Letter of Transmittal

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From: Team 5

Subject: 2022-2023 AIChE Student Design Competition Problem Statement & Rules

The following information is that of the final conclusions and analysis regarding the requested additional processing unit for the purification of pyoil at the Bali Global Petrochemical Plant. This project has been fully analyzed at the preliminary design stage, with economic and design conclusions.

Included in this report are the Process Flow Diagram, Piping and Instrumentation Diagram, Process Safety Elements, Capital Cost, and Variable Cost Estimates. Within this documentation, technical and economic analysis will be discussed. The technical discussion includes the design basis and assumptions used to complete the project. The economic discussion includes the cost equations and assumptions utilized to complete this analysis. Recommendations for further analysis, including action items and design recommendations are included in the conclusion section of the report. All reference tables, constants, and equations are listed within the report, or in the attached appendices.

It is believed that this project is ready to move into the detailed design stage and should continue to be analyzed for further accuracy of results. In parallel, an economic analysis should be completed after establishing feed prices, and values associated with the products of the distillation towers.

The proposed project will purify the required mass flow rate of Pyoil from the pyrolizer in preparation for the steam cracking units. This is based off a total capital cost of $13.04MM and a total annual operating cost of $908.5M for fixed and variable operating cost.

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Engineer 1, Engineering 2, Engineer 3, Engineer 4

Team Number: 5

Executive Summary

Global Petrochemicals (GP) currently is a leading manufacturer of basic chemicals and high-performance polymers.[1] As a result, the company has a substantial customer base. In recent years, the customer base has begun demanding GP to act in reducing the amount of plastic waste worldwide as well as produced by the company. [1] This pressure has led GP to make a commitment to produce 10% of its virgin resin-quality plastics from recovered plastic in the near future. [1] Currently, this is an ambitious goal that requires significant applications of new process technology and major investments. With the additional process unit in the following documentation, GP would establish efficient and effective plastic waste collection, while also utilizing this plastic waste to create the basic chemicals that are in demand from GP's customer base. A preliminary project design has been completed for this unit.

The additional process unit would be a Pyoil Purification Unit that purifies a pyrolysis oil stream to meet the specifications needed for downstream steam cracker units. This unit would consist of an adsorption process to dehydrate and remove heavy metal contaminants from the pyrolysis oil, before moving into a small distillation column to separate the lightest components into a fuel gas stream. Then, the bottoms from the first column move into a second, larger column to separate into Light Cut, Medium Cut, and Heavy Cut streams. The gas would be fed directly to the ethylene plant to be used as a fuel gas, while the cuts would be cooled and stored in tankage. Finally, the Light and Medium Cuts would be pumped from tankage to separate steam crackers for further processing, while the Heavy Cut would be sold as an asphalt component.

The design process is dependent upon the adsorption and regeneration towers to send continuous flow to the distillation columns, as well as the distillation columns to separate the process fluid into on-spec streams that can be cracked by the steam crackers. There is also a significant dependence upon the predistillation column coolers to keep the feed stream at a consistent temperature that keeps the distillation columns at steady state. Finally, the post distillation coolers are heavily relied upon to maintain safe temperatures in the tanks.

This project will not bring any new safety concerns into the petrochemical site that are not already present. All materials, except for the adsorbent material, can be found within the current site. The group added multiple relief options throughout the unit to prevent over pressurizing any piece of equipment to failure. Each relief option does not relieve to the atmosphere but does either relieve to flare or another treated exhaust stream. Based on current analysis, there are not any additional significant safety concerns that will be added to the site with the addition of this unit.

A service factor of 97% was assumed and utilized for this project. The total expected capital cost, variable cost, and fixed costs were determined and are reported. It was concluded that the project will cost $13.04MM in capital, with an expected $233M in variable costs, and an annual $675.4M in fixed costs. A complete economic analysis should be performed when values can be assigned to the cost of the unit feed and the product stream values.

In summary, the preliminary design analysis demonstrated that the development of a Pyoil Purification Unit is feasible. While the unit will require substantial efforts in the detailed design stage to maintain safety and environmentally friendly, the unit can be accomplished at the desired mass flow rate and produce all the requested product streams and at their desired specifications.

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Brief Process Description

Plastic waste is a global issue. Plastic materials are produced at a rapid rate, yet there are few ways to reuse the material. GP aims to recover plastic and create usable material from pyrolyzed oil. This unit would take in the heated plastic waste as pyrolyzed oil into the unit as feedstock. The feed enter the unit at 100 °F and 52,432 lb/hr. [1] The lighter component compositions in the feed were specified and utilized for the simulation. A purification unit was designed in order to remove any impurities that could enter in the feed stream. Therefore, an adsorption tower was designed with two different adsorption beads to remove the various contaminants and water. The adsorption towers were designed to be 50 ft tall and 10 ft in diameter. Trays were designed and priced to ensure the adsorbent beads do not mix and/or get crushed from the mass above. Once exiting the adsorption towers, the stream flowed to the initial distillation tower which was designed and sized to be a two-tray tower to remove the lightest (fuel gas) components from the distillation feed. The initial distillation tower is 15 ft tall and 2.75 ft in diameter. Once exiting the initial distillation tower, the feed moved into a second distillation tower to separate into the three cut components. This was the primary, larger distillation column at 58 ft tall and 9.25 ft in diameter. Each stream off the second column was cooled before being pumped to tankage at the required specifications in Figure 1. The product streams were labeled as the Py Gas, Pyoil Light Cut, Pyoil Medium Cut, and Pyoil Heavy Cut. End boiling points were required and achieved for both the Pyoil Light Cut and the Pyoil Medium Cut, therefore the steam cracking downstream units should properly crack these materials.[1]. The Pyoil Gas was sent to the ethylene product directly to be used as a fuel gas, and the bottoms, or Heavy Cut was sent to tankage to be either sold as an asphalt mixing component, or for other purposes.

Process Details

The required specifications for the products leaving the purification unit are detailed in Table 1. As a result, the designed process does meet all the specifications listed.

Purification Unit Product Stream Name	Py Gas	Pyoil Light Cut	Pyoil Medium Cut	Pyoil Heavy Cut
Steam Cracker Feed Name	Olefin-Rich Vapor, fed directly to the Ethylene Plant	Naphtha (Cracking Furnace Feedstock)	Gas Oil (Cracking Furnace Feedstock)	Not suitable for Steam Cracker
End Boiling Point (EBP)[Note 1]	N/A	392° F (200° C)	752° F (400° C)	--
Temperature	107° F	100° F	100° F	100° F
Pressure	Minimum 2.4 psig	70 psig	70 psig	50 psig

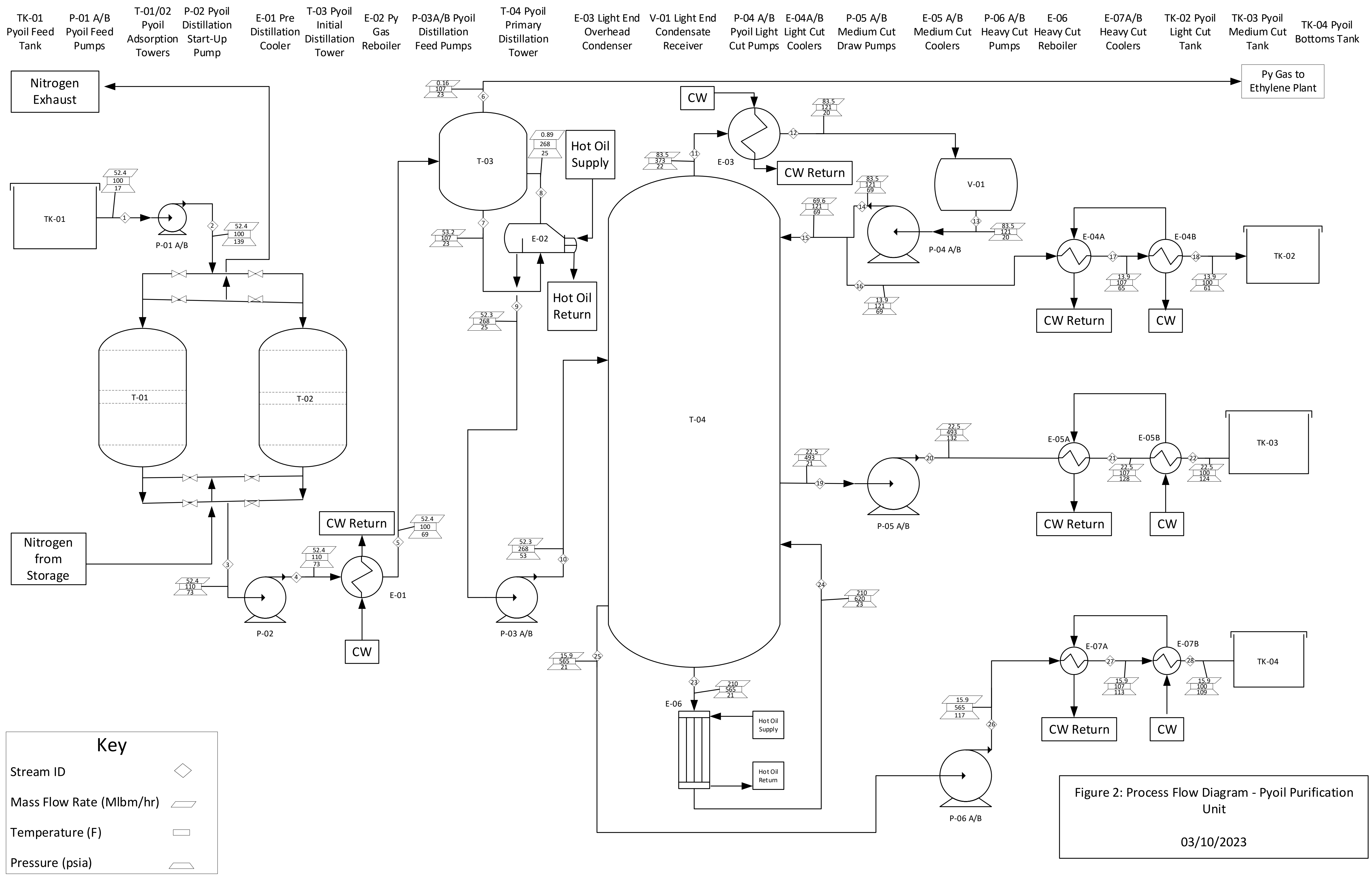
Note 1: ASTM D86 at 760 mm Hg, LV%

Figure 1: Global Petrochemical's Pyoil Purification Process Product Specification [1]

Feed enters the purification unit from the feed tank (TK-01). This tank has a fixed roof, similar to the product tanks that will be discussed later. The process fluid is pumped from TK-01 to the adsorption columns. Feed will only move through one of these towers as the other one is simultaneously regenerated. The pump in between the adsorption columns and the initial distillation column is only anticipated to be turned on for the startup process and was not duplicated. Due to the exothermic nature of the adsorption process, the feed will gain 10 °F – 20 °F before entering the initial distillation tower. The adsorption tower could potentially heat the feed beyond 100 °F, which is the maximum entering temperature for the distillation columns to ensure proper separation; therefore, a process control scheme was set up to cool the pre-primary distillation feed as necessary. Multiple distillation column setups were evaluated, these included swaged columns, multiple columns, single columns, and others, but the optimized design was that of a dual distillation setup. The first distillation column was smaller, only separating out the Py Gas without a condenser reflux. The Py Gas is directly sent to the ethylene plant to be used as a fuel gas. Then, the bottoms product from the initial distillation tower is sent to a larger, secondary distillation column. The secondary distillation column was designed to pull the Pyoil Light Cut as the distillate product, the Medium Cut as a side draw, and the Heavy Cut as the bottoms. Once leaving the secondary distillation columns, all streams were cooled to the temperatures as required in Table 1 before entering separate fixed roofs tanks. The process is illustrated in Figure 2, the process flow diagram.

Process Flow Diagram

Figure 2: Process Flow Diagram of Pyoil Purification Process



Material Balance

Each numbered stream in Figure 2 is specified in Table 1, the process stream table, with its respective temperature, pressure, mass flow rate, and with individual component flow rates.

