LETTER OF TRANSMITTAL

Date: March 10, 2023

To: Joe Machado, Senior Advisor, Alliance to End Plastic Waste

          Greg Yeo, Retired Chief Engineer, ExxonMobil

From: Connor Allen, Madison Neighbors, Hannah Nicolaysen, Christopher Spellings

Subject: Global Petrochemicals – Pyrolysis Oil Purification Unit Preliminary Design and Feasibility Study

Enclosed you will find the Pyrolysis Oil Purification Unit preliminary design for Global Petrochemicals. This includes an estimation of cost and process safety description of the design. The location of this project is Southeast Asia. The purification unit is comprised of both distillation and adsorption. This unit is capable of fractionating and decontaminating 4,800 bbl/day of pyrolysis oil. The distillation design is comprised of two distillation towers in series to produce four product cuts. These four cuts are Olefin-Rich Vapor (Pygas), Light Cut (Naphtha), Medium Cut (Gas Oil), and a Heavy Cut. The adsorption design utilizes eight columns to provide continuous operation with proper contaminant protection. The adsorption process decontaminates the Light and Medium Cuts which are then sent to storage and ultimately to the adjacent ethylene plant as steam cracker feedstock. This design also includes storage capacity for one week's worth of volume for raw feed and all liquid products. The Pygas is sent directly to the adjacent ethylene plant, whereas the Heavy Cut is not suitable for cracking, therefore it is disposed appropriately. Lastly, in compliance with environmental emission regulations, a vapor recovery unit was designed for recovering off-gas hydrocarbon vapors from storage facilities. To meet process safety codes and standards, a control scheme was implemented which includes alarms for pressures on each distillation tower as well as fail-safe control valves in applicable locations to minimize the likelihood of overpressure. For process safety, pressure relief devices were sized/designed for the columns in the distillation unit. The design was estimated to have a capital cost estimate of $28.4 million. Also, annual estimations of variable operating costs and fixed operating costs were determined to be $3.64 million and $4.68 million, respectively.

To overcome the second challenge within this project, a "cold eyes analysis" was performed on the sorting facility operation in Bali, Indonesia to provide an innovational framework for the increase in quantity, quality, and affordability of the plastic waste collection upstream of the designed purification unit.

Sincerely,

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Global Petrochemicals Indonesia

Pyrolysis Oil Purification Unit Preliminary Design

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EXECUTIVE SUMMARY

The objective of this study was to design a process unit that fractionates and purifies a stream of pyrolysis oil derived from pyrolyzed plastic waste to meet the specifications for feedstock to an adjacent ethylene plant for Global Petrochemicals. The location of this study was in Indonesia, Southeast Asia. The purpose for the study emerges from the idea that the economy surrounding plastics and their usage should be circular. The collection of plastic waste should be managed in a way that reduces environmental impacts and creates a closed loop for the recycle and reuse of plastics. To address the problem of plastics waste management, the design of a purification unit achieves the goal of successfully adding to the growing topic of a circular economy. Additionally, it diversifies the portfolio of feed sources to Global Petrochemicals' ethylene plant. This design prioritizes several important design considerations which includes process safety, environmental impacts, energy efficiency, and flexibility.

The designed purification unit incorporates both distillation and adsorption. This unit is capable of fractionating and decontaminating 4,800 bbl/day of pyrolysis oil with an overall yield of 86% by mass. The distillation design is comprised of two distillation columns in series with a system of eleven heat exchangers; two of which utilize heat integration to minimize utility usage. This process produces four specified cuts, which are Pygas (Olefin-Rich Vapor), Light Cut (Naphtha), Medium Cut (Gas Oil), and Heavy Cut. The adsorption design utilizes eight columns to provide continuous operation with proper contaminant protection. The adsorption process decontaminates the Light and Medium Cuts which are then sent to storage and ultimately to the adjacent ethylene plant as steam cracker feedstock. The Pygas is sent directly to the ethylene plant from the distillation with no further processing. Since the Heavy Cut is not suitable for cracking, it is appropriately disposed of through options, such as, selling it to a third party as a contaminated atmospheric residue or sending it to a third party for waste management. The designed purification unit includes one week of storage for all liquid products and feeds with proper emission control measures by incorporating a designed vapor recovery unit to recover overhead hydrocarbon vapors from storage vessels. The capital cost of this project was estimated to be $28.4 million with annual estimations of variable and fixed operating costs of $3.64 million and $4.68 million, respectively. These costs include all installed equipment with additional added costs for contingency and potential fees. Also accounted for in these estimates is the cost of operational labor, with an estimated requirement of 16 operators, based on the size of the unit.

A "cold eyes analysis" was also performed on the sorting facility operation in Bali, Indonesia to provide innovative ideas that increase the quantity, quality, and affordability gaps of collecting plastic waste. To ensure its long-term contribution to the business, the utilization of tourism and technology were determined to be primary avenues to close these gaps. Additionally, coupled with the implementation of waste banks, local communities could be provided with additional sources of income to help grow the local economy, ultimately leading to better waste stewardship.

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The purpose of the process is to separate and purify a pyrolysis oil (Pyoil) feedstock from a pyrolysis reactor in East Java, Indonesia to provide a clean feed stream to Global Petrochemicals' ethylene plant. The process to accomplish this is twofold. The first is to take the pyrolysis oil feedstock and fractionate it into four "cuts" based on the end boiling point (ASTM D86 EBP) of the hydrocarbons in the stream.[1] The second portion is to purify two of the cuts by removing present contaminants, such as chlorides and metals, using adsorption so that the ethylene plant equipment is not damaged by this product as feedstock. The first product of the unit is a non-condensable hydrocarbon stream called Pygas (Olefin-Rich Vapor) that is fed directly to the ethylene plant. The second product is the Light Cut (Naphtha), which has an EBP of 392 °F and is fully treated for contaminants and sent to the ethylene plant as a "Cracking Furnace Feedstock." The third product is the Medium Cut (Gas Oil), with an EBP of 620 °F, and is also fully treated against contaminants before being sent to the ethylene plant as "Cracking Furnace Feedstock." The final product stream is the Heavy Cut, which does not have an effective EBP, and is not suitable for cracking, therefore it is disposed of in an appropriate manner.[1]

The process begins with a pyrolysis oil feedstock that comes from the pyrolysis plant and is fed into feed tank storage, TK-101, which has a one-week volume capacity. The hydrocarbon mixture then leaves the tank and is pumped using P-101 A/B as raw material input into the process. From here, the Pyoil stream passes through a preheater train, E-101 – E-103, that preheats/super-heats the feed before entering the first distillation column, T-101. The heat exchangers, E-101 and E-102, work on the principle of heat integration and are used to cool the Medium and Heavy Cut product streams. E-103 is a heat exchanger that utilizes hot oil, to make up for any heat not provided previously, to reach a desired feed temperature of 420 °F.

The super-heated Pyoil feed enters the first trayed distillation column, T-101, which operates at column top pressure and temperature of 5.9 psig and 337 °F, respectively. This column utilizes fixed valve rectifying and bubble cap stripping sections to separate the process components into three fractions as Pygas (stream 9), Light Cut (stream 13), and a bottoms liquid (stream 17). Additionally, it should be noted that excess water is removed from the process with V-101 configured as a 3-phase separator. This water is then sent to wastewater treatment (Sour Water, stream 10). The Pygas is sent directly to the ethylene plant, at 107 °F, while the Light Cut is further cooled and sent to the adsorption unit for contamination treatment. The bottoms liquid, however, is drawn off the bottom of T-101 as stream 17 and pumped using P-103 A/B through E-107, which heats the process stream to a temperature of 600 °F. This super-heated fluid is then fed directly into the second trayed column as T-102, which operates at a column top pressure and temperature of 5.9 psig and 539 °F, respectively. This second column contains internals identical to the first column to separate the process components into two fractions as a Medium Cut (stream 24) and Heavy Cut (stream 29). The Medium Cut is further cooled and sent to the adsorption unit, while the Heavy Cut is cooled and sent to tankage (TK-102) before proper disposal. Once a week, TK-102 is emptied using P-106A/B. The Heavy Cut is to be trucked/shipped to either a refinery for profit, or to be treated at cost with a waste management company, both of which are located in Singapore.[2,3,4]

Next, the cooled Light Cut, at 100 °F, is passed through the adsorbers T-103 – T-106. These four adsorbers purify the stream of contaminants, with T-103 and T-104 removing metals using PuriCycle HP, and T-105 and T-106 removing chlorides via adsorbent PuriCycle H. The reason for four adsorbers (two of each type) is so the parallel columns, to those processing, can be regenerated simultaneously. Therefore, the stream will pass through one metal and one chloride treatment adsorber. Once the stream has been purified, it is deposited into TK-103, which holds one week's worth of storage. From there, P-108 A/B pumps it, at 70 psig, to the ethylene plant.

The cooled Medium Cut, at 100 °F, has a similar experience to the Light Cut. Four adsorbers are utilized to purify the stream of contaminants, with T-107 and T-108 removing metals with PuriCycle HP, and T-109 and T-110 removing chlorides with PuriCycle H. There are four Medium Cut absorbers, thereby allowing continuous operation. Therefore, the stream will only pass through one metal and one chloride adsorber. Once the stream has been treated, it is deposited into TK-104, which holds a week's worth of storage. From there, P-109 A/B pumps it, at 70 psig, to the ethylene plant.

One important note is that TK-101 and TK-103 are attached to a vapor recovery unit that takes overhead hydrocarbon vapors from both storage tanks and sends it to V-103 to knockout undesirable liquid. From there, it is compressed through C-101 A/B to 17.1 psia, then combines with the Pygas product from V-101 and creates stream 39. This combined stream is sent as the final Pygas product to the ethylene plant. The vapor recovery unit ultimately mitigates emissions by recovering the hydrocarbon vapors produced from storing the components in atmospheric storage.

To maintain consistent operation of the adsorbers, a hot nitrogen stream was used to regenerate any adsorbent that is "spent." The regeneration stream operates as a closed loop, to minimize the cost and waste of nitrogen. This nitrogen comes from the plant's utility stream at battery limits but must be pressurized and heated to provide adequate regeneration to the adsorbers. This is done by utilizing C-102, E-112, and E-114 which pressurizes and heats the nitrogen to 36 psia and 500 °F, respectively. The nitrogen then passes through any regenerating beds. After regeneration, the spent nitrogen combines into stream 66 and passes through E-112, thereby cooling the stream down while exchanging heat with the compressed nitrogen stream heading to regeneration. After passing through E-112, the partly cooled nitrogen is then further cooled to 100 °F using cooling water in E-113. This steam then passes through a knockout vessel, V-104, where condensed hydrocarbons and metal/chloride contaminants exit in stream 70. This knockout liquid is then sent to a waste processing facility.[3,4] The now "scrubbed" nitrogen then exits the top of the vessel as stream 69 and combines with any new make-up nitrogen before being recompressed in C-102 and restarting the cycle.

As per *Table 1*, the total utility consumption can be seen. This is inclusive of both the distillation and adsorption systems within the purification unit process.

Table 1. Utility Consumption

Utility Consumption		
Electricity	11,940,000	kW-hr
Fuel Gas (Hot Oil)	126,100	MBTU HHV
Cooling Water	153,100	MBTU

A visual layout of the process description can be found in the process flow diagrams (PFD) for both distillation and adsorption, *Figures 1 and 2*, respectively. Within the PFDs, the temperature, pressure, and mass flow rate are displayed for each labeled stream. If more information about a particular stream is desired, then it can be seen in *Tables 2-1 – 2-4*, which is the itemized stream summary tables. The labels of major equipment also show up on the PFD, where more detailed information such as sizing parameters, design pressures, and design temperatures can be found in the sized equipment list, *Table 3*.

Process Flow Diagrams for Process and Utilities Areas: Distillation

TK-101 Pyoil Feed Storage Tank | **P-101 A/B** Pyoil Feed Pump | **E-101** Pyoil Feed Preheater 1 | **E-102** Pyoil Feed Preheater 2 | **E-103** Pyoil Feed Preheater 3 | **T-101** Light Cut Column | **E-104** Light Cut Condenser | **V-101** Light Cut Reflux Drum, 3 Phase Separator | **P-102 A/B** Light Cut Reflux Pump | **E-105** Light Cut Tower Thermosyphon Reboiler | **E-106** Light Cut Cooler | **P-103 A/B** Medium Cut Tower Feed Pump | **E-107** Medium Cut Tower Feed Super-Heater

T-102 Medium Cut Column | **E-108** Medium Cut Condenser | **V-102** Medium Cut Reflux Drum | **P-104 A/B** Medium Cut Reflux Pump | **E-109** Medium Cut Tower Thermosyphon Reboiler | **P-105 A/B** Heavy Cut Pump | **E-110** Heavy Cut Cooler | **E-111** Medium Cut Cooler | **TK-102** Heavy Cut Storage Tank | **P-106 A/B** Heavy Cut Disposal Pump | **V-103** Vapor Recovery Scrubber/Knockout Drum | **C-101 A/B** Vapor Recovery Compressor | **P-107 A/B** Vapor Recovery Knockout Pump

KEY

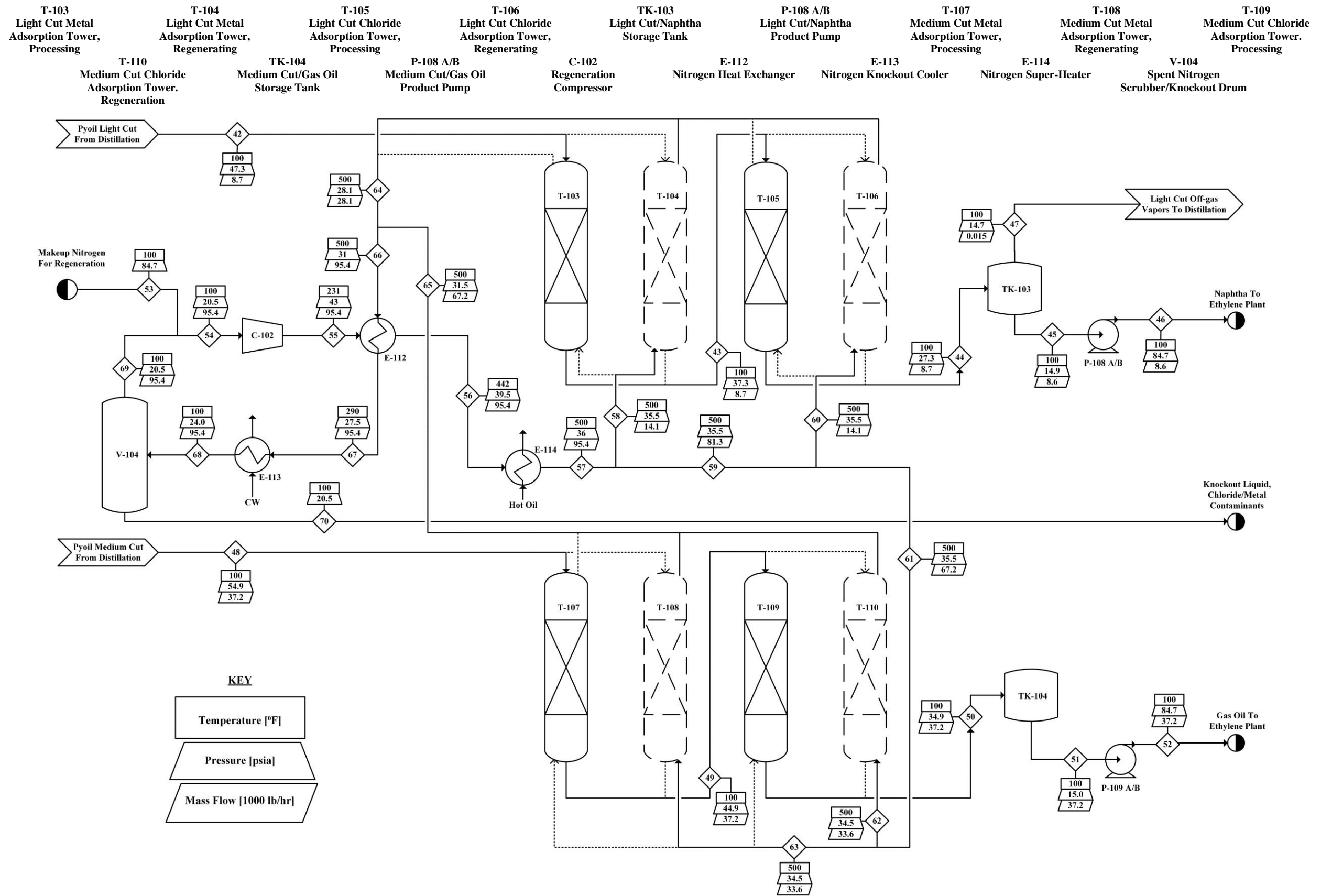|	
Temperature [°F]	
Pressure [psia]	
Mass Flow [1000 lb/hr]	

