.65 MM	($22.32) MM	0	1	1
2A	$34.61 MM	($27.24) MM	0	0	2
2B	$46.19 MM	($33.70) MM	0	1	2
2C	$33.67 MM	($28.15) MM	0	0	2
2D	$14.64 MM	($16.29) MM	0	0	1
2E	$29.80 MM	($27.03) MM	1	1	1
2F	$11.83 MM	($14.87) MM	1	1	0
2G	$27.37 MM	($21.60) MM	0	1	1
2H	$29.44 MM	($24.20) MM	0	1	1

This analysis led to the conclusion that five 5 MMSCFD units, three 2.5 MMSCFD units, and one 500 MSCFD unit would be fabricated in year 0. The units fabricated are shown below in Table 16.

Table 16: Unit Capacities

Unit Size	Unit Name
500 MSCFD	A1
2.5 MMSCFD	B1
2.5 MMSCFD	B2
2.5 MMSCFD	B3
5 MMSCFD	C1
5 MMSCFD	C2
5 MMSCFD	C3
5 MMSCFD	C4
5 MMSCFD	C5

After units were determined, a sensitivity analysis was run to determine the optimum time for a well to produce, while maximizing NPV. The graphs of the varying times v. NPV are shown below for Group 1, Group 3, and Group 4.

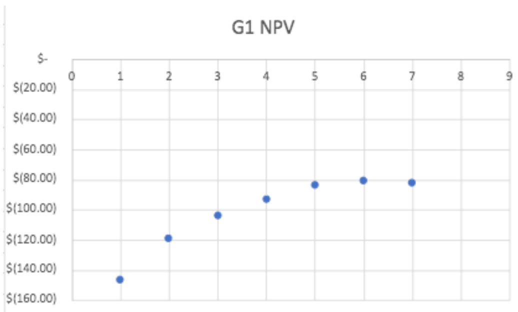

Figure 10: Group 1 NPV

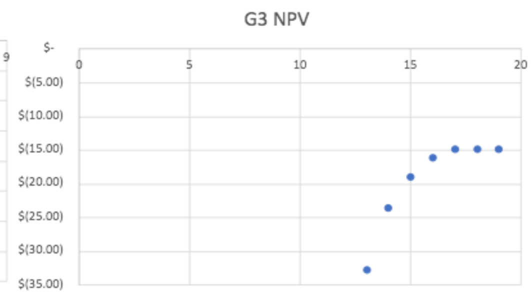

Figure 11: Group 3 NPV

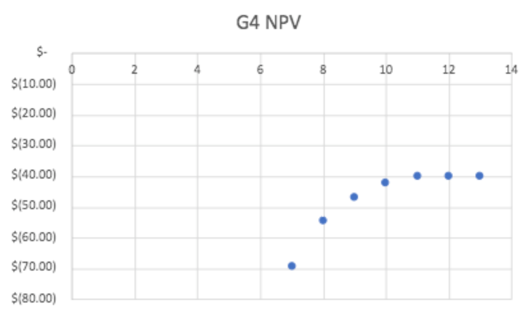

Figure 12: Group 4 NPV

Group 2 was the last group of wells to be put into production from the field as well as optimized. From the graphs above, the implementation of Group 2 into the cash flow sheet moved closer to the present the maximum NPV value of the project due to its additional production capacity. This was completed by trial and error of moving the starting of Groups 3 and 4 as well as the beginning of Group 2.

Once the groups, the best cases, and the number of each unit to be fabricated had been determined, a map of the field was created, and the distances between all wells were calculated. The distances between wells were used to minimize the trucking costs. This allowed each well to be moved the shortest distance, while still optimizing production and following best case scenarios. After determining the redeployment schedule, a sensitivity analysis was run to determine the units' duration on each well, in order to maximize production. It was found that the units would stay on the Group 1 wells from years 1-4, Group 2 wells from years 5-8, Group 3 wells from years 9-12, and Group 4 wells from years 13-20. The deployment and redeployment schedule of each unit is shown below in Figure 13.

Project Planner

LEGEND: Permiting | Construction | Group 1 | Group 2 | Group 3 | Group 4 | Turnaround

ACTIVITY	PLANNED PRODUCTION START	PLAN DURATION	0		1	2	3	4	5	6
A1	YR 1	19 Years	Permitting	Construction		WELL 1B		WELL 1B	WELL 2F	
B1	YR 2	20 Years	Permitting	Construction		WELL 1B		WELL 1B	WELL 2H	
B2	YR 3	20 Years	Permitting	Construction		WELL 1E		WELL 1E	WELL 2F	
B3	YR 4	21 Years	Permitting	Construction		WELL 1E		WELL 1E	WELL 1D	
C1	YR 5	22 Years	Permitting	Construction		WELL 2C		WELL 2C	WELL 2D	
C2	YR 6	23 Years	Permitting	Construction		WELL 2C		WELL 2C	WELL 2H	
C3	YR 7	24 Years	Permitting	Construction		WELL 1F		WELL 1F	WELL 1D	
C4	YR 8	25 Years	Permitting	Construction		WELL 1F		WELL 1F	WELL 2A	
C5	YR 9	26 Years	Permitting	Construction		WELL 1F		WELL 1F	WELL 2A	
C6	SPARE		Permitting	Construction		SPARING		SPARING	SPARING	

Figure 13: Project Schedule

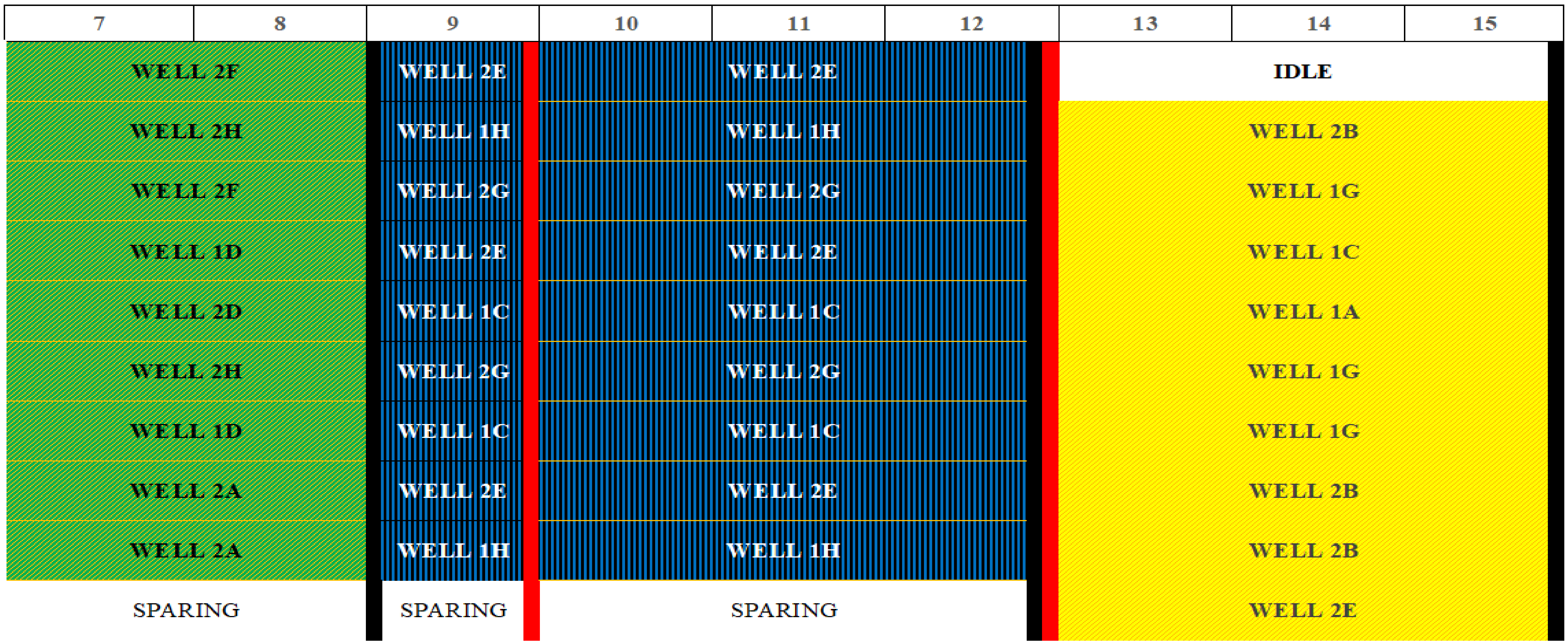

Figure 13: Cont'd

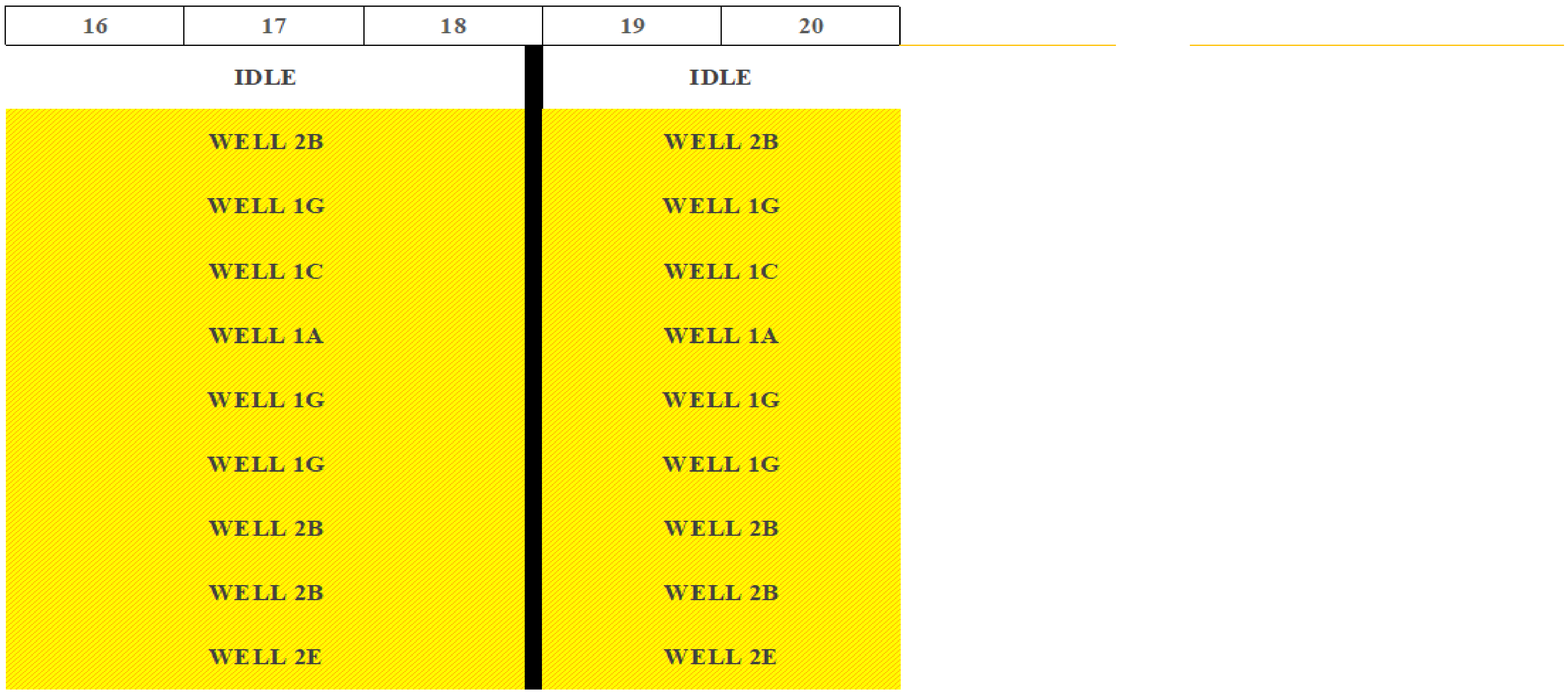

Figure 13: Cont'd

The groupings approach of scheduling wells was conducted to maximize capacity at the central plant, optimizing the number of units needed to be produced, allowing the retirement of unit A1 to occur at the end of year 12. Retirement of all other units will occur at the end of the project's life.

After distinguishing the deployment and redeployment schedule, spare units were considered. Due to the turnaround being every three years, it was determined that only one spare unit would be needed on hand. It was then decided that the unit would be a 5 MMSCFD unit, due to the high capacity and flexible placement, allowing for it to replace any size unit at any well and allowing for that well to maintain maximum capacity. Assuming the redeployment of the units to Group 4 wells, the spare C6 unit could be used on well 2E to continue production, after its completion of the third round of wells, maximizing NPV for the project.

Expected Plot Layout

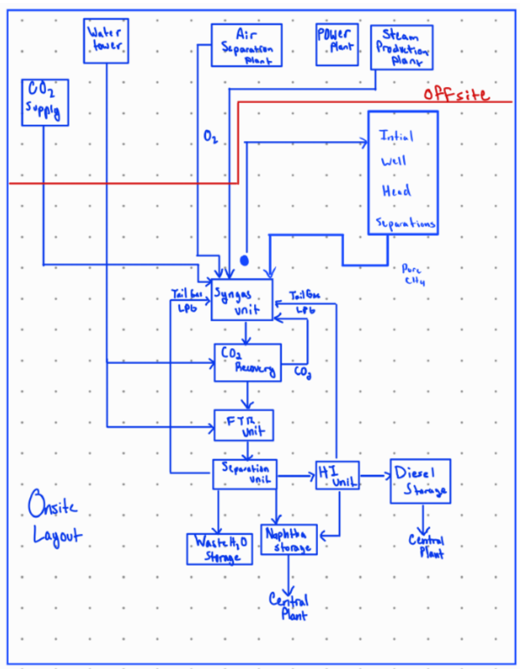

Figure 14: Expected Plot Layout Drawing

Summary of Project NPV and Sensitivities

NPV

This project is based on a 20-year project evaluation life. The project was assumed to be under permitting applications and construction during 2022, with production beginning at the first of 2023. Depreciation was accounted for using straight-line depreciation over a 7-year depreciation period. Per the project statement, taxes were accounted for at 20% taxable income. The discount rate was assumed to be 8% and the service factor was given as 80%. It was also assumed that escalation could be accounted for under the washout assumption.

Cash Flow				0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Actual Year			2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042
End of Year			0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
Yearly Production (MMSCF)																							
	1	1A														1312.2	1312.2	853.0	554.4	360.4	234.2	152.3	99.0
	1	1B		872.8	872.8	567.3	368.8																
	1	1C										2442.6	2442.6	1587.7	1032.0	670.8	436.0	283.4	184.2	119.7	77.8	50.6	32.9
	1	1D						2174.5	2174.5	1413.4	918.7												
	1	1E		960.7	960.7	624.4	405.9																
	1	1F		4303.2	4303.2	2797.1	1818.1																
	1	1G														3374.1	3374.1	2193.1	1425.5	926.6	602.3	391.5	254.5
	1	1H										2067.7	2067.7	1344.0	873.6								
	1	2A						2682.6	2682.6	1743.7	1133.4												
	1	2B														3579.3	3579.3	2326.6	1512.3	983.0	638.9	415.3	269.9
	1	2C		2552.7	2552.7	1659.2	1078.5																
	1	2D						1121.3	1121.3	728.8	473.7												
	1	2E										2318.2	2318.2	1506.8	979.4	636.6	413.8	269.0	174.8	113.6	73.9	48.0	31.2
	1	2F						839.2	839.2	545.5	354.6												
	1	2G										1774.2	1774.2	1153.2	749.6								
	1	2H						1894.5	1894.5	1231.4	800.4												
		Total	0.000	8689.3	8689.3	5648.1	3671.2	8712.1	8712.1	5662.9	3680.9	8602.6	8602.6	5591.7	3634.6	9573.1	9115.5	5925.1	3851.3	2503.3	1627.2	1057.7	687.5
Revenue			0.000	57.006	57.006	37.054	24.085	57.156	57.156	37.151	24.148	56.437	56.437	36.684	23.845	62.804	59.802	38.871	25.266	16.423	10.675	6.939	4.510
+Salvage Cost															0.746								17.341
-Utilities			0.000	-17.477	-17.477	-11.360	-7.384	-17.523	-17.523	-11.390	-7.403	-17.302	-17.302	-11.247	-7.310	-19.254	-18.334	-11.917	-7.746	-5.035	-3.273	-2.127	-1.383
-Trucking																							
	62	1A		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	-0.174	-0.174	-0.113	-0.073	-0.048	-0.031	-0.020	-0.013
	21	1B		-0.039	-0.039	-0.025	-0.017	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	17	1C		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	-0.089	-0.089	-0.058	-0.037	-0.024	-0.016	-0.010	-0.007	-0.004	-0.003	-0.002	-0.001
	73.4	1D		0.000	0.000	0.000	0.000	-0.341	-0.341	-0.222	-0.144	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	6.2	1E		-0.013	-0.013	-0.008	-0.005	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	91	1F		-0.837	-0.837	-0.544	-0.354	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	59	1G		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	-0.425	-0.425	-0.277	-0.180	-0.117	-0.076	-0.049	-0.032
	89.3	1H		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	-0.395	-0.395	-0.256	-0.167	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	85.8	2A		0.000	0.000	0.000	0.000	-0.492	-0.492	-0.320	-0.208	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	84.7	2B		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	-0.648	-0.648	-0.421	-0.274	-0.178	-0.116	-0.075	-0.049
	2.4	2C		-0.013	-0.013	-0.009	-0.006	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	42	2D		0.000	0.000	0.000	0.000	-0.101	-0.101	-0.065	-0.043	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	98.8	2E		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	-0.049	-0.049	-0.032	-0.021	-0.013	-0.009	-0.006	-0.004	-0.002	-0.002	-0.001	-0.001
	13	2F		0.000	0.000	0.000	0.000	-0.023	-0.023	-0.015	-0.010	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	87	2G		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	-0.330	-0.330	-0.214	-0.139	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
	36	2H		0.000	0.000	0.000	0.000	-0.146	-0.146	-0.095	-0.062	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
		Total	0.000	-0.902	-0.902	-0.586	-0.381	-1.103	-1.103	-0.717	-0.466	-0.862	-0.862	-0.560	-0.364	-1.285	-1.272	-0.827	-0.537	-0.349	-0.227	-0.148	-0.096
-Catalyst Cost			-1.790			-1.790			-1.790			-1.790			-1.790			-1.790			-1.790		
-Operating Costs			-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920	-4.920

Figure 15: Cash Flow

-Operating Labor		-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316	-0.316
-Unit Redployment					End	End		End		End	End			End			End			End		
3	1A													-0.948								
	1B					-0.306								End								
	1C													-0.7988								
	1D									-2.205												
	1E					-0.6098																
	1F					-4.668																
	1G																					
	1H													-2.2965								
	2A									-2.6715												
	2B																					
	2C					-2.574								-2.8305								
	2D									-0.582												
	2E																					
	2F									-0.7224												
	2G																					
	2H					-8.1578				-0.4329												
	Total	0	0	0	0	-16.316	0	0	0	-6.6138	0	0	0	-6.8738	0	0	0	0	0	0	0	0
-Salvage Removal Cost																0						-0.225
-Depreciation			-25.8389	-25.8389	-25.8389	-25.839	-25.83892	-25.8389	-25.8389			0	0	0	0	0	0	0				
-Writeoff		0																				
Taxable Income		-7.0258	7.553027	7.553027	-7.7568	-31.07	7.455805	5.66562	-6.02981	4.4296	31.2473	33.0375	19.6419	3.017	37.0292	35.7068	19.101791	11.74732	5.80329	0.1495	-0.5717	14.912
-Tax @ 20%		1.4052	-1.51061	-1.51061	1.55136	6.214	-1.491161	-1.13312	1.20596	-0.8859	-6.2495	-6.6075	-3.92838	-0.6034	-7.40584	-7.1414	-3.8203582	-2.34946	-1.1607	-0.03	0.1143	-2.982
Net Income		-5.6206	6.042422	6.042422	-6.20544	-24.856	5.964644	4.5325	-4.82385	3.54368	24.9979	26.43	15.7135	2.4136	29.6233	28.5654	15.2814328	9.397855	4.64263	0.1196	-0.4573	11.93
+Depreciation		0	25.83892	25.83892	25.8389	25.839	25.83892	25.8389	25.8389	0	0	0	0	0	0	0	0	0	0	0	0	0
+Writeoff		0																				
-Working Capital																						
-Fixed Capital (Equipment)																						
	500 MSCFD	-7.461																				
	2.5 MMSCFD	-44.16																				
	5 MMSCFD	-112.4																				
	Total	-164																				
-Spared Fix Capital	5 MMSCFD	-16.89																				
Cash Flow		-186.5	31.8813	31.8813	19.6335	0.983	31.80356	30.3714	21.015	3.5437	24.998	26.43	15.7135	2.4136	29.623	28.565	15.281433	9.39785	4.6426	0.12	-0.457	11.93
Discounted Cash Flow (8%)		-186.5	29.5198	27.3331	15.5857	0.7225	21.64497	19.1391	12.262	1.9145	12.505	12.242	6.73927	0.9585	10.892	9.7254	4.8173449	2.74314	1.2548	0.03	-0.106	2.559
Constant Dollar Rate		1	0.97087	0.9426	0.91514	0.8885	0.862609	0.83748	0.8131	0.7894	0.7664	0.7441	0.72242	0.7014	0.681	0.6611	0.6418619	0.62317	0.605	0.587	0.5703	0.554
Yr		0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
ROR		8.53%																				
NPV		$5.99	MM																			

Figure 15: Cont'd

Sensitivity Analysis

An economic sensitivity analysis was run on the project to determine the best and worst case scenarios. The best-case scenario included reducing the capital cost to 80% of the expected value, operating cost and utilities cost to 90% of the expected value, increasing revenue by 30%, and trucking cost to $1.19/mile drive (95% of expected cost). Trucking cost was reduced to account for the slight possibility of trucking advancements that could result in a lower cost. The worst case included increasing capital cost to 130% of the expected cost, operating cost and utilities cost to 105%, decreasing revenue by 30%, and the trucking cost to $6.50/mile driven. The trucking cost was increased, accounting for realistic values, given by a ChampionX Consulting Group vendor quote. The operating cost and utilities cost varied with room for fluctuation, but under the assumption that they would be reduced over time. The capital cost varied, based on historical data[30]. The revenue varied, based on the WTI price data collected over the last 10 years[31]. A summary of the sensitivity analysis and quantitative are shown below in Table 17.