Table 1: Process Stream Table

Stream Number		1	2	3	4	5	6	7
Description	Parameter Units	From TK1	To Ads	From Ads	To E-01	To T-03	Py Gas	T-03 Btm
Vapor Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.000	0.0000
Pressure	Psia	16.64	139.0	72.52	72.52	68.52	23.00	23.00
Temperature	°F	100.0	100.0	110.0	110.0	100.0	107.0	107.0
Molar Enthalpy	Mbtu/lbmol	-166.3	-166.2	-165.3	-165.3	-166.2	-17.68	-160.8
Density (Actual)	lb/ft3	48.03	48.07	47.79	47.79	48.05	0.1408	47.68
Molecular Weight		182.0	182.0	182.0	182.0	182.0	36.65	177.0
Total Molar Flow Rate	lbmol/hr	288.1	288.1	288.1	288.1	288.1	4.431	300.3
Volumetric Flow Rate (Actual)	bbl/day	4666	4663	4690	4690	4665	4930	4765
Total Mass Flow Rate	Mlb/hr	52.43	52.43	52.43	52.43	52.43	0.1624	53.16
Component Mass Flowrates								
Water	Mlb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Nitrogen	Mlb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Hydrogen	Mlb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	Mlb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	Mlb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methane	Mlb/hr	0.0052	0.0052	0.0052	0.0052	0.0052	0.0050	0.0024
Ethane	Mlb/hr	0.0524	0.0524	0.0524	0.0524	0.0524	0.0350	0.0793
Ethylene	Mlb/hr	0.0262	0.0262	0.0262	0.0262	0.0262	0.0205	0.0319
Propane	Mlb/hr	0.1521	0.1521	0.1521	0.1521	0.1521	0.0397	0.2944
Propene	Mlb/hr	0.1363	0.1363	0.1363	0.1363	0.1363	0.0423	0.2650
n-Butane	Mlb/hr	0.1888	0.1888	0.1888	0.1888	0.1888	0.0126	0.3064
13-Butadiene	Mlb/hr	0.0367	0.0367	0.0367	0.0367	0.0367	0.0030	0.0618
n-Pentane	Mlb/hr	0.0682	0.0682	0.0682	0.0682	0.0682	0.0012	0.0905
n-Hexane	Mlb/hr	0.2097	0.2097	0.2097	0.2097	0.2097	0.0010	0.2439
NBP[0]171*	Mlb/hr	0.1006	0.1006	0.1006	0.1006	0.1006	0.0004	0.1140
NBP[0]202*	Mlb/hr	0.1258	0.1258	0.1258	0.1258	0.1258	0.0002	0.1371
NBP[0]226*	Mlb/hr	0.1396	0.1396	0.1396	0.1396	0.1396	0.0002	0.1487
NBP[0]246*	Mlb/hr	0.2470	0.2470	0.2470	0.2470	0.2470	0.0002	0.2592
NBP[0]277*	Mlb/hr	0.3906	0.3906	0.3906	0.3906	0.3906	0.0001	0.4028
NBP[0]302*	Mlb/hr	1.363	1.363	1.363	1.363	1.363	0.0003	1.393
NBP[0]325*	Mlb/hr	2.694	2.694	2.694	2.694	2.694	0.0003	2.735
NBP[0]352*	Mlb/hr	2.767	2.767	2.767	2.767	2.767	0.0002	2.794
NBP[0]375*	Mlb/hr	3.911	3.911	3.911	3.911	3.911	0.0001	3.938
NBP[0]400*	Mlb/hr	4.012	4.012	4.012	4.012	4.012	0.0001	4.029
NBP[0]426*	Mlb/hr	3.595	3.595	3.595	3.595	3.595	0.0000	3.605
NBP[0]452*	Mlb/hr	3.268	3.268	3.268	3.268	3.268	0.0000	3.274
NBP[0]479*	Mlb/hr	3.352	3.352	3.352	3.352	3.352	0.0000	3.356
NBP[0]504*	Mlb/hr	3.720	3.720	3.720	3.720	3.720	0.0000	3.722
NBP[0]530*	Mlb/hr	3.986	3.986	3.986	3.986	3.986	0.0000	3.988
NBP[0]555*	Mlb/hr	3.888	3.888	3.888	3.888	3.888	0.0000	3.889
NBP[0]580*	Mlb/hr	3.204	3.204	3.204	3.204	3.204	0.0000	3.204
NBP[0]608*	Mlb/hr	2.635	2.635	2.635	2.635	2.635	0.0000	2.636
NBP[0]631*	Mlb/hr	2.093	2.093	2.093	2.093	2.093	0.0000	2.093
NBP[0]659*	Mlb/hr	1.457	1.457	1.457	1.457	1.457	0.0000	1.457
NBP[0]683*	Mlb/hr	1.293	1.293	1.293	1.293	1.293	0.0000	1.293
NBP[0]710*	Mlb/hr	0.7728	0.7728	0.7728	0.7728	0.7728	0.0000	0.7728
NBP[0]736*	Mlb/hr	0.7100	0.7100	0.7100	0.7100	0.7100	0.0000	0.7100
NBP[0]761*	Mlb/hr	0.6194	0.6194	0.6194	0.6194	0.6194	0.0000	0.6194
NBP[0]787*	Mlb/hr	0.4951	0.4951	0.4951	0.4951	0.4951	0.0000	0.4951
NBP[0]821*	Mlb/hr	0.7160	0.7160	0.7160	0.7160	0.7160	0.0000	0.7160

Stream Number		8	9	10	11	12	13	14
Description	Parameter Units	T-03 boil up	To P-03	To T-04	T-04 OH Vap	To V-01	To P-04	Reflux pre-split
Vapor Fraction		1.000	0.0000	0.0000	1.000	0.0000	0.0000	0.0000
Pressure	Psia	25.00	25.00	53.04	22.00	20.00	20.00	68.72
Temperature	°F	267.8	267.8	267.8	372.7	121.0	121.0	121.0
Molar Enthalpy	Btu/lbmol	-27.59	-152.1	-152.0	-81.87	-115.7	-115.7	-115.7
Density (Actual)	lb/ft3	0.1740	43.55	43.56	0.3335	44.84	44.84	44.86
Molecular Weight		53.40	184.3	184.3	127.3	127.3	127.3	127.3
Total Molar Flow Rate	lbmol/hr	16.61	283.7	283.7	655.8	655.8	655.8	655.8
Volumetric Flow Rate (Actual)	bbl/day	21790	5130	5129	1070000	7956	7956	7952
Total Mass Flow Rate	lb/hr	0.8869	52.27	52.27	83.46	83.46	83.46	83.46
Component Mass Flowrates								
Water	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Nitrogen	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Hydrogen	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methane	lb/hr	0.0022	0.0002	0.0002	0.0013	0.0013	0.0013	0.0013
Ethane	lb/hr	0.0618	0.0175	0.0175	0.1048	0.1048	0.1048	0.1048
Ethylene	lb/hr	0.0261	0.0058	0.0058	0.0345	0.0345	0.0345	0.0345
Propane	lb/hr	0.1820	0.1123	0.1123	0.6739	0.6739	0.6739	0.6739
Propene	lb/hr	0.1710	0.0940	0.0940	0.5642	0.5642	0.5642	0.5642
n-Butane	lb/hr	0.1302	0.1762	0.1762	1.057	1.057	1.057	1.057
13-Butadiene	lb/hr	0.0281	0.0337	0.0337	0.2021	0.2021	0.2021	0.2021
n-Pentane	lb/hr	0.0235	0.0670	0.0670	0.4019	0.4019	0.4019	0.4019
n-Hexane	lb/hr	0.0353	0.2087	0.2087	1.252	1.252	1.252	1.252
NBP[0]171*	lb/hr	0.0137	0.1003	0.1003	0.6016	0.6016	0.6016	0.6016
NBP[0]202*	lb/hr	0.0116	0.1256	0.1256	0.7535	0.7535	0.7535	0.7535
NBP[0]226*	lb/hr	0.0092	0.1395	0.1395	0.8368	0.8368	0.8368	0.8368
NBP[0]246*	lb/hr	0.0123	0.2469	0.2469	1.481	1.481	1.481	1.481
NBP[0]277*	lb/hr	0.0124	0.3904	0.3904	2.343	2.343	2.343	2.343
NBP[0]302*	lb/hr	0.0298	1.363	1.363	8.179	8.179	8.179	8.179
NBP[0]325*	lb/hr	0.0410	2.694	2.694	16.16	16.16	16.16	16.16
NBP[0]352*	lb/hr	0.0275	2.766	2.766	16.53	16.53	16.53	16.53
NBP[0]375*	lb/hr	0.0266	3.911	3.911	22.10	22.10	22.10	22.10
NBP[0]400*	lb/hr	0.0177	4.012	4.012	9.593	9.593	9.593	9.593
NBP[0]426*	lb/hr	0.0100	3.595	3.595	0.5500	0.5500	0.5500	0.5500
NBP[0]452*	lb/hr	0.0057	3.268	3.268	0.0335	0.0335	0.0335	0.0335
NBP[0]479*	lb/hr	0.0036	3.352	3.352	0.0038	0.0038	0.0038	0.0038
NBP[0]504*	lb/hr	0.0024	3.720	3.720	0.0006	0.0006	0.0006	0.0006
NBP[0]530*	lb/hr	0.0016	3.986	3.986	0.0001	0.0001	0.0001	0.0001
NBP[0]555*	lb/hr	0.0009	3.888	3.888	0.0000	0.0000	0.0000	0.0000
NBP[0]580*	lb/hr	0.0004	3.204	3.204	0.0000	0.0000	0.0000	0.0000
NBP[0]608*	lb/hr	0.0002	2.635	2.635	0.0000	0.0000	0.0000	0.0000
NBP[0]631*	lb/hr	0.0001	2.093	2.093	0.0000	0.0000	0.0000	0.0000
NBP[0]659*	lb/hr	0.0000	1.457	1.457	0.0000	0.0000	0.0000	0.0000
NBP[0]683*	lb/hr	0.0000	1.293	1.293	0.0000	0.0000	0.0000	0.0000
NBP[0]710*	lb/hr	0.0000	0.7728	0.7728	0.0000	0.0000	0.0000	0.0000
NBP[0]736*	lb/hr	0.0000	0.7100	0.7100	0.0000	0.0000	0.0000	0.0000
NBP[0]761*	lb/hr	0.0000	0.6194	0.6194	0.0000	0.0000	0.0000	0.0000
NBP[0]787*	lb/hr	0.0000	0.4951	0.4951	0.0000	0.0000	0.0000	0.0000
NBP[0]821*	lb/hr	0.0000	0.7160	0.7160	0.0000	0.0000	0.0000	0.0000

Stream Number		15	16	17	18	19	20	21
Description	Parameter Units	Reflux	LC Draw	LC Mid Cool	To TK-02	MC Draw	MC Pumped	MC Mid Cool
Vapor Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Pressure	Psia	68.72	68.72	64.72	60.72	21.15	131.8	127.8
Temperature	°F	121.0	121.0	107.0	100.0	493.2	493.2	107.0
Molar Enthalpy	Btu/lbmol	-115.7	-115.7	-116.6	-117.0	-134.0	-134.0	-179.2
Density (Actual)	lb/ft3	44.86	44.86	45.26	45.45	37.42	37.52	48.48
Molecular Weight		127.3	127.3	127.3	127.3	196.6	196.6	196.6
Total Molar Flow Rate	lbmol/hr	546.5	109.3	109.3	109.3	114.5	114.5	114.5
Volumetric Flow Rate (Actual)	bbl/day	6628	1324	1314	1308	2570	2563	1984
Total Mass Flow Rate	lb/hr	69.55	13.91	13.91	13.91	22.51	22.51	22.51
Component Mass Flowrates								
Water	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Nitrogen	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Hydrogen	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methane	lb/hr	0.0011	0.0002	0.0002	0.0002	0.0000	0.0000	0.0000
Ethane	lb/hr	0.0874	0.0175	0.0175	0.0175	0.0000	0.0000	0.0000
Ethylene	lb/hr	0.0288	0.0058	0.0058	0.0058	0.0000	0.0000	0.0000
Propane	lb/hr	0.5616	0.1123	0.1123	0.1123	0.0000	0.0000	0.0000
Propene	lb/hr	0.4702	0.0940	0.0940	0.0940	0.0000	0.0000	0.0000
n-Butane	lb/hr	0.8809	0.1762	0.1762	0.1762	0.0000	0.0000	0.0000
13-Butadiene	lb/hr	0.1684	0.0337	0.0337	0.0337	0.0000	0.0000	0.0000
n-Pentane	lb/hr	0.3349	0.0670	0.0670	0.0670	0.0000	0.0000	0.0000
n-Hexane	lb/hr	1.0434	0.2087	0.2087	0.2087	0.0000	0.0000	0.0000
NBP[0]171*	lb/hr	0.5013	0.1003	0.1003	0.1003	0.0000	0.0000	0.0000
NBP[0]202*	lb/hr	0.6279	0.1256	0.1256	0.1256	0.0000	0.0000	0.0000
NBP[0]226*	lb/hr	0.6973	0.1395	0.1395	0.1395	0.0000	0.0000	0.0000
NBP[0]246*	lb/hr	1.2343	0.2469	0.2469	0.2469	0.0000	0.0000	0.0000
NBP[0]277*	lb/hr	1.9521	0.3904	0.3904	0.3904	0.0000	0.0000	0.0000
NBP[0]302*	lb/hr	6.8158	1.363	1.363	1.363	0.0000	0.0000	0.0000
NBP[0]325*	lb/hr	13.47	2.693	2.693	2.693	0.0006	0.0006	0.0006
NBP[0]352*	lb/hr	13.78	2.756	2.756	2.756	0.0108	0.0108	0.0108
NBP[0]375*	lb/hr	18.42	3.684	3.684	3.684	0.2259	0.2259	0.2259
NBP[0]400*	lb/hr	7.994	1.599	1.599	1.599	2.379	2.379	2.379
NBP[0]426*	lb/hr	0.4583	0.0917	0.0917	0.0917	3.408	3.408	3.408
NBP[0]452*	lb/hr	0.0279	0.0056	0.0056	0.0056	3.092	3.092	3.092
NBP[0]479*	lb/hr	0.0032	0.0006	0.0006	0.0006	3.009	3.009	3.009
NBP[0]504*	lb/hr	0.0005	0.0001	0.0001	0.0001	3.017	3.017	3.017
NBP[0]530*	lb/hr	0.0001	0.0000	0.0000	0.0000	2.696	2.696	2.696
NBP[0]555*	lb/hr	0.0000	0.0000	0.0000	0.0000	1.975	1.975	1.975
NBP[0]580*	lb/hr	0.0000	0.0000	0.0000	0.0000	1.127	1.127	1.127
NBP[0]608*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.5958	0.5958	0.5958
NBP[0]631*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.3431	0.3431	0.3431
NBP[0]659*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.1801	0.1801	0.1801
NBP[0]683*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.1386	0.1386	0.1386
NBP[0]710*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0758	0.0758	0.0758
NBP[0]736*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0666	0.0666	0.0666
NBP[0]761*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0567	0.0567	0.0567
NBP[0]787*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0447	0.0447	0.0447
NBP[0]821*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0639	0.0639	0.0639