Figure 1. Distillation Process Flow Diagram

Figure 2. Adsorption Process Flow Diagram

Stream Summary Table as Material/Energy Balances

Table 2-1. Stream Summary Table, Streams 1-18

Stream Number	1*	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Stream Description	Raw Pyrolysis Feed Oil	Pyoil Feed Exiting TK-101 Entering P-101A/B	Pyoil Feed Exiting P-101A/B Entering E-101	Pyoil Feed Exiting E-101 Entering E-102	Pyoil Feed Exiting E-102 Entering E-103	Superheated Pyoil Feed Exiting E-103 Entering T-101	Column Top Vapors Exiting T-101 Entering E-104	Liquid/Vapor Exiting E-104 Entering V-101	Pygas Vapors Exiting V-101	Sour Water Exiting V-101	Hydrocarbon Liquid Exiting V-101 Entering P-102A/B	Reflux liquid Exiting P-102A/B Entering T-101	Light Cut Distillate Exiting P-102A/B Entering E-106	Cooled Light Cut Exiting E-106 Sent to Adsorption	Column Sump Liquid Exiting T-101 Entering E-105	Boilup Exiting E-105 Entering T-101 Bottoms	Bottom Column Hydrocarbon Liquid Exiting T-101 Entering P-103A/B	Hydrocarbon Liquid Exiting P-103A/B Entering E-107
Pressure [psia]	18.0	15.0	128	121	114	107	20.6	17.1	17.1	19.7	19.7	64.3	64.3	47.3	26.5	24.3	26.5	78.6
Temperature [°F]	100	100	100	347	403	420	337	107	107	107	107	107	107	100	469	500	500	500
Phase	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Vapor	Liquid/Vapor	Vapor	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid/Vapor	Liquid	Liquid
Molar Enthalpy [Btu/lbmole]	-165,000	-166,000	-166,000	-141,000	-134,000	-132,000	-72,000	-100,000	-18,200	-122,000	-106,000	-106,000	-106,000	-107,000	-134,000	-122,000	-143,000	-143,000
Molar Flow [lbmole/hr]	297.0	295.8	295.8	295.8	295.8	295.8	142.4	142.4	9.9	--	132.4	59.3	73.1	73.1	394.7	394.7	212.7	212.7
Mass Enthalpy [Btu/lb]	-907	-907	-907	-771	-734	-723	-635	-863	-392	-6,800	-898	-898	-898	-902	-698	-630	-675	-675
Mass Flow [lb/hr]	54,100	54,000	54,000	54,000	54,000	54,000	16,100	16,100	463	--	15,700	7,030	8,660	8,660	76,100	76,100	44,900	44,900
Component Mass Fraction																		
Nitrogen	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Water	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methane	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0002	0.0002	0.0059	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ethane	0.0010	0.0008	0.0008	0.0008	0.0008	0.0008	0.0032	0.0032	0.0822	0.0000	0.0009	0.0009	0.0009	0.0009	0.0000	0.0000	0.0000	0.0000
Ethylene	0.0005	0.0004	0.0004	0.0004	0.0004	0.0004	0.0014	0.0014	0.0404	0.0000	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.0000	0.0000
Propane	0.0029	0.0028	0.0028	0.0028	0.0028	0.0028	0.0122	0.0122	0.1937	0.0000	0.0068	0.0068	0.0068	0.0068	0.0000	0.0000	0.0000	0.0000
Propylene	0.0026	0.0024	0.0024	0.0024	0.0024	0.0024	0.0105	0.0105	0.1829	0.0000	0.0055	0.0055	0.0055	0.0055	0.0000	0.0000	0.0000	0.0000
Butane	0.0036	0.0035	0.0035	0.0035	0.0035	0.0035	0.0184	0.0184	0.1302	0.0000	0.0152	0.0152	0.0152	0.0152	0.0000	0.0000	0.0000	0.0000
C4 Olefins (1-Butene)	0.0062	0.0061	0.0061	0.0061	0.0061	0.0061	0.0310	0.0310	0.2533	0.0000	0.0244	0.0244	0.0244	0.0244	0.0000	0.0000	0.0000	0.0000
1,3-Butadiene	0.0007	0.0007	0.0007	0.0007	0.0007	0.0007	0.0035	0.0035	0.0281	0.0000	0.0028	0.0028	0.0028	0.0028	0.0000	0.0000	0.0000	0.0000
Pentane	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0074	0.0074	0.0193	0.0000	0.0070	0.0070	0.0070	0.0070	0.0000	0.0000	0.0000	0.0000
Hexane	0.0040	0.0040	0.0040	0.0040	0.0040	0.0040	0.0237	0.0237	0.0215	0.0000	0.0238	0.0238	0.0238	0.0238	0.0000	0.0000	0.0000	0.0000
C6-C7 (NBP 150-249 °F)	0.0090	0.0090	0.0090	0.0090	0.0090	0.0090	0.0542	0.0542	0.0193	0.0000	0.0552	0.0552	0.0552	0.0552	0.0000	0.0000	0.0000	0.0000
C8-C9 (NBP 250-349 °F)	0.1093	0.1094	0.1094	0.1094	0.1094	0.1094	0.5726	0.5726	0.0208	0.0000	0.5889	0.5889	0.5889	0.5889	0.0418	0.0418	0.0179	0.0179
C10-C12 (NBP 350-449 °F)	0.2644	0.2647	0.2647	0.2647	0.2647	0.2647	0.2616	0.2616	0.0025	0.0000	0.2693	0.2693	0.2693	0.2693	0.3998	0.3998	0.2665	0.2665
C13-C16 (NBP 450-549 °F)	0.2608	0.2610	0.2610	0.2610	0.2610	0.2610	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2871	0.2871	0.3140	0.3140
C17-C20 (NBP 550-649 °F)	0.2208	0.2209	0.2209	0.2209	0.2209	0.2209	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1880	0.1880	0.2658	0.2658
C20-C24 (NBP 650-749 °F)	0.0789	0.0789	0.0789	0.0789	0.0789	0.0789	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0589	0.0589	0.0950	0.0950
C25+ (NBP 750+ °F)	0.0339	0.0339	0.0339	0.0339	0.0339	0.0339	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0243	0.0243	0.0408	0.0408
Additional Properties																		
Mass Density [lb/ft³]	47.99	48.05	48.08	41.16	39.41	38.86	0.29	4.23	0.13	62.10	44.62	44.62	44.62	44.84	37.92	1.05	37.80	37.80
Actual Volume Flow [ft³/hr]	1,130	1,120	1,120	1,310	1,370	1,390	56,100	3,820	3,470	--	351	157	194	193	2,010	72,600	1,190	1,190
Molecular Weight	182.0	182.6	182.6	182.6	182.6	182.6	113.4	113.4	46.6	18.0	118.4	118.4	118.4	118.4	192.7	192.7	211.0	211.0
True VP at 100 °F [psia]	18.0	14.7	14.7	14.7	14.7	14.7	35.2	35.2	292.0	--	15.8	15.8	15.8	15.8	0.0	0.0	0.0	0.0
Heat Capacity [Btu/lb-°F]	0.480	0.480	0.480	0.625	0.659	0.669	0.534	0.495	0.405	1.031	0.501	0.501	0.501	0.497	0.696	0.666	0.707	0.707
Viscosity [cP]	1.57	1.59	1.59	0.31	0.24	0.22	0.01	0.52	0.01	0.63	0.52	0.52	0.52	0.55	0.19	0.18	0.18	0.18

* Indicates that these streams are, during normal operation, non-steady-state or dynamic and that the flow rates provided here are averaged values spread across time for mass balance purposes.

Table 2-2. Stream Summary Table, Streams 19-35

Stream Number	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33*	34*	35*
Stream Description	Superheated Hydrocarbon Liquid Exiting E-107 Entering T-102	Column Top Vapors Exiting T-102 Entering E-108	Condensate Exiting E-108 Entering V-102	Hydrocarbon Liquid Exiting V-102 Entering P-104A/B	Reflux liquid Exiting P-104A/B Entering T-102	Medium Cut Distillate Exiting P-104A/B Entering E-101	Medium Cut Exiting E-106 Entering E-111	Cooled Medium Cut Exiting E-111 Sent to Adsorption	Column Sump Liquid Exiting T-102 Entering E-109	Boilup Exiting E-109 Entering T-102 Bottom	Heavy Cut Liquid Exiting T-102 Bottom Entering E-102	Heavy Cut Exiting E-102 Entering P-105A/B	Heavy Cut Exiting P-105A/B Entering E-110	Cooled Heavy Cut Exiting E-110 Entering TK-102	Heavy Cut Exiting TK-102 Entering P-106A/B	Heavy Cut Exiting P-106A/B Sent to Disposal	Pyoil Feed Off-gas Vapors Exiting TK-101 to VRU
Pressure [psia]	71.6	20.6	17.1	19.3	78.9	78.9	61.9	54.9	26.7	24.0	26.7	21.3	44.3	27.3	15.0	64.7	14.7
Temperature [°F]	600	539	457	457	457	457	120	100	703	725	725	367	367	100	100	100	100
Phase	Liquid	Vapor	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid/Vapor	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Vapor
Molar Enthalpy [Btu/lbmole]	-127,000	-109,000	-139,000	-139,000	-139,000	-139,000	-178,000	-180,000	-159,000	-133,000	-161,000	-242,000	-242,000	-289,000	-289,000	-289,000	-16,400
Molar Flow [lbmole/hr]	212.7	243.0	243.0	243.0	54.4	188.6	188.6	188.6	167.1	167.1	24.1	24.1	24.1	24.1	24.1	24.1	1.2
Mass Enthalpy [Btu/lb]	-602	-552	-706	-706	-706	-706	-904	-914	-523	-439	-506	-760	-760	-905	-905	-905	-448
Mass Flow [lb/hr]	44,900	47,900	47,900	47,900	10,700	37,200	37,200	37,200	50,700	50,700	7,700	7,700	7,700	7,700	7,700	7,700	46
Component Mass Fraction																	
Nitrogen	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Water	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0566
Ethane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1861
Ethylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1272
Propane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1804
Propylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1900
Butane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0687
C4 Olefins (1-Butene)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1436
1,3-Butadiene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0159
Pentane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0078
Hexane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0077
C6-C7 (NBP 150-249 °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0066
C8-C9 (NBP 250-349 °F)	0.0179	0.0216	0.0216	0.0216	0.0216	0.0216	0.0216	0.0216	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0069
C10-C12 (NBP 350-449 °F)	0.2665	0.3216	0.3216	0.3216	0.3216	0.3216	0.3216	0.3216	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0024
C13-C16 (NBP 450-549 °F)	0.3140	0.3789	0.3789	0.3789	0.3789	0.3789	0.3789	0.3789	0.0016	0.0016	0.0005	0.0005	0.0005	0.0005	0.0005	0.0005	0.0001
C17-C20 (NBP 550-649 °F)	0.2658	0.2719	0.2719	0.2719	0.2719	0.2719	0.2719	0.2719	0.3781	0.3781	0.2366	0.2366	0.2366	0.2366	0.2366	0.2366	0.0000
C20-C24 (NBP 650-749 °F)	0.0950	0.0060	0.0060	0.0060	0.0060	0.0060	0.0060	0.0060	0.5100	0.5100	0.5250	0.5250	0.5250	0.5250	0.5250	0.5250	0.0000
C25+ (NBP 750+ °F)	0.0408	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1103	0.1103	0.2379	0.2379	0.2379	0.2379	0.2379	0.2379	0.0000
Additional Properties																	
Mass Density [lb/ft³]	34.31	0.41	38.56	38.56	38.56	38.56	48.07	48.59	35.00	0.75	34.84	45.27	45.27	52.07	52.07	52.07	0.09
Actual Volume Flow [ft³/hr]	1,310	117,000	1,240	1,240	278	964	774	765	1,450	68,000	221	170	170	148	148	148	502
Molecular Weight	211.0	197.2	197.2	197.2	197.2	197.2	197.2	197.2	303.5	303.5	318.8	318.8	318.8	318.8	318.8	318.8	36.7
True VP at 100 °F [psia]	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	747.3
Heat Capacity [Btu/lb-°F]	0.771	0.635	0.686	0.686	0.686	0.686	0.489	0.478	0.794	0.730	0.803	0.617	0.617	0.470	0.470	0.470	0.410
Viscosity [cP]	0.12	0.01	0.21	0.21	0.21	0.21	1.44	1.74	0.09	0.07	0.07	0.53	0.53	8.40	8.40	8.40	0.01

* Indicates that these streams are, during normal operation, non-steady-state or dynamic and that the flow rates provided here are averaged values spread across time for mass balance purposes.

Table 2-3. Stream Summary Table, Streams 36-52

Stream Number	36*	37*	38*	39*	40*	41*	42	43	44	45*	46*	47*	48	49	50	51*	52*
Stream Description	Combined Off-gas Vapors From Storage Entering V-103	Scrubbed Hydrocarbon Vapors Exiting V-103 Entering C-101A/B	Compressed Light Hydrocarbon Vapors Exiting C-101A/B	Combined Pygas Sent to Ethylene Plant	VRU Knockout Liquid Exiting V-103 Entering P-107A/B	VRU Knockout Liquid Exiting P-107A/B Entering TK-101	Light Cut From Distillation Entering T-103 or T-104 For Demetalization	Demetallized Light Cut Entering T-105 or T-106 For Dechlorination	Decontaminated Light Cut Entering TK-103	Light Cut Exiting TK-103 Entering P-108A/B	Light Cut Exiting P-108A/B Sent to Ethylene Plant as Naphtha	Light Cut Off-gas Vapors Exiting TK-103 to VRU	Medium Cut From Distillation Entering T-107 or T-108 For Demetalization	Demetallized Medium Cut Entering T-109 or T-110 For Dechlorination	Decontaminated Medium Cut Entering TK-104	Medium Cut Exiting TK-104 Entering P-109A/B	Medium Cut Exiting P-109A/B Sent to Ethylene Plant as Gas Oil
Pressure [psia]	14.7	14.2	17.1	17.1	14.9	44.0	47.3	37.3	27.3	14.9	84.7	14.7	54.9	44.9	34.9	15.0	84.7
Temperature [°F]	100	100	113	107	100	100	100	100	100	100	100	100	100	100	100	100	100
Phase	Vapor	Vapor	Vapor	Vapor	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Vapor	Liquid	Liquid	Liquid	Liquid	Liquid
Molar Enthalpy [Btu/lbmole]	-16,800	-16,800	-16,600	-18,000	-153,000	-153,000	-107,000	-107,000	-107,000	-107,000	-107,000	-18,400	-180,000	-180,000	-180,000	-180,000	-180,000
Molar Flow [lbmole/hr]	1.6	1.6	1.6	11.5	--	--	73.1	73.1	73.1	72.8	72.8	0.32	188.6	188.6	188.6	188.6	188.6
Mass Enthalpy [Btu/lb]	-434	-434	-428	-396	-907	-907	-902	-902	-902	-902	-902	-390	-914	-914	-914	-914	-914
Mass Flow [lb/hr]	61	61	61	523	--	--	8,660	8,660	8,660	8,640	8,640	15	37,200	37,200	37,200	37,200	37,200
Component Mass Fraction																	
Nitrogen	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Water	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methane	0.0435	0.0435	0.0435	0.0102	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0102	0.0000	0.0000	0.0000	0.0000	0.0000
Ethane	0.1590	0.1590	0.1590	0.0911	0.0008	0.0008	0.0009	0.0009	0.0009	0.0029	0.0029	0.1197	0.0000	0.0000	0.0000	0.0000	0.0000
Ethylene	0.1045	0.1045	0.1045	0.0478	0.0004	0.0004	0.0003	0.0003	0.0003	0.0010	0.0010	0.0593	0.0000	0.0000	0.0000	0.0000	0.0000
Propane	0.1844	0.1844	0.1844	0.1926	0.0033	0.0033	0.0068	0.0068	0.0068	0.0174	0.0174	0.2101	0.0000	0.0000	0.0000	0.0000	0.0000
Propylene	0.1887	0.1887	0.1887	0.1835	0.0029	0.0029	0.0055	0.0055	0.0055	0.0145	0.0145	0.2066	0.0000	0.0000	0.0000	0.0000	0.0000
Butane	0.0847	0.0847	0.0847	0.1250	0.0052	0.0052	0.0152	0.0152	0.0152	0.0305	0.0305	0.1083	0.0000	0.0000	0.0000	0.0000	0.0000
C4 Olefins (1-Butene)	0.1724	0.1724	0.1724	0.2439	0.0086	0.0086	0.0244	0.0244	0.0244	0.0508	0.0508	0.2184	0.0000	0.0000	0.0000	0.0000	0.0000
1,3-Butadiene	0.0191	0.0191	0.0191	0.0271	0.0010	0.0010	0.0028	0.0028	0.0028	0.0060	0.0060	0.0251	0.0000	0.0000	0.0000	0.0000	0.0000
Pentane	0.0107	0.0107	0.0107	0.0183	0.0021	0.0021	0.0070	0.0070	0.0070	0.0116	0.0116	0.0127	0.0000	0.0000	0.0000	0.0000	0.0000
Hexane	0.0110	0.0110	0.0110	0.0203	0.0068	0.0068	0.0238	0.0238	0.0238	0.0328	0.0328	0.0116	0.0000	0.0000	0.0000	0.0000	0.0000
C6-C7 (NBP 150-249 °F)	0.0096	0.0096	0.0096	0.0182	0.0156	0.0156	0.0552	0.0552	0.0552	0.0707	0.0707	0.0098	0.0000	0.0000	0.0000	0.0000	0.0000
C8-C9 (NBP 250-349 °F)	0.0100	0.0100	0.0100	0.0195	0.1812	0.1812	0.5889	0.5889	0.5889	0.5456	0.5456	0.0074	0.0216	0.0216	0.0216	0.0216	0.0216
C10-C12 (NBP 350-449 °F)	0.0023	0.0023	0.0023	0.0025	0.2713	0.2713	0.2693	0.2693	0.2693	0.2162	0.2162	0.0007	0.3216	0.3216	0.3216	0.3216	0.3216
C13-C16 (NBP 450-549 °F)	0.0001	0.0001	0.0001	0.0000	0.2241	0.2241	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3789	0.3789	0.3789	0.3789	0.3789
C17-C20 (NBP 550-649 °F)	0.0000	0.0000	0.0000	0.0000	0.1859	0.1859	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2719	0.2719	0.2719	0.2719	0.2719
C20-C24 (NBP 650-749 °F)	0.0000	0.0000	0.0000	0.0000	0.0643	0.0643	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0060	0.0060	0.0060	0.0060	0.0060
C25+ (NBP 750+ °F)	0.0000	0.0000	0.0000	0.0000	0.0265	0.0265	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Additional Properties																	
Mass Density [lb/ft³]	0.10	0.10	0.11	0.13	43.09	43.09	44.84	44.84	44.84	44.84	44.84	0.12	48.59	48.59	48.59	48.59	48.59
Actual Volume Flow [ft³/hr]	629	629	537	4,010	--	--	193	193	193	193	193	128	765	765	765	765	765
Molecular Weight	38.8	38.8	38.8	45.5	168.8	168.8	118.4	118.4	118.4	118.7	118.7	47.1	197.2	197.2	197.2	197.2	197.2
True VP at 100 °F [psia]	650.6	650.6	650.6	340.3	14.9	14.9	15.8	15.8	15.8	14.7	14.7	262.6	0.0	0.0	0.0	0.0	0.0
Heat Capacity [Btu/lb-°F]	0.408	0.408	0.415	0.406	0.482	0.482	0.497	0.497	0.497	0.497	0.497	0.400	0.478	0.478	0.478	0.478	0.478
Viscosity [cP]	0.01	0.01	0.01	0.01	1.22	1.22	0.55	0.55	0.55	0.55	0.55	0.01	1.74	1.74	1.74	1.74	1.74