Table 17: Sensitivity Analysis

	Best Case	Expected Case	Worst Case
ROR	23.37%	8.53%	-11.14%
NPV	$156.97 MM	$5.99 MM	($187.07) MM

After conducting a sensitivity analysis, the standard deviation was calculated using the propagation of error method with values from the best and worst case scenarios. This allowed the conclusion that the economic aspect of the project is heavily dependent upon the price of oil. Using the 95% confidence limits, the ROR of the expected case was found to range between 1.97% and 19.03%.

Engineering Calculations, Computer Simulation Outputs

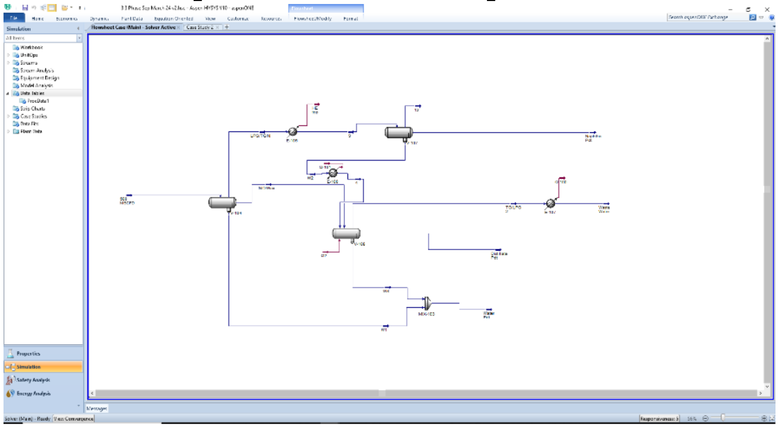

Figure 16: Aspen HYSYS Separation (Other sizes included in Appendix)

HEX Coolers for 500																								
Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*
401,170	441,286	343.9	80	100	120	84.4	0.9	150	38.7	0.24036	40	3.3444	0.2745	-0.0472	4584.332	1	1	0.96	1.21	2.17	22004.79	37408.15	62471.61

Table 18: HEX Sizing Calculation (Full Sizing included in Appendix)

Liquid Phase			
V-101 to Condensate			
Nominal D	Inner Dia(in)	Q (bbl/d)	V (ft/s)
0.125	0.269	29	4.774958
0.25	0.364	29	2.607783
0.375	0.493	29	1.421609
0.5	0.622	29	0.893086
0.75	0.824	29	0.508885
1	1.049	29	0.313995
1.25	1.38	29	0.181433
1.5	1.61	29	0.133298
Liquid Cal	D*5.61458/(24*60*60)/(PI()/4*(Q/12)^2) = V		

Gas Phase				
V-101 to E-101				
Nominal D	Inner Dia(in)	Q (MMCFD)	P(psia)	V (ft/s)
0.125	0.269	0.2957	246	519.1127
0.25	0.364	0.2957	246	283.5068
0.375	0.493	0.2957	246	154.5512
0.5	0.622	0.2957	246	97.09245
0.75	0.824	0.2957	246	55.32378
1	1.049	0.2957	246	34.13621
1.25	1.38	0.2957	246	19.72459
1.5	1.61	0.2957	246	14.49154
	Gas Cal	(0.75*(Q*1000000/24)/(D^2*P)) =V		

3 Phase Calculation				
R-101 to V-101				
	Vapor	Liq	Aq	
Frac	0.558	0.0237	0.4183	0.442
N				0
V(m/s)	0.1			0.075416
Q(m^3/s)	0.000153556	6.39444E-05	5.19E-05	0.000116
D (m)	0.044216853			0.044217
				1.740827
Based on flow regime in GPSA				1.75in

Two Phase Gas/Liq Flow Pipe Dia						
Stream Na	QL (m3/s)	QG (m3/s)	Inner Dia (	VsG	VsL	Inner Dia (In)
E-103 to V	2.85404E-05	0.001040318	0.00635	32.8495	0.901204	0.25
E-103 to V	2.85404E-05	0.001040318	0.0127	8.212375	0.225301	0.5
E-103 to V	2.85404E-05	0.001040318	0.01905	3.649944	0.100134	0.75
E-103 to V	2.85404E-05	0.001040318	0.0254	2.053094	0.056325	1
E-103 to V	2.85404E-05	0.001040318	0.0381	0.912486	0.025033	1.5
E-103 to V	2.85404E-05	0.001040318	0.0508	0.513273	0.014081	2
E-103 to V	2.85404E-05	0.001040318	0.0635	0.328495	0.009012	2.5
VsG=QG/(	QG/(PI()/4*D^2)		VsL=QL/(PI()/4*D^2)			

Figure 17: Pipe Sizing Calculation Charts (Full Piping Summary in Appendix)

Reactor Operating Temperature	466.811886	K	
T3	0.755590114		
T4	1.007031008		
alpha	0.936538837		
n	Wn	Mn	Mn*n
5	0.015491348	0.048822	0.244108
6	0.017409899	0.045723	0.274339
7	0.019022555	0.042822	0.299751
8	0.020360413	0.040104	0.320833
9	0.021451857	0.037559	0.338031
10	0.022322775	0.035175	0.351755
11	0.02299676	0.032943	0.362375
12	0.023495301	0.030853	0.370231
13	0.02383795	0.028895	0.375631
14	0.024042486	0.027061	0.378854
15	0.02412506	0.025344	0.380155
16	0.024100326	0.023735	0.379765
17	0.023981572	0.022229	0.377894
18	0.02378083	0.020818	0.37473
19	0.023508986	0.019497	0.370447
20	0.023175872	0.01826	0.365198
21	0.02279036	0.017101	0.359123
22	0.022360441	0.016016	0.352348
23	0.021893304	0.014999	0.344987
24	0.021395405	0.014048	0.337142
25	0.020872528	0.013156	0.328902
26	0.020329851	0.012321	0.320351
27	0.019771991	0.011539	0.31156
28	0.019203061	0.010807	0.302595
29	0.018626713	0.010121	0.293514
30	0.018046179	0.009479	0.284366
31	0.017464313	0.008877	0.275197
32	0.01688362	0.008314	0.266047
33	0.016306296	0.007786	0.256949
34	0.015734252	0.007292	0.247935
35	0.015169142	0.006829	0.23903
36	0.01461239	0.006396	0.230257
37	0.014065212	0.00599	0.221635
38	0.013528634	0.00561	0.21318
39	0.013003515	0.005254	0.204905
40	0.01249056	0.004921	0.196822
41	0.011990342	0.004608	0.18894
42	0.01150331	0.004316	0.181265
43	0.011029803	0.004042	0.173804
44	0.010570068	0.003785	0.16656
45	0.010124263	0.003545	0.159535
46	0.009692471	0.00332	0.152731
47	0.00927471	0.00311	0.146148
48	0.008870937	0.002912	0.139785
49	0.00848106	0.002727	0.133642

Figure 18: Polymath Code and ASF Distribution (Complete results and code in Appendix)

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Appendix

Figure 19: ASF Distribution at Optimum Energy Efficiency

Reactor Operating Temperature	466.81189	K
T3	0.75559011	
T4	1.00703101	
alpha	0.93653884	

n	Wn	Mn	Mn*n
5	0.01549135	0.048822	0.244108
6	0.0174099	0.045723	0.274339
7	0.01902255	0.042822	0.299751
8	0.02036041	0.040104	0.320833
9	0.02145186	0.037559	0.338031
10	0.02232277	0.035175	0.351755
11	0.02299676	0.032943	0.362375
12	0.0234953	0.030853	0.370231
13	0.02383795	0.028895	0.375631
14	0.02404249	0.027061	0.378854
15	0.02412506	0.025344	0.380155
16	0.02410033	0.023735	0.379765
17	0.02398157	0.022229	0.377894
18	0.02378083	0.020818	0.37473
19	0.02350899	0.019497	0.370447
20	0.02317587	0.01826	0.365198
21	0.02279036	0.017101	0.359123
22	0.02236044	0.016016	0.352348
23	0.0218933	0.014999	0.344987
24	0.0213954	0.014048	0.337142
25	0.02087253	0.013156	0.328902
26	0.02032985	0.012321	0.320351
27	0.01977199	0.011539	0.31156
28	0.01920306	0.010807	0.302595
29	0.01862671	0.010121	0.293514
30	0.01804618	0.009479	0.284366
31	0.01746431	0.008877	0.275197
32	0.01688362	0.008314	0.266047
33	0.0163063	0.007786	0.256949
34	0.01573425	0.007292	0.247935
35	0.01516914	0.006829	0.23903
36	0.01461239	0.006396	0.230257
37	0.01406521	0.00599	0.221635
38	0.01352863	0.00561	0.21318
39	0.01300351	0.005254	0.204905
40	0.01249056	0.004921	0.196822
41	0.01199034	0.004608	0.18894
42	0.01150331	0.004316	0.181265
43	0.0110298	0.004042	0.173804
44	0.01057007	0.003785	0.16656
45	0.01012426	0.003545	0.159535
46	0.00969247	0.00332	0.152731
47	0.00927471	0.00311	0.146148
48	0.00887094	0.002912	0.139785
49	0.00848106	0.002727	0.133642
50	0.00810494	0.002554	0.127715
51	0.0077424	0.002392	0.122002
52	0.00739324	0.00224	0.1165
53	0.00705721	0.002098	0.111205
54	0.00673406	0.001965	0.106113
55	0.0064235	0.00184	0.101219
56	0.00612523	0.001724	0.096519
57	0.00583896	0.001614	0.092008
58	0.00556435	0.001512	0.087681
59	0.00530108	0.001416	0.083533
60	0.00504881	0.001326	0.079557
61	0.00480721	0.001242	0.07575
62	0.00457595	0.001163	0.072106
63	0.00435467	0.001089	0.06862
64	0.00414306	0.00102	0.065285
65	0.00394076	0.000955	0.062097
66	0.00374746	0.000895	0.059051
67	0.00356281	0.000838	0.056142
68	0.00338652	0.000785	0.053364
69	0.00321824	0.000735	0.050712
70	0.00305769	0.000688	0.048182
71	0.00290456	0.000645	0.045769
72	0.00275854	0.000604	0.043468
73	0.00261936	0.000565	0.041275
74	0.00248674	0.00053	0.039185
75	0.0023604	0.000496	0.037194
76	0.00224008	0.000464	0.035298
77	0.00212553	0.000435	0.033493
78	0.00201649	0.000407	0.031775
79	0.00191274	0.000382	0.03014
80	0.00181403	0.000357	0.028585
81	0.00172014	0.000335	0.027105
82	0.00163087	0.000313	0.025699
83	0.001546	0.000294	0.024361
84	0.00146533	0.000275	0.02309
85	0.00138868	0.000257	0.021882
86	0.00131585	0.000241	0.020735
87	0.00124668	0.000226	0.019645
88	0.00118098	0.000211	0.018609
89	0.0011186	0.000198	0.017627
90	0.00105939	0.000185	0.016693
91	0.00100318	0.000174	0.015808
92	0.00094984	0.000163	0.014967
93	0.00089923	0.000152	0.01417
94	0.00085122	0.000143	0.013413
95	0.00080568	0.000134	0.012696
96	0.0007625	0.000125	0.012015
97	0.00072155	0.000117	0.01137
98	0.00068272	0.00011	0.010758
99	0.00064592	0.000103	0.010178
100	0.00061104	9.63E-05	0.009629

FTR Products		Mass Fraction	Mole Fraction
C21-C200		0.601056741	0.268052409
C21-C215		0.601056741	0.268052409
C5-C200		0.95416073	0.767892219
C1-C4		0.04583927	0.232107781
Naptha (C5-C10)		0.116058847	0.250204935
Diesel (C11-C20)		0.237045142	0.249634875
C1		0.043784301	0.20723909
C2		0.000934397	0.008289564
C3		0.00063717	0.008289564
C4		0.000483402	0.008289564
HI unit		Mass Fraction	Mole Fraction
C1		0.006010567	
C2		0.003005284	
C3		0.021036986	
C4		0.021036986	
Naptha (C5-C10)		0.150264185	
Diesel		0.399702733	
FTR C_P Fractions		Mass Fraction	Mole Fraction
C1		0.043784301	0.20723909
C2		0.000934397	0.008289564
C3		0.00063717	0.008289564
C4		0.000483402	0.008289564
	5	0.015491348	0.048821508
	6	0.017409899	0.045723239
	7	0.019022555	0.042821589
	8	0.020360413	0.040104081
	9	0.021451857	0.037559029
	10	0.022322775	0.03517549
	11	0.02299676	0.032943212
	12	0.023495301	0.030852598
	13	0.02383795	0.028894656
	14	0.024042486	0.027060967
	15	0.02412506	0.025343647
	16	0.024100326	0.02373531
	17	0.023981572	0.022229039
	18	0.02378083	0.020818359
	19	0.023508986	0.019497201
	20	0.023175872	0.018259886
C21-C25		0.109312037	0.075320051
C26-C29		0.077931615	0.044788611
C30-C35		0.099603802	0.048578116
C36-C47		0.141885277	0.054897199
C48+		0.172324009	0.044468432