Stream Number		22	23	24	25	26	27	28
Description	Parameter Units	To Tk-03	T-04 bttms	T-04 Boil Up	HC Draw	To P-06	HC Inter	To Tk-04
Vapor Fraction		0.0000	0.0518	1.000	0.0000	0.0000	0.0000	0.0000
Pressure	Psia	123.8	21.00	23.00	21.00	116.7	112.7	108.7
Temperature	°F	100.0	564.7	620.4	564.7	564.7	107.0	100.0
Molar Enthalpy	Btu/lbmol	-179.9	-148.0	-116.4	-166.5	-166.5	-239.8	-240.6
Density (Actual)	lb/ft^3	48.65	36.88	0.5063	38.06	38.13	50.59	50.77
Molecular Weight		196.6	233.3	233.3	264.8	264.8	264.8	264.8
Total Molar Flow Rate	lbmol/hr	114.5	901.7	901.7	59.88	59.88	59.88	59.88
Volumetric Flow Rate (Actual)	bbl/day	1977	119700	1776000	1781	1778	1340	1335
Total Mass Flow Rate	lb/hr	22.51	210.4	210.4	15.85	15.85	15.85	15.85
Component Mass Flowrates								
Water	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Nitrogen	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Hydrogen	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methane	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ethane	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ethylene	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Propane	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Propene	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
n-Butane	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
13-Butadiene	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
n-Pentane	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
n-Hexane	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]171*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]202*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]226*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]246*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]277*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]302*	lb/hr	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]325*	lb/hr	0.0006	0.0002	0.0002	0.0000	0.0000	0.0000	0.0000
NBP[0]352*	lb/hr	0.0108	0.0047	0.0047	0.0000	0.0000	0.0000	0.0000
NBP[0]375*	lb/hr	0.2259	0.1460	0.1460	0.0017	0.0017	0.0017	0.0017
NBP[0]400*	lb/hr	2.379	2.397	2.397	0.0336	0.0336	0.0336	0.0336
NBP[0]426*	lb/hr	3.408	5.505	5.505	0.0945	0.0945	0.0945	0.0945
NBP[0]452*	lb/hr	3.092	8.086	8.086	0.1711	0.1711	0.1711	0.1711
NBP[0]479*	lb/hr	3.009	12.99	12.99	0.3426	0.3426	0.3426	0.3426
NBP[0]504*	lb/hr	3.017	21.33	21.33	0.7024	0.7024	0.7024	0.7024
NBP[0]530*	lb/hr	2.696	31.15	31.15	1.290	1.290	1.290	1.290
NBP[0]555*	lb/hr	1.975	36.50	36.50	1.913	1.913	1.913	1.913
NBP[0]580*	lb/hr	1.127	31.31	31.31	2.077	2.077	2.077	2.077
NBP[0]608*	lb/hr	0.5958	23.45	23.45	2.040	2.040	2.040	2.040
NBP[0]631*	lb/hr	0.3431	15.94	15.94	1.750	1.750	1.750	1.750
NBP[0]659*	lb/hr	0.1801	8.633	8.633	1.277	1.277	1.277	1.277
NBP[0]683*	lb/hr	0.1386	6.004	6.004	1.154	1.154	1.154	1.154
NBP[0]710*	lb/hr	0.0758	2.667	2.667	0.6970	0.6970	0.6970	0.6970
NBP[0]736*	lb/hr	0.0666	1.824	1.824	0.6434	0.6434	0.6434	0.6434
NBP[0]761*	lb/hr	0.0567	1.171	1.171	0.5627	0.5627	0.5627	0.5627
NBP[0]787*	lb/hr	0.0447	0.6750	0.6750	0.4504	0.4504	0.4504	0.4504
NBP[0]821*	lb/hr	0.0639	0.6299	0.6299	0.6521	0.6521	0.6521	0.6521

Sized Equipment List

Each piece of equipment was sized according to the requirements from the simulated process. Additional details regarding the simulated process can be found within Appendix B. Further descriptions as to equipment sizing, with equipment information, can be found in the following sized equipment tables.

Tanks and vessels were sized according to the required amount of volume that would be required or produced for one week. Therefore, the feed tank was sized to accommodate 7 days of accumulated feed material, and the product tanks were sized to hold 7 days of accumulated product material. Upon determination of required volume, the height and diameter of the vessel was optimized according to structural stability and costing.[2] The material of the components was determined by the composition as it enters the vessel. The heavy metals in the feed stock could be corrosive for carbon steel, so a stainless-steel clad carbon steel material was utilized to mitigate equipment damage and potential leakage pre-adsorption tower.

Table 2: Tanks and Vessel Sizing

	TK-01	TK-02	TK-03	TK-04	V-01
Description	Pyoil Feed Tank	Pyoil Light Cut Tank	Pyoil Medium Cut Tank	Pyoil Bottoms Tank	PyGas Overhead Drum
Height (ft)	70	40	40	40	10.5
Diameter (ft)	59	41	50	41	3.7
Orientation	Vertical	Vertical	Vertical	Vertical	Horizontal
Pressure (psig)	55	125	125	125	100
Temperature (°F)	150	150	150	150	320
MOC	SS clad CS	CS	CS	CS	CS

The towers required extensive detailed equipment sizing for both the external shell, and internal trays. First, the adsorption towers (T-01 and T-02) were designed to accommodate the adsorbent volume and required feed volume. The decision to stack two different types of adsorbents within each adsorbing column was made to conserve space in the processing site, minimizing capital costs. Once the volume was determined, the height and diameter were optimized with various tower heuristic such as, L/D and below a certain height to withstand high wind speeds, to create a strong tower structure[2].

The distillation towers were designed in Aspen HYSYS, the process simulation detailed in Appendix B. A two-column distillation design (T-03 and T-04) was created to optimize the Light and Medium Cuts. The current distillation sizing was optimized by iterating different variables, such as temperature, tray number, column diameter, feed tray, and others, to get the highest usable product yield, with the smallest equipment costs. Upon the finalized simulation report, heuristics were utilized to further determine the design sizing for the towers.[3] Final tower sizing results are detailed in Table 3.

Table 3: Tower Sizing

	T-01	T-02	T-03	T-04
Description	Pyoil Adsorption Tower	Pyoil Adsorption Tower	Pyoil Initial Distillation Tower	Pyoil Primary Distillation Tower
Diameter (ft)	10	10	2.75	9.25
Height (ft)	50	50	15	58
Pressure (psig)	125	125	75	73
Temperature (°F)	175	175	-	-
Number Trays	4	4	2	19
Type Trays	Support & Distribution	Support & Distribution	Bubble Cap	Sieve
MOC Tower	SS Clad CS	SS Clad CS	SS Clad CS	SS Clad CS
MOC Trays	SS	SS	SS	SS

Pump sizing was determined from the amount of pressure drop and gain while moving through each piece of equipment. Pumps were sized according to the pressure difference between the suction pressure and the discharge pressure to determine the size of pump itself and the size of the driving motor. Heuristics were then applied to the calculated operating size.[4] Each pump is detailed in Table 4.

Table 4: Pump Sizing

	P-01 A/B	P-02	P-03 A/B	P-04 A/B	P-05 A/B	P-06 A/B
Description	Pyoil Feed Pumps	Pyoil Distillation Start-Up Pump	Pyoil Distillation Feed Pumps	Pyoil Light Cut Pumps	Medium Cut Draw Pumps	Heavy Cut Pumps
Flow (gpm)	136	136	150	233	75	55
Type-Drive	Electric Motor	Electric Motor	Electric Motor	Electric Motor	Electric Motor	Electric Motor
Temperature (°F)	150	160	318	171	543	615
Discharge Pressure (psig)	174	88	88	105	167	152
ΔP (psi)	118	18	36	45	95	101
Shaft Power (kW)	15	5	5	10	10	5
MOC	SS	SS	SS	CS	CS	CS

Heat exchangers were designed according to the Logarithmic Mean Temperature Difference (LMTD) method.[3] The first heat exchanger was designed to cool a maximum of 150 °F material down to 107 °F, as required for the distillation columns. This was oversized, as the anticipated temperature entering the exchanger is only 110 °F, but the control scheme is designed to limit the amount of cooling to a minimum in an effort to minimize utility usage. The reboilers were sized from the information found in Appendix B, the HYSYS simulation. The smallest reboiler is a kettle reboiler while the large reboiler is a thermosyphon. All the other heat exchangers were typical shell and tube heat exchangers, Finally, the product coolers were designed to remove heat from the product streams, in accordance with the specifications as found in Table 1.

Table 5: Heat Exchanger Sizing

	E-01	E-02	E-03	E-04A	E-04B	E-05A	E-05B	E-06	E-07A	E-07B
Duty (MMBtu/hr)	1.3	4.67	22.2	0.101	0.049	5.19	0.076	28	5.05	0.0531
Shell										
Temp. (°F)	200	420	380	172	160	545	160	671	671	160
Pres. (psig)	88	60	55	105	101	106	102	58	97	93
Phase	L	L-V	L-V	L	L	L	L	L-V	L	L
MOC	CS	CS	CS	CS	CS	CS	CS	CS	CS	CS
Tube										
Temp. (°F)	150	800	160	157	144	160	150	800	160	150
Pres. (psig)	70	230	70	70	70	70	70	230	70	70
Phase	L	L	L	L	L	L	L	L	L	L
MOC	CS	CS	CS	CS	CS	CS	CS	CS	CS	CS

Economics

All economic estimates were calculated from the bare module cost estimating equation, as found in Appendix A. A service factor of 97% was used for each variable and utility cost.[2] Overall, a total capital cost estimate of $13.04MM was calculated. A total operating cost of $908.5M was determined with a variable cost of $233M (~26%) and a fixed operating cost of $675.4M (~74%) of the total operating cost.

Capital Cost Estimate

The capital costs were calculated utilizing the bare module cost method.[3] This method accounts for the physical equipment and installation costs for each piece of equipment.[3] Each equation had a factor in place to estimate pressure factors, material considerations, and overall size. Additionally, a Chemical Engineering Plant Cost Index (CEPCI) ratio was utilized to scale up each capital cost to 2021 rates.[5] All values were found during the simulated process unit or during sizing calculations. The total capital cost was $13.04MM, and it was broken down in the pie chart in Figure 3, and in Table 6.