* Indicates that these streams are, during normal operation, non-steady-state or dynamic and that the flow rates provided here are averaged values spread across time for mass balance purposes.

Table 2-4. Stream Summary Table, Streams 53-70

Stream Number	53	54	55	56	57	58	59	60	61	63	63	64	65	66	67	68	69	70
Stream Description	Makeup Nitrogen For Regeneration	Nitrogen Entering C-102	Compressed Nitrogen Exiting C-102 Entering E-112	Partially Heated Nitrogen Exiting E-112 Entering E-114	Superheated Nitrogen Exiting E-114 Used For Regeneration	Hot Nitrogen From Header Entering T-103 or T-104 For Desorption	Hot Nitrogen Header Downstream of T-103 and T-104	Hot Nitrogen From Header Entering T-105 or T-106 For Desorption	Hot Nitrogen Header Downstream of T-105 and T-106	Hot Nitrogen From Header Entering T-109 or T-110 For Desorption	Hot Nitrogen From Header Entering T-107 or T-108 For Desorption	Spent Regeneration Nitrogen From T-103, T-104 T-105, T-106	Spent Regeneration Nitrogen From T-107, T-108 T-109, T-110	Combined Spent Nitrogen Entering E-112	Partially Cooled Nitrogen/ Contaminate condensate Exiting E-112 Entering E-113	Cooled Nitrogen/ Contaminate condensate Exiting E-113 Entering V-104	Dried Nitrogen Sent to C-101	Contaminate Knockout Liquid Sent to External
Pressure [psia]	84.7	20.5	43.0	39.5	36.0	35.5	35.5	35.0	35.0	34.5	34.5	31.5	31.5	31.0	27.5	24.0	20.5	20.5
Temperature [°F]	100	100	231	442	500	500	500	500	500	500	500	500	500	500	290	100	100	100
Phase	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Vapor	Liquid/Vapor	Liquid/Vapor	Vapor	Liquid
Molar Enthalpy [Btu/lbmole]	143	156	1,070	2,580	3,010	3,010	3,010	3,010	3,010	3,010	3,010	3,010	3,010	3,010	1,500	156	156	--
Molar Flow [lbmole/hr]	--	3,404	3,404	3,404	3,404	502	2,902	502	2,400	1,200	1,200	1,004	2,400	3,404	3,404	3,404	3,404	--
Mass Enthalpy [Btu/lb]	5.1	5.6	38	92	107	107	107	107	107	107	107	107	107	107	53	5.6	5.6	--
Mass Flow [lb/hr]	--	95,400	95,400	95,400	95,400	14,100	81,300	14,100	67,200	33,600	33,600	28,100	67,200	95,400	95,400	95,400	95,400	--
Component Mass Fraction																		
Nitrogen	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	--
Water	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Methane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Ethane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Ethylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Propane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Propylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Butane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C4 Olefins (1-Butene)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
1,3-Butadiene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Pentane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Hexane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C6-C7 (NBP 150-249 °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C8-C9 (NBP 250-349 °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C10-C12 (NBP 350-449 °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C13-C16 (NBP 450-549 °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C17-C20 (NBP 550-649 °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C20-C24 (NBP 650-749 °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
C25+ (NBP 750+ °F)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	--
Additional Properties																		
Mass Density [lb/ft³]	0.40	0.10	0.16	0.11	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.08	0.08	0.08	0.10	0.11	0.10	--
Actual Volume Flow [ft³/hr]	--	997,000	587,000	834,000	975,000	144,000	831,000	144,000	687,000	344,000	344,000	334,000	798,000	1,130,000	996,000	852,000	997,000	--
Molecular Weight	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	28.01	--
True VP at 100 °F [psia]	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Heat Capacity [Btu/lb-°F]	0.251	0.250	0.253	0.258	0.259	0.259	0.259	0.259	0.259	0.259	0.259	0.259	0.259	0.259	0.254	0.250	0.250	--
Viscosity [cP]	0.02	0.02	0.02	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.02	0.02	0.02	--

Table 3. Equipment Summary Table

Equipment Summary Table

Heat Exchangers	E-101	E-102	E-103	E-104	E-105	E-106	E-107	E-108	E-109	E-110	E-111	E-112	E-113	E-114					
Type	Hairpin	Multiple Pipe	Double Pipe	Fixed Tube	Fixed Tube	Double Pipe	Multiple Pipe	Fixed Tube	Fixed Tube	Multiple Pipe	Multiple Pipe	Fixed Tube	Fixed Tube	Multiple Pipe					
Area [ft^2]	3,514	454	70	560	477	40	457	221	4,364	428	300	7,615	2,518	278					
Duty [MMBtu/hr]	7.36	1.96	0.61	4.02	4.80	0.028	3.30	7.39	4.24	1.12	0.360	5.14	4.56	1.44					

Shell																			
Design Temperature [ºF]	397	453	802	387	802	147	802	589	802	157	157	340	157	550					
Design Pressure [psig]	163	156	280	56	280	120	280	56	280	120	120	78	120	75					
Phase	Liquid	Liquid	Liquid	Partial Condensing	Liquid	Liquid	Liquid	Condensing	Liquid	Liquid	Liquid	Vapor	Liquid	Vapor					
MOC	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S	S.S	S.S					

Tube																			
Design Temperature [ºF]	507	775	470	157	550	157	650	157	775	417	170	492	340	802					
Design Pres. [psig]	114	62	149	120	62	100	114	120	62	80	107	66	63	280					
Phase	Liquid	Liquid	Liquid	Liquid	Partial Vaporization	Liquid	Liquid	Liquid	Partial Vaporization	Liquid	Liquid	Partial Condensing	Partial Condensing	Liquid					
MOC	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.					

Tanks/Vessels/Towers	TK-101	TK-102	TK-103	TK-104	V-101	V-102	V-103	V-104	T-101	T-102	T-103	T-104	T-105	T-106	T-107	T-108	T-109	T-110	
Design Temperature [ºF]	150	150	150	150	157	507	150	150	550	775	150	150	150	150	150	150	150	150	
Design Pressure [psig]	1.0	1.0	1.0	1.0	52	52	50	59	59	60	59	81	81	81	81	88	88	88	
Orientation	Vertical	Vertical	Vertical	Vertical	Horizontal	Horizontal	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	
MOC	S.S. Clad	S.S. Clad	S.S. Clad	S.S. Clad	S.S.	S.S.	S.S. Clad	S.S. Clad	S.S.	S.S.	S.S. Clad	S.S. Clad	S.S. Clad	S.S. Clad	S.S. Clad	S.S. Clad	S.S. Clad	S.S. Clad	

Size																			
Height/Length [ft]	58	29	32	51	12	18	5.0	23.75	82	76	30	30	30	30	30	30	30	30	
Diameter [ft]	72	37	40	63	4.0	6.0	2.0	9.5	5.0	5.0	10	10	10	10	16	16	16	16	
Internals	Fixed Roof	Fixed Roof	Fixed Roof	Fixed Roof	3-phase Separation with Light Liquid Weir, Demister	--	Demister	Demister	One Tray Pass, 37 Trays, Fixed Valve Rectifying, Bubble Cap Stripping, Bottoms Baffle	One Tray Pass, 34 Trays, Fixed Valve Rectifying, Bubble Cap Stripping, Bottoms Baffle	PuriCycle HP	PuriCycle HP	PuriCycle H	PuriCycle H	PuriCycle HP	PuriCycle HP	PuriCycle H	PuriCycle H	

Pumps/Compressors	P-101 A	P-101 B	P-102 A	P-102 B	P-103 A	P-103 B	P-104 A	P-104 B	P-105 A	P-105 B	P-106 A	P-106 B	P-107 A	P-107 B	P-108 A	P-108 B	P-109 A	P-109 B	C-101 A	C-101 B	C-102
Volumetric Flow [gal/min]	140	140	44	44	148	148	155	155	21	21	276	276	19	19	193	193	763	764	4,400	4,400	122,000
Fluid Density [lb/gal]	6.42	6.42	5.97	5.97	5.05	5.05	5.15	5.15	6.05	6.05	6.96	6.96	5.76	5.76	5.99	5.99	6.50	6.5	0.013	0.013	0.013
Power (shaft) [hp]	16	16	2.6	3.6	7.6	7.6	9.1	9.1	0.9	0.9	15	15	1.0	1.0	15	15	49	50	1.2	1.2	1600
Efficiency (Fluid Power/Shaft Power)	57.5%	57.5%	44.0%	44.0%	59.0%	59.0%	59.0%	59.0%	31.5%	31.5%	65.0%	65.0%	31.5%	31.5%	62.0%	62.0%	74.0%	74.0%	70.0%	70.0%	77.0%
Type	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal
MOC	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	S.S.	C.S.
Drive	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Explosion Proof Electric	Steam Turbine	Enclosed Electric

Capital Cost Estimate

The capital cost of each piece of equipment was determined by utilizing the capital equipment-costing program, CAPCOST. This included using a series of equations found in Turton, et al.[5]

$$\log C_p^0 = K_1 + K_2 \log A + K_3 (\log A)^2 \tag{1}^5$$

The first intermediate calculation included determining the initial purchased cost using *Equation 1*, with coefficients K_1, K_2 and K_3, and a capacity parameter (A). These coefficients are found in the literature and are equipment specific. The capacity factor is also equipment specific as it utilizes the area (m^2) of a heat exchanger while conversely using the volume (m^3) of a distillation tower. The Turton et al. textbook employs SI units for costing correlations; however, for reporting purposes, the final solutions of this project are reported in field units.

$$C_p = C_p^0 F_M F_P \tag{2}^5$$

$$\log F_P = C_1 + C_2 \log P_{Design} + C_3 (\log P_{Design})^2 \tag{3}^5$$

Next, the purchased cost was determined by *Equation 2* as it considers the material of construction (MOC). For this project, the MOC was chosen to be 316 stainless steel for most equipment while stainless steel clad was chosen in cases less prone to corrosion effects to reduce costs. The F_M factor is where this is considered and can be found in literature tables that include equipment and MOC specific identification numbers.[5] The pressure factor, F_P, is found using *Equation 3* which utilizes design pressure rather than operating pressure.

$$C_{BM} = B_1 C_p^0 + B_2 C_P \tag{4}^5$$

The bare module cost, or installed cost, was determined by applying *Equation 4* where B_1 and B_2 are literature constants, while the initial purchased cost (C_p^0) and purchased cost (C_P) were determined earlier. For the sake of preliminary design cost estimations, aspects such as piping, valves, and instrumentation, as seen on the P&ID, are not individually considered. Within the installed cost, these items are indirectly accounted for.

After the installed cost was determined for each piece of equipment, the Chemical Engineering Plant Cost Index (CEPCI) was utilized. This is a critical step regarding economics to ensure accuracy in reporting costs. The factors used to calculate the bare module cost as well as the pressure factor are obtained from literature reports based on 2001 U.S. markets.

$$Cost\ in\ Year\ A = Cost\ in\ Year\ B \left[\frac{CEPCI\ in\ Year\ A}{CEPCI\ in\ Year\ B}\right] \tag{5}^5$$

By applying *Equation 5* as well as a CEPCI value of 394.3 for 2001 and 776.3 for Dec. 2021, the installed costs were adjusted to a 2021 U.S. dollars basis.[6] The costs were then multiplied with a location factor of 0.87.[7] This location factor accounts for the difference in the cost of equipment

in Singapore with reference to the U.S. market. These values are reported in *Table 4* as the installed costs for each respective piece of major equipment.

$$C_{TM} = 1.18 \sum_i C_{BM,i} \qquad (6)^5$$

After obtaining the adjusted installed costs, the Guthrie's method was utilized which is represented as *Equation 6*. This method was chosen over the grassroots method because it was determined that the purification unit was a moderate expansion to the existing ethylene plant which falls under the criteria listed for use of the Guthrie's method. Seen within the 1.18 factor for Guthrie's method is 15% for contingency and 3% for fees. The finalized capital cost estimation utilizing the Guthrie's method was found to be around $28.4 million which can be seen in *Table 4*. Conversely, the grassroots method assumes there is no infrastructure in place and that it will be a from-scratch build including roadways, operator space, and landscape clearing.

Table 4. Estimated Capital Cost of Major Equipment

Capital Cost of Major Equipment	
Towers/Adsorbers (Including Internals)	
T-101	$784,000
T-102	$749,000
T-103	$788,000
T-104	$788,000
T-105	$788,000
T-106	$788,000
T-107	$2,391,000
T-108	$2,391,000
T-109	$2,391,000
T-110	$2,391,000
Heat Exchangers	
E-101	$678,000
E-102	$144,000
E-103	$35,000
E-104	$202,000
E-105	$203,000
E-106	$31,000
E-107	$145,000
E-108	$169,000
E-109	$483,000
E-110	$136,000
E-111	$96,000
E-112	$639,000
E-113	$348,000
E-114	$90,000

Phase Separators/Reflux Drums	
V-101	$71,000
V-102	$168,000
V-103	$20,000
V-104	$490,000
Pumps (Including Drivers)	
P-101 A/B	$105,000
P-102 A/B	$48,000
P-103 A/B	$72,000
P-104 A/B	$75,000
P-105 A/B	$42,000
P-106 A/B	$103,000
P-107 A/B	$42,000
P-108 A/B	$94,000
P-109 A/B	$198,000
Compressors (Including Drivers)	
C-101 A/B	$5,000
C-102	$2,710,000
Storage Tanks	
TK-101	$887,000
TK-102	$273,000
TK-103	$310,000
TK-104	$682,000
Total Capital Cost	**$24,043,000**
Total Capital Cost Plus Fees and Contingency, C_{TM}	**$28,400,000**

As per *Figure 3*, the adsorption columns will be the largest capital cost of this project. This is mainly due to the number of columns required to maintain a continuous operation. There are eight columns in total which includes metal and chloride adsorption towers for both the Light Cut and Medium Cut products from distillation. It should also be noted that the adsorption tower costs do not include the adsorbents, BASF PuriCycle H and PuriCycle HP, as the approximation of these costs will show up in the fixed cost estimate. On the contrary, the two distillation towers, as well as the pump and compressor cost approximations, include the costs of the trays and drivers, respectively.