Figure 20: Empirical Fit of Specific Heat Capacity to Temperature

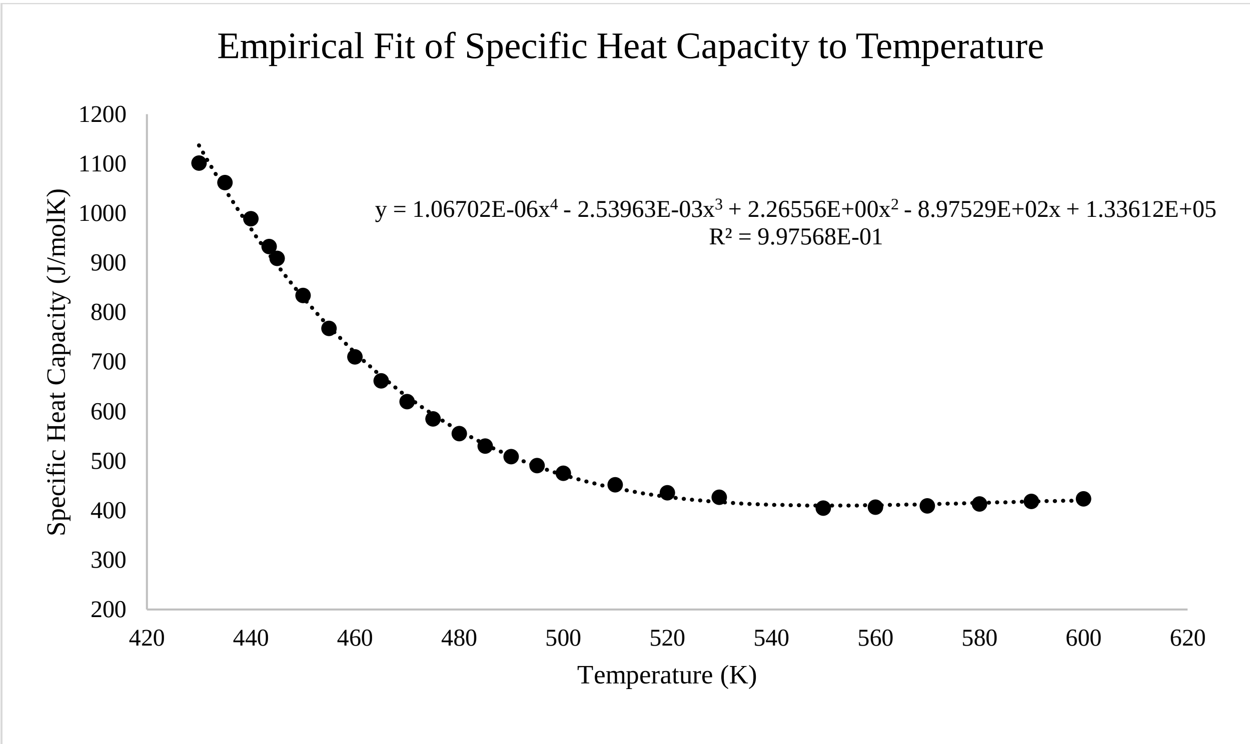


Figure 21: Average Chain Length

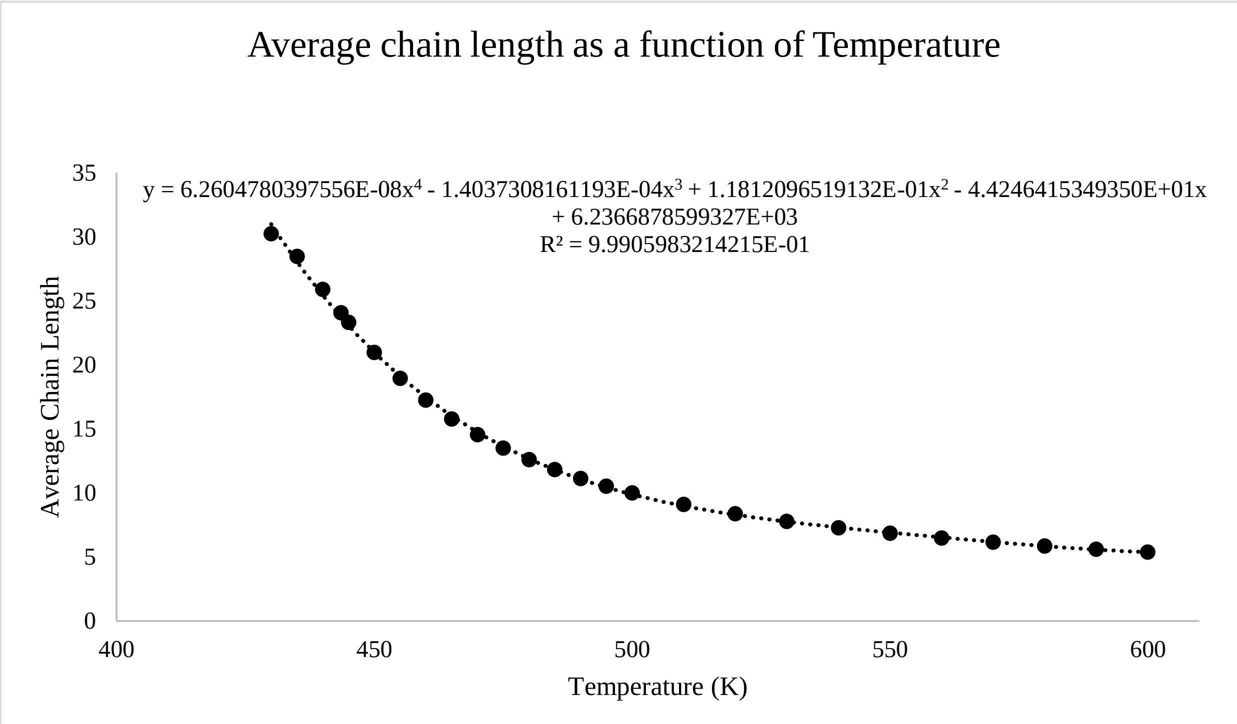


Figure 22: FTR Temperature Profile

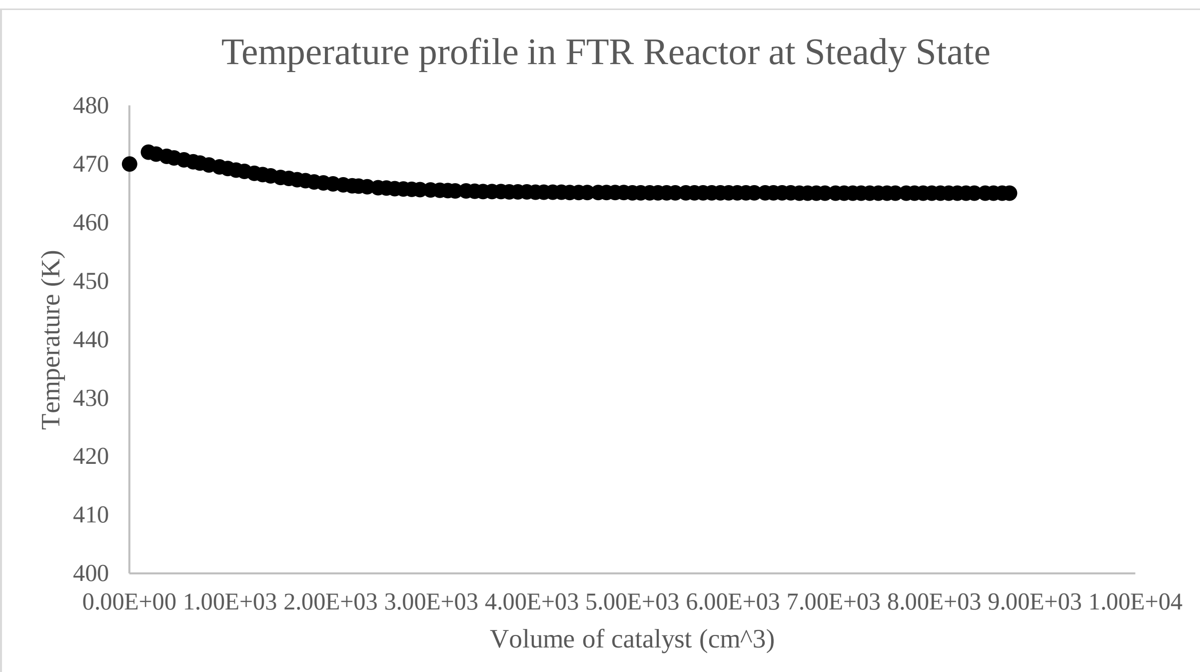


Figure 23: Price per Barrel v. Temperature

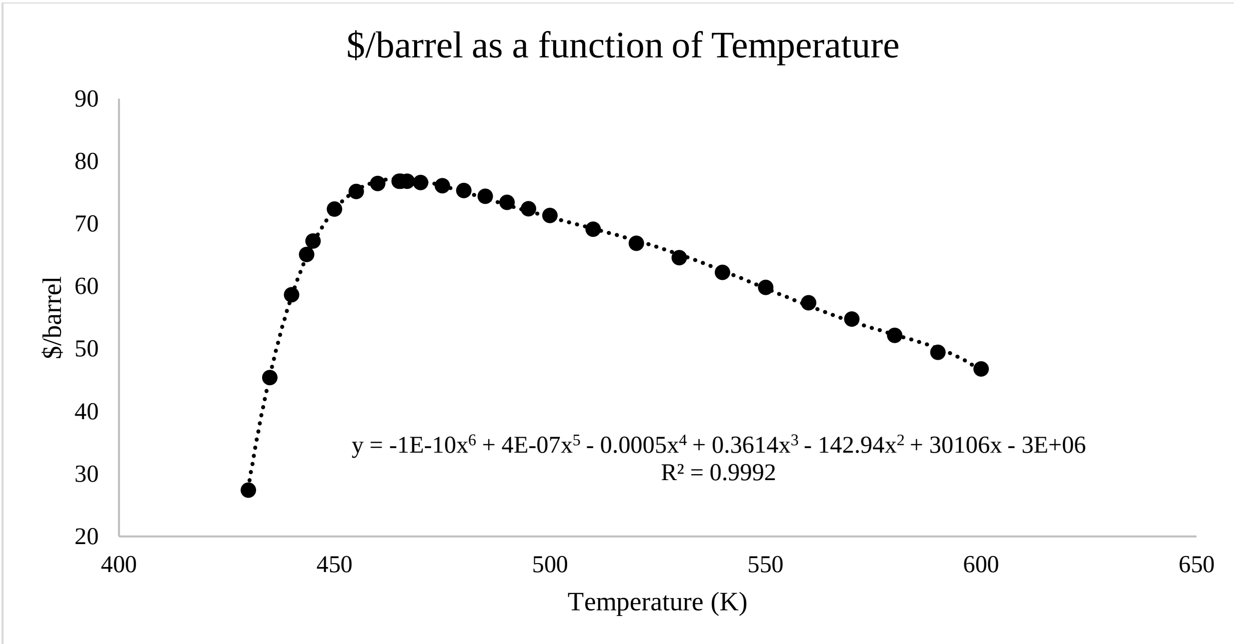


Figure 24: Price per Pound v. Temperature

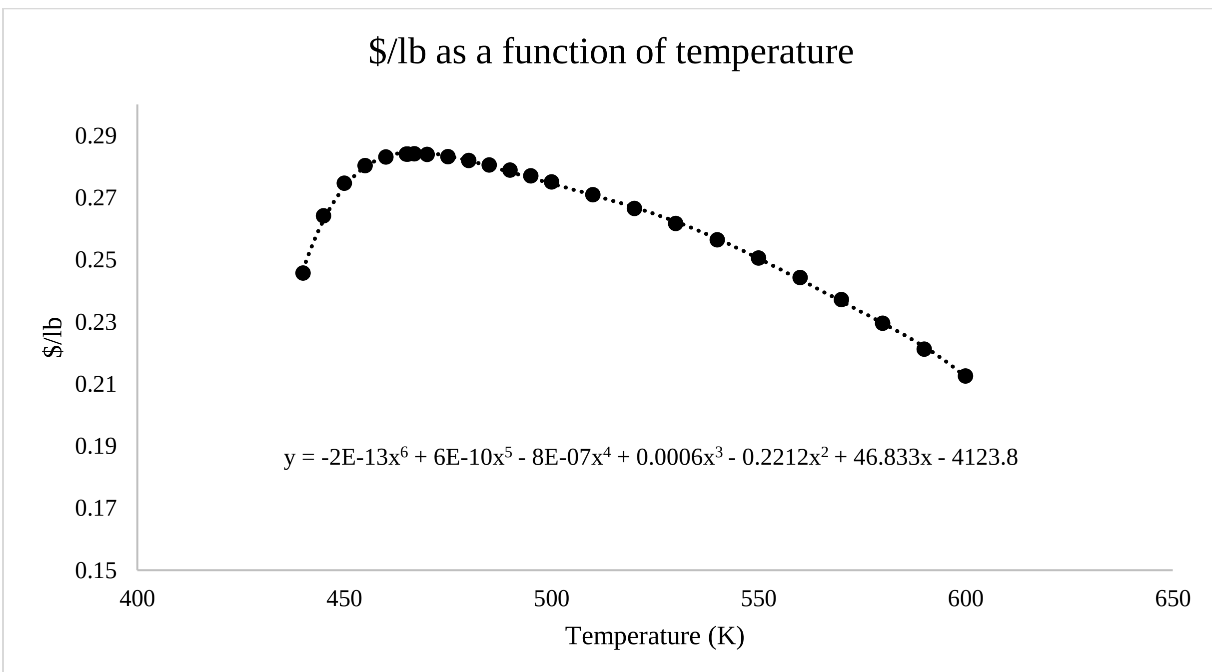


Figure 25: Value of Hydrocarbons

Value of Hydrocarbons per Barrel

		Mass Balance	100 lb basis	$	density	$	Value without HI
	LPG (C3-C4	0.001120572	0.1120572			0	0
	Naptha	0.266323032	26.632303	75	45	0.296823983	7.905106
	Diesel	0.636747875	63.674788	90	53	0.302424436	19.25681
						Total Revenue	27.16192
			1 Barrel Basis				Value with HI
	LPG (C3-C4	0.001120572	0.3029369			0	0
	Naptha	0.266323032	71.998149	75	45	0.296823983	21.37078
	Diesel	0.636747875	172.13933	90	53	0.302424436	52.05914
4.53	Cost/Barrel						-0.14024
							73.28968

Component	lb/barrel pure	contribution	Volume contribution (barrel/lb)	Volume Fraction	
					0.87911
					0.780671
C3	201.601764	0.128454505	7.784857384	0.432556	0.670332
C4	206.77104	0.099953529	10.00464926	0.555896	0.535819
Naptha	252.553	29.31100988	0.034116873	0.001896	
Diesel	297.553	70.53349324	0.014177662	0.000788	
C21-C25	276.0393384	30.17442246	0.033140651	0.001841	
C26-C29	279.140904	21.7539016	0.045968765	0.002554	
C30-C35	281.8978512	28.07809782	0.035614948	0.001979	
C36-C47	284.9994168	40.43722131	0.024729691	0.001374	
C48+	289.1348376	49.82487423	0.020070297	0.001115	
	Total	270.3414286	17.99732553		

Figure 26: Polymath FTR Reactor Report and Code

POLYMATH Report
Ordinary Differential Equations

Calculated values of DEQ variables

	Variable	Initial value	Minimal value	Maximal value	Final value
1	a	1.26E-11	1.26E-11	1.102362	1.102362
2	A_c	7.917298	7.917298	7.917298	7.917298
3	alpha	2.25E-05	2.25E-05	2.25E-05	2.25E-05
4	Area_Costing	1.26E-08	1.26E-08	1102.362	1102.362
5	Area_Cross_Max	1.008E+04	1.008E+04	1.008E+04	1.008E+04
6	beta_0	0.0015927	0.0015927	0.0015927	0.0015927
7	C_CH4	45.66084	45.36797	45.7775	45.36797
8	C_CO	30.24453	30.22142	30.25376	30.22142
9	C_CO2	49.45079	49.40747	49.46809	49.40747
10	C_H2	28.66103	28.64352	28.66802	28.64352
11	C_H2O	34.51579	34.45562	34.53981	34.45562
12	C_HC	630.7407	615.6069	672.1558	672.1558
13	C_N2	29.10142	29.07925	29.11029	29.07925
14	C_T	1816.673	1593.838	1816.673	1593.838
15	Coe	0.001	0.001	0.001	0.001
16	D	3.175	3.175	3.175	3.175
17	D_p	0.0015875	0.0015875	0.0015875	0.0015875
18	delta	-2.	-2.	-2.	-2.
19	epsilon	-0.5964935	-0.5964935	-0.5964935	-0.5964935
20	F_CH4	5.4981	5.4981	5.4981	5.4981
21	F_CO	17.5571	0.4875536	17.5571	0.4875536
22	F_CO2	0.372987	0.372987	0.372987	0.372987
23	F_CO_0	17.5571	17.5571	17.5571	17.5571
24	F_H2	35.2691	0.0649022	35.2691	0.0649022
25	F_H2_0	35.2691	35.2691	35.2691	35.2691
26	F_H2O	0.067721	0.067721	17.13727	17.13727
27	F_H2O_0	0.067721	0.067721	0.067721	0.067721
28	F_HC	0	0	1.065105	1.065105
29	F_N2	0.102693	0.102693	0.102693	0.102693
30	FT	58.8677	24.72861	58.8677	24.72861
31	G	4.206738	4.206738	4.206738	4.206738
32	g_c	1.	1.	1.	1.
33	Hrx	-1.633E+05	-1.633E+05	-1.633E+05	-1.633E+05
34	k	0.0173	0.0173	0.0173	0.0173
35	k2	4.512	4.512	4.512	4.512
36	L	1.263E-08	1.263E-08	1105.175	1105.175
37	mu	1.693E-05	1.693E-05	1.693E-05	1.693E-05

38	n	14.75298	14.28982	16.02616	16.02616
39	p	1.	0.9247946	1.	0.9247946
40	P	29.8	27.55888	29.8	27.55888
41	P_0	29.8	29.8	29.8	29.8
42	P_CO	8.887753	0.5433557	8.887753	0.5433557
43	P_H2	17.85392	0.0723304	17.85392	0.0723304
44	phi	0.4	0.4	0.4	0.4
45	Qg	1768.621	2.418058	1919.583	2.418058
46	Qr	1377.573	2.425646	1926.525	2.425646
47	R	8.314	8.314	8.314	8.314
48	r_CO	-0.0108315	-0.011756	-1.481E-05	-1.481E-05
49	rho_0	12.81	12.81	12.81	12.81
50	rho_d	0.8	0.8	0.8	0.8
51	T	470.	465.0088	471.9925	465.0088
52	T1	0.9411827	0.8494184	0.9799316	0.8494184
53	T2	1.117569	1.037873	1.348869	1.348869
54	T_0	470.	470.	470.	470.
55	Ta	465.	465.	465.	465.
56	Tubes	1000.	1000.	1000.	1000.
57	U	2.187E+06	2.187E+06	2.187E+06	2.187E+06
58	V	1.0E-07	1.0E-07	8750.	8750.
59	Vol_flow	2600.	2600.	2600.	2600.
60	X	0	0	0.9722304	0.9722304
61	y_a0	0.2982467	0.2982467	0.2982467	0.2982467

Differential equations

1 d(X)/d(V) = -1*(r_CO)/F_CO_0

-1*(r_CO)/F_CO_0 #1/cm^3 cat

2 d(p)/d(V) = -1.5*(alpha/(2*p))*(T/T_0)*(1+epsilon*X)

1/cm^3, 1.5 is factor adjustment for liquid pressure drop

3 d(T)/d(V) = (Qg-Qr)/(C_T)

K/cm^3

Explicit equations

1 n = 6.2604780397556*(10^-8)*T^4-1.4037308161193*(10^-4)*T^3+1.1812096519132*(10^-1)*T^2-4.4246415349350*10*T+6.2366878599327*10^3

dimensionless

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

dimensionless

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

dimensionless

4 k = 0.0173

```
   gmol CO / hr cm^3 cat*atm^2
5  k2 = 4.512
   atm ^-1
6  Coe = 1/1000
7  F_CO_0 = Coe*17557.1
   100 #moles/hr, from SU and appropriate simulation
8  D = 2.54*(20/16)
   cm, basis
9  P_0 = 29.8
   atm
10 P = P_0*p
   atm
11 phi = 0.4
   dimensionless
12 rho_d = 0.8
   g/cm^3 bulk density
13 A_c = 3.14159*0.25*D^2
   cm^2
14 D_p = 2.54/(16*100)
   m, catalyst diameter
15 g_c = 1
   dimensionless
16 mu = 1.693*10^-5
   1 #kg/m*s, from SU and appropriate simulation
17 L = V/A_c
   cm
18 rho_0 = 12.81
   kg/m^3, from SU and appropriate simulation
19 Vol_flow = 1.625*(10^4)*0.16
   kg/m^3
20 F_H2O_0 = Coe*67.721
   100 #moles/hr, from SU and appropriate simulation
21 delta = -2
   dimensionless, from kinetics
22 F_N2 = Coe*102.693
   moles/hr, from SU and appropriate simulation
23 F_CO2 = Coe*372.987
   moles/hr, from SU and approprpriate simulation
24 F_HC = X*(1/n)*F_CO_0
   moles/hr
25 F_H2_0 = Coe*35269.1
```

```
200#moles/hr, from SU and appropriate simulation
26 F_CO = F_CO_0*(1-X)
moles/hr
27 F_H2O = F_CO_0*X+F_H2O_0
moles/hr
28 F_H2 = F_H2_0-(2*n+1)*(1/n)*F_CO_0*X
moles/hr
29 F_CH4 = Coe*5498.1
moles/hr, from SU and appropropriate simulation
30 y_a0 = F_CO_0/(F_CO_0+F_H2_0+F_H2O_0+F_N2+F_CO2+F_CH4)
dimensionless, from SU and appropropriate simulation
31 FT = F_HC+F_CO+F_H2+F_H2O+F_CO2+F_CH4+F_N2
moles
32 R = 8.314
J/mol*K
33 G = Coe*rho_0*(Vol_flow/A_c)
1 #kg/s*m^2, from SU and appropriate simulation
34 P_H2 = P*F_H2/FT
atm
35 epsilon = delta*y_a0
36 beta_0 = (1/100)*(1/101325)*(G*(1-phi)/(rho_0*g_c*D_p*(phi^3)))*((150*(1-phi)*mu/D_p)+1.75*G)
atm/cm
37 C_HC = 1.06702*10^(-6)*T^4-2.53963*10^(-3)*T^3+2.26556*T^2-8.97529*(10^2)*T+1.33612*10^5
J/mol*K
38 C_CH4 = R*(1.702 + (9.081*10^-3) * ((T)) - (2.164*10^-6) * ((T)^2))
39 C_H2 = R*(3.249 + (0.422*10^-3) * ((T)) - (0.083*10^-9) * ((T)^2))
40 C_CO2 = R*(5.457 + (1.045 * 10^-3) * ((T)) - (1.157*10^-9) * ((T)^2))
41 C_H2O = R*(3.47 + (1.45*10^-3) * ((T)) + (0.121*10^-9) * ((T)^2))
42 C_CO = R*(3.376 + (0.557*10^-3) * ((T)) - (0.031*10^-9) * ((T)^2))
43 C_N2 = R*(3.208 + (0.593*10^-3)*T+0.030*(10^5)*T^-2)
44 Hrx = -1*(70200)*2.326
J/gmol CO
45 P_CO = P*F_CO/FT
atm
46 r_CO = -1*(k*T1*P_H2*P_CO)/((1+k2+T2*P_CO)^2)
gmol CO/(hr*cm^3cat)
47 Qg = r_CO*Hrx
J/hr*cm^3 cat
48 T_0 = 470
K, basis
49 Ta = 465
```

K, basis

50 U = (2.044175*10*1000)*0.385*((G*3600/10)^0.8)/D^0.2

J/hr*(m^2)*K

51 a = (1/(100^2))*V/(0.25*D)

m^2,

52 Qr = U*a*(T-Ta)/V

J/hr*cm^3 cat

53 alpha = 2*beta_0/(A_c*P_0*(1-phi))

1/cm^3, adjusted alpha parameter

54 C_T = F_HC*C_HC+F_H2O*C_H2O+F_CO2*C_CO2+F_CO*C_CO+F_N2*C_N2+F_H2*C_H2+F_CH4*C_CH4

J/hr*K

55 Tubes = 1/Coe

56 Area_Cross_Max = Tubes*D*D

57 Area_Costing = Tubes*3.14159*D*L*(1/100)^2

m^2; area of tubes for costing

General

Total number of equations	60
Number of differential equations	3
Number of explicit equations	57
Elapsed time	1.157 sec
Solution method	RKF_45
Step size guess. h	0.000001
Truncation error tolerance. eps	0.000001

Data file: c:\users\lewel\downloads\pbr_final!.pol

FTR PolyMath Code

#Kinetics

r_CO=-1*(k*T1*P_H2*P_CO)/((1+k2+T2*P_CO)^2) #gmol CO/(hr*cm^3cat)