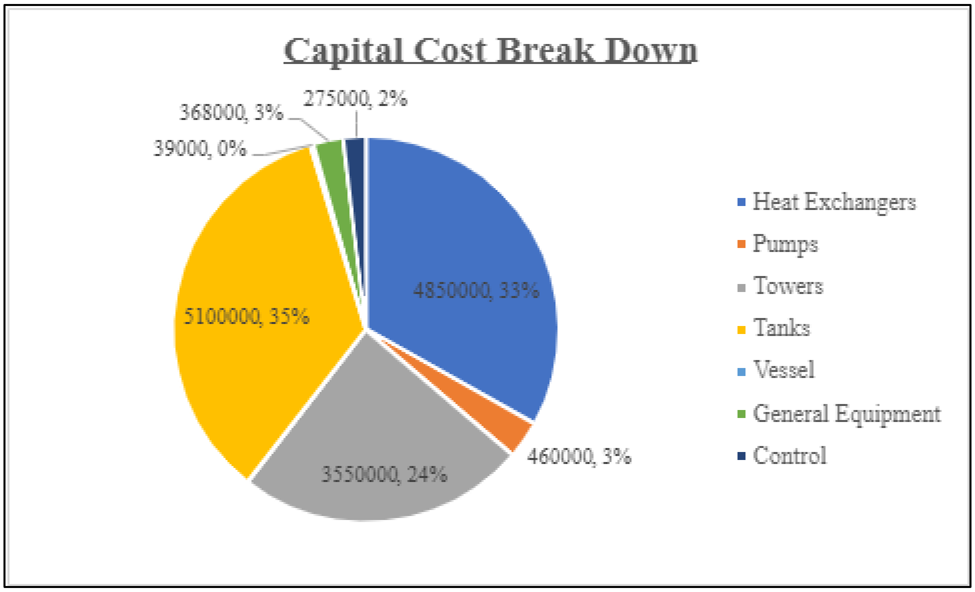


Figure 3 Capital Cost Break Down

Table 6: Summary of Capital Costs

Total Equipment Capital Costs	
Equipment	**Cost**
Heat Exchangers	$4.85 MM
Pumps	$460 M
Towers	$3.55 MM
Tanks	$5.10 MM
Vessel	$39 M
General Equipment	$368 M
Control	$275 M
Total	$13.04MM

Heat exchangers were priced using an increased fouling factor to account for optimized performance after a layer of fouling has occurred. These could be upsized in the future for a higher capital cost, but a lower risk to prevent sending hot liquid to tankage. The most expensive exchangers were the primary product cooler for the Light and Medium Cuts. While these were expensive, the heat exchangers could not be adjusted as the final temperature is a requirement to meet product specifications.

Table 7: Heat Exchanger Capital Cost

Capital Cost - HEX	
Exchanger Number	**Cost**
E-01	$512 M
E-02	$119.3 M
E-03	$367.2 M
E-04A	$1.14 MM
E-04B	$652.2 M
E-05A	$191.8 M
E-05B	$107.9 M
E-06	$327 M
E-07 A	$1.16 MM
E-07 B	$268 M

Pump capital cost was found by utilizing the pressure increase across the pump. Carbon steel was chosen as all heavy metal corrosion components should be adsorbed in the adsorption towers. Larger pumps have a larger cost; therefore, the feed pumps had the highest capital cost compared to the smaller product stream pumps. Each pump was spared with an identical sized sparing pump, with the only exception being the primary distillation tower feed pump (P-02). Pump P-02 is only intended for startup use and will not be used during daily operations, therefore it is not economical to spare. Sparing pumps allow for cheaper management of pump maintenance than shutting down a section or full unit, for mechanical issues involving pumps. If capital costs needed to be cut, it is suggested that spare pumps are downsized to run 75% - 90% of the desired volumetric flow.

Table 8: Pump Capital Cost

Capital Cost - Pumps	
Pump Number	**Cost**
P-01A	$74.7 M
P-01B	$74.7 M
P-02	$31.6 M
P-03A	$31.6 M
P-03B	$31.6 M
P-04A	$47.4 M
P-04B	$47.4 M
P-05A	$41.8 M
P-05B	$41.8 M
P-06A	$33.5 M

Tower costing has been further explained in the appendices. The adsorption towers are further costed in Appendix A, while the distillation towers are further costed in Appendix B. Each tower capital costs includes the cost of column internals, extra height for vapor disengagement and liquid level, and column skirts.[3]

Table 9: Tower Capital Cost

Capital Cost - Towers	
Tower Number	**Cost**
T-01	$1.06 MM
T-02	$1.06 MM
T-03	$162 M
T-04	$1.27 MM

Tank capital cost was found using the desired 7 day holding volume, and the anticipated pressure and temperatures within the tank. Larger tanks were more expensive than the smaller tanks as more material and structural components are desired. Therefore, the feed tank was the most expensive as it holds the highest volume and is made of stainless-steel clad carbon steel. Since the feed has not yet been treated in the adsorber at this stage, it is crucial that stainless steel is used to prevent significant corrosion from the potential water and heavy metals in the feed material. All product tanks are carbon steel as there should not be any contaminants after separation.

Vessel costing was performed similarly to the tank costing.[2] The condensing vessel required was relatively small and inexpensive, but still required to hold a liquid head above the reflux pump.

Table 10: Tank & Vessel Capital Cost

Capital Cost - Tanks/Vessels	
Tank Number	**Cost**
TK-01	$2.32 MM
TK-02	$796 M
TK-03	$998 M
TK-04	$985.3 M
V-01	$39.3 M

General equipment capital costs and sensor costs were assumed by utilizing the lengths of piping, number of control valves, and number of relief valves.[3] Pressure and required size were taken into consideration.

These general equipment and sensor capital costs are only 5% of the overall capital costs as shown in the pie chart shown in Figure 3.

Table 11: Capital Cost for Various Equipment

Capital Cost - Other	
General Equipment	**Cost**
Piping	$218 M
Control Valves	$50 M
Relief Devices	$100 M
Controls	**Cost**
Sensors	$275M

Variable Cost Estimate

Annual variable costs were associated with the utility costs of cooling water for heat exchangers and electric power required of pumps. Hot oil was assumed to be provided as a utility at no cost, due to heat integration in another process within the Global Petrochemical Bali plant. To perform a full economic analysis in the future, the cost associated with hot oil should be found and factored into the final project economics. Cooling water for the feed heat exchangers and product heat exchangers was found using a 97% service factor, implying the plant was functional and using utilities for 97% of the year. Cooling water was priced utilizing the given unit price (per million BTU) as was the electricity price (per kW-hr). At large, the annual operating costs were only 7% of the total capital costs associated with a new unit. These yearly costs may be minimized by adding controls onto slightly upsized heat exchangers, minimizing the amount of cooling water used, while still cooling down the feed steam sufficiently.

Table 12: Variable Operating Cost

VARIABLE OPERATING COST		
Utility	**Equipment**	**Cost**
Cooling Water (Unit Price 0.5 $/MMBtu)	E-01	$5.6M
	E-03	$94.4M
	E-04A	$430
	E-04B	$210
	E-05A	$22.1M
	E-05B	$325

	E-07A	$21.5M
	E-07B	$230
	TOTAL	$144.5M
Electric (Unit Price 0.25 $/kW-hr)	P-01	$28.0M
	P-02	$9.3M
	P-03	$9.3M
	P-04	$18.6M
	P-05	$14.0M
	P-06	$9.3M
	TOTAL	$88.5M
OVERAL TOTAL		$233M

In addition to pricing the correlated utility requirements, the annual utility usage was also calculated to ensure that the current site would have the information needed to potentially upsize the utility systems prior to starting up the new unit. Again, a service factor of 97% was used. In total, it is highly recommended that an additional utility unit be constructed or upsizing the current utility units to accommodate the additional load.

Table 13: Utility Usage per Equipment

UTILITY USAGE		
Utility	**Equipment**	**Annual Consumption**
Cooling Water	E-01	11,046 MMBtu
	E-03	188,700 MMBtu
	E-04A	858.2 MMBtu
	E-04B	413 MMBtu
	E-05A	44,100 MMBtu
	E-05B	647.5 MMBtu
	E-07A	42,900 MMBtu
	E-07B	451.2 MMBtu

	TOTAL	566,800 MMBtu
Electric	P-01	13.2 kW
	P-02	4.39 kW
	P-03	4.39 kW
	P-04	8.77 kW
	P-05	6.58 kW
	P-06	4.39 kW
	TOTAL	41.7 kW
Hot Oil	E-02	39.7 MMBtu
	E-06	238,000 MMBtu
	TOTAL	277,600 MMBtu

Fixed Cost Estimate

Fixed costing was performed considering two main aspects: labor cost and maintenance cost. Labor cost was calculated to consider operational personnel on site to monitor and maintain the new Pyoil Purification Process unit during operational hours. The labor cost was created assuming a $60 dollar/hour wage for the plant operators working 3 12-hour shifts a week.[3] This was estimated based on previous experience working with refinery operators. This was found to be $675.4M annually, but this amount could potentially be adjusted from the site's current operations staff and the variance in regional competitive pay for this position. If this new process unit is added onto an existing unit, then existing operators could be given additional responsibilities for the new equipment. Unfortunately, this would come with its own safety hazards as fatigued operators cannot perform at their highest level of insight, knowledge, and ethical decision making, which is vital when working at petrochemical sites. Maintenance cost was calculated using a heuristic of 6% of the annual operating costs (sum of labor cost and variable operating cost estimates).[3] While this number could be lower if preventative measures are minimized, setting a larger amount for the annual maintenance cost will lessen the risk of experiencing a significant shutdown (that can often be associated with a lack of maintenance). Initially expecting to include replaced adsorbent beads in the fixed operating costs, it was assumed that the adsorbent could be completely regenerated through a hot nitrogen processing so no extra costing was considered in adsorbent replacement. Finally, further contact with the adsorbent manufacturer should be utilized to estimate this cost prior to finalizing the project economics.

Table 14: Fixed Operating Cost

FIXED OPERATING COST	
Labor Cost	$624 M

Maintenance Cost	$51.4 M
TOTAL	$675.4M

Process Safety

Safety was a key principle during development of the preliminary design for the Pyoil Purification Unit. Analysis was performed for the environmental impacts, inherent safety techniques, pressure relief, and other facets of process safety to mitigate risks during the preliminary design. It is further advised that a more in-depth process hazard analysis should be fully developed by subject matter experts alongside the detailed designing of the process plant. In terms of risk management, the adsorption towers must be working and adsorbing the heavy metals out of the system prior to the distillation columns, or it will result in a very dangerous scenario. Contaminants from the feed stream should be fully adsorbed into the adsorption columns without leaking harmful contaminants into downstream processes. It would significantly affect the mechanical integrity of the downstream (carbon steel) equipment, and potentially contaminate both the steam cracking units as well as the bottoms material. Additionally, the adsorption unit also needs to provide a continuous distillation feed stream. This will be accomplished with a fully developed standard operating procedure, and a well understood lock out tag out process. Overall, the process safety of this unit has been taken into consideration but should be evaluated by experts in parallel with the detailed design process and project economic analysis.

Environmental Impacts

For the Pyoil Purification Unit, the lighter components are most volatile in comparison to other components within the process. Therefore, the amount of each component can be simulated and calculated for the distillate product to determine how much of each component would be released if there was a potential release of material.

In normal operation, the flow rate for the Py Gas stream is 162.4 lb/hr. The following components account for 98.5% of the Py Gas stream: methane[6], ethane[7], ethylene[8], propane[9], propene[10], butane[11], 1,3-butadiene[12], n-pentane[13], and n-hexane.[14] These components are highly flammable and/or explosive. Pentane and hexane may cause narcotic effects, irritation of the eye and throat, and toxic when inhaled. When released, hexane causes toxic effects to the surrounding aquatic life with long lasting effects. Therefore, the valve on the vapor outlet of the initial distillation column and the bottoms product fails to open to the primary distillation column. It will vent to flare. This prevents overpressurizing in the initial distillation column.

The outlet vapor of the primary distillation column must be considered as well. The mass flow of the distillate stream was 83.5 Mlb/hr. This stream will also contain light components that are volatile. The components in the stream are the same as the Py Gas stream, accounting for 5.14% of the mass flow. In addition, the stream also contains pseudo components NBP 171 - NBP 452 to simulate the stream such as heptane[15], octane[16], nonane[17], pyrene[18], undecane[19], dodecane[20], and tridecane[21] with similar boiling points to the listed normal boiling points in the distillate stream. The pseudo components with the greatest amount of flow were NBP 171 - NBP 400 which accounted for 91.5% of the stream. The components are highly flammable and toxic to aquatic environments with long lasting effects. Therefore, a relief valve was sized to accommodate the vapor flow out of the tower in the event of a fire to prevent loss of containment. The relief device will send flammable gas to be fully combusted to protect the environment and the system. A rupture disk was added in parallel to the relief valve to handle larger upsets in the system. The components in the stream do not have halogenated hydrocarbons in it. However, if there were to be any materials in the substance that are halogenated hydrocarbons, it will be less than the Best Available Control Technology (BACT) for releases which are 250 ton/year of uncontrolled release.[22] Using Good Engineering Practice (GEP), the column does minimize the amount of uncontrolled release by using a relief valve and sending the gaseous vapors to the flare instead of releasing the flammable gases to the atmosphere.