Displayed in both *Table 4* and *Figure 3*, the costs of the atmospheric storage tanks for the Pyoil feedstock, Light Cut, Medium Cut, and Heavy Cut have been included. To help reduce emissions, a vapor recovery unit (VRU) was implemented to recover the overhead vapors of TK-101 and TK-103. This includes the capital expenses of C-101 A/B, V-103, and P-107 A/B which creates a small up-front expense, especially in relation to the overall project. This allows the

company to remain in compliance with environmental emission regulations. Lastly, it was assumed there was an existing onsite flare; therefore, the capital cost was not estimated.[1]

Capital Costs Breakdown

- Pumps, 2.4%
- Drivers, 2.1%
- Phase Separators,...
- Distillation Columns, 6.4%
- Storage Tanks, 9.0%
- Compressors, 10.0%
- Heat Exchangers, 14.1%
- Adsorption Columns, 52.9%
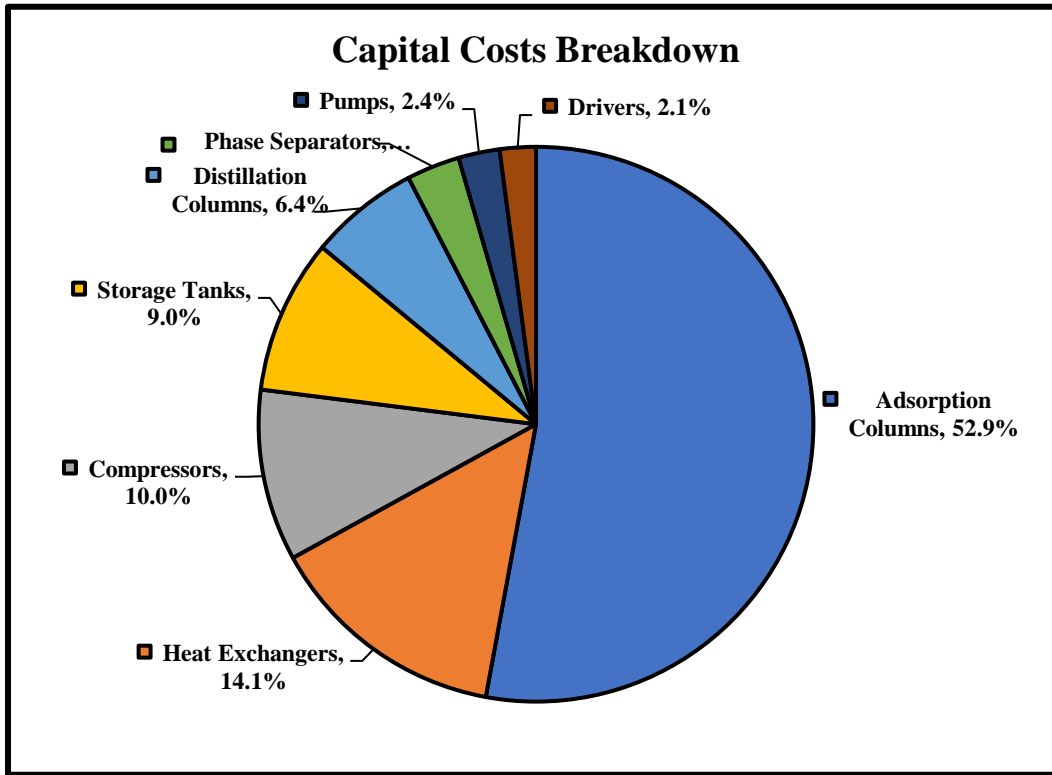
Figure 3. Capital Cost Breakdown Pie Chart

Variable Cost Estimate

The variable operating cost of this project was estimated based on the consumption rates of available utilities and can be subject to changes in production rates, such as periods of upscaling or downtime. Each was approximated using the provided energy costs which are location specific and can be seen in *Table 5*. Based on the location of study, the energy rates are considered high in comparison to typical energy expenses due to the main source of fuel being imported LNG.[1]

Table 5. Utility Cost Per Unit of Energy

Utility Costs[1]		
Utility	**Cost**	**Unit**
Electricity	$0.25	Per kW-hr
Fuel Gas (Hot Oil)	$15.00	Per MBtu HHV
Cooling Water	$0.50	Per MBtu

The heat duties for the heat exchangers were pulled directly from Aspen HYSYS. These values were used to determine the expense estimations for both hot oil and cooling water. *Table 6* displays these totals and separates them based on equipment type and operation. One key feature to mention is that a 97% service factor was utilized when determining operating

expenses. This accounts for downtime, turnaround, and maintenance. The energy requirement for each pump was found by calculating the purchased horsepower which includes motor efficiency. This value was converted to kilowatt-hours.

Table 6. Estimated Annual Variable Operating Costs

Annual Variable Operating Costs	
Distillation	
Heat Exchangers	
Cooling Water	$55,000
Fuel Gas (Hot Oil)	$1,651,000
Pumps	
Electricity	$69,000
Compressors	
Electricity	$2,400
Adsorption	
Heat Exchangers	
Cooling Water	$19,000
Fuel Gas (Hot Oil)	$184,000
Pumps	
Electricity	$16,000
Compressors	
Electricity	$2,683,000
Total Annual Variable Operating Cost	**$4,679,000**

     Figure 4 is a visual breakdown of the utilities used and their associated costs. While *Table 6* is an itemized list as per equipment type and process unit, *Figure 4* is a summation of each utility type. Electricity will be the largest utility expense which stems mainly from the nitrogen regeneration compressor requirement. This is due to the capacity of the compressor itself, and it is costed as if it is operating 24/7 for the desorption of four beds simultaneously. This is considered the worst-case scenario. However, it should be noted that it is unlikely all beds will be regenerated at the same time, meaning that the compressor will likely not be in constant operation. The reason for costing it this way was due to lack of information on regeneration cycle times; therefore, this should be further explored in the detailed design phase. Aside from electricity, the hot oil was estimated as the second largest expense.

Variable Operating Costs Breakdown
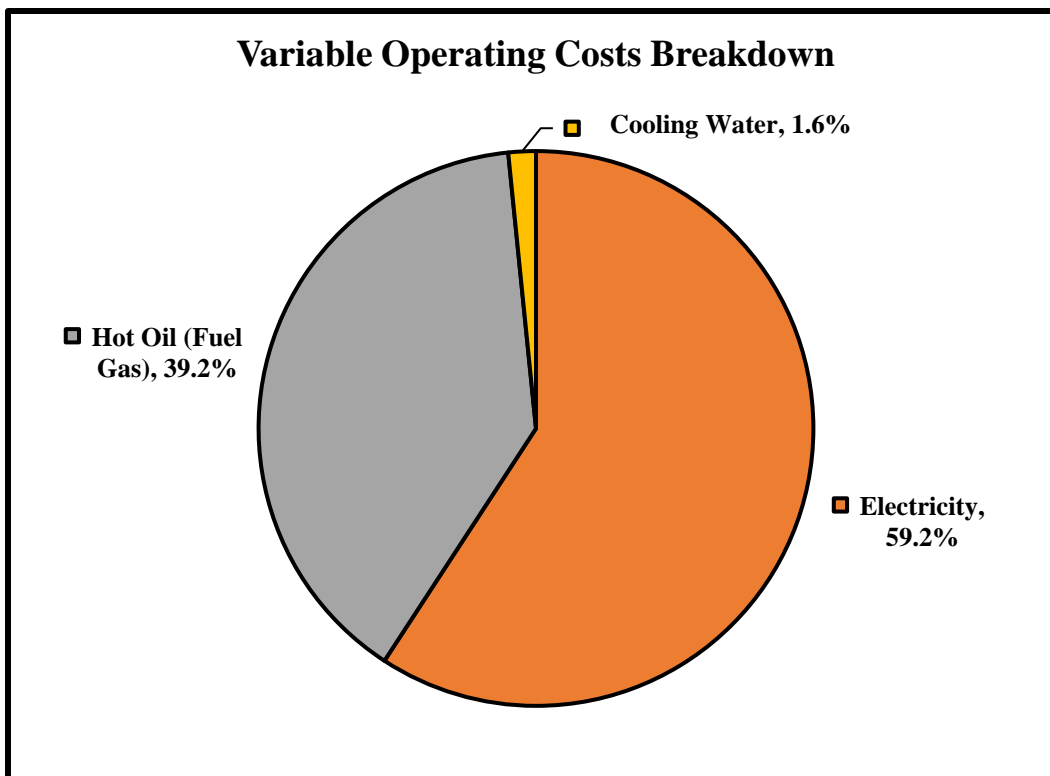
Figure 4. Variable Operating Costs Breakdown Pie Chart

Fixed Cost Estimate

Many expenditures estimated as fixed operating costs were obtained by equations described in Turton et al., which are functions of operating labor and/or capital investment.[5] The number of operators was estimated at being 16 operators, based on the number of major equipment present in the purification unit. With this, it was assumed the location of the plant was to be built in Indonesia where the average annual salary for a petrochemical plant operator is $15,900.[1,8] Therefore, the total cost for operational labor was determined to be $254,000, seen in *Table 7.* The 'Cats and Chems Allowance' portion is the estimated cost of the adsorbents. The desired adsorbent used in the unit for the removal of chlorides was selected to be a zeolite-based adsorbent in comparison to other types such as aluminum oxide (Alumina). Furthermore, it was decided that 13X zeolite can successfully remove the undesired contaminants.[9,10] From this decision, a cost for 13X zeolite adsorbent was acquired from vendors at $2.00 per kg.[11,12] The bulk density for the adsorbent was used with previous column volume calculations to determine the required mass of the adsorbent. This mass value was then used in conjunction with the sourced cost to find the total cost of the adsorbent required for each of the four beds meant to treat chlorides. For the treatment of metal contaminants, it was also determined that a zeolite-based adsorbent was desired in comparison to others on the market; therefore, the same process was used to estimate the cost of these beds. The reason for estimating the cost of adsorbents in this fashion is due to the inability to get in contact with BASF directly. If this project were taken into the detailed design phase and ultimately pursued, direct communication with the manufacturer, BASF, would be required for more accurate cost points. The breakdown of all estimated fixed operating expenses is itemized in *Table 7.*

Table 7. Estimated Annual Fixed Operating Costs

Annual Fixed Operating Costs	
Maintenance	$1,702,000
Cats and Chems Allowance	$1,288,000
Administration	$300,000
Operational Labor	$254,000
Direct Supervisory/Clerical Labor	$46,000
Lab Charges	$38,000
Total Annual Fixed Operating Cost	$3,628,000

Figure 5 visually represents the distribution of the fixed operating expenses within the purification unit. Maintenance and adsorbents are observed to be the largest expenses. Maintenance will generally be high in this area to include regular calibration of the pressure relief valves, implementation of new work orders, and third-party requirements. The adsorbents come with a large price tag; however, they should have a relatively long lifespan due to the quantity of contamination expected, and the assumed ability to extend lifetime through regeneration.

Fixed Operating Costs Breakdown

- Direct Supervisory/Clerical Labor, 1.3%
- Lab Charges, 1.0%
- Operational Labor, 7.0%
- Administration, 8.3%
- Maintenance, 46.9%
- Cats and Chems Allowance, 35.5%
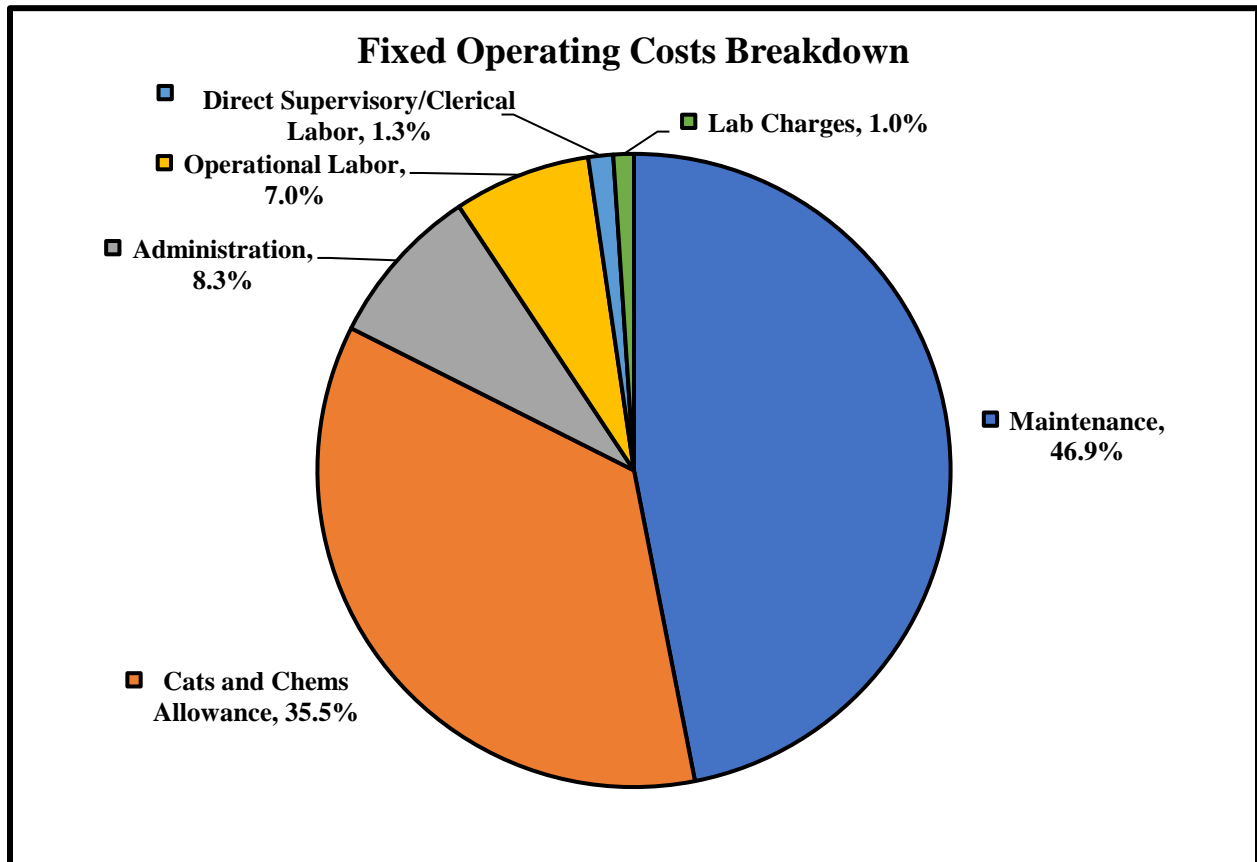
Figure 5. Fixed Operating Costs Breakdown Pie Chart

Minimizing Environmental Impacts

Following the Clean Air Act Amendments from 1970 to 1990, the Environmental Protection Agency (EPA) established the Prevention of Significant Deterioration (PSD) program which requires the instillation of Best Available Control Technology (BACT) to control or limit emissions generated by a unit.[13] In 2011, the EPA signed an agreement with the Indonesian Ministry of the Environment to collaborate and set standards to improve air quality in Indonesia.[14] In compliance with this agreement, the use of fixed roof tanks requires a vapor recovery unit (VRU) that can capture 95% of hydrocarbon emissions that accumulate within these tanks. With the implementation of a VRU, the overall performance of this design is optimized by sending the recovered emissions to the ethylene plant without further treatment. An onsite flare is assumed to be available for the safe removal of excess or hazardous waste products in the event of over-pressurization or during startups and shutdowns.[1] When the flare is in use, the waste products are burned and vented to the atmosphere in the form of carbon dioxide which is less harmful to the environment. A properly designed and operated flare should achieve a combustion efficiency of 98%.[15]

To reduce the risk of loss of containment, stainless steel was selected for the process unit as the material of construction due to its capability to operate under high temperatures and pressures as well as its resistance to corrosion. This is due to the presence of chloride contaminants in the process.[16] *Table 8* lists and compares the compatibility of organic chlorides with carbon steel and stainless steel. From these ratings, it was determined that stainless steel was superior to carbon steel due to its compatibility with said organic chloride compounds.