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

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

k=0.0173 #gmol CO / hr cm^3 cat*atm^2

k2=4.512 #atm ^-1

#Partial Pressures

P_CO=P*F_CO/FT #atm

P_H2=P*F_H2/FT #atm

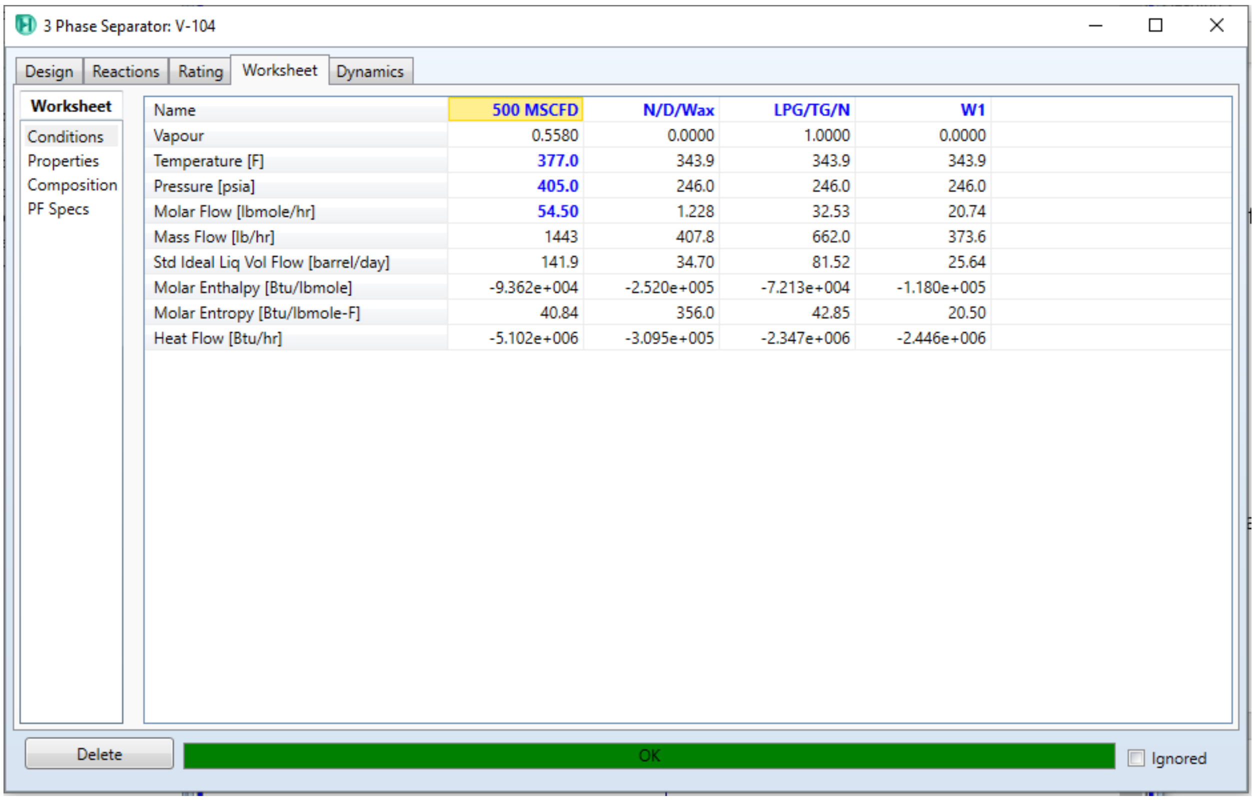
```
#Mass Balance
d(X) / d(V) = -1*(r_CO)/F_CO_0#-1*(r_CO)/F_CO_0 #1/cm^3 cat
X(0) = 0 #dimensionless

L = V/A_c #cm

#Bounds of integration
V(0) = 0.0000001 #cm^3 cat
V(f) = 8750# cm^3 cat

#Pressure Drop
d(p) / d(V) = -1.5*(alpha/(2*p))*(T/T_0)*(1+epsilon*X) #1/cm^3, 1.5 is factor adjustment for liquid pressure drop
p(0) = 1 #dimensionless
P=P_0*p #atm

        #catalyst parameters
alpha= 2*beta_0/(A_c*P_0*(1-phi)) #1/cm^3, adjusted alpha parameter
phi=0.4 #dimensionless
rho_d=0.8 #g/cm^3 bulk density
beta_0=(1/100)*(1/101325)*(G*(1-phi)/(rho_0*g_c*D_p*(phi^3)))*((150*(1-phi)*mu/D_p)+1.75*G)#atm/cm, GARRISON PLEASE CHECK

D_p=2.54/(16*100) #m, catalyst diameter GARRISON PLEASE CHECK
g_c=1 #dimensionless
mu= 1.693*10^-5 #1 #kg/m*s, will get from SU and appropriate simulation
G= Coe*rho_0*(Vol_flow/A_c)#1 #kg/s*m^2, will get from SU and appropriate simulation
rho_0 = 12.81 #kg/m^3 will get from SU and appropriate simulation
```

```
Vol_flow =1.625*(10^4)*0.16 #kg/m^3

y_a0=F_CO_0/(F_CO_0+F_H2_0+F_H2O_0+F_N2+F_CO2+F_CH4) #dimensionless, will get
from SU and appropropriate simulation

delta=-2 #dimensionless, from kinetics

epsilon=delta*y_a0


#Mole Flows

FT=F_HC+F_CO+F_H2+F_H2O+F_CO2+F_CH4+F_N2 #moles


F_HC=X*(1/n)*F_CO_0 #moles/hr

F_CO=F_CO_0*(1-X) #moles/hr

F_H2O=F_CO_0*X+F_H2O_0 #moles/hr

F_H2=F_H2_0-(2*n+1)*(1/n)*F_CO_0*X #moles/hr

F_CH4=Coe*5498.1 #moles/hr, from SU and appropropriate simulation

F_CO2=Coe*372.987 #moles/hr, from SU and appropropriate simulation

F_N2=Coe*102.693 #moles/hr, from SU and appropriate simulation


#Average Chain Length

n=6.2604780397556*(10^-8)*T^4-1.4037308161193*(10^-4)*T^3+1.1812096519132*(10^-
1)*T^2-4.4246415349350*10*T+6.2366878599327*10^3 #dimensionless


#Energy Balance

d(T) / d(V) = (Qg-Qr)/(C_T) #K/cm^3

T(0) = 470 #473.15 #K, Will get from SU

#T=467.4957

R=8.314 #J/mol*K

Qr=U*a*(T-Ta)/V # J/hr*cm^3 cat

Qg=r_CO*Hrx # J/hr*cm^3 cat


U=(2.044175*10*1000)*0.385*((G*3600/10)^0.8)/D^0.2 #J/hr*(m^2)*K
```

```
#Heat Capacities ()

C_T=F_HC*C_HC+F_H2O*C_H2O+F_CO2*C_CO2+F_CO*C_CO+F_N2*C_N2+F_H2*C_H2+F_CH4*C_CH4 #J/hr*K

C_HC=1.06702*10^(-6)*T^4-2.53963*10^(-3)*T^3+2.26556*T^2-8.97529*(10^2)*T+1.33612*10^5#J/mol*K

C_CH4 =  R*(1.702 + (9.081*10^-3) * ((T)) - (2.164*10^-6) * ((T)^2))

C_H2 =  R*(3.249 + (0.422*10^-3) * ((T)) - (0.083*10^-9) * ((T)^2))

C_CO2 =  R*(5.457 + (1.045 * 10^-3) * ((T)) - (1.157*10^-9) * ((T)^2))

C_H2O =  R*(3.47 + (1.45*10^-3) * ((T)) + (0.121*10^-9) * ((T)^2))

C_CO =  R*(3.376 + (0.557*10^-3) * ((T)) - (0.031*10^-9) * ((T)^2))

C_N2 = R*(3.208 + (0.593*10^-3)*T+0.030*(10^5)*T^-2)


Hrx=-1*(70200)*2.326 #J/gmol CO


#Reactor Parameters

F_CO_0=Coe*17557.1 #100 #moles/hr, from SU and appropriate simulation

F_H2_0= Coe*35269.1 #200#moles/hr, from SU and appropriate simulation

F_H2O_0= Coe*67.721#100 #moles/hr, from SU and appropriate simulation

T_0=470 #473.15 #K, basis

Ta=465#451.5 #K, basis

D=2.54*(20/16) #cm, basis

a=(1/(100^2))*V/(0.25*D) #m^2,

A_c=3.14159*0.25*D^2 #cm^2

P_0= 29.8 #atm

Area_Cross_Max=Tubes*D*D

Area_Costing=Tubes*3.14159*D*L*(1/100)^2 #m^2; area of tubes for costing

Tubes=1/Coe

Coe = 1/1000
```

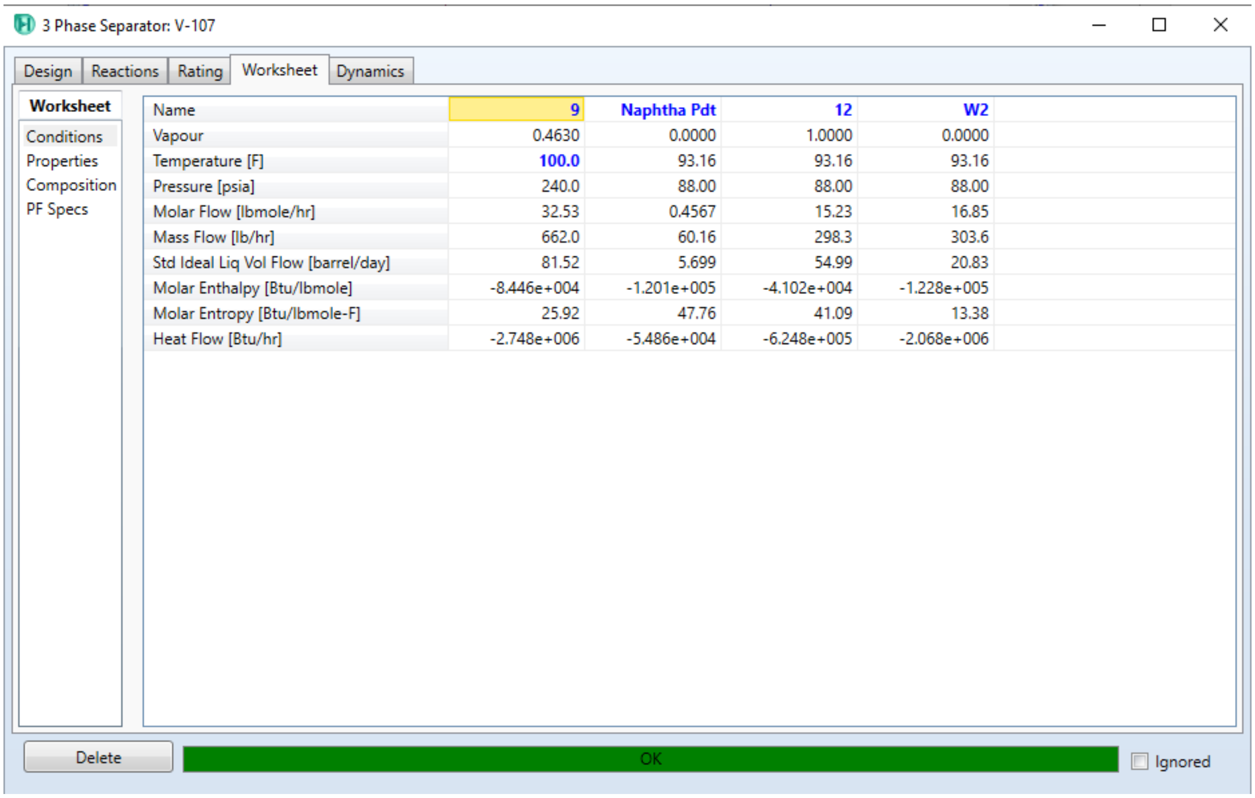
Figure 27: Aspen HYSYS Separators

500 MSCFD V-101



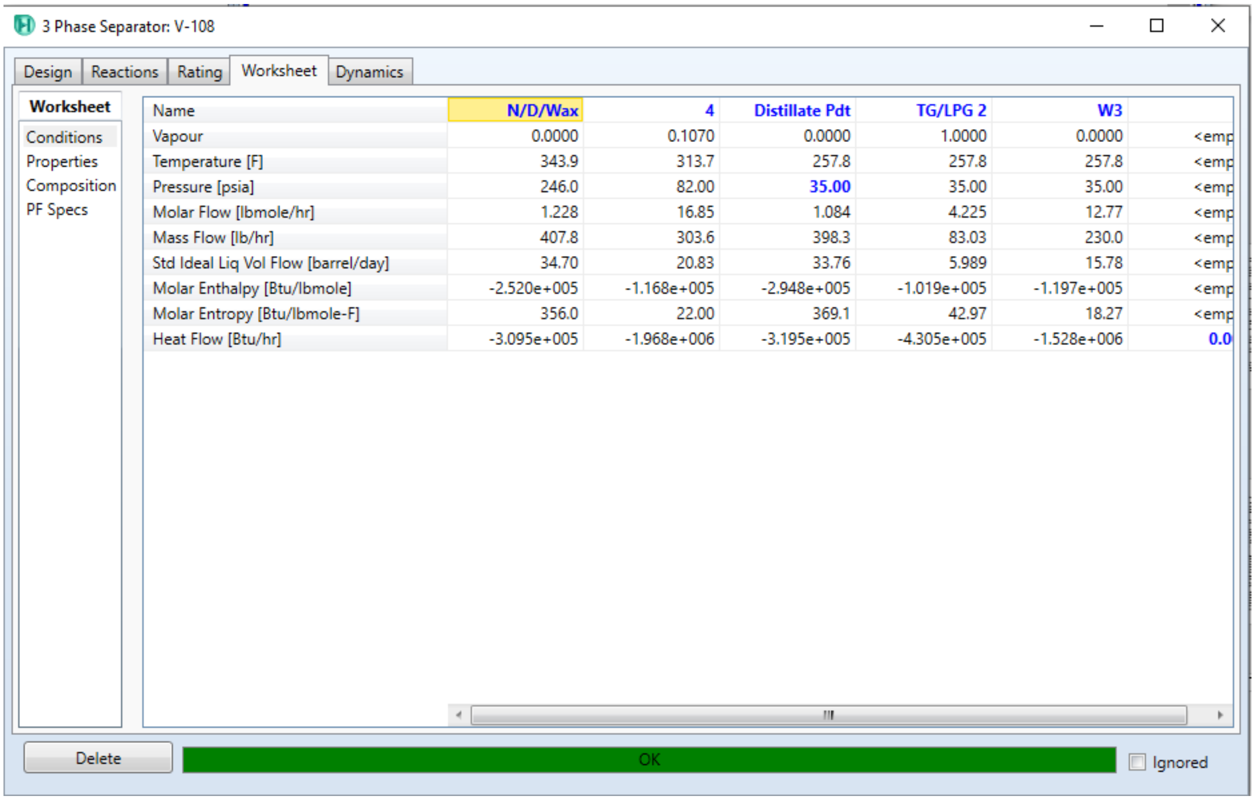
Name	500 MSCFD	N/D/Wax	LPG/TG/N	W1
Vapour	0.5580	0.0000	1.0000	0.0000
Temperature [F]	377.0	343.9	343.9	343.9
Pressure [psia]	405.0	246.0	246.0	246.0
Molar Flow [lbmole/hr]	54.50	1.228	32.53	20.74
Mass Flow [lb/hr]	1443	407.8	662.0	373.6
Std Ideal Liq Vol Flow [barrel/day]	141.9	34.70	81.52	25.64
Molar Enthalpy [Btu/lbmole]	-9.362e+004	-2.520e+005	-7.213e+004	-1.180e+005
Molar Entropy [Btu/lbmole-F]	40.84	356.0	42.85	20.50
Heat Flow [Btu/hr]	-5.102e+006	-3.095e+005	-2.347e+006	-2.446e+006

500 MSCFD V-102



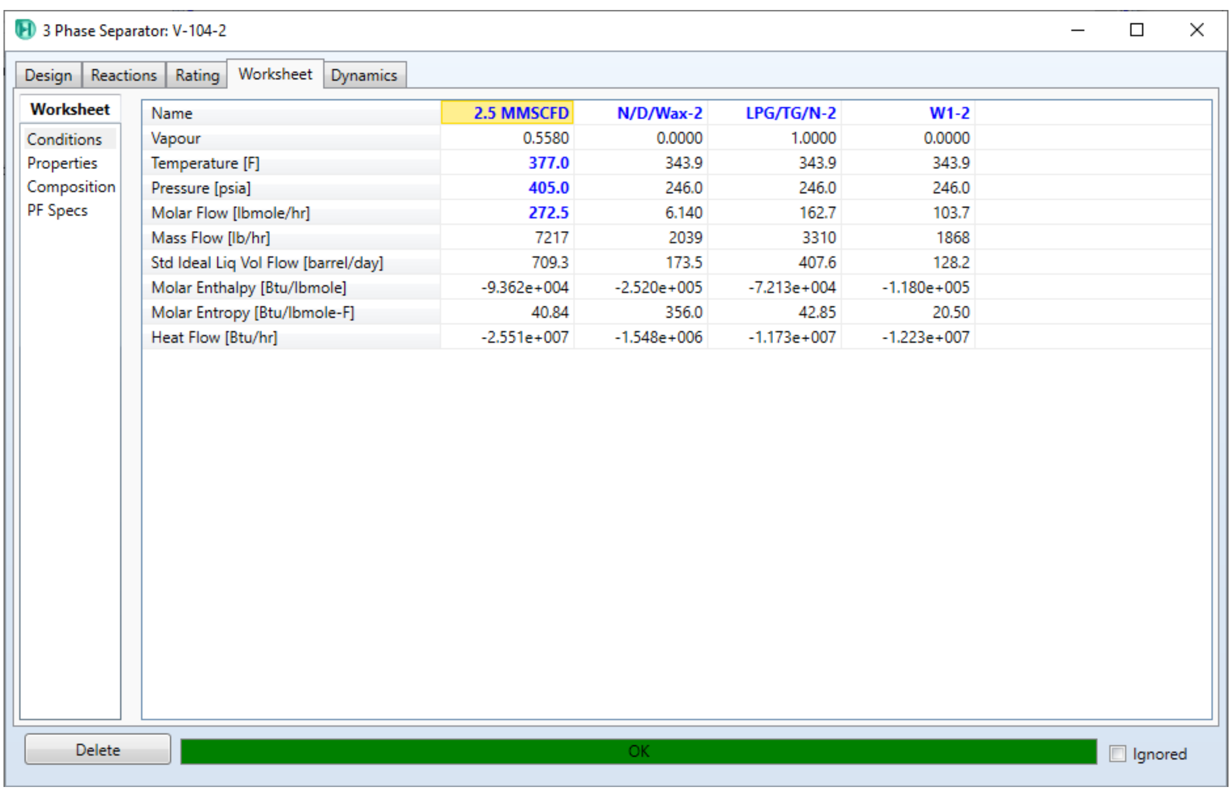
Name	9	Naphtha Pdt	12	W2
Vapour	0.4630	0.0000	1.0000	0.0000
Temperature [F]	100.0	93.16	93.16	93.16
Pressure [psia]	240.0	88.00	88.00	88.00
Molar Flow [lbmole/hr]	32.53	0.4567	15.23	16.85
Mass Flow [lb/hr]	662.0	60.16	298.3	303.6
Std Ideal Liq Vol Flow [barrel/day]	81.52	5.699	54.99	20.83
Molar Enthalpy [Btu/lbmole]	-8.446e+004	-1.201e+005	-4.102e+004	-1.228e+005
Molar Entropy [Btu/lbmole-F]	25.92	47.76	41.09	13.38
Heat Flow [Btu/hr]	-2.748e+006	-5.486e+004	-6.248e+005	-2.068e+006

500 MSCFD V-103



Name	N/D/Wax	4	Distillate Pdt	TG/LPG 2	W3	
Vapour	0.0000	0.1070	0.0000	1.0000	0.0000	<emp
Temperature [F]	343.9	313.7	257.8	257.8	257.8	<emp
Pressure [psia]	246.0	82.00	35.00	35.00	35.00	<emp
Molar Flow [lbmole/hr]	1.228	16.85	1.084	4.225	12.77	<emp
Mass Flow [lb/hr]	407.8	303.6	398.3	83.03	230.0	<emp
Std Ideal Liq Vol Flow [barrel/day]	34.70	20.83	33.76	5.989	15.78	<emp
Molar Enthalpy [Btu/lbmole]	-2.520e+005	-1.168e+005	-2.948e+005	-1.019e+005	-1.197e+005	<emp
Molar Entropy [Btu/lbmole-F]	356.0	22.00	369.1	42.97	18.27	<emp
Heat Flow [Btu/hr]	-3.095e+005	-1.968e+006	-3.195e+005	-4.305e+005	-1.528e+006	0.0

2500 MSCFD V-201



Name	2.5 MMSCFD	N/D/Wax-2	LPG/TG/N-2	W1-2
Vapour	0.5580	0.0000	1.0000	0.0000
Temperature [F]	377.0	343.9	343.9	343.9
Pressure [psia]	405.0	246.0	246.0	246.0
Molar Flow [lbmole/hr]	272.5	6.140	162.7	103.7
Mass Flow [lb/hr]	7217	2039	3310	1868
Std Ideal Liq Vol Flow [barrel/day]	709.3	173.5	407.6	128.2
Molar Enthalpy [Btu/lbmole]	-9.362e+004	-2.520e+005	-7.213e+004	-1.180e+005
Molar Entropy [Btu/lbmole-F]	40.84	356.0	42.85	20.50
Heat Flow [Btu/hr]	-2.551e+007	-1.548e+006	-1.173e+007	-1.223e+007