Although the Heavy Cut components are not as volatile as the lighter components in this process, the Heavy Cut environmental concerns must be considered. The Heavy Cut components consist of large hydrocarbon chains such as C11 - C29. The combination of this organic material and remaining Pyoil released into the environment will poison the soil, pollute the groundwater, and harm the nearby plants and animals if not processed correctly. Within this mixture, there is a potential for residual heavy metals if the purification is not working properly. The heavy metals will be in the Heavy Cut and contribute to the toxicity level if there was an environmental exposure. Therefore, the method best used to get rid of the Heavy Cut would be by using the process called Stabilization/Solidification.[23] This is performed by sealing in the waste using a binder. Stabilization/Solidification prevents heavy cuts from leaching into the environment. The cement will absorb the contaminates thereby encapsulating the contaminants within the heavy components. This allows for the Heavy Cut to be used in a large quantity while not polluting the environment. There are additional options to consider such as first using a solvent such as toluene to separate any residual oil and/or in combination with a centrifuge before using a binder.[23] Both solvent and centrifuge are expensive options, however, they both help with oil recovery that can be recycled back into the feed stream to be further processed. There are many ways to dispose of the Heavy Cut, however, each has its own disadvantages.

Inherent Safety

Specific health hazards for the major components found in the streams are listed in Table 15. This includes the OSHA Personnel Exposure Limit Concentration, the NFPA Diamond Classification, and the lethal concentration statistics.

Table 15: Health Hazards

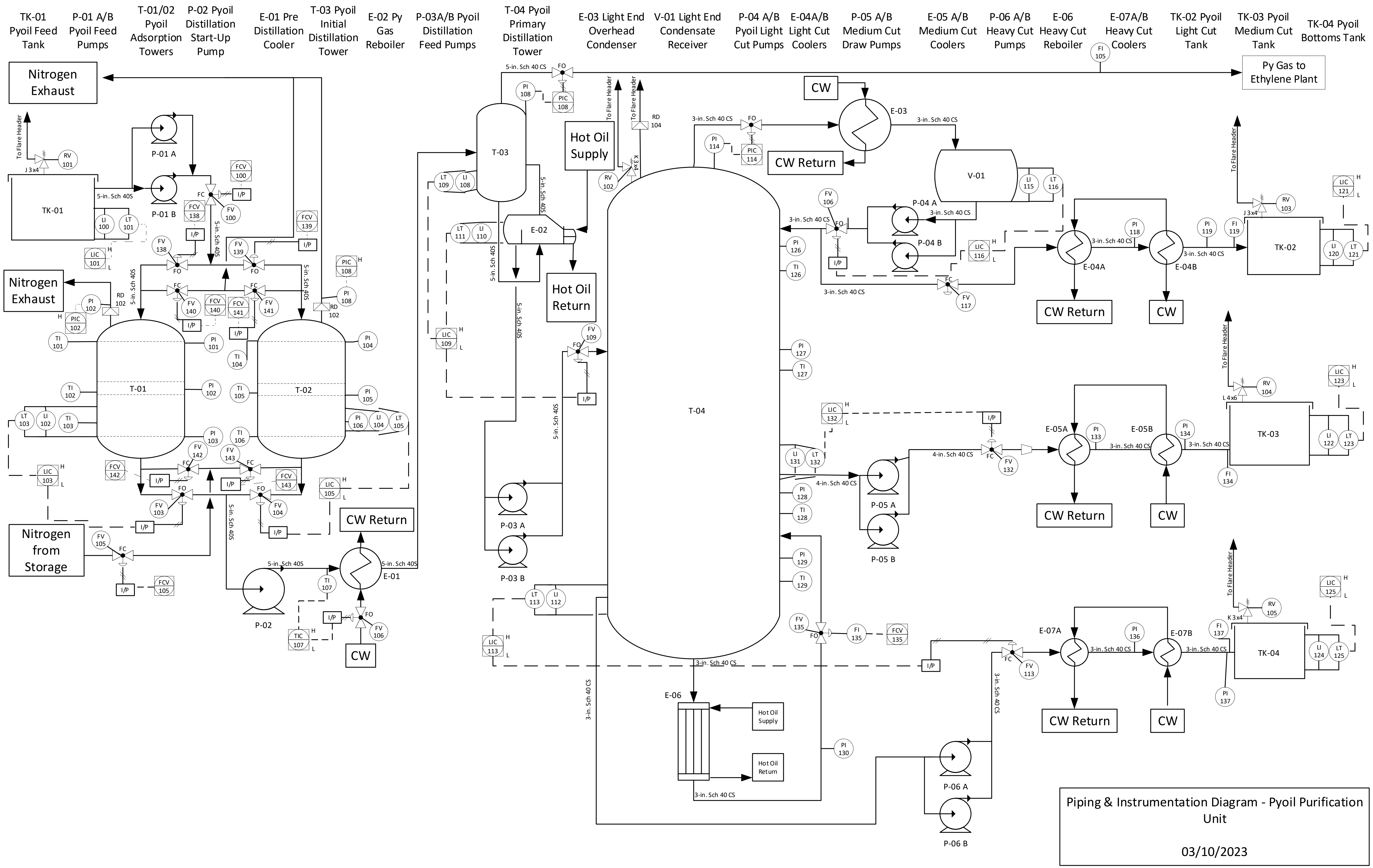
Chemicals	OSHA PEL Concentration	NFPA Diamond Classification	Lethal Concentration (LC50)
Methane[6]	1000 ppm[24]	Health - 2 Fire - 4 Instability - 0	Inhalation/mouse3 26 gm/m^3 (2 hours)
Ethane[25]	1000 ppm	Health - 1 Fire - 4 Instability - 0	Inhalation/mouse 658 mg/l (4 hours)
Ethylene[24]	-No occupational limits -Decreases the amount of available oxygen	Health - 2 Fire - 4 Instability - 2	-No occupational limits -Decreases the amount of available oxygen

Propane[9]	1000 ppm	Health - 2 Fire - 4 Instability - 0	>800000 ppm (15 min)
Propene[10]	500 ppm	Health - 1 Fire - 4 Instability - 1	>65000 ppm (4 hour)
Butane[11]	800 ppm	Health - 1 Fire - 4 Instability - 0	Rat 658,000 mg/m^3[26]
1,3-Butadiene[12]	5 ppm	Health - 2 Fire - 4 Instability – 2 Special – no water, will react violently	Rat 285,000 mg/m^3[27] (4 hours)
Pentane[13]	1000 ppm	Health - 1 Fire - 4 Instability - 0	Rat 364,000 mg/m^3[28] (4 hours)
Hexane[29]	500 ppm	Health - 0 Fire - 3 Instability - 0	Rat 77000 ppm (1 hour)
Heptane[30]**	500 ppm	Health - 1 Fire - 3 Instability - 0	Rat 103,000 mg/m^3/4h (4 hours)
n-Octane[31]**	500 ppm	Health - 1 Fire - 3 Instability - 0	Rat 118,000 mg/m3/4hr
n-Nonane[32]**	200 ppm (TWA)	Health - 1 Fire - 3 Instability - 0	Rat 3,200 ppm/4h
Pyrene[33]**	0.2 mg/m^3	Health - 2 Fire - 1 Instability - 0	Rat 170 mg/m^3[34]
N-undecane[35]**	200 ppm (TWA)	Health - 1 Fire - 2 Instability – 0	Rat > 442 ppm/8H
Dodecane[36]**	20mL intermittent 96 hours	Health - 1 Fire - 2 Instability - 0	Rat > 142 ppm/8H
Tri-decane**	0.0073 ppm[37]	Health - 0 Fire - 2 Instability - 0	Rat > 41 ppm/8H[38]

Estimated using pseudo components

Piping and Instrumentation Diagram

. Figure4 5: Piping and Instrumentation Diagram for Pyoil Purification Unit



Control systems are used in petrochemical processing units to automate the process while minimizing human error. These systems also maintain optimized product rates that are within specifications. In terms of process safety, controls are used to maintain the process in operable conditions without putting the unit or personnel at risk when working with volatile components, high pressures, high temperatures, or other hazardous conditions. In the preliminary Pyoil Purification Unit, control schemes were designed in order to control major equipment and streams throughout the processing unit. The completed piping and instrumentation diagram (P&ID) is shown in Figure 5.

All tanks, vessels, and towers in the process received a level indicator to maintain a safe and operable level within each type of process vessels. A high and low alarm was installed to warn of any risk of high and low liquid levels within the vessels.

The process feed and product start or end at their designated tank before exiting the unit. The process will be controlled through flow controllers that will be using gate valves to move feed through the process. Flow indicators and control valves can be found in the feed tank and adsorption process where proper liquid flow control is necessary to control the filtration rate and the production rate for the remaining process. The flow control valves manipulate the liquid flow rate through equipment, which in the end determines the rate of production of the process. Flow controllers were placed on leading and discharging lines from major process equipment (vessels, pumps, distillation column) to fully process leading and lagging control on the flow, these flow rates can also be used to maintain a mass balance of the unit when operating.

The Pyoil Purification Unit contains hydrocarbon compositions and lighter gases that require certain boiling temperatures and pressures that are established to produce the Pyoil Gas, Light Cut, Medium Cut, and Heavy Cuts products. Pressure and temperature indicators were placed throughout the adsorption columns (T-01 and T-02) to monitor the exothermic process that occurs as the feed product passes through the adsorption column, monitoring for large temperature increases or overpressure scenarios. This is also utilized in the distillation column to monitor the separation that is occurring in the column.

Pressure controllers were also used to control the Pyoil Gas flow rate leaving the distillation column (T-04). Automated pressure controls account for adjustments in the pressure of the column. The automation also reduced the chance for the column to overpressure. Overpressure often occurs from various incidents. For this reason, pressure indicators are often used to signal an overpressure incident in conjunction with rupture disks. In the P&ID, the pressure controller was used to signal a high-pressure alarm when the relief device or rupture disk had lifted or burst.

Pressure controllers are also utilized throughout the light, medium, and Heavy Cut heat exchangers to monitor that the products are coming in as per the specific requirements from GP. Pressure and temperature have a proportional correlation, so pressure control was utilized due to the higher precision within pressure sensors.

Upon review of the preliminary design, a detailed design should be produced with additional controls to continue automation, maintain safety regulations per GP's standards, and optimize usability for operations staff. High and low alarms should also be utilized as per the standard operating procedures and the hazard and operability study to complete the instrumentation and controls of the Pyoil Purification Unit.

The piping was designed according to the volumetric flow rate of the streams (Q). Using the Q value obtained from the simulation report, the diameter of piping was determined with the fluid velocity of 1 meter/second.[3] This is a common heuristic for petrochemical processes. The pricing per foot of pipe was then established using Equation 1, which was used to determine the capital cost of piping material alone.

The choice of using Schedule 40 Steel before the distillation column and Schedule 40 Carbon steel after was due to the corrosive nature of the feed stream prior to distillation. Schedule 40 is a common piping material for petrochemical processes and was verified for components of the Pyoil.

[Equation 1][3]

$$PC_{pipe}\left[\frac{\$}{ft\ of\ pipe}\right] = 10 * d_{pipe}[inch] + 2 * d_{pipe}^{1.4}[inch]$$

Pressure Relief Valve Sizing

In an event of overpressure, it is crucial to have a relief device in place to prevent a loss of containment. Therefore, due to the large flow rate going into T-03 and T-04 a relief device was calculated to accommodate the largest amount of mass flow that the column could potentially see. The relief device must be sized to the worst-case scenario.[39] Therefore, the relief valve was designed to accommodate a fire. The total tower liquid height surpassed 15 ft. Therefore, using the heuristic that the fire only reaches 25 ft high the tower's wetted area could be calculated.[39] The wetted area (A_w) was calculated using the circumference of the tower at a height of 15 ft assuming a 10 ft skirt.[39] The wetted area accounts for the piping in contact with the fire as well as the thermosiphon's head and bottom where the fire can contact the process fluid and vaporize it. Assuming the worst-case scenario, there would be no insulation; the correction factor (F) is then 1.[39] In Equation 2, the heat duty for the fire ($\dot{Q_{fire}}$) as the worst-case scenario can be found.