Table 8. Chemical Compatibility of MOCs with Organic Chlorides

Chemical Compatibility of Organic Chlorides[17,18]		
Compound	Carbon Steel	Stainless Steel
Chloromethane	D	A
Dichloromethane	B	B
Trichloromethane	B	A
Tetrachloromethane	B	A
Chloroethane	D	A
Dichloroethane	D	B
Trichloroethane	B	B
Tetrachloroethane	B	B
Chloroethylene (Vinyl Chloride)	B	A
Dichloroethylene	B	A
Trichloroethylene	B	B
Ethylene Chloride	D	B
Ethylene Dichloride	A	B
Trichloropropane	A	A
Chloropentane	A	A
Benzyl Chloride	D	C
Chlorobenzene	B	A

A = Excellent, B = Good, C = Fair, D = Severe Effect

TK-101
Pyoil Feed
Storage Tank
72' D x 58' H
S.S. Clad
42,000 bbl

P-101 A/B
Pyoil Feed Pump
Centrifugal, S.S.
140 gpm / 16 hp
MAWP: 163 psig

E-101
Pyoil Feed
Preheater 1
Hairpin, S.S.
3,514 ft²
MAWP: 163 psig (shell)
114 psig (tube)

E-102
Pyoil Feed
Preheater 2
Multiple Pipe, S.S.
454 ft²
MAWP: 156 psig (shell)
62 psig (tube)

E-103
Pyoil Feed
Preheater 3
Double Pipe, S.S.
70 ft²
MAWP: 280 psig (shell)
149 psig (tube)

T-101
Light Cut Column
5' D x 82' H, S.S.
Rectifying: Fixed Valve
Stripping: Bubble Cap
MAWP: 60 psig @ 550 ºF

E-104
Light Cut Condenser
Fixed Tube, S.S.
560 ft²
MAWP: 56 psig (shell)
120 psig (tube)

V-101
Light Cut
Reflux Drum
3 Phase Separator, S.S.
4' D x 12' H
MAWP: 52 psig @ 157 ºF

P-102 A/B
Light Cut
Reflux Pump
Centrifugal, S.S.
44 gpm / 4 hp
MAWP: 100 psig

E-105
Light Cut Tower
Thermosyphon Reboiler
Fixed Tube, S.S.
477 ft²
MAWP: 280 psig (shell)
62 psig (tube)

E-106
Light Cut Cooler
Double Pipe, S.S.
40 ft²
MAWP: 120 psig (shell)
100 psig (tube)

P-103 A/B
Medium Cut Tower
Feed Pump
Centrifugal, S.S.
148 gpm / 8 hp
MAWP: 114 psig

E-107
Medium Cut Tower Feed
Super-Heater
Multiple Pipe, S.S.
457 ft²
MAWP: 280 psig (shell)
114 psig (tube)

T-102
Medium Cut Column
5' D x 76' H, S.S.
Rectifying: Fixed Valve
Stripping: Bubble Cap
MAWP: 59 psig @ 775 ºF

V-102
Medium Cut
Reflux Drum
6' D x 8' H, S.S.
MAWP: 52 psig @
507 ºF

E-108
Medium Cut Condenser
Fixed Tube, S.S.
221 ft²
MAWP: 56 psig (shell)
120 psig (tube)

P-104 A/B
Medium Cut
Reflux Pump
Centrifugal, S.S.
155 gpm / 9 hp
MAWP: 114 psig

E-109
Medium Cut Tower
Thermosyphon Reboiler
Fixed Tube, S.S.
4,364 ft²
MAWP: 280 psig (shell)
62 psig (tube)

P-105 A/B
Heavy Cut Pump
Centrifugal, S.S.
21 gpm / 1 hp
MAWP: 80 psig

E-110
Heavy Cut Cooler
Multiple Pipe, S.S.
428 ft²
MAWP: 120 psig (shell)
80 psig (tube)

E-111
Medium Cut Cooler
Multiple Pipe, S.S.
300 ft²
MAWP: 120 psig (shell)
107 psig (tube)

TK-102
Heavy Cut
Storage Tank
37' D x 29' H
S.S. Clad
5,500 bbl

P-106 A/B
Heavy Cut
Disposal Pump
Centrifugal, S.S.
276 gpm / 15 hp
MAWP: 112 psig
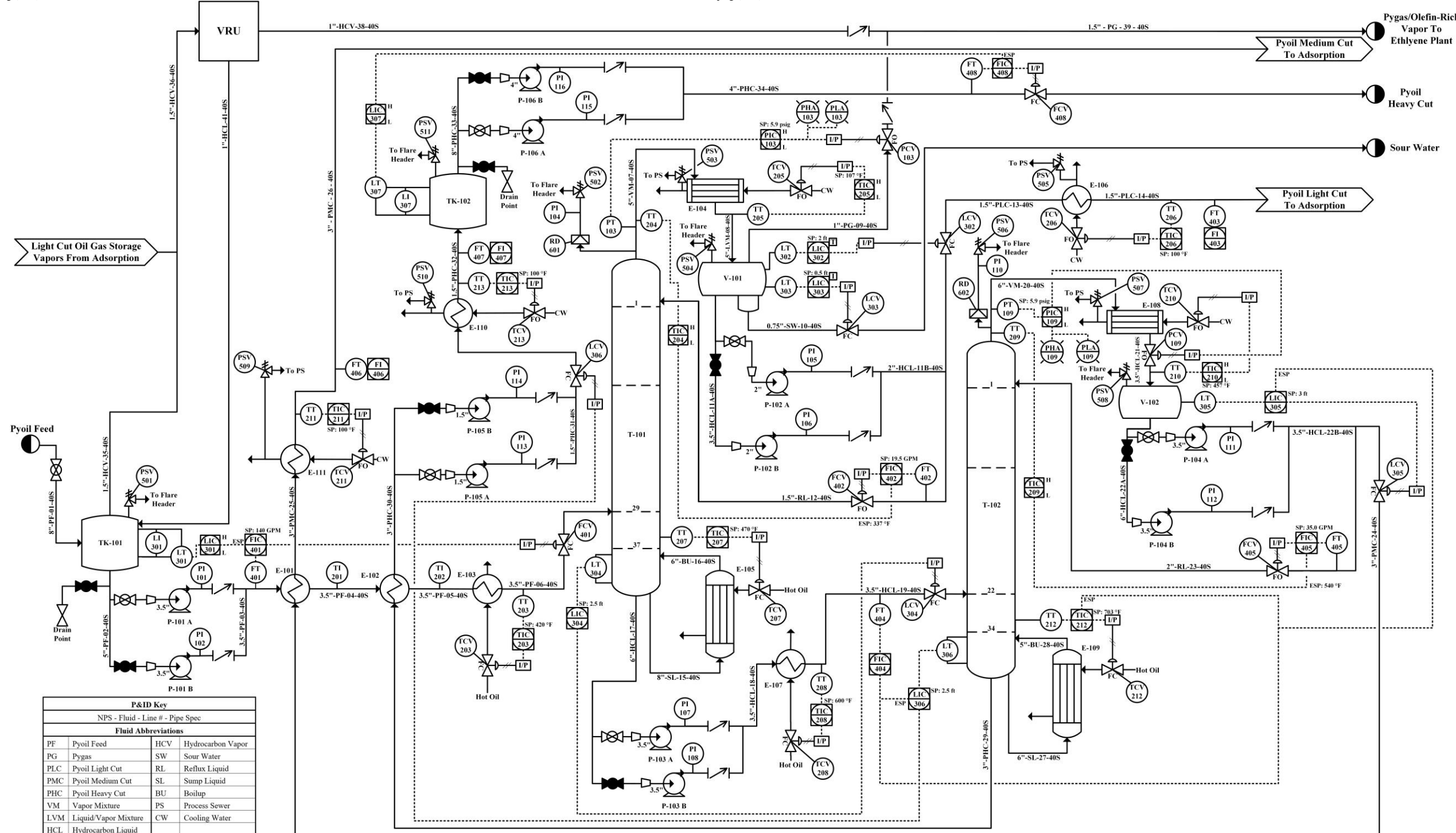
P&ID Key

NPS - Fluid - Line # - Pipe Spec

Fluid Abbreviations

PF	Pyoil Feed	HCV	Hydrocarbon Vapor
PG	Pygas	SW	Sour Water
PLC	Pyoil Light Cut	RL	Reflux Liquid
PMC	Pyoil Medium Cut	SL	Sump Liquid
PHC	Pyoil Heavy Cut	BU	Boilup
VM	Vapor Mixture	PS	Process Sewer
LVM	Liquid/Vapor Mixture	CW	Cooling Water
HCL	Hydrocarbon Liquid		

Figure 6. Distillation Piping and Instrumentation Diagram

Pressure Relief Valve Sizing

Pressure relief is a vital aspect of process unit safety, so proper sizing and relieving considerations are crucial. Therefore, to properly size the pressure relief systems of T-101 and T-102, multiple relief scenarios were considered and simulated. These scenarios were selected by their likelihood of occurrence, and HYSYS Safety Analysis was used to determine the composition, properties, and relieving conditions for each case. These simulations utilized equations and assumptions from the ASME VIII and API 520 codes and standards.[19] For the relief system, a flare was assumed to be onsite and available for the venting and combustion of any process fluids.[1] Additionally, for each column, the process fluid was determined to be fouling and corrosive, which warranted the placement of a rupture disk upstream of the pressure safety valve. Stainless steel was chosen as the MOC for PSVs to avoid the corrosion of the relieving devices and increase reliability of the relief system. Using the maximum allowable working pressure (MAWP) as the set pressure, and the ASME 3% and 10% rules, the built-up backpressure provided by the common relief header was low enough that a conventional valve may be used.

Beginning with T-101, *Table 9* below lists three common scenarios, with the worst case bolded.

Table 9: T-101 Pressure Relief Scenarios

Scenario	Molecular Weight	Relieving Phase	Flow Rate Required (lb/hr)	Orifice Area Calculated (in²)
Fire	125.1	Vapor	34,770	3.799
Cooling Water Failure	127.6	Vapor	36,510	4.300
Reflux Failure	**127.6**	**Vapor**	**36,510**	**4.300**

From these cases, a reflux failure was determined to be the worst scenario. The orifice area was calculated to be 4.300 in², which would fit in the standard valve size of 4.340 in² (size N), but would use 99.41% of its relief capacity, which is higher than the maximum preferred value of 95%.[20] Therefore, the next standard size orifice, 6.380 in² (size P), was selected. This lowers the used relief capacity to 67.63%, which is safer and more flexible. From this orifice size, a standard 4x6 inch valve body was chosen.

The same approach was taken for T-102, and the results are tabulated in *Table 10*.

Table 10: T-102 Pressure Relief Scenarios

Scenario	Molecular Weight	Relieving Phase	Flow Rate Required (lb/hr)	Orifice Area Calculated (in²)
Fire	206.3	Vapor	44,370	3.920
Cooling Water Failure	211.0	Vapor	46,590	4.493
Reflux Failure	**211.0**	**Vapor**	**46,590**	**4.493**

Similar to the pressure relief in T-101, the reflux failure was determined to be the worst scenario. The orifice area calculated was 4.493 in^2, so a 6.380 in^2 (size P) is recommended with a capacity usage of 70.42%. Also, for this case, a 4x6 inch standard valve body was chosen.

Though rupture disks were placed upstream of the PSV, due to process conditions, they were not directly sized and should be analyzed further in the detailed design phase.

Failure Rate Analysis

With Indonesia, Southeast Asia experiencing high humidity, salinity, and vibrational shocks from earthquakes, reliability of process controls is of great concern because these factors may increase instrument failure rates. To understand and manage this risk, the average annual failure rate of the automated controllers and alarmed indicators for both distillation columns are tabulated in *Table 11*. It is also important to note that increased preventative maintenance is required to maintain greater reliability due to the mentioned environmental concerns.

Table 11. Annual Failure Rates for Various Process Instruments

Annual Failure Rates of Automatic Controllers and Alarmed Indicators for T-101 & T-102	
Type	Failure Rate (Per Year)[21]
Temperature Controller	0.097
Flow Rate Controller	1.07
Liquid Level Controller	1.15
Pressure Controller	0.19
Temperature Indicator	0.54
Flow Rate Indicator	9.29
Liquid Level Indicator	0.10
Pressure Indicator	0.033

Personnel Exposure Risk

In compliance with OSHA's Hazard Communication Standard (HCS), available information about the hazardous substances within this design are compiled in *Tables 12, 13,* and *14*.[22] Employees must be trained to properly identify and understand the risks associated with these materials to establish protective measures against them. *Table 12* displays the chemical exposure limits for each component within an eight-hour time-weighted average. For most substances listed, the proper protective equipment of eyewear, chemical-resistant gloves, and appropriate footwear are required. In the case of exposure to asphyxiants, air-fed or self-contained breathing respirators are required.[23]

Table 12. OSHA Chemical Exposure Limits for Process Components

OSHA Chemical Exposure Limits (8-hour TWA)[23]	
Component	**Concentration (ppm)**
Nitrogen	Asphyxiant
Hydrogen	Asphyxiant
Carbon Monoxide	35
Carbon Dioxide	5,000
Methane	-
Ethane	800
Ethylene	Asphyxiant
Propane	1,000
Propylene	500
Butane	800
C_4 Olefins	250
1,3-Butadiene	1
Pentane	1,000
Hexane	500
C_6+	800

TWA: Total Weight Average

Table 13 exhibits the National Fire Protection Association (NFPA) ratings for process components to provide emergency personnel with a general identification of the hazards and their severity in the event of an incident. Most components within this process pose severe flammability hazards, in which appropriate fire-fighting tactics must be used. There are also various health risks if exposed; therefore, the proper protective equipment must be worn.

Table 13. NFPA Ratings for Process Components

NFPA Ratings[23]				
Component	**Health**	**Flammability**	**Reactivity**	**Special Hazard**
Nitrogen	-	-	-	SA
Hydrogen	-	4	-	-
Carbon Monoxide	3	4	-	-
Carbon Dioxide	2	-	-	SA
Methane	1	4	-	-
Ethane	1	4	-	-
Ethylene	1	4	-	-
Propane	2	4	-	-
Propylene	1	4	-	-
Butane	1	4	-	-
C_4 Olefins	2	4	2	-
1,3-Butadiene	2	4	2	-
Pentane	3	4	-	-
Hexane	3	3	1	-
C_6+	2	4	-	-

SA: Simple Asphyxiant Gas

In the event of exposure through inhalation, consumption, or dermal contact, *Table 14* conveys the lethal dose and concentration limits for the components of this process. This toxicological information determines the amount of material needed to kill 50% of the testing species. For gaseous substances, this amount is determined over an observed time interval. In terms of human limits to these components, approximations and safety factors must be utilized.

Table 14. Lethal Dose/Concentration Limits for Process Components

Lethal Dose/Concentration Limits[23]				
Component	**Species**	**LD50/LC50**	**Dose [ppm]**	**Exposure [hr]**
Nitrogen	-	-	-	-
Hydrogen	-	-	-	-
Carbon Monoxide	Rat	LC50	3,760	1
Carbon Dioxide	-	-	-	-
Methane	Rat	LD50 LC50	2,000 20,000	- 2
Ethane	Rat	LC50	658	4
Ethylene	-	-	-	-
Propane	-	-	-	-
Propylene	-	-	-	-
Butane	Rat	LC50	658	4
C_4 Olefins	Rat	LC50	285	4
1,3-Butadiene	Rat	LC50	128,000	4
Pentane	Rat (Oral) Rabbit (Dermal) Rat (Inhalation)	LD50 LC50	>2,000 3,000 364	- - 4
Hexane	Rat (Oral) Rabbit (Dermal)	LD50	>15,000 3,350	- -
C_{6+}	Rat	LC50	490	4

Atmospheric Detonation of Distillation Inventory

Since this design involves hydrocarbon processing, it is important to highlight the risks of an incidental explosion caused by loss of containment. In the worst-case scenario, the instant release of flammable vapors into the atmosphere will undergo a flash vaporization and form a flammable vapor cloud which, when ignited, results in an atmospheric detonation. Therefore, to determine the damage caused by this explosion, a TNT mass equivalency calculation was performed on the largest distillation column in this design, T-101, to gauge the consequence of such an event.

$$m_{TNT} = \frac{\eta \, m \, \Delta H_C}{E_{TNT}}$$

(7)[24]

The explosion efficiency, η, for hydrocarbons varies between 1% and 10% due to the influence of multiple factors, such as composition reactivity, incomplete air/vapor mixing, and the location of where the explosion occurs.[24] Additionally, the release of vapors into the atmosphere would create an unconfined, as opposed to a confined, explosion in this worst-case scenario. Therefore, due to these factors, the explosion efficiency remains within the aforementioned range where an efficiency of 2% was chosen.