2500 MSCFD V-202

3 Phase Separator: V-107-2

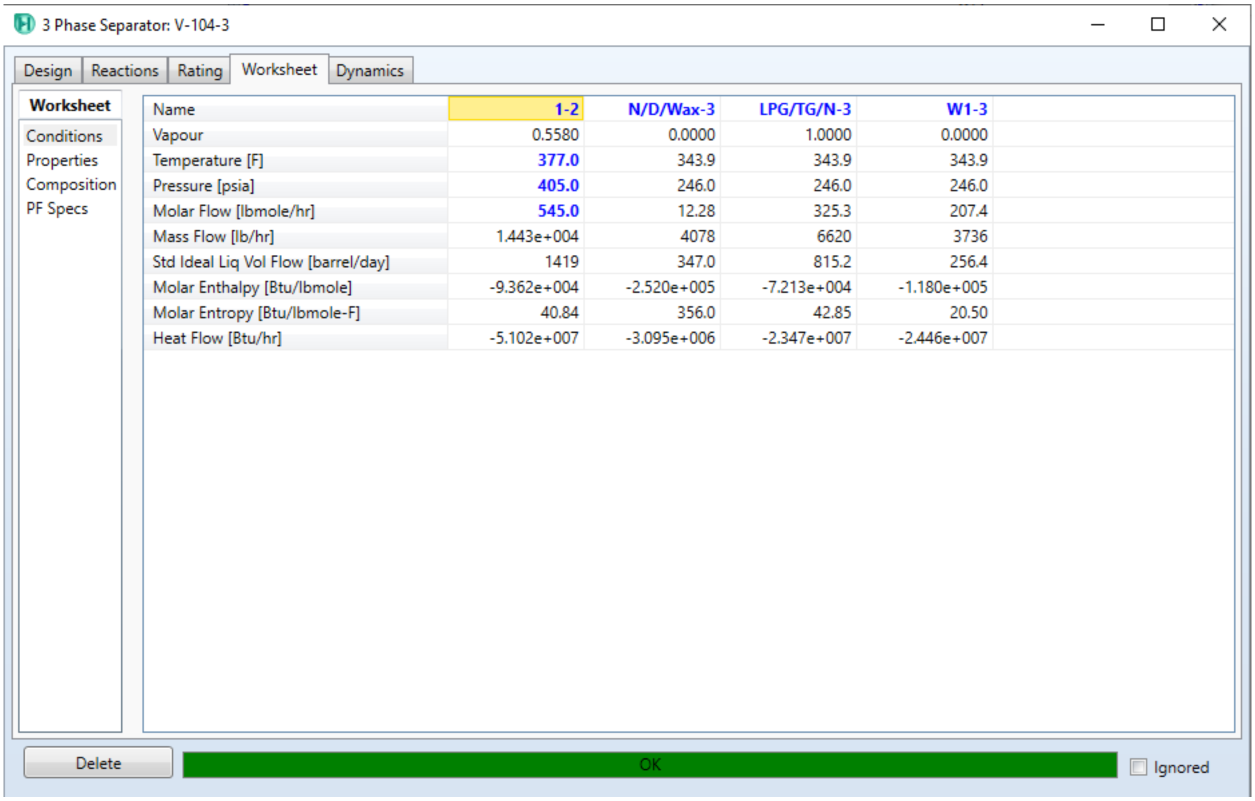
Name	9-2	Naphtha Pdt-2	12-2	W2-2
Vapour	0.4630	0.0000	1.0000	0.0000
Temperature [F]	100.0	93.16	93.16	93.16
Pressure [psia]	240.0	88.00	88.00	88.00
Molar Flow [lbmole/hr]	162.7	2.284	76.15	84.24
Mass Flow [lb/hr]	3310	300.8	1491	1518
Std Ideal Liq Vol Flow [barrel/day]	407.6	28.49	274.9	104.1
Molar Enthalpy [Btu/lbmole]	-8.446e+004	-1.201e+005	-4.102e+004	-1.228e+005
Molar Entropy [Btu/lbmole-F]	25.92	47.76	41.09	13.38
Heat Flow [Btu/hr]	-1.374e+007	-2.743e+005	-3.124e+006	-1.034e+007

2500 MSCFD V-203

3 Phase Separator: V-108-2

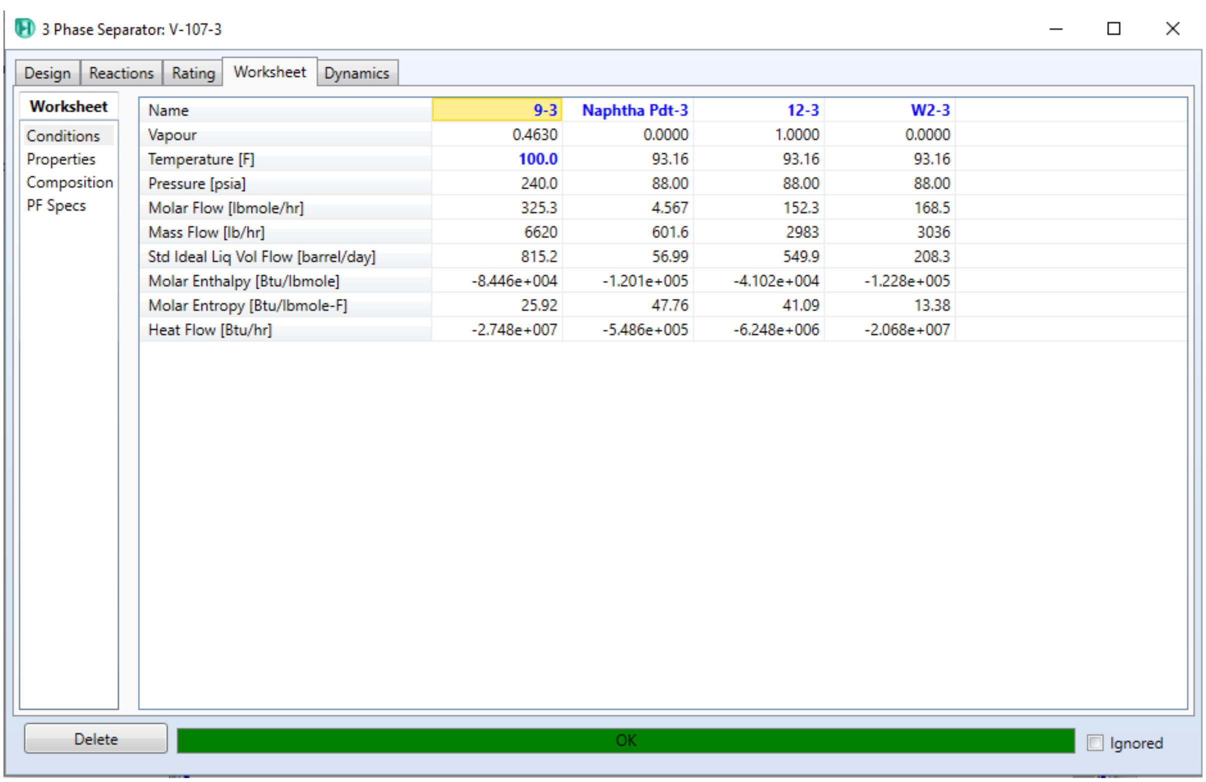
Name	N/D/Wax-2	W2-2	Distillate Pdt-2	TG/LPG 2-2	W3-2	Q
Vapour	0.0000	0.0000	0.0000	1.0000	0.0000	<emp
Temperature [F]	343.9	93.16	257.8	257.8	257.8	<emp
Pressure [psia]	246.0	88.00	35.00	35.00	35.00	<emp
Molar Flow [lbmole/hr]	6.140	84.24	5.419	21.13	63.83	<emp
Mass Flow [lb/hr]	2039	1518	1992	415.2	1150	<emp
Std Ideal Liq Vol Flow [barrel/day]	173.5	104.1	168.8	29.94	78.90	<emp
Molar Enthalpy [Btu/lbmole]	-2.520e+005	-1.228e+005	-2.948e+005	-1.019e+005	-1.197e+005	<emp
Molar Entropy [Btu/lbmole-F]	356.0	13.38	369.1	42.97	18.27	<emp
Heat Flow [Btu/hr]	-1.548e+006	-1.034e+007	-1.597e+006	-2.153e+006	-7.638e+006	5.000e+

5000 MSCF V-301



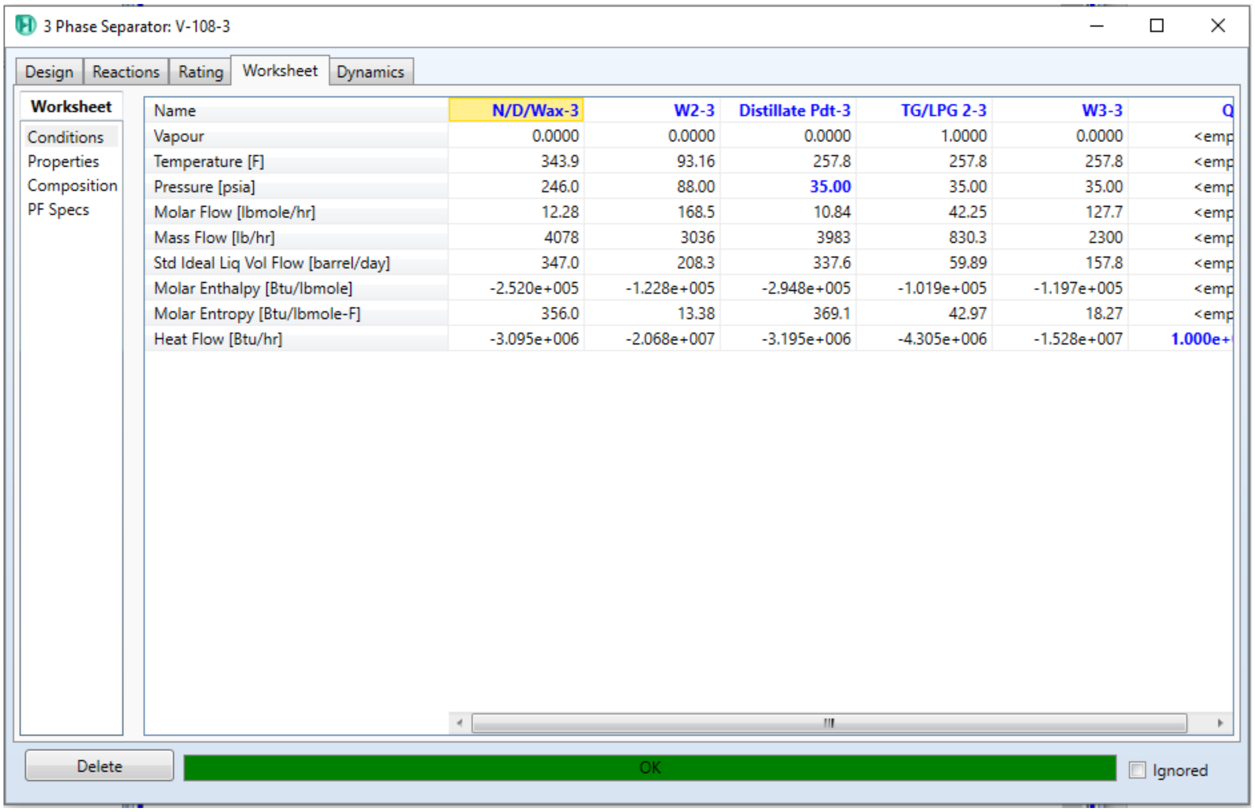
Name	1-2	N/D/Wax-3	LPG/TG/N-3	W1-3
Vapour	0.5580	0.0000	1.0000	0.0000
Temperature [F]	377.0	343.9	343.9	343.9
Pressure [psia]	405.0	246.0	246.0	246.0
Molar Flow [lbmole/hr]	545.0	12.28	325.3	207.4
Mass Flow [lb/hr]	1.443e+004	4078	6620	3736
Std Ideal Liq Vol Flow [barrel/day]	1419	347.0	815.2	256.4
Molar Enthalpy [Btu/lbmole]	-9.362e+004	-2.520e+005	-7.213e+004	-1.180e+005
Molar Entropy [Btu/lbmole-F]	40.84	356.0	42.85	20.50
Heat Flow [Btu/hr]	-5.102e+007	-3.095e+006	-2.347e+007	-2.446e+007

5000 MSCF V-302



Name	9-3	Naphtha Pdt-3	12-3	W2-3
Vapour	0.4630	0.0000	1.0000	0.0000
Temperature [F]	100.0	93.16	93.16	93.16
Pressure [psia]	240.0	88.00	88.00	88.00
Molar Flow [lbmole/hr]	325.3	4.567	152.3	168.5
Mass Flow [lb/hr]	6620	601.6	2983	3036
Std Ideal Liq Vol Flow [barrel/day]	815.2	56.99	549.9	208.3
Molar Enthalpy [Btu/lbmole]	-8.446e+004	-1.201e+005	-4.102e+004	-1.228e+005
Molar Entropy [Btu/lbmole-F]	25.92	47.76	41.09	13.38
Heat Flow [Btu/hr]	-2.748e+007	-5.486e+005	-6.248e+006	-2.068e+007

5000 MSCF V-303


3 Phase Separator: V-108-3

Design | Reactions | Rating | Worksheet | Dynamics

Worksheet: Conditions, Properties, Composition, PF Specs

Name	N/D/Wax-3	W2-3	Distillate Pdt-3	TG/LPG 2-3	W3-3	Q
Vapour	0.0000	0.0000	0.0000	1.0000	0.0000	<emp
Temperature [F]	343.9	93.16	257.8	257.8	257.8	<emp
Pressure [psia]	246.0	88.00	35.00	35.00	35.00	<emp
Molar Flow [lbmole/hr]	12.28	168.5	10.84	42.25	127.7	<emp
Mass Flow [lb/hr]	4078	3036	3983	830.3	2300	<emp
Std Ideal Liq Vol Flow [barrel/day]	347.0	208.3	337.6	59.89	157.8	<emp
Molar Enthalpy [Btu/lbmole]	-2.520e+005	-1.228e+005	-2.948e+005	-1.019e+005	-1.197e+005	<emp
Molar Entropy [Btu/lbmole-F]	356.0	13.38	369.1	42.97	18.27	<emp
Heat Flow [Btu/hr]	-3.095e+006	-2.068e+007	-3.195e+006	-4.305e+006	-1.528e+007	1.000e+

Delete | OK | Ignored

Figure 28: 3-Phase Separator Sizing

Example 4: Size a horizontal high pressure separator for the following conditions:

Gas Flow Rate:	10.0 MMSCFD
Operating Pressure:	800 psig
Condensate Load:	500 bbl/day
Water Load:	100 bbl/day

From Figure 4A, at 800 psig operating pressure, a 20" x 10' horizontal separator will handle 10.2 MMSCFD operating ½ full of liquid. Where three phase operation is required in a horizontal separator, the liquid section should be ½ full, otherwise the level control action becomes too critical.

From Table 4B, the liquid capacity will be:

$$W = \frac{1440\ (v)}{t} = \frac{1440\ (1.80)}{5.0} = 518 \text{ bbl/day}$$

Therefore, the 20" x 10' separator will not handle the combined liquid load of 500 + 100 = 600 bbl/day. Five minute retention time is used as a conservative figure without any additional information.

From Table 4B, a separator with more settling volume is a 24" x 10'. Its liquid capacity is:

$$W = \frac{1440\ (2.63)}{5.0} = 757 \text{ bbl/day}$$

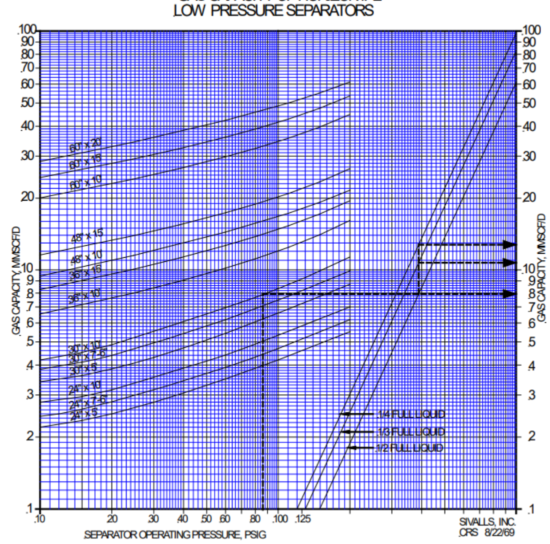
The gas capacity of a 24" x 10' separator at 800 psig is 15.0 MMSCFD.

SETTLING VOLUMES OF STANDARD HORIZONTAL HIGH PRESSURE SEPARATORS, 230 PSIG THROUGH 2000 PSIG W.P. **

Size	Settling Volume, bbl*		
Dia. x Length	½ Full	1/3 Full	¼ Full
12 ¾" x 5'	0.38	0.22	0.15
12 ¾" x 7 ½'	0.55	0.32	0.21
12 ¾" x 10'	0.72	0.42	0.28
16" x 5'	0.61	0.35	0.24
16" x 7 ½'	0.88	0.50	0.34
16" x 10'	1.14	0.66	0.44
20" x 5'	0.98	0.55	0.38
20" x 7 ½'	1.39	0.79	0.54
20" x 10'	1.80	1.03	0.70
24" x 5'	1.45	0.83	0.55
24" x 7 ½'	2.04	1.18	0.78
24" x 10'	2.63	1.52	1.01
24" x 15'	3.81	2.21	1.47
30" x 5'	2.43	1.39	0.91
30" x 7 ½'	3.40	1.96	1.29
30" x 10'	4.37	2.52	1.67
30" x 15'	6.30	3.65	2.42
36" x 7 ½'	4.99	2.87	1.90
36" x 10'	6.38	3.68	2.45
36" x 15'	9.17	5.30	3.54
36" x 20'	11.96	6.92	4.63
42" x 7 ½'	6.93	3.98	2.61
42" x 10'	8.83	5.09	3.35
42" x 15'	12.62	7.30	4.83
42" x 20'	16.41	9.51	6.32
48" x 7 ½'	9.28	5.32	3.51
48" x 10'	11.77	6.77	4.49
48" x 15'	16.74	9.67	6.43
48" x 20'	21.71	12.57	8.38
54" x 7 ½'	12.02	6.87	4.49
54" x 10'	15.17	8.71	5.73
54" x 15'	12.49	12.40	8.20
54" x 20'	27.81	16.08	10.68
60" x 7 ½'	15.05	8.60	5.66
60" x 10'	18.93	10.86	7.17
60" x 15'	26.68	15.38	10.21
60" x 20'	34.44	19.90	13.24

*Based on 1000 psig W.P. Separator.
**Standard working pressures available are 230, 500, 600, 1000, 1200, 1440, 1500, and 2000 psig.