[Equation 2][39]2

$$\dot{Q_{fire}} = 34{,}500 \cdot F \cdot A_w^{0.82}$$

Using Equation 3, the mass flow rate relieved ($\dot{m_{relief}}$) was found using $\dot{Q_{fire}}$, density of the liquid (ρ_L), density of the vapor (ρ_v), and heat of vaporization (λ).[39]

[Equation 3][39]

$$\dot{m_{relief}} \geq \frac{\dot{Q_{fire}}}{\lambda} \cdot \left(1 - \frac{\rho_v}{\rho_L}\right)$$

The Maximum Allowable Working Pressure (MAWP) can be found by using the heuristic to add 50 psig to the operating pressure to calculate the design pressure.[3] Then, the overpressure in psia (P_o) can be calculated using Equation 4, and the choked flow (P) can be found using the heat capacity ratio (γ) in Equation 5.[39]

[Equation 4] [39]

$$P_o = 1.1 \cdot MAWP$$

[Equation 5] [39]

$$P = P_o \cdot \left(\frac{2}{\gamma+1}\right)^{\frac{(\gamma+1)}{\gamma}}$$

Then, using a backpressure factor (K_b) is 1, C_o as 0.975 for a conventional relief valve, gravitational constant (g_c), molecular weight (MW), universal gas constant (R), operating temperature (T_o), compressibility factor (z), and heat capacity ratio (γ) in Equation 6 to find the orifice area (A).[39]

[Equation 6] [39]

$$A = \frac{\dot{m}}{C_o \cdot K_b \cdot P_o} / \sqrt{\frac{\gamma \cdot g_c \cdot MW}{R \cdot T_o \cdot z} \cdot \left(\frac{2}{\gamma+1}\right)^{\frac{\gamma+1}{\gamma-1}}}$$

Using the orifice area calculated, the area will be rounded up to the next common size.[39] The same process using Equation 6 was used to determine the orifice areas of the TK-01, TK-02, TK-03, and TK-04. The wetted area was assumed to be the circumference of each tank multiplied by the height of the liquid assuming the tank is half-filled.[39] Then, for each tank, the wetted area was used to calculate and determine the necessary relief valve orifice area for the process vessel.

Failure Rate Analysis

The failure rate of a system is the number of failure events divided by the total elapsed operating time during which the events occurred.[40] This cannot be calculated prior to startup but can be within the first year.

Instead, the failure rate should be minimized through the Process Hazard Analysis (PHA) process, specifically by identifying the distillation column vulnerabilities, or failure modes. These include unanticipated corrosion, design faults such as excessive flooding or weeping during operations, external events such as failures of other operating units, fire/explosions, human error in operation, impurities in the feed stream that are not removed during the adsorption process, maintenance faults in heat exchanger cleaning or pump maintenance, overheating of feeds, over pressurization and potential tray uplift, or structural failures. Preventing failure of the column will be reliant upon operations and implementing engineers alike and should be discussed in the detailed design stage.[41]

Personnel Exposure Risk

Personnel exposure risk was designated to include all risks that each employee is exposed to on the site. These risks include hazardous chemicals, extreme temperatures, and significantly above atmospheric pressures.[42]

Within this process unit, hazardous chemicals can be found within every process fluid. Compounds such as H_2S and CO are present, and while typically contained, should be monitored with both personal and stationary gas monitors.[42] Individuals exposed to any amount of hazardous chemicals should report and document these exposures to prevent over exposures to hazardous substances.[42]

Extreme temperatures and pressures can also be found in the heat exchangers, towers, and tanks across the process unit. While these temperatures and pressures are continuously monitored, there is always a risk of loss of containment during a unit upset. This risk should be minimized by limiting personnel time in the unit, keeping office spaces within blast-resistant buildings, and by keeping relief devices properly maintained.[42]

Finally, overall personnel risk should be minimized with professional training of all employees prior to unit startup. Then, this training should be kept up to date about start-ups, shutdowns, unit upsets, and steady state operations. Daily steady state operations should be routinely audited by subject matter experts to ensure standard operating procedures are continually followed. Failure to do so will result in unit upsets, therefore maintaining properly written procedures is necessary.

After unit start up, monitoring should occur for carcinogenic compounds and other hazardous substances.[42] This monitoring should be supervised by an industrial hygiene professional and reported to proper authorities.

Atmospheric Detonation of Distillation Inventory

The atmospheric detonation of distillation inventory was estimated using the TNT Equivalency method. This method equated the energy of a combustible fuel to an equivalent mass of TNT. Then, the mass of TNT is equated to a level of damage to common structure, process equipment, and humans.[43]

First, the equivalent mass of TNT was calculated using Equation 7.[43] In this equation, $m_{TNT}$is the equivalent mass of TNT, η is the empirical explosion efficiency, m is the mass of the hydrocarbon in the system, is the energy of explosion of the flammable gas, and $E_{TNT}$the energy of explosion of TNT. For the purposes of this calculation, the explosion efficiency was assumed to be 2%, the mass of hydrocarbon was calculated utilizing the molecular weight, molecular flow of hydrocarbon, and an assumed time before flow shutoff of 10 minutes.[43] The molecular weight of the material was found using the column simulation: 184 lbmol/lb. A worst-case scenario of 30 Mlb/hr from the relief valve scenario was used. Then, the energy of explosion was found using a known energy of explosion for a similar molecular weight hydrocarbon.[43] The energy of explosion for TNT was provided at 2,016 BTU/lb.[43]

[Equation 7][43] *3*

$$m_{TNT} = \frac{\eta * m * \Delta H_c}{E_{TNT}}$$

This equation found that around 21 Mlb of TNT was equivalent to the worst-case scenario mass inside the column. Then, this value of equivalent TNT was used to find the overpressure estimate from the explosion, with respect to a predetermined distance from the ground-zero point of explosion *(r*), Equation 8[43]. Multiple *r* values were used to find the detonation impacts at different distances away from the explosion.

[Equation 8][43]*4*

$$z_e = \frac{r}{m_{TNT}^{1/3}}$$

In conjunction with the scaled overpressure equation found as Equation 9, the scaled overpressure at each distance away from the explosion was found.[43]

This equation found that around 21 Mlb of TNT was equivalent to the worst-case scenario mass inside the column. Then, this value of equivalent TNT was used to find the overpressure estimate from the explosion, with respect to a predetermined distance from the ground-zero point of explosion *(r)*, Equation 8.[43] Multiple *r* values were used to find the detonation impacts at different distances away from the explosion.

The scaled overpressure values were used to find the actual overpressure that would occur during a potential explosion from Equation 10.[43] This final overpressure value was found with respect to the overpressure above atmospheric pressure ($\frac{p_o}{p_a}$) that would be experienced in an explosive event and the expected atmospheric pressure in the area of detonation (*EAP*).[43]

[Equation 10]5[43]

$$Overpressure = EAP \cdot \frac{p_0}{p_a}$$

Each overpressure value was correlated to an approximate damage estimate for the affected area.[3] These correlated *r* values, z_e values, and the p_0/p_a values as well as their damage estimates are found in Table 16.

Table 16: TNT Equivalency Effects at Varying Radii from Explosion Center

Distance from Point of Explosion (r{mi.})	**Scaled Distance (z_e{ft/lb$^{(1/3)}$})**	**Scaled Overpressure, $\frac{p_o}{p_a}$**	**Over-pressure psig**	**Damage Estimate[43]**
0.25	19	0.0915	1.345	Steel frames of buildings distorted, fastenings fail, wood panels buckle, panels blow in
0.5	38	0.0440	0.647	Large and small windows shatter, minor damage to house structures
0.75	57	0.0291	0.428	Limited structural damage
1.00	76	0.0218	0.320	0.95 probability of no serious damage at or below this value

While this is a large distance to prevent significant property damage, this could be minimized by reducing the amount of material refluxing within the tower. This is suggested to be examined during the detailed design stage. Additionally, multiple layers of protection should be discussed to prevent atmospheric releases of large amounts of process material. This would further prevent an explosion of this magnitude.

Hazard and Operability Study

Working with hazardous fluids, high pressures, high temperature, and large equipment within this process design will present various risk that can occur throughout the process. A Hazard and Operability Analysis (HAZOP) shown in Table 17 represents a systematic technique to examine risk causes, consequences, and action required.

Table 17: Hazard and Operability Study[43]

Project Name: Pyoil Purification Unit				**Date: 3/10/23**		Page 1 of 1
Process: Primary Distillation Column						
Reference Drawing: Pyoil Purification Unit PFD and P&ID				**Section: T-04 Primary Distillation Column of Pyoil Purification Unit**		
Item	*Study Node*	*Process Parameters*	*Deviation*	*Possible Causes*	*Possible Consequences*	*Action Required*
1A	T-04	Flow	NO	• Blockage in pipes • Closed control valves • Valve failure • Pump failure	• Column dries out • Off specification product • Potentially dangerous products • No operation	• Check control valves & pipes • Emergency shut down • Bypass necessary equipment • Install low level alarms
			LESS	• Blockage in pipes • Closed control valves • Valve failure • Pump failure	• Column dries out • Off specification product	• Check control valves & pipes • Emergency shut down • Bypass necessary equipment • Install low level alarms
			MORE	• Fully opened control valve • Control valve failure (fail open) • Increased pump capacity	• Flooding in column • Change in product specification • Temperature decrease • Increased level in liquid bottoms	• Check maintenance procedures • Install/Check controls • Reduce feed being processed
1B		Level	LOW	• Leakage in pipe • Clogged pipe or equipment	• Decreased level in column • Back flow of material	• Check maintenance • Install low level alarm
			HIGH	• Output blockage in pipe • Closed output control valve • High reflux rate	• Overpressure reflux drum • Condensed liquids flow back into distillation	• Install high liquid level alarm • Check maintenance procedures
1C		Temperature	LOW	• Higher cooling water temperature • Reboiler corrosion	• Off specification products • Not enough product production • Pulling vacuum on column • Flooding on trays	• Increase pressure • Check reboiler health • Install low temperature alarm
			HIGH	• Increased column pressure • Hotter feed inlet • Lower cooling water flow	• Off specification products • Too much gas production/loss of profit • Weeping of trays	• Decrease column pressure • Install high temperature alarm • Check maintenance procedures
1D		Pressure	LOW	• Decreased flow rate • Temperature increase • Change in feed composition • Tray damage	• Pulling vacuum • Extremely off specification products • Not enough product production • Column possibly collapsing on itself	• Install low pressure alarm • Check bottoms flow controller • Decrease temperature
			HIGH	• Increased flow rate • Decrease feed temperature • Gas control valve blockage	• Risk of over pressuring incident • Potentially dangerous products • Risk of loss of containment	• Install a high-pressure alarm • Check top control valve • Check maintenance procedures

Recommendations for Improvement of Bali Sorting Facility

Overall, the Bali Sorting Facility could utilize detailed process controls to optimize the current process. These controls would allow multiple trains to switch on and off depending on waste levels within the facility. Controls would also assist drivers to know which dumpsters to stop at, and which section to take. The group concluded three key recommendations for the Bali Sorting Facility: move the facility to a multi train system, utilize dumpster availability and sectioning, and begin river and ocean waste collection. These three recommendations would each assist the others in closing the gaps defined in further sections, in conjunction with additional efforts put in place, this would significantly increase the efficiency of the sorting facility while minimizing the costs associated with the facility.

Multiple sorting trains should be started up and utilized within the facility. These specific trains could be switched on and off depending on the inputs of the system. Additionally, this method would allow the facility to trial robotic and autonomous trains at facilities. After experimenting and collecting data, each facility could decide if the cost of the more expensive trains is worth the upfront capital cost. This proposal can be seen in Figure 5, below. Each color/pattern could be a different type of train, sorting as needed materials.

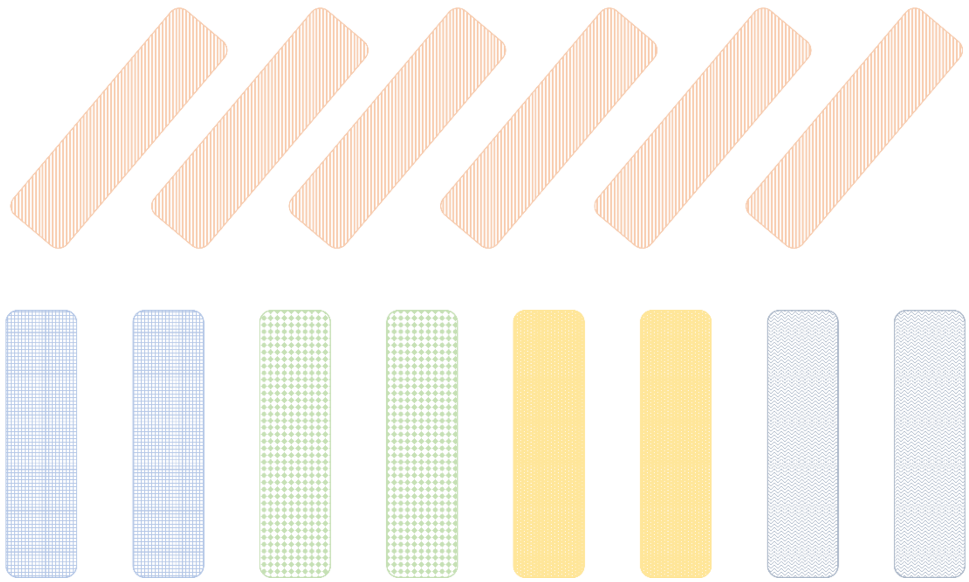


Figure 56: Sorting Train Configuration Proposal

Additionally, increasing dumpster availability and sectioning would allow larger, fewer dumpsters in neighborhoods. These dumpsters would have at least two sections, one for inorganic waste, and another for organic waste. Specializing these dumpsters would allow ease of access for drivers to pick up the waste and spend less time collecting, and more time driving. Finally, larger dumpsters would allow the bike paths and frequency to be altered. More collections would be performed during the tourist season and less collections during rush hour. A proposed system for this dumpster can be seen in Figure 6. Each section of the dumpster would be removable for easy dumping, and each section would fit within a larger bin to keep the sections together and accessible.