The approach taken to calculate the mass of the hydrocarbon, m, within the column was to separate the components into two categories: heavy and light. The mass of the heavy species within the column (hexane – C_{6+}) was determined by multiplying the volume of the liquid within the column by its density. The volume of liquid was estimated by assuming a tray liquid level of 3 in with a 4 ft bottom column liquid level.[25] The mass of the vapor-phase, light components (methane – pentane), within the column was found by multiplying the volume of the vapor present within the column by the density of said vapor. It was then assumed that, as a worst-case scenario, the heavy components will have a heat of combustion value, ΔH_C, equivalent to hexane, and the light components will have a heat of combustion equivalent to methane. From Crowl et al., the heats of combustion and energy of exploding TNT, $E_{TNT} = 2{,}016\ Btu/lb$, were found. [24] Using *Equation 7* and the respective calorific values, an equivalent mass of TNT was calculated for both species and added together for a total equivalent mass of **1.04 tTNT**. Values used for calculations and final values are shown below in *Table 15*.

Table 15. TNT Equivalency Calculation Values for T-101

TNT Equivalency Calculation Values for T-101					
Component	Volume [ft^3]	Density [lb/ft^3]	Mass [lb]	ΔH_C [Btu/lbmole]	m_{TNT} [tons]
Lights as Vapor	1,355	0.603	817.1	1,866	0.087
Heavies as Liquid	255	39.04	9,965	8,967	0.951

Figure 7 visually represents what the consequences would be for a detonation of this magnitude of TNT.[26] It would result in a moderate blast damage radius (5 psi) of 50 m with a light blast damage radius (1 psi) of 120 m. It should also be noted that the radius for thermal radiation, which results in third degree burns, is estimated at being 50 m. An incident such as this would have a significant impact on surrounding production units with a high likelihood of injuries and/or deaths.
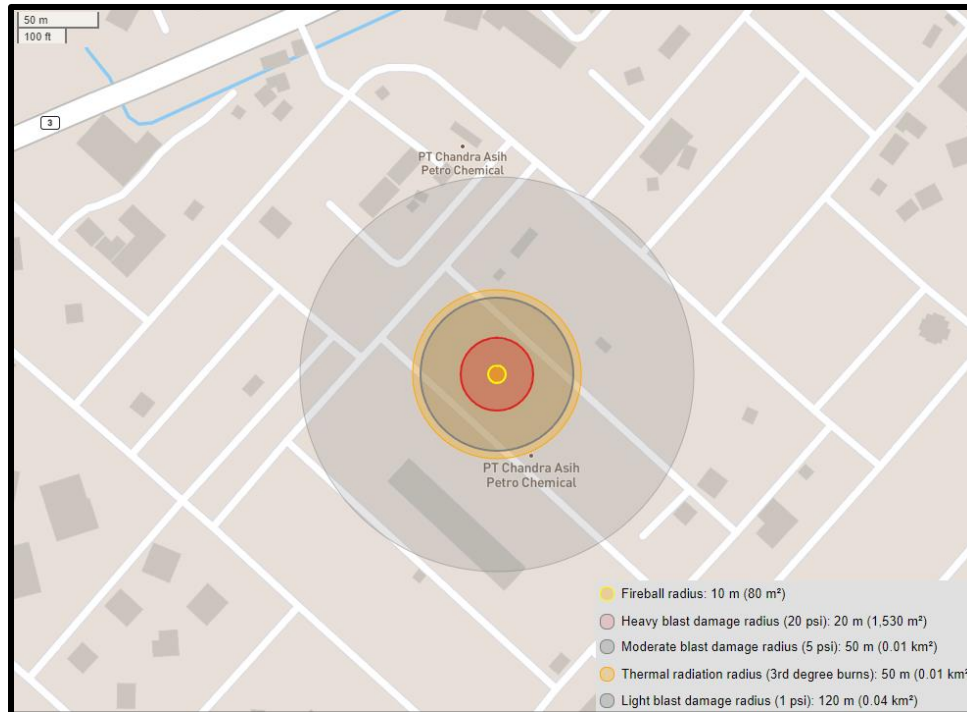
Figure 7. Simulated Blast Radius Based on TNT Equivalency Calculations[26]

Hazard and Operability Study (HAZOP) of The Largest Distillation Column

A Hazard and Operability (HAZOP) study was performed to identify all potential hazards and their consequences within this design. This Process Hazard Analysis (PHA) method follows OSHA's Process Safety Management (PSM) standard which requires employers to recognize and assess the potential hazards in a process that could result in damage to workers, property, or the environment.[27] The main goal of PSM is to prevent the loss of containment of highly hazardous chemicals and to reduce the risk of fires/explosions as they have the largest consequences. *Table 16* exhibits specific risks and the potential consequences on the surroundings.

Table 16. HAZOP Summary

HAZOP Summary[28]						
Hazard	**Equipment Damage**	**Environmental Compliance**	**Loss of Life**	**Disruption of Other Business Units**	**Legal/ PR**	**Community Impact**
Over-pressurization	High	High	Med	Med	High	Med
Fire/Explosion	High	High	High	High	High	High
Loss of Containment	High	High	High	High	High	High
Liquid Slug Flow to Flare	High	High	Med	Med	Med	Med
Loss of Cooling Water	High	High	Med	High	Med	Med

Valve Failure/Seal and Flange Leaks	High	High	Med	Med	Med	Med
Vibrations of Piping and Equipment	High	High	Med	Med	Med	Med
Coke Deposition	High	High	Med	Med	Med	Med
Impurities in Products	Med	Med	Low	Low	Low	Low

RECOMMENDATIONS FOR IMPROVEMENT OF THE BALI SORTING FACILITY

Closing The Quantity Gap

The production of the pyrolysis unit is heavily reliant on the quantity of supply plastic waste. Without a consistent feedstock, the pyrolysis unit will be economically unfeasible. It is crucial that there is a mechanism in place to ensure that the readily available plastic waste is making it to the already in place sorting facility located in Bali, Indonesia. There are currently many plans in place aiming to increase the quantity of waste being collected; some being educational programs, having "trash-pickers," and a household participation cost lower than the global waste management benchmark.[1] Although these have been helpful, it still misses the mark in appropriately incentivizing households and businesses. A possible improvement to increase and stabilize the continuous supply of waste would be to introduce the idea of transitioning the sorting facility into an integrated sorting and collection waste bank. It would allow locals to exchange their waste for monetary reward that is based on the product as well as its weight in kilograms. The client will record their contributions, which could allow for tax breaks in the future. Once the waste is properly sorted by either the customer or the facility, an additional route will be implemented for truckers to haul the flexible plastic waste to recyclers, aggregators, and now also the pyrolysis unit directly. It is also important to examine the activeness of aggregators compared to the less active waste banks in the informal sector. With junk shops handling over 17 tons of waste per day, Global Petrochemicals should consider tendering with aggregators and contracting them to provide recyclables to the pyrolysis unit. This will create a higher source of income for aggregators in the informal sector while giving them a stronger sense of importance within the waste management system.[29]

Closing The Quality Gap

One of the main concerns at the pyrolysis unit is contamination. This stream will ultimately feed into the purification unit where it is impossible to eliminate every known contaminant. Specifically, PVC and metals pose concerns if the contaminants were to make it to the steam cracker as they are detrimental and could shut down the operation if seen in large quantities. For this reason, it is imperative to rid the waste of contamination at the source: households specifically. To encourage households in joining the push to segregate waste, an all-inclusive app should be created. The app would be a 'check-in' platform that allows the residents to keep track of their waste management progress. This would include a feature to scan and track the contributions to the waste bank. Upon meeting progress checkpoints by tallying points for

various activities, the user will receive coupons or vouchers to various local establishments. Users would also get automated graphics that can be shared on social media to encourage others to join. Points can be earned by going to the waste bank, referring others, joining another user in an app reported waste zone, reporting a heavy influx of waste, or spreading the word on popular social media platforms such as Facebook and Instagram. This would be utilizing one of the most powerful aspects of human life today: social media. Whatever is 'trending' tends to be what people are chasing after, so this would allow proper waste management to become 'cool.'

More specifically, waste banks can also provide an educational opportunity to teach communities and raise public awareness about sorting their waste properly. In doing so, they encourage customers to pre-sort their waste before selling it as it will save time and can easily be sent to the pyrolysis unit. With the implementation of this app, residents take on a personal responsibility to achieve certain goals for rewards and, in return, will decrease the risk of exposure to sharp materials or toxic substances to waste bank employees during the sorting and handling process. However, if the waste is not pre-sorted by the customer, it will be carefully sorted before sending it on to be utilized at the pyrolysis unit.

Closing the Affordability Gap

Tourism has become a major economic factor in Bali, Indonesia with urbanization introducing hotels, resorts, and restaurants that cater to foreigners. This brings millions of people in per year which is a massive contribution to the inappropriate disposal of waste. For this reason, it would be a good approach to appeal to tourists and get them involved in the local cleanup efforts. Through partnership with local resorts, one way would be to have a waste management fund in which resort-goers have the option to contribute monetary amounts (nearest dollar round-up, resort fees, suggested contributions) or provide their personal time. With the main accumulations of waste being on the shorelines and in the oceans, water-related tourist attractions could allow waste cleanup in conjunction with resort activities. One example is a resort-offered canoeing session. While the tourists are experiencing the planned canoeing trip, they would be offered a basket and a 'litter-picker' that ensures no direct contact will be made with the left-behind waste. Once the tourists return with the collected waste, they will hand it over to the resort. It is then the resort's duty to take it to the waste bank. Any profit from the waste bank will be split evenly between the tourists and the resort. Tourists can receive monetary compensation in the form of a room credit, gift card, or voucher of their choice. By utilizing the tourists, it will reduce the cost of collection by using already in place operations to collect the waste and, in turn, reduce the requirement of route drivers. This also reduces the wear and tear on the vehicles.

Most residents take their trash to the waste bank due to the personal expense for standard waste management and the opportunity for larger monetary gain. However, households have the option for a waste pick-up service if they pay a fee. This fee will be tailored to pay the drivers; however, the residents will still receive a small cut of what the profit is at the waste bank. The fee is to encourage residents to take the waste to the waste bank themselves, which would lower the cost of the waste collection and the need to operate six days/week. The waste bank operation will overall decrease the current sorting facility cost, which in turn allows them to capitalize on

more revenue in the end. The cost to buy the waste from clients will be less expensive than the collection procedure and, when it is pre-sorted, it will lower the cost and necessity of employees at the facility.

CONCLUSIONS

The purpose of this study was to design a unit for Global Petrochemicals that will purify a stream of pyrolysis oil derived from pyrolyzed plastic waste to meet the specifications for feedstock to an ethylene plant (or steam cracker) located in Southeast Asia. This process perpetuates the idea of a circular economy for plastics through Chemical Recycling which eliminates plastic waste by reusing, creating, and selling new plastic products. This design achieves the sustainability goal of a circular economy by prioritizing process safety, energy efficiency, minimal environmental impacts, maximized reliability, and viable economic performance. Through the incorporation of two distillation towers and eight adsorption columns, this unit fractionates and decontaminates 4,800 bbl/day of pyrolysis oil and produces four specified product-cuts that are sent to the ethylene plant or appropriately disposed of. Tankage was also implemented for one week's worth of storage for all liquid feeds and products. This unit also incorporates a vapor recovery unit for proper emission control. The capital cost of this project was estimated to be $28.4 million with annual estimations of variable and fixed operating costs of $3.64 million and $4.68 million, respectively.

In addition to this design, a "cold eyes analysis" was performed on the sorting facility operation in Bali, Indonesia to generate innovative ideas that would increase the quantity, quality, and affordability gaps of plastic waste collection. Community participation and the quality of the sorted plastic waste are improved through the utilization of technology and tourism which creates a sense of fulfillment for residents and tourists, alike. Additionally, with the implementation of waste banks, local communities are provided with an additional source of income to help grow the local economy while improving overall waste management in the process.

APPENDICES

Appendix A: Adsorption Section Detail

Adsorption takes place after the fractionation unit and is focused on the removal of contaminants such as chlorides and metals. The adsorption unit was designed with eight beds/columns. Half of these columns are dedicated to the decontamination of the Light Cut, while the remaining four decontaminate the Medium Cut. This is to fulfill the requirement of separate processing for each product stream.[1] Additionally, it was assumed that the process of adsorption would be exothermic; therefore, the unit would treat process fluid at near ambient conditions to improve performance.[30] The design utilizes two columns for decontamination of each stream at any point in time, where one utilizes the BASF PuriCycle HP to first remove the metals, and then the second contains BASF PuriCycle H to remove chlorides.[31] The other two towers on each treatment stream are designed identically to the processing columns and are placed in a parallel configuration; however, they are designed to regenerate the adsorbent while

the parallel columns are processing. This includes the assumption that the adsorbents are regenerable by utilizing hot nitrogen at 500 ⁰F in a temperature swing operation. With this unit process configuration, the adsorption towers will be equipped to handle a continuous supply of the Pyoil cuts recovered via distillation without interrupting production. Furthermore, continuous operation in this manner eliminates the need for intermediates.

While processing, the pressure drop across each adsorption tower was assumed to be 10 psi.[30] This equates to a total pressure drop of 20 psi for the process stream, as it will only flow through two towers. The Light and Medium Cut are first treated for metals followed by treatment of chlorides. This order of treatment was assumed based on anecdotal experience that metals as contaminants have the tendency to poison/foul adsorbents used for the removal of chlorides. However, this assumption should be verified in the detailed design by direct interaction with vendors. Another key design consideration is that many of the metal contaminants will leave the distillation unit through the Heavy Cut rather than through the valued Light and Medium Cuts. This is due to the volatility of the metal contaminants present; therefore, it should be noted that, due to this phenomenon, the number of metal contaminants present in the Light and Medium Cuts will be less than what was assumed for this preliminary design. This will increase the breakthrough time for the adsorption of metals allowing for higher reliability with uninterrupted, continuous operation. Upon leaving the chloride adsorption tower, each treated stream is sent to atmospheric tank storage and then ultimately pumped to the ethylene plant.

Stemming from the assumption of regenerable adsorbents, the necessity to recycle and decontaminate nitrogen arose. It was assumed that nitrogen is readily available at battery limits from a process header. This nitrogen utility was assumed to be available at 100 ⁰F and 70 psig as make-up nitrogen is required. Additionally, it was decided that a closed loop regeneration process would be ideal to minimize costs and emissions.[32] Within the closed nitrogen cycle, the nitrogen is compressed and heated to 500 ⁰F and sent to a common header for the beds that are regenerating. After flowing through the regenerating beds, the spent nitrogen stream exits to a return common header. Spent nitrogen is then cooled using a heat integrated exchanger with the compressed nitrogen (E-112) and cooled fully with cooling water, in E-113, before passing through a knockout drum/scrubber (V-104) to remove condensed contaminate liquids. Within the knockout drum, the volatile organic compounds (hydrocarbons), chlorides, and metals are condensed and drawn off as waste. The waste stream will be disposed of to one of the two treatment center options in Singapore where NSL OilChem handles shipping in a turn-key operation while ECO Special Waste Management does not.[3,4] The scrubbed nitrogen is then reused for further regeneration.

Aside from just comfortably handling normal, continuous operation, the adsorption setup can also reliably operate during periods of unusually high levels of contamination. The feedstock for the purification unit can have great variability in levels of contamination due to upstream sorting/quality control before entering the pyrolizer. The feedstock is something that cannot be controlled and, to accommodate this variability, the choice of using a dual bed configuration was seen as beneficial. To mitigate effects of large amounts of contaminants, the cycle times on regeneration/adsorption can be decreased to ensure proper product specifications. Another

consideration is to have back-up adsorbents readily available to replace the beds quickly in the unlikely case of over-contamination.

The material of construction chosen for the adsorption towers is stainless steel clad, while the associated pumps, heat exchangers and compressors are stainless steel. These were chosen as opposed to carbon steel because of the environmental and process conditions. The location of the unit is very close to the ocean, which produces a saline environment that is known to be excessively corrosive to carbon steel.[33] Also, the contaminants within the process, specifically chlorides, have been proven to be corrosive to carbon steel. Stainless steel also mitigates the environmental risks by reducing the chance for loss of containment. A full list of the chemical compatibility comparisons between stainless steel and carbon steel can be found in the earlier section of *Minimizing Environmental Impacts*.

Lastly, the adsorbers were strategically placed at the end of the process (after distillation) to meet criteria in treating the Light and Medium Cut separately and guarantee they meet the ethylene plant required specifications.[1] This selection of treating the contaminated streams after fractionation was due to the unnecessary waste and adsorbent expenditure that would result from treating the Heavy Cut in addition to the Light and Medium Cut. This, combined with the one-week product holdup, validates that the ethylene plant will be able to confidently rely on the designed purification process as a reliable feedstock.