GAS CAPACITY OF HORIZONTAL LOW PRESSURE SEPARATORS



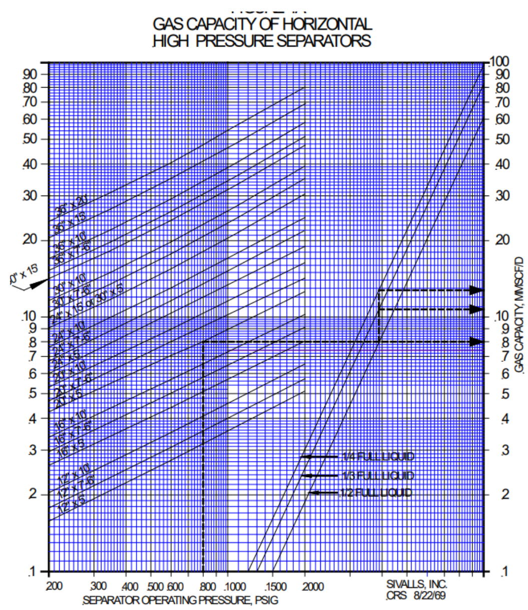


Figure 29: Heat Exchanger Sizing Tables

Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*I	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*	Scaled TM
84,518	92,970	257.8	75	115	115	80.8	0.9	90	14.2	0.24036	40	4.0336	0.2341	0.0497	42082.73	1	1	0.96	1.21	2.17	201997.1	343395.1	573469.76	$188,571.00
HEX Coolers for 500																								
Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*I	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*	
401,170	441,286	343.9	80	100	120	84.4	0.9	150	38.7	0.24036	40	3.3444	0.2745	-0.0472	4584.332	1	1	0.96	1.21	2.17	22004.79	37408.15	62471.6057	
HEX Heater for 500																								
Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*I	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*	
100,000	110,000	353	94.1	353	315.3	114.8	0.9	60	17.7	0.9999	221.2	3.3444	0.2745	-0.0472	4107.923	1	1	0.96	1.21	2.17	19718.03	33520.65	55979.4873	

Air Coolers for 2500																								
Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*I	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*	Scaled TM
422,594	464,854	257.8	75	115	115	80.8	0.9	90	71	0.24036	40	4.0336	0.2341	0.0497	42082.73	1	1	0.96	1.21	2.17	201997.1	343395.1	573469.76	688163.7123
Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*I	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*	
2,005,851	2,206,436	343.9	80	100	120	84.4	0.9	150	193.6	0.24036	40	4.8306	-0.8509	0.3187	35602.03	1	1	0.96	1.21	2.17	170889.8	290512.6	485156.037	
Air Coolers for 5000																								
Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*I	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*	
845,189	929,708	257.8	75	115	115	80.8	0.9	90	142.1	0.24036	40	4.0336	0.2341	0.0497	58586.13	1	1	0.96	1.21	2.17	281213.4	478062.9	798364.975	
HEX Coolers for 5000																								
Duty (Btu/hr)	Duty (Btu/hr) OSIZE	Thi (F)	Tci (F)	Tho (F)	Tco (F)	ΔTlm (F)	F	U (Btu/hr*	A (ft^2)	p (Btu/lb*I	ΔT (F)	K1	K2	K3	Cpo	FP	FM	B1	B2	Fbm	CBM	TM	TM(2022)*	
4,011,704	4,412,874	343.9	80	100	120	84.4	0.9	150	387.2	0.24036	40	4.8306	-0.8509	0.3187	57945.44	1	1	0.96	1.21	2.17	278138.1	472834.8	789634.038	

Figure 30: Purchased Cost Scale Curve For HEX

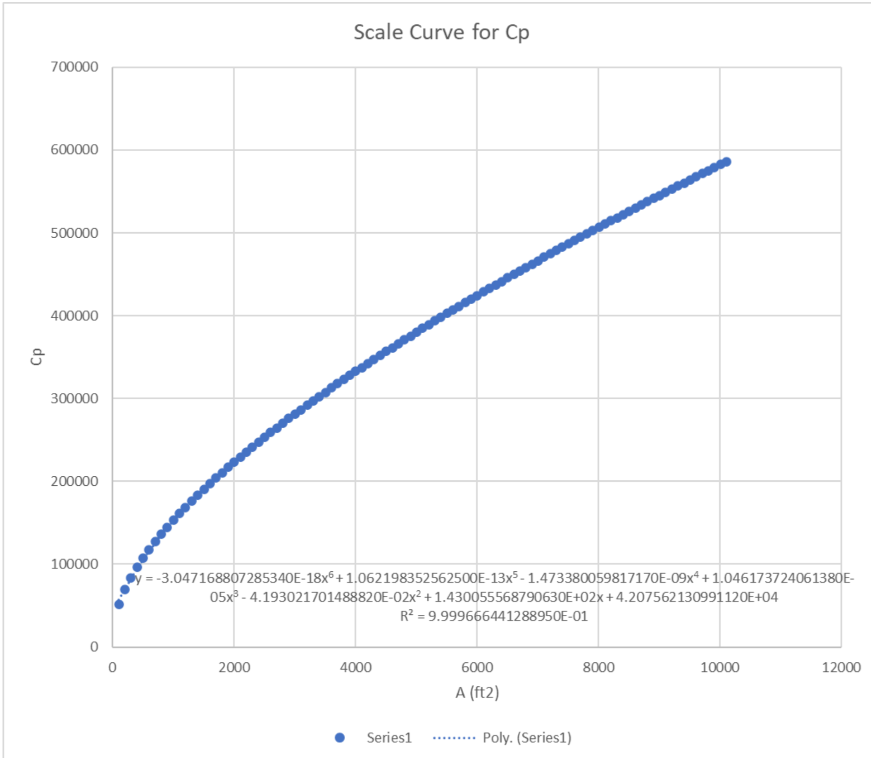


Figure 31: Piping Tables

Gas Pipe Diameter Table

		500					2500					5000		
V-101 to E-101					V-101 to E-101					V-101 to E-101				
Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)
0.75	0.824	0.2957	246	55.32378	0.75	0.824	1.4785	246	276.6189	0.75	0.824	2.957	246	553.2378
1	1.049	0.2957	246	34.13621	1	1.049	1.4785	246	170.681	1	1.049	2.957	246	341.3621
1.25	1.38	0.2957	246	19.72459	1.25	1.38	1.4785	246	98.62297	1.25	1.38	2.957	246	197.2459
1.5	1.61	0.2957	246	14.49154	1.5	1.61	1.4785	246	72.45769	1.5	1.61	2.957	246	144.9154
2	2.067	0.2957	246	8.791952	2	2.067	1.4785	246	43.95976	2	2.067	2.957	246	87.91952
2.5	2.469	0.2957	246	6.162034	2.5	2.469	1.4785	246	30.81017	2.5	2.469	2.957	246	61.62034
3	3.068	0.2957	246	3.990759	3	3.068	1.4785	246	19.9538	3	3.068	2.957	246	39.90759
3.5	3.548	0.2957	246	2.984001	3.5	3.548	1.4785	246	14.92001	3.5	3.548	2.957	246	29.84001
4	4.026	0.2957	246	2.317494	4	4.026	1.4785	246	11.58747	4	4.026	2.957	246	23.17494
V-102 to Tailgas					V-102 to Tailgas					V-102 to Tailgas				
Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)
1	1.049	0.1384	88	44.66347	1	1.049	0.692	88	223.3174	1	1.049	1.384	88	446.6347
1.25	1.38	0.1384	88	25.80746	1.25	1.38	0.692	88	129.0373	1.25	1.38	1.384	88	258.0746
1.5	1.61	0.1384	88	18.96058	1.5	1.61	0.692	88	94.80292	1.5	1.61	1.384	88	189.6058
2	2.067	0.1384	88	11.5033	2	2.067	0.692	88	57.51651	2	2.067	1.384	88	115.033
2.5	2.469	0.1384	88	8.062343	2.5	2.469	0.692	88	40.31171	2.5	2.469	1.384	88	80.62343
3	3.068	0.1384	88	5.221469	3	3.068	0.692	88	26.10735	3	3.068	1.384	88	52.21469
3.5	3.548	0.1384	88	3.904237	3.5	3.548	0.692	88	19.52119	3.5	3.548	1.384	88	39.04237
4	4.026	0.1384	88	3.032186	4	4.026	0.692	88	15.16093	4	4.026	1.384	88	30.32186
5	5.047	0.1384	88	1.929465	5	5.047	0.692	88	9.647323	5	5.047	1.384	88	19.29465
V-103 to E-102					V-103 to E-102					V-103 to E-102				
Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)
0.75	0.824	3.84E-02	35	50.5044	0.75	0.824	1.92E-01	35	252.522	0.75	0.824	3.84E-01	35	505.044
1	1.049	3.84E-02	35	31.16252	1	1.049	1.92E-01	35	155.8126	1	1.049	3.84E-01	35	311.6252
1.25	1.38	3.84E-02	35	18.00634	1.25	1.38	1.92E-01	35	90.0317	1.25	1.38	3.84E-01	35	180.0634
1.5	1.61	3.84E-02	35	13.22915	1.5	1.61	1.92E-01	35	66.14574	1.5	1.61	3.84E-01	35	132.2915
2	2.067	3.84E-02	35	8.026065	2	2.067	1.92E-01	35	40.13032	2	2.067	3.84E-01	35	80.26065
2.5	2.469	3.84E-02	35	5.625245	2.5	2.469	1.92E-01	35	28.12622	2.5	2.469	3.84E-01	35	56.25245
3	3.068	3.84E-02	35	3.643115	3	3.068	1.92E-01	35	18.21558	3	3.068	3.84E-01	35	36.43115
3.5	3.548	3.84E-02	35	2.724058	3.5	3.548	1.92E-01	35	13.62029	3.5	3.548	3.84E-01	35	27.24058
4	4.026	3.84E-02	35	2.115612	4	4.026	1.92E-01	35	10.57806	4	4.026	3.84E-01	35	21.15612
Syngas to Reactor					Syngas to Reactor					Syngas to Reactor				
Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)
4	4.026	1.18E+00	35	65.00046	4	4.026	5.90E+00	35	325.0023	4	4.026	11.8	35	650.0046
5	5.047	1.18E+00	35	41.3616	5	5.047	5.90E+00	35	206.808	5	5.047	1.18E+01	35	413.616
6	6.065	1.18E+00	35	28.64194	6	6.065	5.90E+00	35	143.2097	6	6.065	1.18E+01	35	286.4194
8	7.981	1.18E+00	35	16.54053	8	7.981	5.90E+00	35	82.70264	8	7.981	1.18E+01	35	165.4053
10	10.02	1.18E+00	35	10.4937	10	10.02	5.90E+00	35	52.46849	10	10.02	1.18E+01	35	104.937
12	11.94	1.18E+00	35	7.390185	12	11.94	5.90E+00	35	36.95093	12	11.94	1.18E+01	35	73.90185
14	13.12	1.18E+00	35	6.120633	14	13.12	5.90E+00	35	30.60316	14	13.12	1.18E+01	35	61.20633
Pump Reactor out					Pump Reactor out					Pump Reactor out				
Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)
4	4.026	1.148	35	63.23774	4	4.026	5.74	35	316.1887	4	4.026	11.48	35	632.3774
5	5.047	1.148	35	40.23993	5	5.047	5.74	35	201.1997	5	5.047	11.48	35	402.3993
6	6.065	1.148	35	27.86521	6	6.065	5.74	35	139.326	6	6.065	11.48	35	278.6521
8	7.981	1.148	35	16.09197	8	7.981	5.74	35	80.45986	8	7.981	11.48	35	160.9197
10	10.02	1.148	35	10.20912	10	10.02	5.74	35	51.04561	10	10.02	11.48	35	102.0912
12	11.94	1.148	35	7.189774	12	11.94	5.74	35	35.94887	12	11.94	11.48	35	71.89774
14	13.12	1.148	35	5.954649	14	13.12	5.74	35	29.77325	14	13.12	11.48	35	59.54649
E-103 in					E-103 in					E-103 in				
Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)	Nominal D	Inner Dia(	Q (MMCFD	P(psia)	V (ft/s)
3	3.068	0.5207	35	49.39225	3	3.068	2.6035	35	246.9613	3	3.068	5.207	35	493.9225
3.5	3.548	0.5207	35	36.93196	3.5	3.548	2.6035	35	184.6598	3.5	3.548	5.207	35	369.3196
4	4.026	0.5207	35	28.68283	4	4.026	2.6035	35	143.4142	4	4.026	5.207	35	286.8283
5	5.047	0.5207	35	18.25168	5	5.047	2.6035	35	91.25842	5	5.047	5.207	35	182.5168
6	6.065	0.5207	35	12.63886	6	6.065	2.6035	35	63.19431	6	6.065	5.207	35	126.3886
8	7.981	0.5207	35	7.298858	8	7.981	2.6035	35	36.49429	8	7.981	5.207	35	72.98858
10	10.02	0.5207	35	4.630566	10	10.02	2.6035	35	23.15283	10	10.02	5.207	35	46.30566
12	11.94	0.5207	35	3.261076	12	11.94	2.6035	35	16.30538	12	11.94	5.207	35	32.61076
14	13.12	0.5207	35	2.700859	14	13.12	2.6035	35	13.50429	14	13.12	5.207	35	27.00859

Liquid Pipe Diameter Table

500				2500				5000			
V-101 to Condensate				V-101 to Condensate				V-101 to Condensate			
Nominal D	Inner Dia(	Q (bbl/d)	V (ft/s)	Nominal D	Inner Dia(	Q (bbl/d)	V (ft/s)	Nominal D	Inner Dia(	Q (bbl/d)	V (ft/s)
0.125	0.269	29	4.774958	0.125	0.269	145	23.87479	0.125	0.269	290	47.74958
0.25	0.364	29	2.607783	0.25	0.364	145	13.03891	0.25	0.364	290	26.07783
0.375	0.493	29	1.421609	0.375	0.493	145	7.108047	0.375	0.493	290	14.21609
0.5	0.622	29	0.893086	0.5	0.622	145	4.465431	0.5	0.622	290	8.930862
V-101 to V-103				V-101 to V-103				V-101 to V-103			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
0.125	0.269	40	6.586149	0.125	0.269	200	32.93075	0.125	0.269	400	65.86149
0.25	0.364	40	3.596941	0.25	0.364	200	17.98471	0.25	0.364	400	35.96941
0.375	0.493	40	1.960841	0.375	0.493	200	9.804203	0.375	0.493	400	19.60841
0.5	0.622	40	1.231843	0.5	0.622	200	6.159215	0.5	0.622	400	12.31843
0.75	0.824	40	0.70191	0.75	0.824	200	3.509552	0.75	0.824	400	7.019105
1	1.049	40	0.433097	1	1.049	200	2.165485	1	1.049	400	4.33097
V-103 to Distillate				V-103 to Distillate				V-103 to Distillate			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
0.125	0.269	37	6.092188	0.125	0.269	185	30.46094	0.125	0.269	370	60.92188
0.25	0.364	37	3.327171	0.25	0.364	185	16.63585	0.25	0.364	370	33.27171
0.375	0.493	37	1.813778	0.375	0.493	185	9.068888	0.375	0.493	370	18.13778
0.5	0.622	37	1.139455	0.5	0.622	185	5.697274	0.5	0.622	370	11.39455
0.75	0.824	37	0.649267	0.75	0.824	185	3.246336	0.75	0.824	370	6.492672
V-103 to Condensate				V-103 to Condensate				V-103 to Condensate			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
0.125	0.269	17	2.799113	0.125	0.269	85	13.99557	0.125	0.269	170	27.99113
0.25	0.364	17	1.5287	0.25	0.364	85	7.643501	0.25	0.364	170	15.287
0.375	0.493	17	0.833357	0.375	0.493	85	4.166786	0.375	0.493	170	8.333573
V-102 to E-103				V-102 to E-103				V-102 to E-103			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
0.125	0.269	21	3.457728	0.125	0.269	105	17.28864	0.125	0.269	210	34.57728
0.25	0.364	21	1.888394	0.25	0.364	105	9.441971	0.25	0.364	210	18.88394
0.375	0.493	21	1.029441	0.375	0.493	105	5.147207	0.375	0.493	210	10.29441
0.5	0.622	21	0.646718	0.5	0.622	105	3.233588	0.5	0.622	210	6.467176
V-102 to Naphtha				V-102 to Naphtha				V-102 to Naphtha			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
0.125	0.269	6	0.987922	0.125	0.269	30	4.939612	0.125	0.269	60	9.879224
0.25	0.364	6	0.539541	0.25	0.364	30	2.697706	0.25	0.364	60	5.395412
0.375	0.493	6	0.294126	0.375	0.493	30	1.47063	0.375	0.493	60	2.941261
E-102 to Waste Water				E-102 to Waste Water				E-102 to Waste Water			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
0.125	0.269	43	7.080111	0.125	0.269	215	35.40055	0.125	0.269	430	70.80111
0.25	0.364	43	3.866712	0.25	0.364	215	19.33356	0.25	0.364	430	38.66712
0.375	0.493	43	2.107904	0.375	0.493	215	10.53952	0.375	0.493	430	21.07904
0.5	0.622	43	1.324231	0.5	0.622	215	6.621156	0.5	0.622	430	13.24231
0.75	0.824	43	0.754554	0.75	0.824	215	3.772769	0.75	0.824	430	7.545537
Pump reactor in				Pump reactor in				Pump reactor in			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
0.125	0.269	158.4	26.08115	0.125	0.269	792	130.4058	0.125	0.269	1584	260.8115
0.25	0.364	158.4	14.24389	0.25	0.364	792	71.21944	0.25	0.364	1584	142.4389
0.375	0.493	158.4	7.764929	0.375	0.493	792	38.82464	0.375	0.493	1584	77.64929
0.5	0.622	158.4	4.878098	0.5	0.622	792	24.39049	0.5	0.622	1584	48.78098
0.75	0.824	158.4	2.779565	0.75	0.824	792	13.89783	0.75	0.824	1584	27.79565
1	1.049	158.4	1.715064	1	1.049	792	8.57532	1	1.049	1584	17.15064
1.25	1.38	158.4	0.990999	1.25	1.38	792	4.954994	1.25	1.38	1584	9.909988
1.5	1.61	158.4	0.728081	1.5	1.61	792	3.640404	1.5	1.61	1584	7.280808
2	2.067	158.4	0.441723	2	2.067	792	2.208617	2	2.067	1584	4.417234
2.5	2.469	158.4	0.309592	2.5	2.469	792	1.547958	2.5	2.469	1584	3.095916
E-102 CW				E-102 CW				E-102 CW			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
1	1.049	751.6	8.137892	1	1.049	3758	40.68946	1	1.049	7516	81.37892
1.25	1.38	751.6	4.702239	1.25	1.38	3758	23.5112	1.25	1.38	7516	47.02239
1.5	1.61	751.6	3.454707	1.5	1.61	3758	17.27353	1.5	1.61	7516	34.54707
2	2.067	751.6	2.095955	2	2.067	3758	10.47978	2	2.067	7516	20.95955
2.5	2.469	751.6	1.468996	2.5	2.469	3758	7.344982	2.5	2.469	7516	14.68996
3	3.068	751.6	0.951376	3	3.068	3758	4.75688	3	3.068	7516	9.51376
3.5	3.548	751.6	0.71137	3.5	3.548	3758	3.556851	3.5	3.548	7516	7.113702
4	4.026	751.6	0.552478	4	4.026	3758	2.762392	4	4.026	7516	5.524785
5	5.047	751.6	0.351557	5	5.047	3758	1.757787	5	5.047	7516	3.515574
E-103 out				E-103 out				E-103 out			
Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
#REF!	0.493	72.21	3.539808	#REF!	0.493	361.05	17.69904	#REF!	0.493	722.1	35.39808
#REF!	0.622	72.21	2.223785	#REF!	0.622	361.05	11.11892	#REF!	0.622	722.1	22.23785
0.75	0.824	72.21	1.267124	0.75	0.824	361.05	6.335619	0.75	0.824	722.1	12.67124
1	1.049	72.21	0.781848	1	1.049	361.05	3.909242	1	1.049	722.1	7.818483
1.25	1.38	72.21	0.451768	1.25	1.38	361.05	2.258839	1.25	1.38	722.1	4.517678
1.5	1.61	72.21	0.331911	1.5	1.61	361.05	1.659555	1.5	1.61	722.1	3.319111
2	2.067	72.21	0.201369	2	2.067	361.05	1.006845	2	2.067	722.1	2.01369
				V-102 to V-103				V-102 to V-103			
				Nominal D	Inner Dia(	Q (bbl/d)	V=	Nominal D	Inner Dia(	Q (bbl/d)	V=
				0.125	0.269	103.9	17.10752	0.125	0.269	207.8	34.21505
				0.25	0.364	103.9	9.343055	0.25	0.364	207.8	18.68611
				0.375	0.493	103.9	5.093284	0.375	0.493	207.8	10.18657
				0.5	0.622	103.9	3.199712	0.5	0.622	207.8	6.399424
				0.75	0.824	103.9	1.823212	0.75	0.824	207.8	3.646425
				1	1.049	103.9	1.124969	1	1.049	207.8	2.249939