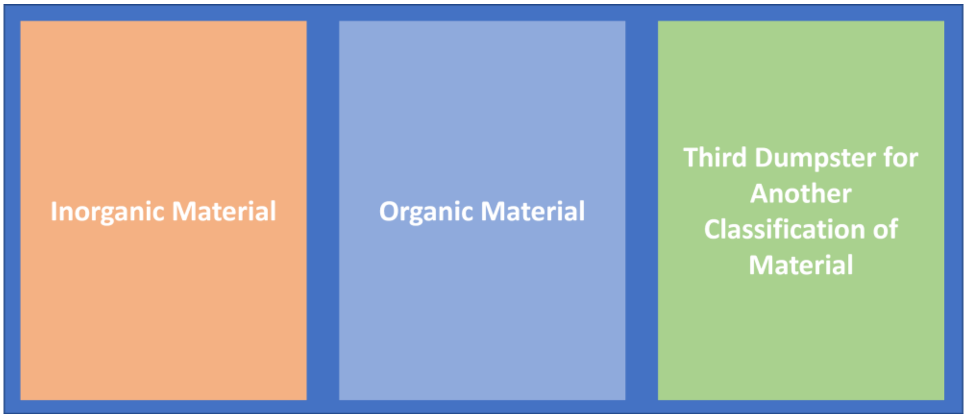


Figure 6 7: Sorted Dumpster Proposal

Finally, starting river and ocean collection would allow waste being removed prior to crowding beaches. Then, river waste collection should begin close to the ocean inlet, these collection sites would have their own employees to move collected waste to separate dumpsters, before being collected by the drivers. These employees would be hired under the collection coordinator and allow the entrance of foreman for each section of drivers, river collections, and shifts for employees.

Quantity Gap

Increasing the throughput of the sorting facility will be required when increasing the collection rate from residential areas. To increase the throughput of the facility, additional trains will be added and utilized when an influx of waste comes in. Shifts on these trains will be adapted to 3, 12-hour shifts per day. This rotation will be maintained by collection bays at the beginning of each train. When a specific train has run all the available feed, the train will be switched off and another switched on in a separate area. Also, dumpsters will be used to collect in a wider number of urban areas, distributed at a consistent rate per population density. Ocean and river collection will assist in a higher amount of feedstock into the facility, and education programs informing children and young adults of the need for sustainable programs. Finally, collection programs will also be added in high waste areas such as store fronts and docks that have large amounts of flexible mono plastic wastes desired for feedstocks.

Quality Gap

Decreasing contamination levels in plastic waste should be done with more training of employees and education systems for the general public. The multi train system would enable an experimental period of robotics to utilize separation. If this proves to be economically feasible, more of this robotics should be used to minimize contamination of plastic streams. Additionally, split dumpsters would enable the public to split their own waste, and in conjunction with education programs in schools and commercial marketing, inform the public on why waste should be split into separate areas. Finally, splitting operations shifts while on the waste lines would increase the quality of performance that is received and improve the contamination levels. Finally, it is suggested that experienced Industrial Engineers are consulted to take a further look at how this sorting process compares to that of other companies in the industry.

Affordability Gap

Decreasing the overall cost per ton of waste and generating additional revenue will be performed by increasing the efficiency of the facility. Multi trains can be optimized by shutting down trains not in use and keeping shift employees moving from area to area as needed. Additionally, dumpster sectioning will decrease the amount of dividing on site and river/ocean collection will be performed to collect a maximized amount of waste with a minimal amount of energy usage. Revenue should be increased by maximizing the splits of materials and selling wastes that are not processed at further downstream facilities. Finally, all process controls will be implemented to tell employees where to focus their attentions while maintaining increased throughput.

Conclusions

In conclusion, the group suggests that GP moves this project into the detailed design stage, while concurrently performing both a detailed design and an economic analysis. The additional process unit will contain two adsorption towers, operating inversely so that one tower is regenerated while the other is processing feed. Then, two distillation columns are utilized. The first separating off the fuel gas from the process fluid, while the second separates the reduced stream into Light, Medium, and Heavy Cuts. After, all the cuts are cooled to spec temperature before being shipped to tankage and off to the steam crackers when available. The group was also able to find a way to sell the bottoms that cannot be cracked in a steam cracker.

In all, to close the gaps within the Bali Sorting Facility, multiple angles will need to be adapted. This includes increasing process controls within the facility that are automated, clearly demonstrating where sorters, drivers, and managers are needed. Additionally, multiple trains will be utilized as well as a new, wider divided dumpster distribution range, and a river and ocean waste collection system. By closing the gaps in the facility, a wider range of citizens will be served, and the facility will become more efficient, increasing revenue and decreasing expenses.

Appendices

Appendix A: Adsorption Section Detail

Design Assumptions

The adsorption unit was developed to treat 100% of the feed material entering the unit. This decision was made to prevent heavy metals venting to atmosphere or exiting the unit in the Heavy Cut stream. While this is a more expensive solution, it demonstrates the company's desire to practice corporate responsibility during the pyrolysis treatment and to minimize capital material costs.

The unit was made using multiple design assumptions. First, it was assumed that each bed of adsorbent was able to undergo complete regeneration. While the group did not determine the regeneration cycle length, the dual tower system (T-01 & T-02) was designed with the intention of regenerating one tower while processing material in the other. Next, it was assumed that each tower could hold two catalyst beds, as shown in Figure 5. This was facilitated by an initial liquid distribution tray on the first bed, followed by the dehydration bed that was supported by a structural tray. Then, another liquid distribution tray was used to redistribute the material over the second adsorbent bed. In total, this minimized the amount of towers and conserved space, but is directly reliant upon the assumption that the beds can be regenerated under identical conditions. It was also assumed that the BASF adsorbent in the problem statement was the required material to use, as found in the problem statement.[1] Finally, it was assumed that the process fluid will gain 10-20°F across each column as the adsorption process is exothermic, and will release extra heat into the process material.[3]

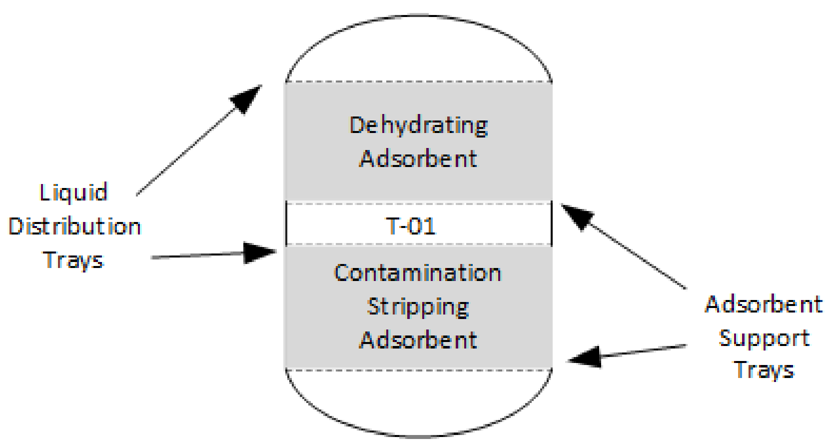


Figure 58: Adsorption Column Internal Diagram

To accomplish continuous flow through the distillation section of the unit, two identical columns and an expert lock out tag out procedure will be developed in future project development. This will seamlessly lock out one column while opening feed flow to the other. Then the locked-out column will be regenerated while the fresh column removes water and heavy metals from the feed. During higher rates of contamination, the beds could have their pressure adjusted with the feed pump (P-01) from the storage tank. This would increase the adsorption rate but accelerate the time before the next regeneration cycle. If this was accounted for prior to significant contamination entering the unit, the regeneration cycle could be adjusted prior to a contaminated feed entering the unit, preventing disruption of the regeneration cycle.

Adsorbents, as requested by the project statement, are BASF's PuriCycle H dehydration material in the initial bed, and BASF's PuriCycle HP heavy metal adsorbent in the second bed.[44] Due to a lack of response from BASF, the group did not have reliable information regarding the adsorption rates of each material, the pressure drops across the material, or the regeneration statistics, it is assumed that the regeneration rate will be dependent upon the contamination of the feed entering the unit, therefore, this will be adjusted over time to maintain continuous flow to the distillation section.

Design Calculations

From the initial project statement, the liquid hourly space velocity (LHSV) was stated at 1 hr^{-1}.[1] This statement was utilized to find the anticipated amount of catalyst in each separate bed. Equation 11 was used, as the flow rate of the total feed material (Q) could be determined by utilizing the desired mass flow rate from the problem statement ($\dot{m}$) multiplied by the density of the feed (ρ), which was found by simulating the process conditions of the feed in Aspen HYSYS. The volumetric flow rate, once divided by the volume of adsorbent being used yields the LHSV for the bed.[45] The actual calculations for a singular adsorbent bed are as follows:

[Equation 11][45] *6*

$$LHSV = \frac{\dot{m} \cdot \rho^{-1}}{V}$$

From the above equation, it was found that each adsorbent bed needed 1,100 ft^3 of material. This volume was utilized to size the adsorption towers (T-01 & T-02). By algebraically utilizing the volume of cylinder equation, Equation 12, and the desired L/D ratio for towers.[45]

[Equation 12][45]*7*

$$volume = \pi \cdot r^2 \cdot h$$

The group strategized that the height of each adsorption tower would be 50 ft. This length accounts for 18 ft of extra dish space at each head of the tower, liquid head at the bottom of the towers, and space for distribution trays. Additionally, the diameter of each tower was found to be 10 ft.

Pricing

Tower pricing was conducted utilizing the total volume of each column. Utilizing the above information, the volume of each adsorption tower was found to be 2300 ft^3. The volume (A), along with the K_1 , K_2, K_3 that can be found in Table 18, were utilized in Equation 13 to find the initial purchase cost estimate of the tower.[3]

[Equation 13][3] 8 $$log\ C_p^0 = K_1 + K_2\ log\ A + K_3(log\ A)^2$$

Then, the purchased cost of \$46.2M was utilized within Equation 14, to find the bare module cost of each tower. This was done using the constants found in Table 18, this table has each value along with the reference. The F_p otherwise known as the pressure factor, was calculated using Equation 15, which has constants that can also be found in Table 18.[3]

[Equation 14][3]9 $$F_P = \frac{\left(\frac{(P+1)D}{1699.2+1.2(P+1)}\right)+0.0038}{0.0067}$$

[Equation 1510][3] $$C_{BM} = C_p^0 * (B_1 + B_2 * F_P * F_M)$$

Table 18: Adsorption Tower Pricing Variables[3]

Variable	Value
P	5 [barg]
S (max allowable stress-CS)	944 [bar]
E (weld efficiency)	0.9
CA (corrosion allowance)	0.00315 [m]
T_{min} (min allowable vessel thickness)	0.0063 [m]
F_m (SS Clad CS[3])	1.8
B_1	2.25
B_2	1.82

The bare module cost of each tower was found to be \$396M. This number was scaled up into 2021 values using the Equation 16, utilizing the 2001 CEPCI value and the 2021 reported CEPCI value.[5]

[Equation 1611][3] $$Updated\ C_{BM} = \frac{2021\ CEPCI}{2001\ CEPCI} \cdot C_{BM}$$

After updating the C_{BM} the total cost for both towers were \$1.63MM. This was exclusively for a stainless-steel clad carbon steel tower, a metallurgy that was chosen based off the corrosion resistance of the material against heavy metal and water corrosion.

Similar costing procedures were used to find the capital cost for the trays within the towers. Specific tray types were not costed, but instead a generalized costing method was used to approximate tray expenses for the project. This was introduced by implementing the cross-sectional area into Equation 15. Different constants, that are found in Table 19, were utilized before finding a base condition price for each individual tray[3]. Then, this number was put into Equation 17, to find the bare module cost of all trays[3].