Appendix B: Distillation Section Detail

The distillation portion of the process was designed to have two fractionation towers, T-101 and T-102 where the MOC is stainless steel. The first tower, T-101, is designed to handle the Pyoil in its entirety, which is preheated to 420 ºF, where it will enter the 37 trayed distillation column at tray 29. The trays of the column are structured such that trays 1-28 are valve trays and trays 29-37 are bubble caps. This column separates the Pyoil feed stream into three product streams: Pygas and Light Cut, that exit the reflux drum, and a mixture of the Medium and Heavy Cut, that exits as the bottoms liquid of the column.

The condenser, E-104, uses cooling water to partially condense the overhead vapor from the column, and the two-phase mixture enters the reflux drum/3-phase separator, V-101. The internals of V-101 can be seen in *Figure 8*. The uncondensed vapor in V-101 is the Pygas stream (or Olefin-Rich Vapor) which is used to control the column pressure. This stream is also drawn off and combines with the Pygas that the vapor recovery unit draws out of TK-101 and TK-103, after which it is sent directly to the ethylene plant without further treatment. The liquid portion of V-101 is the Pyoil Light Cut, which is refluxed back into the column to control temperature and is also recovered as the distillate product. This distillate product is sent to adsorption to remove contaminants before being sent to the ethylene plant. Also, there is the possibility that trace amounts of water can be present in the process from the process feed. T-101 is designed so that the heat in the tower vaporizes all the water, which is condensed in the condenser and separated from the hydrocarbons in V-101. The reflux drum has internal weirs to aid this separation and allows for the removal of the free water that has accumulated. *Figure 8* shows how water will accumulate on the bottom of the 3-phase separator because it is denser than the hydrocarbons.

Note, on the PFD and P&ID included for this project, a visual boot is shown, whereas *Figure 8* displays the actual internals.
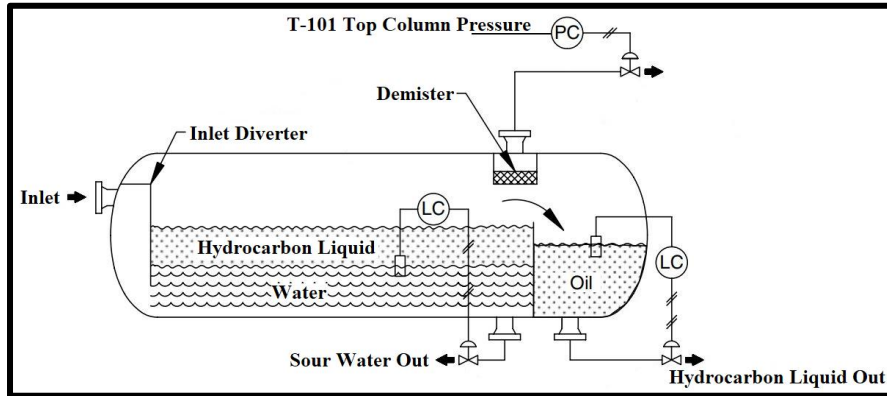
Figure 8. 3-Phase Separator V-101 Internals[34]

A vertical thermosyphon reboiler, E-105, was used to provide the heat and boil-up vapor for T-101. The reboiler operates at a pressure and temperature of 24.3 psia and 500 °F, respectively. The thermosyphon draws liquid off the bottom of the column, seen in *Figure 9*, from a section known as the "sump." This liquid is then sent through E-105, partially vaporizes, and is returned to the column where the liquid portion flows into a partitioned "bottoms" section. These "bottoms" liquids are then drawn off from the column. This stream consists primarily of hydrocarbons with a normal boiling point greater than 392 °F and is pumped through P-103 A/B to a pressure of 78.6 psia.
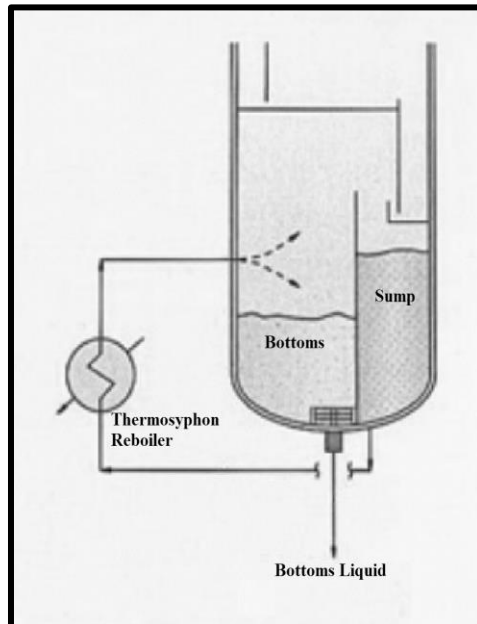
Figure 9. Distillation Partitioned Bottom Column with Baffle Internals[35]

After leaving T-101, the liquid bottoms is preheated via E-107 using hot oil, to a temperature of 600 ⁰F, and fed into the second distillation column, T-102. This tower contains 34 trays, where trays 1 – 21 are valve trays while trays 22 – 34 are bubble cap trays, and the feed

enters the column at tray 22. The goal of T-102 is to separate the mixed feed into the Medium Cut and Heavy Cut. Contrary to T-101, a partially flooded total condenser is utilized where the Medium Cut is recovered as the distillate product and pumped to adsorption for treatment before being sent to the ethylene plant. A vertical thermosyphon reboiler, E-109, is used to provide the heat and boil-up vapor for the column. This is made possible by an identical internal configuration to T-101 and can be observed in *Figure 9*. The bottoms product, Heavy Cut, is drawn off and sent to tankage where it will ultimately be appropriately disposed of. The two determined disposal options for the Heavy Cut are either to be sold to a third-party refinery, such as Singapore Refining Company (SRC), as a contaminated atmospheric residue or to send it to waste treatment facilities in Singapore, such as NSF OilChem or ECO Special.[2,3,4]

The reason for choosing vertical thermosyphon reboilers for both T-101 and T-102 was due to multiple factors. The reboiler is essential for the column to function properly and is the highest operating cost within the distillation process unit, so it must be optimized thoroughly. The reboiler was designed to be a thermosyphon instead of a kettle reboiler, as literature showed it to be common in this process sector and having advantages compared to the frequently modeled kettle reboiler designs. These advantages include lowered capital costs, lower fouling tendencies, and the elimination of a pump.[36,37]

In contrast to the recommended two-tower configuration, a single column configuration was also considered. The single tower was sized and optimized which resulted in 50 theoretical stages. These two arrangements were compared via cost estimation as seen below in *Figure 10*. The two-tower design yielded lower capital, variable, and fixed operating costs. It also provided better separation when compared with a single tower configuration.
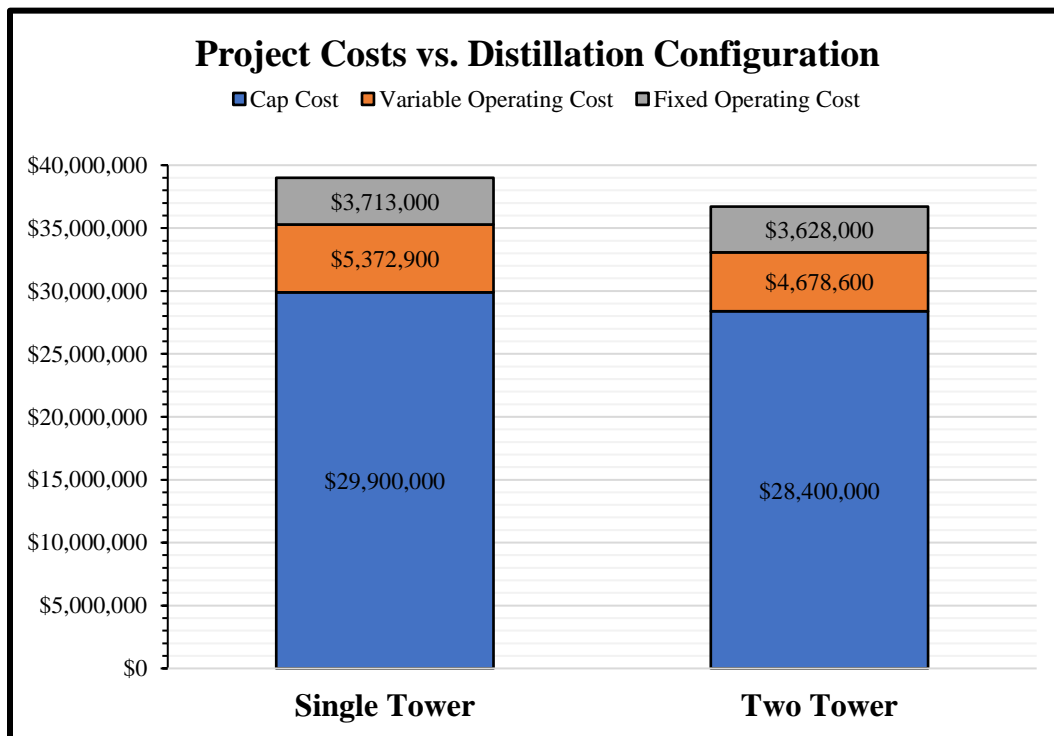
Figure 10: Cost Comparison, Single Tower vs. Two Tower Configuration

This is due to several factors. Firstly, the two-tower configuration has reduced variable operating costs due to the stepwise separation of the cuts, in which T-101 operates at 500 ºF, where the Light Cut and Pygas are drawn off. This means that for T-102, which operates at 725 ºF, the duty is reduced due the lowered mass flow rate. Conversely, a single tower configuration must supply the heat duty required to operate a column in which all the production mass flow rate is processed in one fractionation environment. Additionally, to obtain comparable separation, having one column requires significantly more trays. This leads to a taller column, with the additional capital cost of side strippers and pump-arounds. These side operations are needed to maintain proper separation and column temperature profiles. Apart from just cost considerations, the two-tower design proved advantageous for multiple design considerations. The main one is that it allows for greater ease of control in cases of feedstock variability, since it does not rely on side draws from the tower. This is important because the feedstock is out of the operator's control, as it comes from a pyrolysis unit that processes various types of plastics. Therefore, the feedstock could be an ever-changing percentage of the four cuts.

Column height was of particular concern for the designing of the process, as several factors are at play in the location of study. Firstly, when the L/D ratio becomes larger (20 < L/D < 30), the column design was considered not recommended.[5] This is due to the additional capital cost stemming from the necessity of making the tower larger in diameter or implementing a special support structure. Secondly, having a significantly tall and skinny tower could be a potential risk given the high frequency of seismic events within the region seen below in *Figure 11* as it sits on the infamous fault line known as the "Ring of Fire." From this seismic activity, tsunamis are also of great concern due to their frequency in the region.[38] Aside from seismic activity, the area is also prone to extremely strong winds that pose as another issue for tall and skinny towers.[39] Therefore, splitting the fractionation process into a two-tower configuration allows for the unit to be much more stable, with a more ideal L/D ratio.
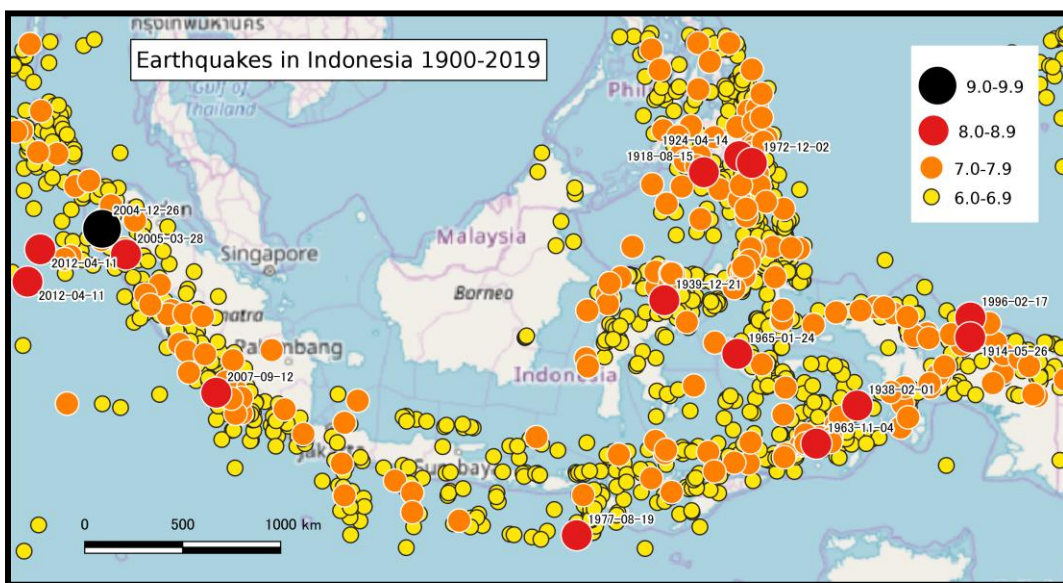

Figure 11. Indonesia Earthquakes 1900-2019[40]

To reduce operating and capital costs, the distillation towers were optimized. This was done through iteration by varying the theoretical number of stages, feed stage location, and feed temperatures. Next, the corresponding N*RR value was found, where N represents the theoretical number of stages and RR represents the reflux ratio. This value was then graphed with respect to the number of theoretical stages as seen for both columns in *Figures 12* and *13* below. This method traditionally yields the best capital cost to operating cost ratio.
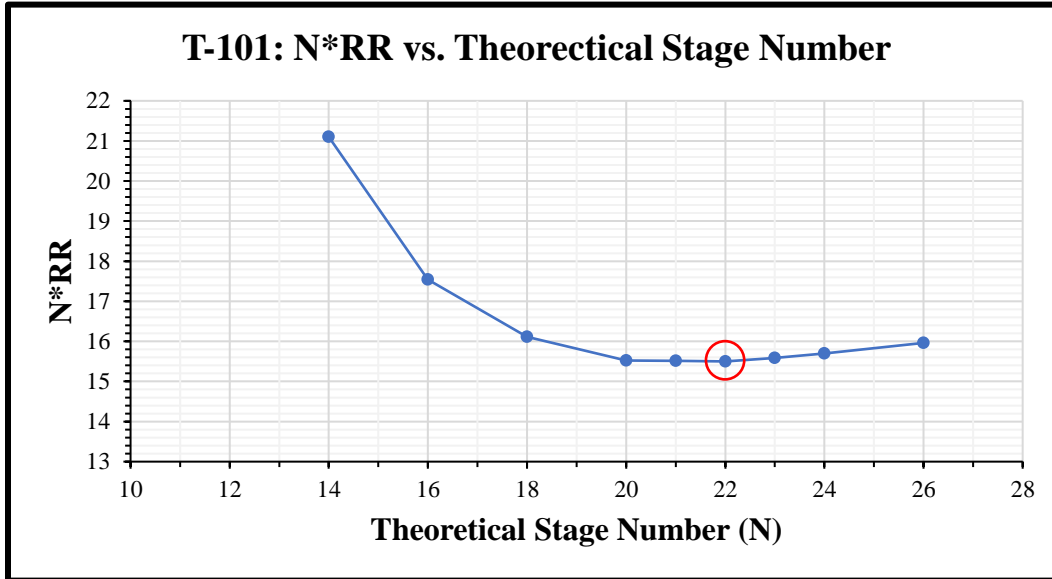
*Figure 12. Optimization of T-101 Using N*RR vs. N Method*
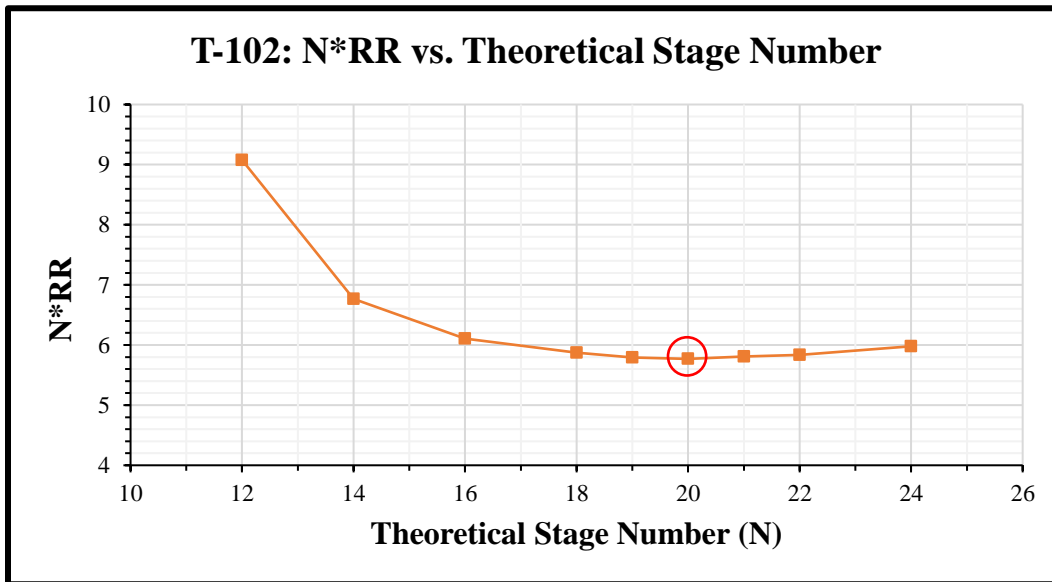
*Figure 13. Optimization of T-102 Using N*RR vs. N Method*

Once the lowest value for N*RR was found, a rigorous method of comparing the molar flow rates of each hypothetical molecule in the cuts to their feed rates was performed to verify that this lowest value had an ideal separation. The Light Cut column, T-101, was optimized by

having 22 theoretical stages with an N*RR of 15.50. The Medium Cut column, T-102, was optimized by having 20 theoretical stages with an N*RR of 5.77.