3-Phase Pipe Diameter Tables

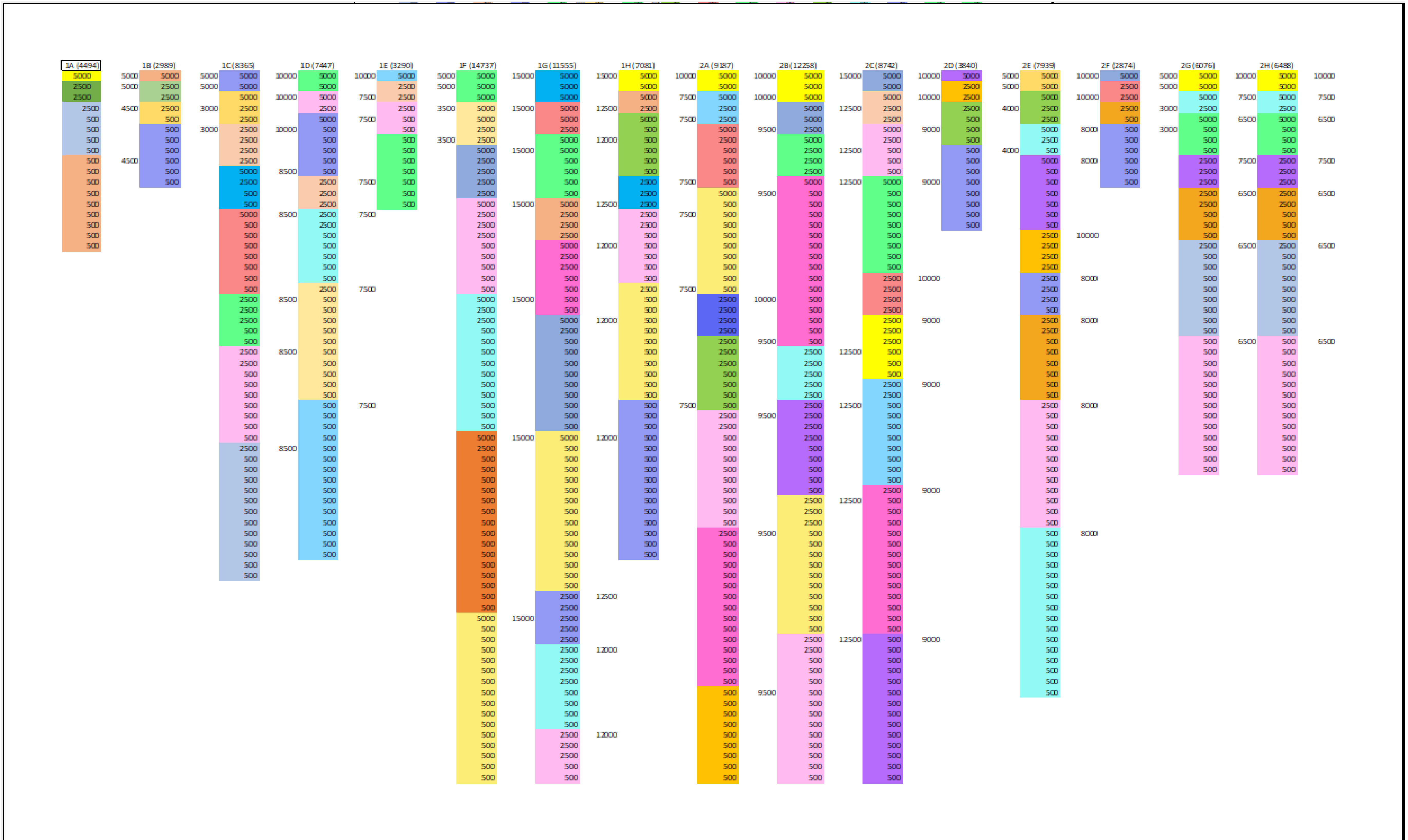
R-101 to V-101					
	Vapor	Liq	Aq		
Frac	0.558	0.0237	0.4183	0.442	
N				0	
V(m/s)	0.1			0.075416	
Q(m^3/s)	0.000154	6.39444E-05	5.18611E-05	0.000116	
D (m)	0.044217			0.044217	1.75in

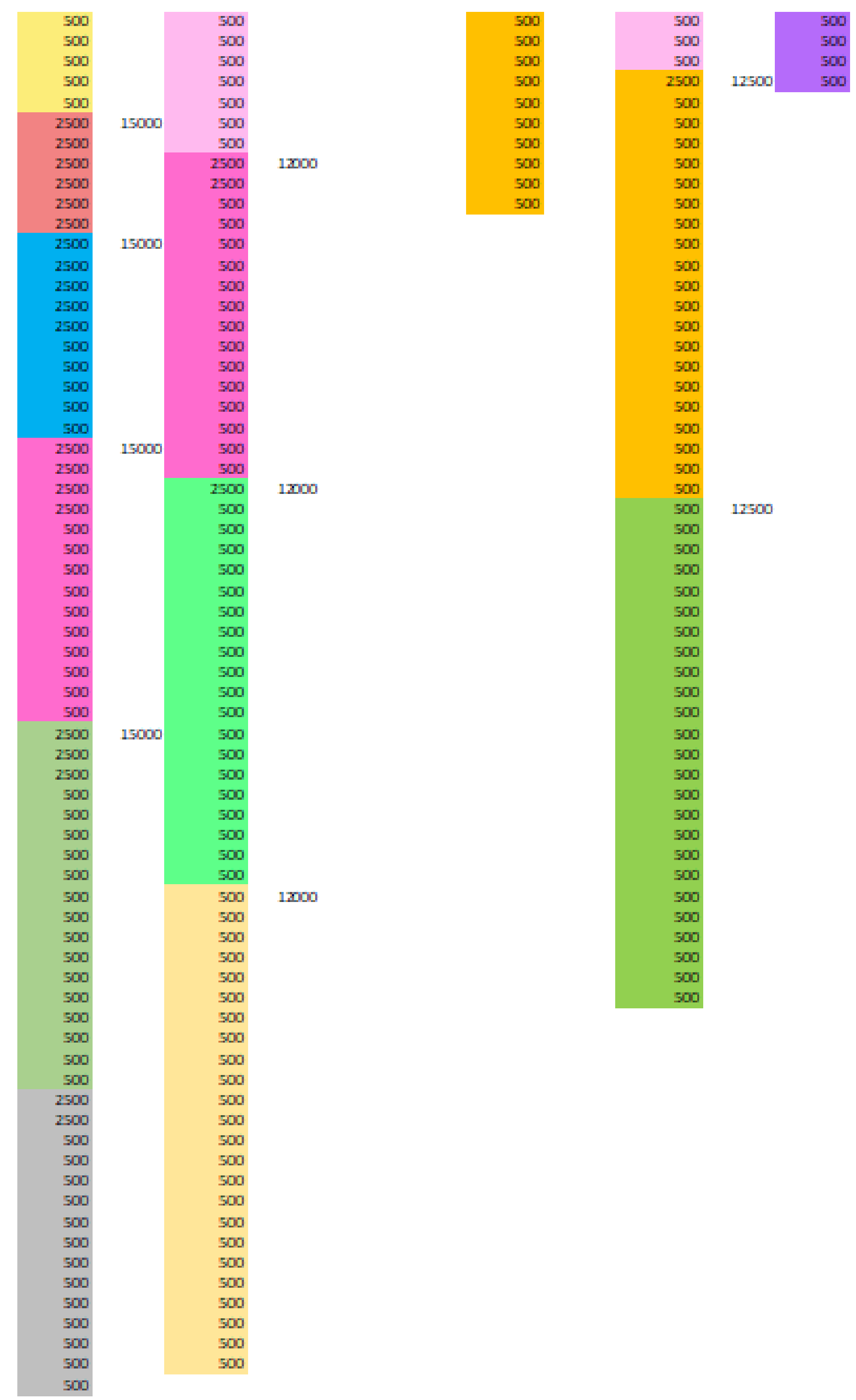
E-101 to V-102					
	Vapor	Liq	Aq		
Frac	0.558	0.0237	0.4183	0.442	
N				0	
V(m/s)	0.1			0.075416	
Q(m^3/s)	0.000154	6.39444E-05	5.18611E-05	0.000116	
D (m)	0.044217			0.044217	1.75in

2-Phase Pipe Diameter Table

Two Phase Gas/Liq Flow Pipe Dia					
Stream Name	QL (m3/s)	QG (m3/s)	Inner Dia (m)	VsG	VsL
E-103 to V-103	2.85404E-05	0.001040318	0.00635	32.84949847	0.901204071
E-103 to V-103	2.85404E-05	0.001040318	0.0127	8.212374619	0.225301018
E-103 to V-103	2.85404E-05	0.001040318	0.01905	3.649944275	0.100133786
E-103 to V-103	2.85404E-05	0.001040318	0.0254	2.053093655	0.056325254
E-103 to V-103	2.85404E-05	0.001040318	0.0381	0.912486069	0.025033446
E-103 to V-103	2.85404E-05	0.001040318	0.0508	0.513273414	0.014081314
E-103 to V-103	2.85404E-05	0.001040318	0.0635	0.328494985	0.009012041

Figure 32: Potential Unit Combinations per Wellhead





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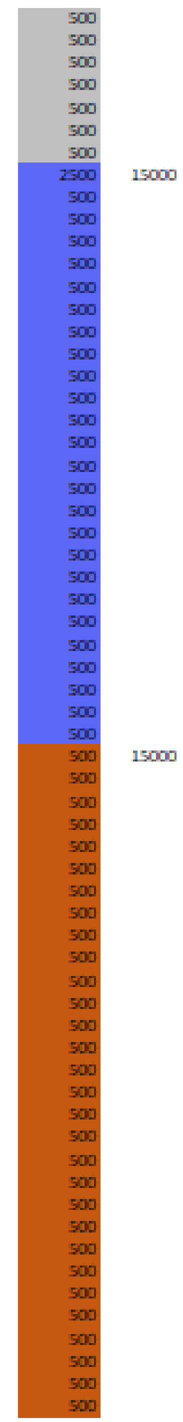
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Figure 33: Oil Field Map

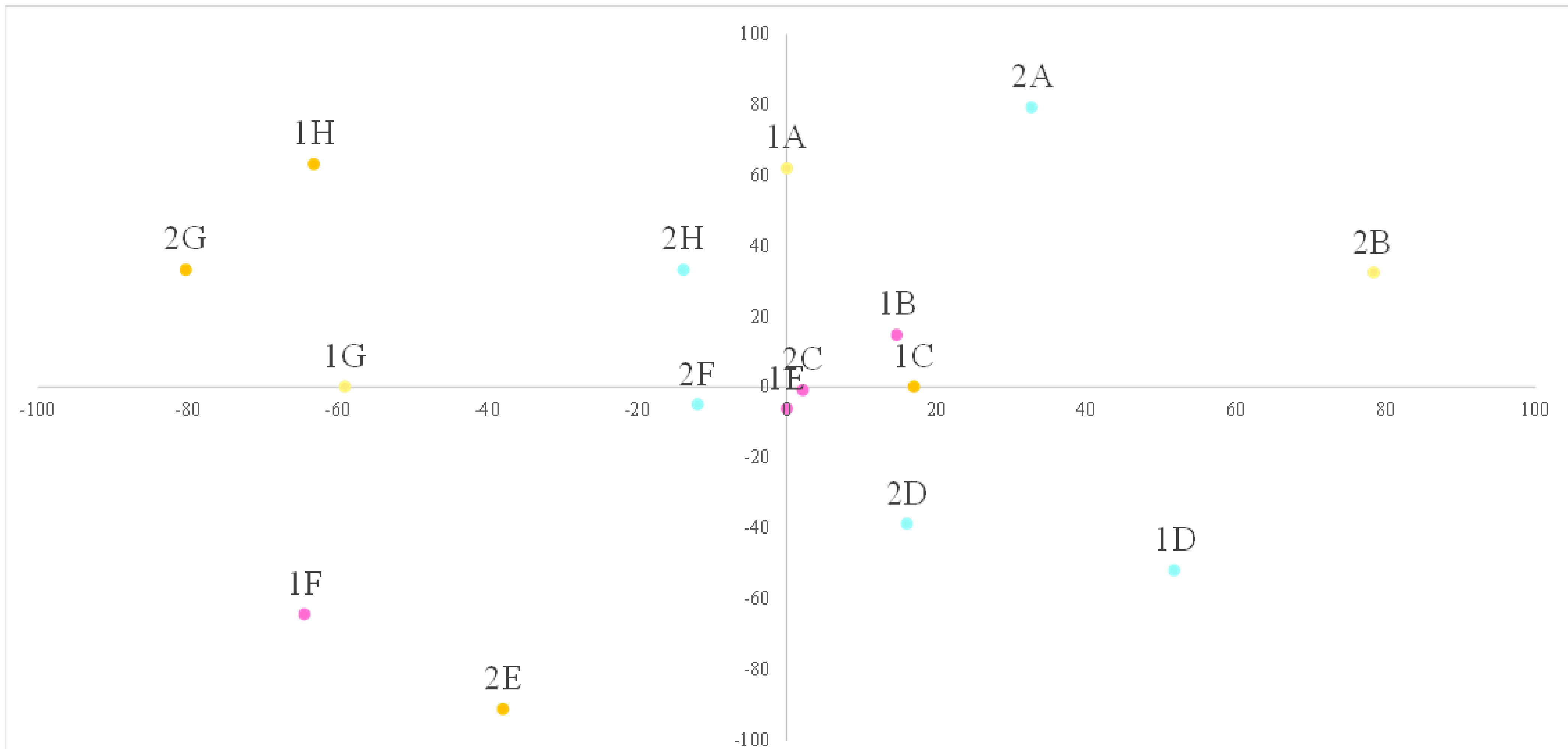


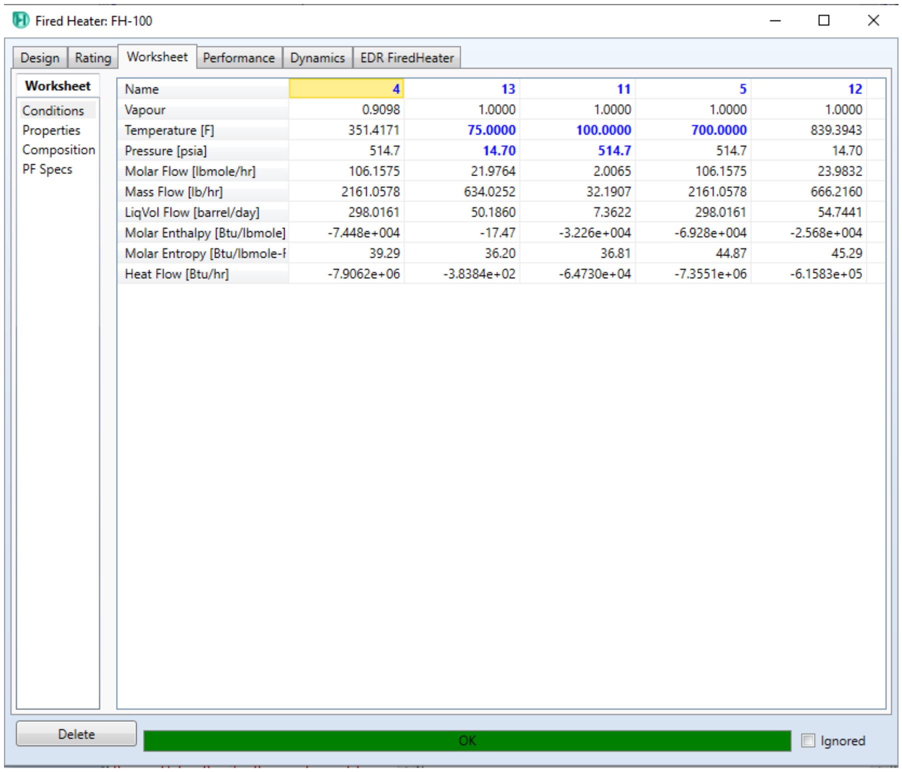
Figure 34: Initial Well Groups on Central Plant Capacity

Well ID														Sum	Total
1A		0	1	0	0	0	0	0	0	0	0	0	0	4494	4,494
1B		1	0	0	0	0	0	0	0	0	0	0	0	2989	2,989
1C		0		1	0	0	0	0	0	0	0	0	0	8365	8,365
1D		0	0	0	1	0	0	0	0	0	0	0	0	7447	7,447
1E		1	0	0	0	0	0	0	0	0	0	0	0	3290	3,290
1F		1	0	0	0	0	0	0	0	0	0	0	0	14737	14,737
1G		0	1	0	0	0	0	0	0	0	0	0	0	11155	11,155
1H		0	0	1	0	0	0	0	0	0	0	0	0	7081	7,081
2A		0	0	0	1	0	0	0	0	0	0	0	0	9187	9,187
2B		0	1	0	0	0	0	0	0	0	0	0	0	12258	12,258
2C		1	0	0	0	0	0	0	0	0	0	0	0	8742	8,742
2D		0	0	0	1	0	0	0	0	0	0	0	0	3840	3,840
2E		0	0	1	0	0	0	0	0	0	0	0	0	7939	7,939
2F		0	0	0	1	0	0	0	0	0	0	0	0	2874	2,874
2G		0	0	1	0	0	0	0	0	0	0	0	0	6076	6,076
2H		0	0	0	1	0	0	0	0	0	0	0	0	6488	6,488
SUM		29758	27907	29461	29836	0	0	0	0	0	0	0	0		
CP Max		30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000		
Yr		0	1	2	3	4	5	6	7	8	9	10	11		
MIN Cost	-1612.614407														

GROUPS
1B,1E,1F,2C
1A,1G,2B
1C,1H,2E,2G
1D,2A,2D,2H,2F

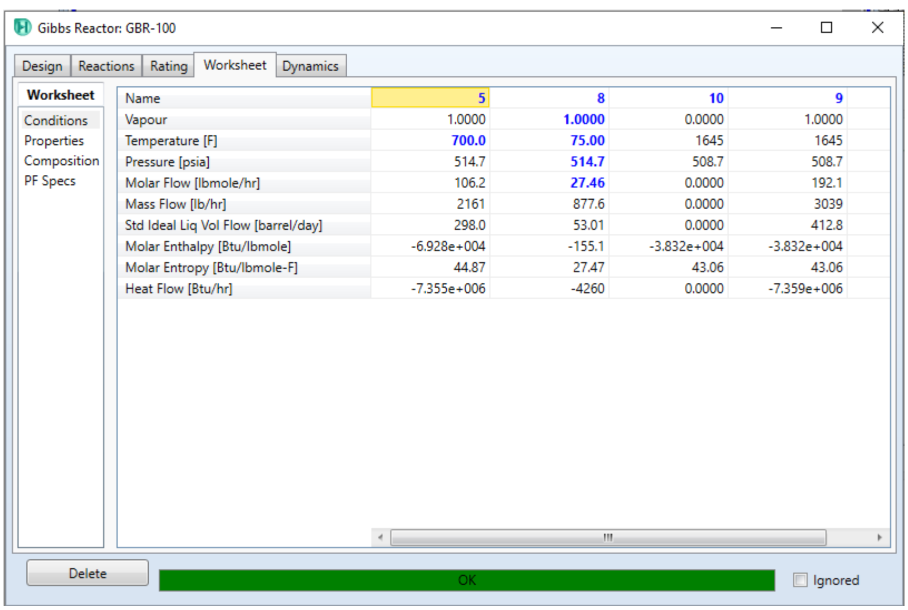
Total Cost												
1A	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86
1B	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26
1C	-114.13	-114.13	-114.13	-114.13	-114.13	-114.13	-114.13	-114.13	-114.13	-114.13	-114.13	-114.13
1D	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11
1E	-55.27	-55.27	-55.27	-55.27	-55.27	-55.27	-55.27	-55.27	-55.27	-55.27	-55.27	-55.27
1F	-176.57	-176.57	-176.57	-176.57	-176.57	-176.57	-176.57	-176.57	-176.57	-176.57	-176.57	-176.57
1G	-194.65	-194.65	-194.65	-194.65	-194.65	-194.65	-194.65	-194.65	-194.65	-194.65	-194.65	-194.65
1H	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11
2A	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71
2B	-152.97	-152.97	-152.97	-152.97	-152.97	-152.97	-152.97	-152.97	-152.97	-152.97	-152.97	-152.97
2C	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71	-117.71
2D	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86	-58.86
2E	-104.12	-104.12	-104.12	-104.12	-104.12	-104.12	-104.12	-104.12	-104.12	-104.12	-104.12	-104.12
2F	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26	-45.26
2G	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11	-94.11
2H	-88.89	-88.89	-88.89	-88.89	-88.89	-88.89	-88.89	-88.89	-88.89	-88.89	-88.89	-88.89