[Equation 1712][3] $$C_{BM} = n \cdot F_{BM} \cdot f_q \cdot C_p$$

Table 19: Adsorption Tray Costing Variables[3]

Variable	**Value**
K_1	3.3322
K_2	0.4838
K_3	0.3434
F_{bm}(SS)	1.8
F_q	1.87
N (number of trays in total)	8

Once the bare module cost was found, the Equation 16, was again utilized to find the updated bare module cost for all trays within both adsorption columns to be $560M.

The last piece to price on the adsorption unit was that of the two adsorbents. These values were estimated utilizing similar adsorbent prices. The prices, along with the density of the similar adsorbent, and the desired volume of each adsorbent were put into Equation 18, to find the total cost of each catalyst required.[3] Values and references for each variable can be found in Table 20.

[Equation 1813][45] $$Bed\,Value \ = \ (Unit\,Value) \cdot (\rho) \cdot (V)$$

Table 20: Adsorbent Costing Assumed Values[46,47]

Variable	Value
Unit Value – Dehydration Adsorbent[46]	1.9 [$/lb]
Unit Value – Heavy Metal Adsorbent[47]	0.45 [$/lb]
Density (p) - Dehydration Adsorbent[46]	48 [lbs/ft^3]
Density (p) - Heavy Metal Adsorbent[47]	124.86 [lbs/ft^3]
Required Volume (V) – Eq. 18	1100 ft^3

Since the values used were already in terms of 2021 values, a CEPCI adjustment was not required. In total, the BASF PuriCycle H adsorbent and the BASF PuriCycle HP adsorbent were estimated to cost $124M in total. This includes adsorbent for both adsorption towers both beds within the towers. The group did not include supplementary adsorbent prices in this scheme as it is believed that the adsorbent will be replaced during a full unit turnaround, so the additional adsorbent (if needed) will be a part of the turnaround budget.

In total, the cost of the towers, internals, and adsorbent materials summed to $2.12MM, the group estimated this to be a fair price due to some minimal industry standard research regarding adsorbent column prices in recent years.

Appendix B: Distillation Section Detail

The distillation process for the Pyoil was designed using a two-tower setup. The first tower is a 2-stage distillation column with no overhead condenser. The tower is sized at a height of 15 ft and a diameter of 2.75 ft. Stainless steel bubble cap trays were selected for the tray type to account for the vapor flow being much lower than the liquid flow. The reboiler for this tower was sized to be a kettle reboiler of 209 ft.^{2} The purpose of this tower is to distill out the light-ends before the primary distillation column. This results in an overhead product stream of Py Gas that is mostly ethane, ethylene, propane, and propene. Lastly, the operating pressure for the column was selected to be 25 psia, resulting in the tower being costed as $162 M.

The primary distillation column was designed as a 58 ft tall, 9.25 ft wide tower with 19 sieve trays. Sieve trays were chosen due to their lower price point to valve trays while still being adequate to maintain proper vapor liquid equilibrium within the column. There is also a side draw on stage the third stage from the bottom. This side draw allows to produce three product streams: the Light Cut stream from the top of the column, the Medium Cut from the third stage, and the Heavy Cut from the tower bottoms. The feed stage for the column was chosen to be the 10th stage from the bottom after iterating for various locations within the Aspen HYSYS modeling software. The side draw stage was selected to align with a liquid product with the specified D86 FBP value. The tower operates at a pressure of 23 psia and a reflux ratio of 5.00. Due to the large boil-up of the tower, the reboiler was sized as a 4750 ft^2 thermosyphon reboiler. These specifications yielded a tower with a cost estimate of $1.27 MM.

The operating pressures were selected by iterating the towers for increased Light and Medium Cut products while trying to maintain a pressure near atmospheric conditions to minimize the inherent risk of pressurized vessels. The material of construction for the towers was chosen as stainless steel clad with full stainless steel trays in order to account for the potential of corrosive, heavy metals passing through the adsorption unit and into the distillation process.

The modeling for the distillation process was done in Aspen HYSYS. This modeling software required several assumptions to be made including a Peng-Robinson property package, no pressure or temperature loss through piping, and program generated pseudo components being used to model the feed. Peng-Robinson was chosen as the property package due to it being particularly accurate for non-polar hydrocarbons, as well as being the most developed package within Aspen HYSYS.[48] In order to account for the idealized piping, the pump calculations were upsized. The pseudo components of the feed were developed from a true boiling point analysis of the feed provided in the project description. When sizing the distillation towers, an assumed tray efficiency of 75% was also used to convert the modeled number of theoretical trays into a usable number of actual trays.

Energy consumption was a priority in the development of the process configuration. The two distillation columns' operating pressures were chosen such that a pump would not be necessary while in steady state operation. The utilities were also limited to cooling water instead of refrigerants and available hot oil instead of steam. Heat integration was also evaluated as a possible way to minimize energy consumption, but no locations were suitable as pre-heating the feed was not compatible with the first column.

To maintain the desired conditions of the distillation towers, pressure controllers and level controllers were installed at the top and bottom of each tower, respectively. These controllers ensure that the liquid levels in the bottom of the towers are maintained and that the operating pressures are held constant.

The addition of the initial distillation tower also serves to keep any water that passes through the adsorber out of the product streams. The tower operates at a temperature of 107 ºF with the reboiler at 267ºF. This temperature profile keeps the water from passing to the primary distillation tower as well as keeping it out of the Py Gas. However, the water can collect in T-03 over time, so it is recommended to drain the liquid from the tower after periods of adsorption difficulty.

T-03:

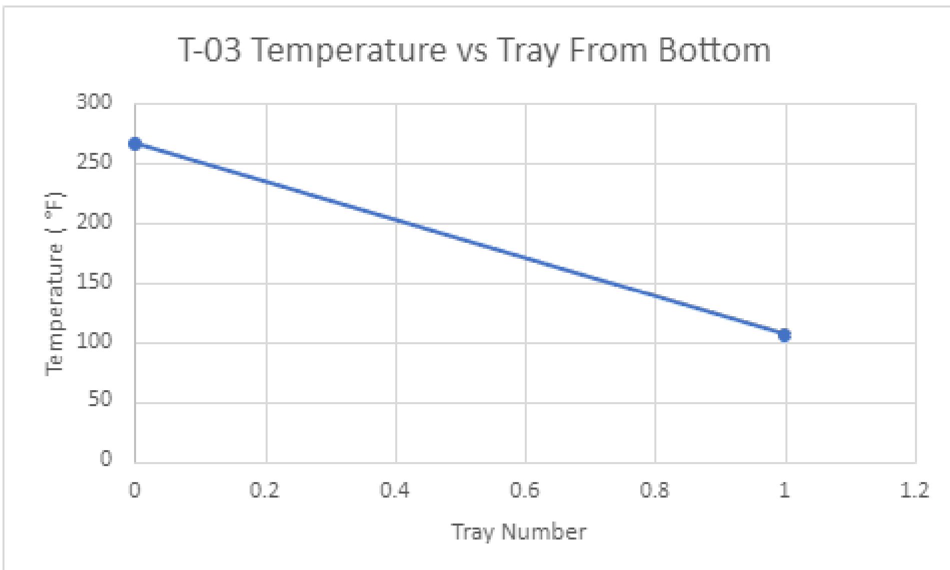


Figure 69: T-03 Temperature vs. Trays (Numbered From Bottom)

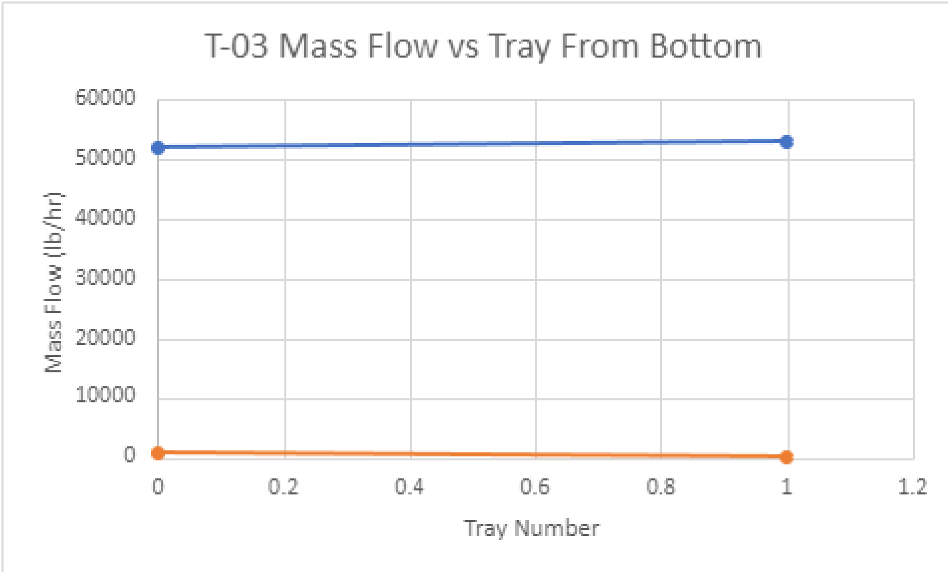


Figure 710: T-03 Mass Flow vs. Tray (Numbered From Bottom)

T-04:

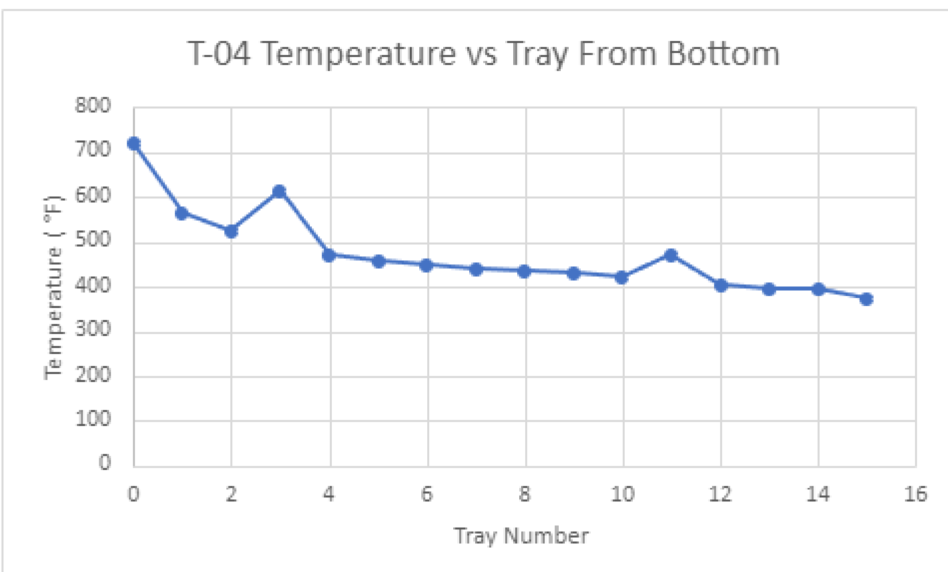


Figure 811: T-04 Temperature vs. Tray (Numbered From Bottom)

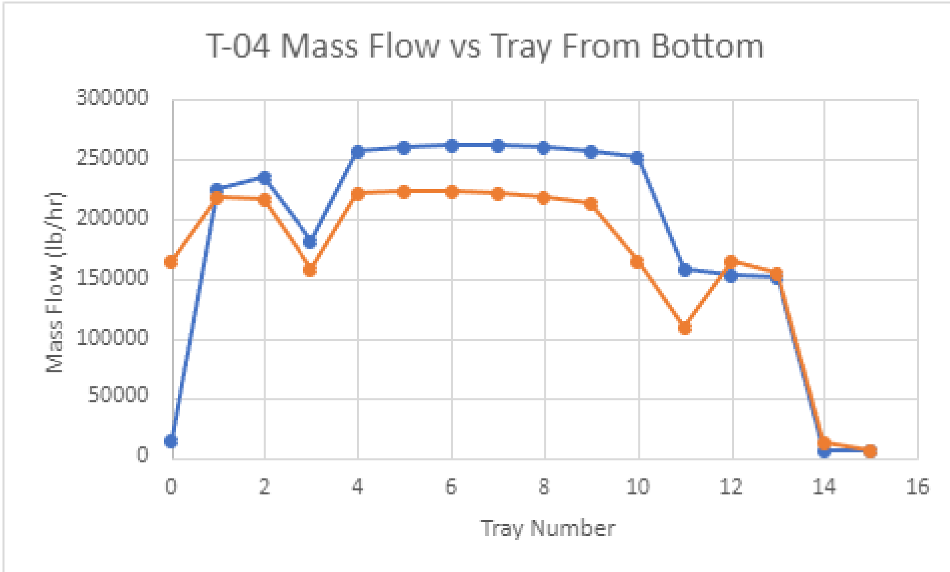


Figure 912: T-04 Mass Flow vs. Tray (Numbered From Bottom)

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