Figure 35: F500 MSCF Furnace



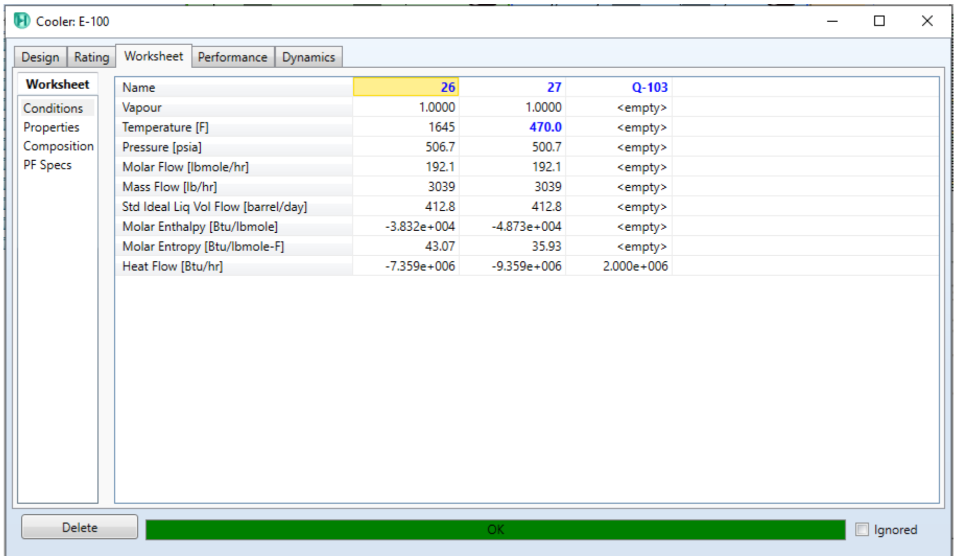
Name	4	13	11	5	12
Vapour	0.9098	1.0000	1.0000	1.0000	1.0000
Temperature [F]	351.4171	75.0000	100.0000	700.0000	839.3943
Pressure [psia]	514.7	14.70	514.7	514.7	14.70
Molar Flow [lbmole/hr]	106.1575	21.9764	2.0065	106.1575	23.9832
Mass Flow [lb/hr]	2161.0578	634.0252	32.1907	2161.0578	666.2160
LiqVol Flow [barrel/day]	298.0161	50.1860	7.3622	298.0161	54.7441
Molar Enthalpy [Btu/lbmole]	-7.448e+004	-17.47	-3.226e+004	-6.928e+004	-2.568e+004
Molar Entropy [Btu/lbmole-F	39.29	36.20	36.81	44.87	45.29
Heat Flow [Btu/hr]	-7.9062e+06	-3.8384e+02	-6.4730e+04	-7.3551e+06	-6.1583e+05

Figure 36: Gibbs reactor 500 MSC



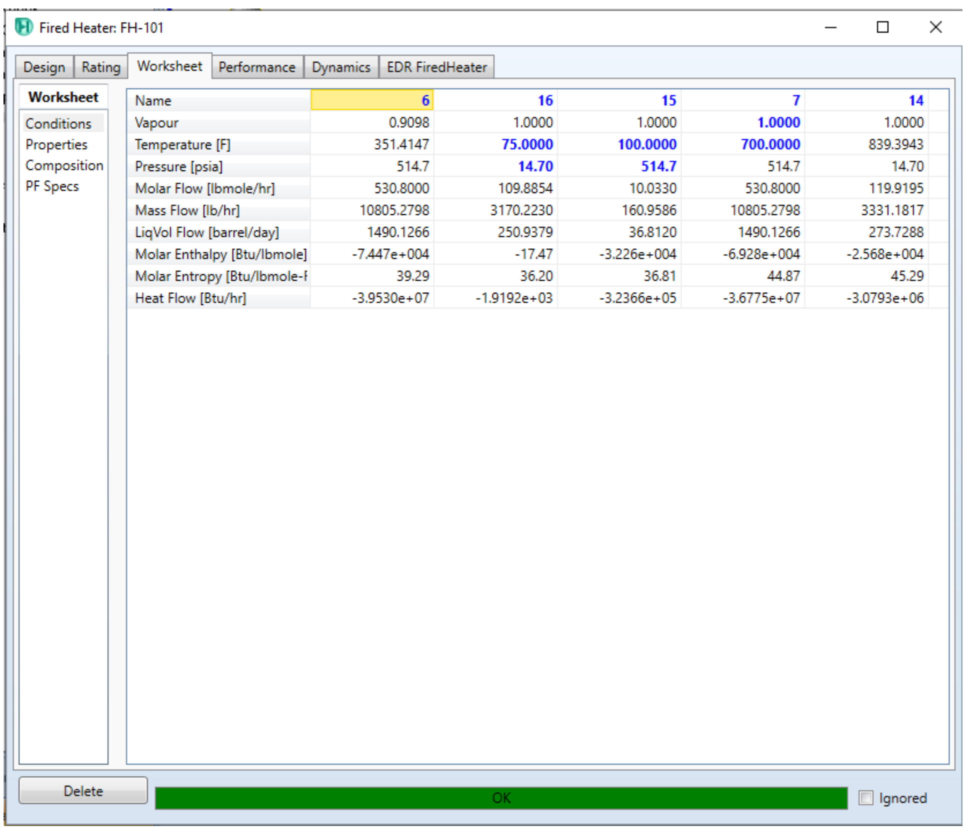
Name	5	8	10	9
Vapour	1.0000	1.0000	0.0000	1.0000
Temperature [F]	700.0	75.00	1645	1645
Pressure [psia]	514.7	514.7	508.7	508.7
Molar Flow [lbmole/hr]	106.2	27.46	0.0000	192.1
Mass Flow [lb/hr]	2161	877.6	0.0000	3039
Std Ideal Liq Vol Flow [barrel/day]	298.0	53.01	0.0000	412.8
Molar Enthalpy [Btu/lbmole]	-6.928e+004	-155.1	-3.832e+004	-3.832e+004
Molar Entropy [Btu/lbmole-F]	44.87	27.47	43.06	43.06
Heat Flow [Btu/hr]	-7.355e+006	-4260	0.0000	-7.359e+006

Figure 37: 500 Cooler



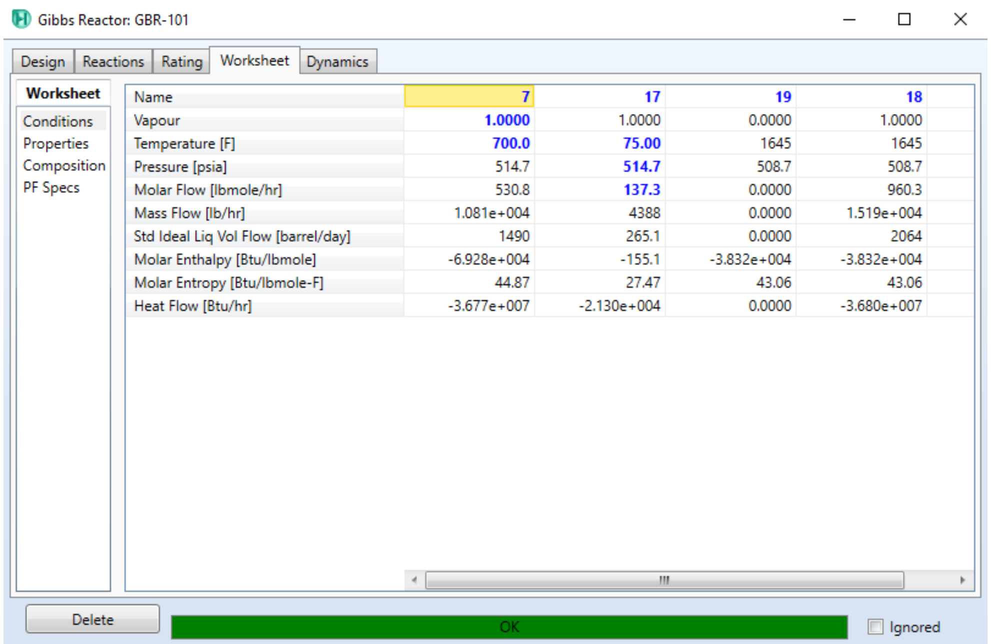
Name	26	27	Q-103
Vapour	1.0000	1.0000	<empty>
Temperature [F]	1645	470.0	<empty>
Pressure [psia]	506.7	500.7	<empty>
Molar Flow [lbmole/hr]	192.1	192.1	<empty>
Mass Flow [lb/hr]	3039	3039	<empty>
Std Ideal Liq Vol Flow [barrel/day]	412.8	412.8	<empty>
Molar Enthalpy [Btu/lbmole]	-3.832e+004	-4.873e+004	<empty>
Molar Entropy [Btu/lbmole-F]	43.07	35.93	<empty>
Heat Flow [Btu/hr]	-7.359e+006	-9.359e+006	2.000e+006

Figure 37: 2500 furnace



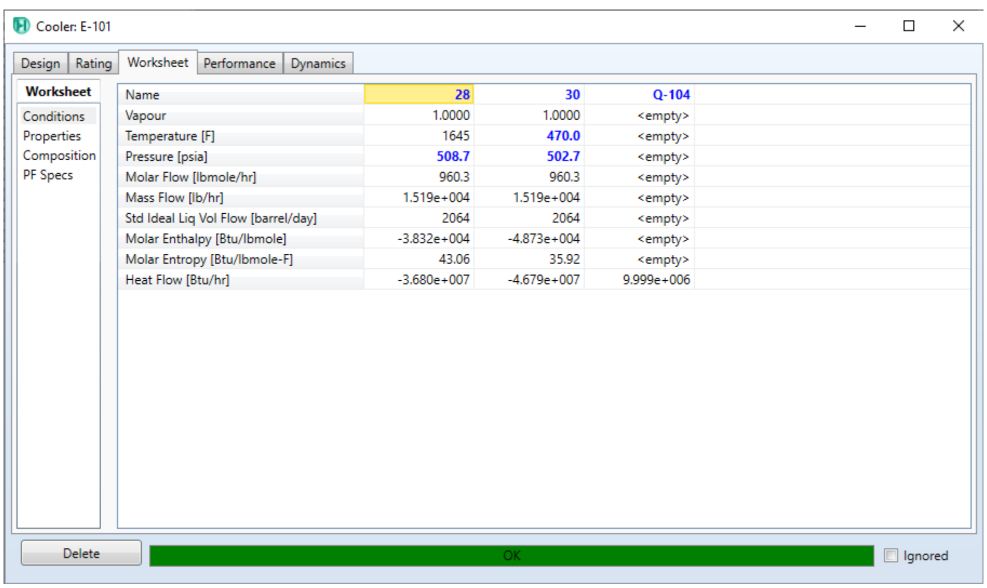
Name	6	16	15	7	14
Vapour	0.9098	1.0000	1.0000	1.0000	1.0000
Temperature [F]	351.4147	75.0000	100.0000	700.0000	839.3943
Pressure [psia]	514.7	14.70	514.7	514.7	14.70
Molar Flow [lbmole/hr]	530.8000	109.8854	10.0330	530.8000	119.9195
Mass Flow [lb/hr]	10805.2798	3170.2230	160.9586	10805.2798	3331.1817
LiqVol Flow [barrel/day]	1490.1266	250.9379	36.8120	1490.1266	273.7288
Molar Enthalpy [Btu/lbmole]	-7.447e+004	-17.47	-3.226e+004	-6.928e+004	-2.568e+004
Molar Entropy [Btu/lbmole-F	39.29	36.20	36.81	44.87	45.29
Heat Flow [Btu/hr]	-3.9530e+07	-1.9192e+03	-3.2366e+05	-3.6775e+07	-3.0793e+06

Figure 38: 2500 Gibbs reactor



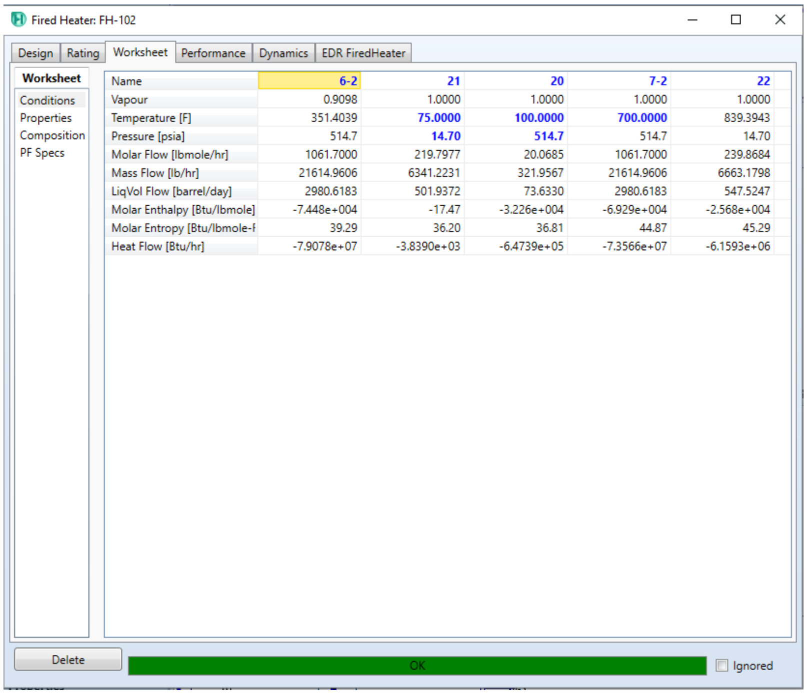
Name	7	17	19	18
Vapour	1.0000	1.0000	0.0000	1.0000
Temperature [F]	700.0	75.00	1645	1645
Pressure [psia]	514.7	514.7	508.7	508.7
Molar Flow [lbmole/hr]	530.8	137.3	0.0000	960.3
Mass Flow [lb/hr]	1.081e+004	4388	0.0000	1.519e+004
Std Ideal Liq Vol Flow [barrel/day]	1490	265.1	0.0000	2064
Molar Enthalpy [Btu/lbmole]	-6.928e+004	-155.1	-3.832e+004	-3.832e+004
Molar Entropy [Btu/lbmole-F]	44.87	27.47	43.06	43.06
Heat Flow [Btu/hr]	-3.677e+007	-2.130e+004	0.0000	-3.680e+007

Figure 39: 2500 cooler



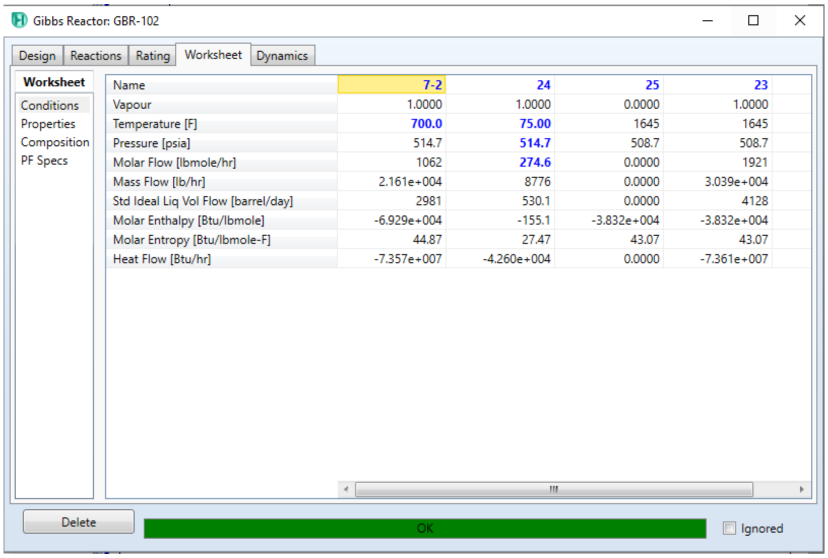
Name	28	30	Q-104
Vapour	1.0000	1.0000	<empty>
Temperature [F]	1645	470.0	<empty>
Pressure [psia]	508.7	502.7	<empty>
Molar Flow [lbmole/hr]	960.3	960.3	<empty>
Mass Flow [lb/hr]	1.519e+004	1.519e+004	<empty>
Std Ideal Liq Vol Flow [barrel/day]	2064	2064	<empty>
Molar Enthalpy [Btu/lbmole]	-3.832e+004	-4.873e+004	<empty>
Molar Entropy [Btu/lbmole-F]	43.06	35.92	<empty>
Heat Flow [Btu/hr]	-3.680e+007	-4.679e+007	9.999e+006

Figure 40: 5000 furnace



Name	6-2	21	20	7-2	22
Vapour	0.9098	1.0000	1.0000	1.0000	1.0000
Temperature [F]	351.4039	75.0000	100.0000	700.0000	839.3943
Pressure [psia]	514.7	14.70	514.7	514.7	14.70
Molar Flow [lbmole/hr]	1061.7000	219.7977	20.0685	1061.7000	239.8684
Mass Flow [lb/hr]	21614.9606	6341.2231	321.9567	21614.9606	6663.1798
LiqVol Flow [barrel/day]	2980.6183	501.9372	73.6330	2980.6183	547.5247
Molar Enthalpy [Btu/lbmole]	-7.448e+004	-17.47	-3.226e+004	-6.929e+004	-2.568e+004
Molar Entropy [Btu/lbmole-F	39.29	36.20	36.81	44.87	45.29
Heat Flow [Btu/hr]	-7.9078e+07	-3.8390e+03	-6.4739e+05	-7.3566e+07	-6.1593e+06

Figure 41: 5000 Gibbs reactor



Name	7-2	24	25	23
Vapour	1.0000	1.0000	0.0000	1.0000
Temperature [F]	700.0	75.00	1645	1645
Pressure [psia]	514.7	514.7	508.7	508.7
Molar Flow [lbmole/hr]	1062	274.6	0.0000	1921
Mass Flow [lb/hr]	2.161e+004	8776	0.0000	3.039e+004
Std Ideal Liq Vol Flow [barrel/day]	2981	530.1	0.0000	4128
Molar Enthalpy [Btu/lbmole]	-6.929e+004	-155.1	-3.832e+004	-3.832e+004
Molar Entropy [Btu/lbmole-F]	44.87	27.47	43.07	43.07
Heat Flow [Btu/hr]	-7.357e+007	-4.260e+004	0.0000	-7.361e+007

Figure 42: 5000 Cooler

Cooler: E-102

Name	29	31	Q-105
Vapour	1.0000	1.0000	<empty>
Temperature [F]	1645	470.0	<empty>
Pressure [psia]	508.7	502.7	<empty>
Molar Flow [lbmole/hr]	1921	1921	<empty>
Mass Flow [lb/hr]	3.039e+004	3.039e+004	<empty>
Std Ideal Liq Vol Flow [barrel/day]	4128	4128	<empty>
Molar Enthalpy [Btu/lbmole]	-3.832e+004	-4.873e+004	<empty>
Molar Entropy [Btu/lbmole-F]	43.07	35.92	<empty>
Heat Flow [Btu/hr]	-7.361e+007	-9.361e+007	2.000e+007

Figure 43: Absorber 500 MSCFD

Column: T-100 / COL1 Fluid Pkg: Basis-1 / Sour SRK

Name	500 Solv Water @COL1	500 SU @COL1	500 Pre FTR HE @COL1	500 Rich Solv @COL1
Vapour	0.0000	1.0000	1.0000	0.0000
Temperature [F]	80.00	470.0	80.15	88.10
Pressure [psia]	500.0	500.0	444.0	445.0
Molar Flow [lbmole/hr]	9500	192.1	129.8	9562
Mass Flow [lb/hr]	1.711e+005	3039	1482	1.727e+005
Std Ideal Liq Vol Flow [barrel/day]	1.174e+004	412.9	294.8	1.186e+004
Molar Enthalpy [Btu/lbmole]	-1.230e+005	-4.871e+004	-1.837e+004	-1.230e+005
Molar Entropy [Btu/lbmole-F]	12.82	35.94	28.48	13.15
Heat Flow [Btu/hr]	-1.169e+009	-9.358e+006	-2.385e+006	-1.176e+009

Figure 44: Absorber 2500 MSCFD

Column: T-101 / COL2 Fluid Pkg: Basis-1 / Sour SRK

Design | Parameters | Side Ops | Internals | Rating | Worksheet | Performance | Flowsheet | Reactions | Dynamics

Worksheet

- Conditions
- Properties
- Compositions
- PF Specs

Name	2.5 MM Solv W @COL2	2.5 MM SU @COL2	2.5 MM Pre-FTR HE @COL2	2.5 MM Rich Solv @COL2
Vapour	0.0000	1.0000	1.0000	0.0000
Temperature [F]	80.00	470.0	80.15	88.09
Pressure [psia]	500.0	500.0	444.0	445.0
Molar Flow [lbmole/hr]	4.750e+004	960.3	649.0	4.781e+004
Mass Flow [lb/hr]	8.557e+005	1.519e+004	7407	8.635e+005
Std Ideal Liq Vol Flow [barrel/day]	5.871e+004	2064	1474	5.930e+004
Molar Enthalpy [Btu/lbmole]	-1.230e+005	-4.871e+004	-1.837e+004	-1.230e+005
Molar Entropy [Btu/lbmole-F]	12.82	35.94	28.48	13.15
Heat Flow [Btu/hr]	-5.844e+009	-4.678e+007	-1.192e+007	-5.879e+009