During this preliminary design study, many assumptions were made with educational and/or literature backing to meet the desired specifications. For modeling purposes, the Peng-Robinson Equation of State (EOS) was used for modeling within the Aspen HYSYS simulation software. This EOS was chosen for its proven success in modeling real, non-polar components.[41] Most hydrocarbons are non-polar and have the heaviest composition presence in the Pyoil feed, while the polar component (water) presence is nearly negligible. Additionally, since the composition of the feed was given in terms of true boiling point (TBP) ranges instead of direct components, modeling the feed required creating "hypothetical" molecules in HYSYS based on the properties and TBP given. Based on hydrocarbon molecule research, it was determined that ~60 hypothetical molecules, with different normal boiling points, would be sufficient to simulate what the true composition would be.[42] For the overall process operation, a 97% service factor was assumed for annual equipment run time, meaning that all equipment was assumed to be shut down for 3% of the year for cleaning, turnarounds, etc. This altered the feed flow rate from 52,432 lb/hr to 54,054 lb/hr.

In relation to the towers, the assumptions made are as follows. Firstly, a 0.1 psi/tray pressure drop was assumed from a heuristic outlined in Turton et. al which provides a mechanism for determining the reboiler pressure and the driving force for vapor flow within the column.[5] The column top pressure was found by assuming a near atmospheric distillation. For the Pygas being sent to the ethylene plant, a parameter of a minimum pressure of 2.4 psig was provided.[1] Using this specification, the pressure in V-101 was taken as 2.4 psig, and a pressure drop of 3.5 psi was assumed through the condenser.[5] From this, the column top pressure of T-101 was then found to be 5.9 psig. Similarly, T-102 was assumed to be slightly above atmospheric distillation to prohibit air ingress into the tower. Therefore, the top column pressure was assumed as 5.9 psig. The tray spacing was also selected based on heuristic values from Turton et al.[5] The simulated tray spacing and subsequent final design were both 2 ft, as it is a common spacing due to maintenance, gives the smallest tower height, and the lowest cost.[5] When modeling the tower through Aspen HYSYS, the theoretical number of stages is what appears in the simulation. This essentially equates to a 100% tray efficiency being assumed, which is not practical. Therefore, the reported number of actual trays in each distillation column on the sized equipment list includes a 60% tray efficiency.[43] To find the height of the columns, the towers were assumed to have a 6 ft section at the base dedicated to liquid disengagement and holdup, with 4 ft space at the top for vapor disengagement. Additionally, it was assumed that both columns sit on a 6 ft skirt.[44]

As previously mentioned, a vertical thermosyphon reboiler was selected over a kettle reboiler design. While this brought advantages, it also resulted in a few challenges that had to be overcome. The primary one is that Aspen HYSYS does not have the capability to directly model a thermosyphon, so a kettle reboiler had to be modeled instead. These two reboilers have fundamentally different operations/setups. In the case of a kettle reboiler, the bottoms are drawn from the reboiler itself. In contrast, a thermosyphon configuration has the bottoms pulled directly

from the column. Without accounting for column internals, this would result in the bottoms stream having different compositions and pressures than if modeled for a kettle reboiler. To mitigate these issues, internals were chosen to mimic the process modeled. The reboiler pressure was modeled in Aspen HYSYS as having the pressure drop according to a 4 ft hydrostatic liquid head that a thermosyphon would have, as the liquid would be taken directly from the bottom of the column and not from the reboiler.

The reflux drums were assumed to be horizontal process vessels with 8-minute holdup and 4-minute surge times according to Svrcek and Monnery.[45,46] An L/D ratio of 3 was also assumed from literature examples, where Svrcek and Monnery state that an ideal range is 1.5 to 3, and heuristic values from Turton et al. give an optimum ratio of 3.[5,45,46] Therefore, 3 was an acceptable value that fulfilled both heuristics. To size the vessel, the separator K value was assumed to be 0.35, and the final diameter and length values were then rounded up to the nearest 3 in.[45,46]

Heat exchangers are also an integral design with several assumptions necessary to accurately model and size them. The first is that the pressure drops across the heat exchangers were assumed to be 3.5 psi as per literature heuristics.[5] Secondly, the overall heat transfer coefficient, U_o, was assumed and averaged from typical ranges found in literature for all necessary fluids.[47] This was a reasonable assumption since the taken value was determined to be representative of the true value, and the alternative was a collective assumption of convective coefficients, thermal conductivity values, and fouling factors, resulting in high uncertainty in a calculated U_o. Thirdly, an approach temperature of 20 ºF was assumed according to Turton et al. to maintain optimal heat transfer between the different mediums.[5] Furthermore, the flow path of each fluid into the heat exchangers (shell or tube side) was selected based on stream properties, composition, and literature recommendations to achieve optimal performance and reduce maintenance.[48] Lastly, all heat exchangers were sized using the LMTD method.

The purpose of the reflux pumps are to transport the liquid exiting the reflux drums back into the columns and to adsorption. They are a vital part of the process and several assumptions had to be made to design them. To determine the pressure at the suction header, the column top pressure was taken from HYSYS, and the pressure drop through the system was traced and calculated. To perform this analysis, the pump was assumed to be at the bottom of the column; and the liquid level of the reflux drum was assumed to be 10 ft above it. Since the pressure of the reflux drum was previously selected, the liquid density was obtained from HYSYS to calculate the hydrostatic pressure on the suction side of the pump and added to this value. Additionally, line losses were assumed on the suction side of the pump as 0.4 psi/100 ft.[5] From these assumptions, suction side pressures were found for both reflux pumps. For the discharge side, the column top pressure was taken from HYSYS, and the hydrostatic pressure and line losses, which were assumed to be 2 psi/100 ft, were added.[5] Furthermore, a pressure drop of 10 psi was assumed for the control valve placed in the line based on Turton et al. heuristics and added to this value.[5] This resulted in the final calculated pressure increase across the pump. For conventional pump sizing, a heuristic of 100 ft between process equipment was utilized for

frictional line loss calculations.[5] Finally, the efficiencies of the reflux and all other pumps were assumed from Peters and Timmerhaus.[49]

The minimization of energy expenditure within a circular plastics economy is of great importance for the sustainability of the sector. Therefore, the two distillation towers were designed and optimized to reduce excessive energy usage. This was first accomplished by utilizing heat integrated heat exchangers. Through extensive research using Turton et al.[5], Katzen et al.[50], and Rossiter[51], this method was found to help reach the highest energy efficiency and minimal energy waste. For this design, the feed is preheated by the heat associated with the fractionated cuts, which allows for the preheating of the feed and the cooling of the products, thus minimizing utilities.

Additionally, the two-tower design minimizes the overall energy consumption by separating the Pygas and Light Cut in the first tower and the Medium and Heavy Cuts in the second tower. The result is that the Pygas and Light Cut aren't excessively heated in the first tower due to the high reboiler temperature needed to achieve effective separation of the Medium and Heavy Cuts, as T-101 only operates at 500 ºF while T-102 operates at 725 ºF. This also saves on cooling water usage as the column condensers do not have to carry a higher heat duty since the overall reflux requirement is minimized. Lastly, as previously mentioned, the N*RR value was used to identify the best balance between the capital and operating costs.

The distillation controls were designed to achieve the desired specifications which are outlined in *Table 17*.

Table 17. Pyoil Product Cuts Specifications

Pyoil Purification Process Products[1]				
Purification Unit Product Stream Name	Pygas	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
Temperature	107 ºF	100 ºF	100 ºF	-
Pressure	Minimum 2.4 psig	70 psig	70 psig	50 psig

To successfully operate the distillation portion of this process and ensure product specification reliability, a comprehensive control scheme was implemented. This includes level control on the storage tanks, with high- and low-level alarms, which work in conjunction with flow controls downstream of each tank to ensure that storage tanks are not allowed to run dry. For instance, when the tank level is registering low then the flow control valve would automatically close. Liquid level control loops are also present on each tower. This is achieved by manipulating the flow rate of the bottoms product leaving the column before it reaches its next destination.

The temperature of the feed into T-101 is directly controlled by a temperature transmitter that is in communication with a handle on the hot oil fed into preheater, E-103. The temperature of the feed into T-102 is also controlled in the same fashion where the hot oil flow into E-107 is

manipulated. Temperature transmitters are also used to control the temperature at the bottom of each tower. This is done by controlling the hot oil supply into the reboilers where more hot oil would induce increased heating, thus higher temperatures in the tower. The column top temperature is also controlled via a temperature transmitter and flow control valve. On both T-101 and T-102, the reflux rate is working in a cascade control loop to maintain a desired temperature at the top of the column. Additionally, there are high- and low-level alarms that indicate the column top temperature, thus alerting operators to possible issues regarding reflux rate. By way of this control, the reflux ratio is maintained as temperature has a direct relation to composition and, with the control loop, it is also related to the flow rate of the reflux itself. E-104, E-106, E-108, E-110, and E-111 all operate using simple temperature feedback control loops. These control loops all manipulate the cooling water supply entering the heat exchanger to maintain and achieve a desired process fluid outlet temperature. The control valves that manage the hot oil utility are set to fail-safe positions which fail close, while any valves controlling cooling water have been configured to fail open.

Column top pressure is crucial to control as overpressure creates a huge concern for vessels. Overpressure events are a leading cause of process industry related incidents and have resulted in many fatalities historically. To ensure regulation and awareness, both a control and alarm system including sirens has been implemented. The pressure in T-101 is regulated by altering the outlet flow rate of the uncondensed Pygas vapors leaving the 3-phase separator, V-101. This is possible because T-101 uses a partial condenser setup. In contrast, T-102 is a total condenser setup, meaning that there is no vapor leaving V-102. For this reason, a flooded condenser design was utilized, E-108, to control the column top pressure. This operates by controlling the flow rate of condensate leaving the condenser, ultimately creating a liquid level inside the condenser that affects the heat transfer area and thus the condensation rate.[52] Additionally, E-108 has a simple temperature feedback control loop that adjusts the flow rate of cooling water based on the outlet temperature of the condensate to maintain proper temperatures.

For additional system responsiveness to a complex and variable feed composition, an advanced cascade control system was implemented. This control accounts for variability in the flow rate of feed into T-102. This variability is then used as an external setpoint to oversee simple feedback control loops used to manage the distillation environment for T-102. This is done by monitoring the feed flow rate into T-102, and if upsets occur then the conditions for the hot oil flow rate into E-109, the rate of bottoms exiting the tower, and the rate of distillate leaving V-102 are all preemptively adjusted.

One important feature of distillation towers is the type of trays they use. Different tray types yield different hydraulics and vapor liquid traffic in the column. Traditionally, sieve trays have been used; however, technological advances have provided different trays, such as fixed valve, moving valve, and bubble cap, that can better handle certain hydraulic conditions. Therefore, to determine the best choice for tray types within each column, vapor liquid hydraulic data from the simulation was used. For both towers, the results indicated low liquid flow and high vapor flows in the rectifying section, with low vapor flow and high liquid flow in the stripping sections. This can be observed in *Figures 14* and *15* below.
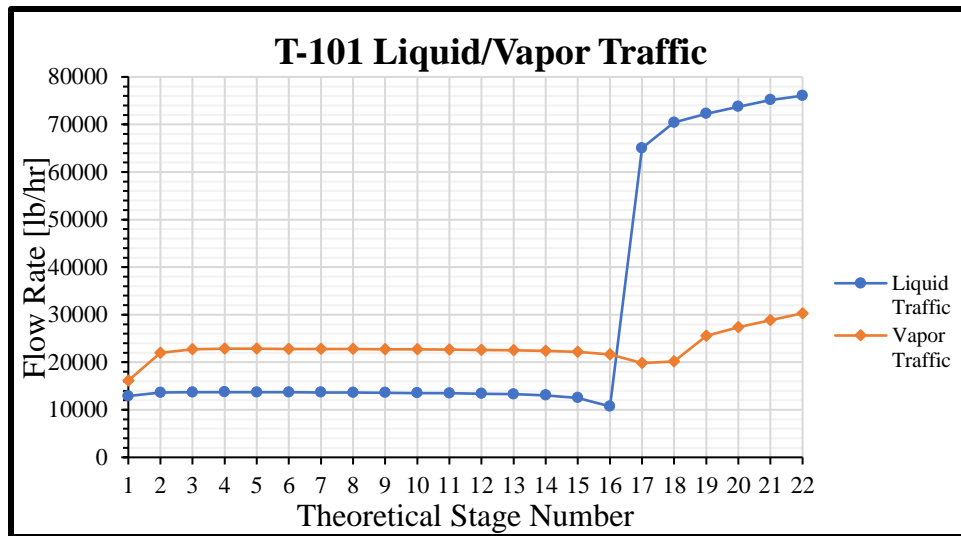
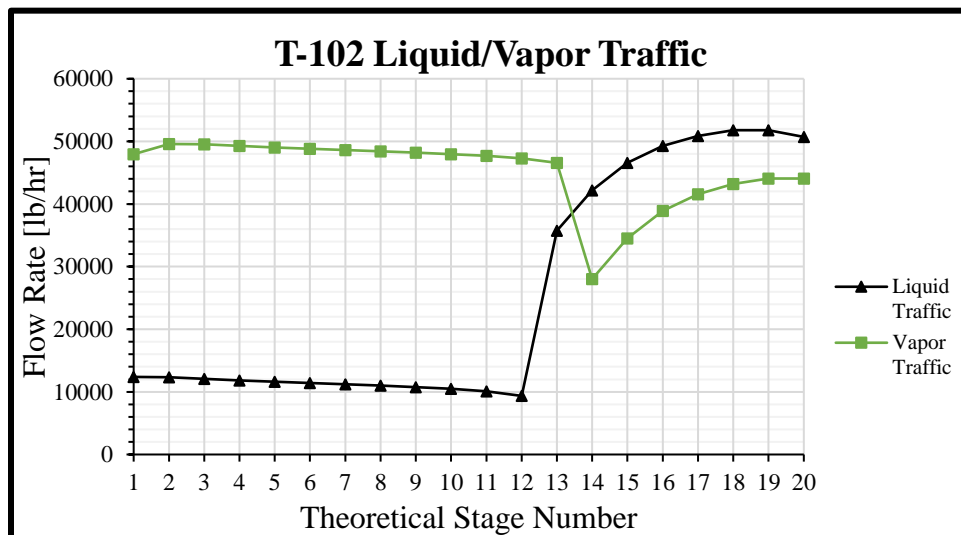*Figure 14. T-101 Vapor and Liquid Traffic*

Figure 15. T-102 Vapor and Liquid Traffic

Due to the low liquid and high vapor rectifying flow, the options were narrowed to either fixed valve, or sieve trays. After comparison, fixed valves were chosen for the rectifying section as they allow for more flexibility in operations, as well as a higher turndown ratio compared to the capabilities of sieve trays.[53] For the stripping section, the high liquid flow rate causes potential concern for weeping to occur, so a tray type was desired that could minimize this potential by better handling low vapor flow rate and considerable turndown. From this, the options were narrowed to either moving valve or bubble cap type trays. With the high potential of fouling within the distillation environment, it was decided that a tray with movable components was susceptible to increased rates of fouling effects and lost efficiency.[54] Therefore, bubble cap trays were selected as the best fit for this operation as they would provide the best flexibility for varying feed compositions and minimize the effects of fouling. Since they are more expensive, making half of the column bubble caps rather than the entire column also helped mitigate costs.[55]

Additionally, all trays within both columns are single pass trays. This was determined through hydraulics data of liquid traffic within the column for each tray type based on flow rates and a desirable amount of weir loading and downcomer flow rate given the geometry of the column internals. Single pass trays were observed to give the best hydraulic performance with proper tray pressure drop and reasonable liquid traffic within both towers.

As mentioned previously, it is important to understand the conditions within the column to accurately model and design a process system. One of these conditions has been mentioned previously, which is the profiles of the liquid and vapor traffic within the columns, found in *Figures 14* and *15* above. The other imperative condition of proper separation is the temperature profiles within the columns, which are found in *Figures 16* and *17* below. An important note about all these profiles is that they are based on the number of theoretical stages in the column. This simulation data was used to determine the desirable internals for the unit.
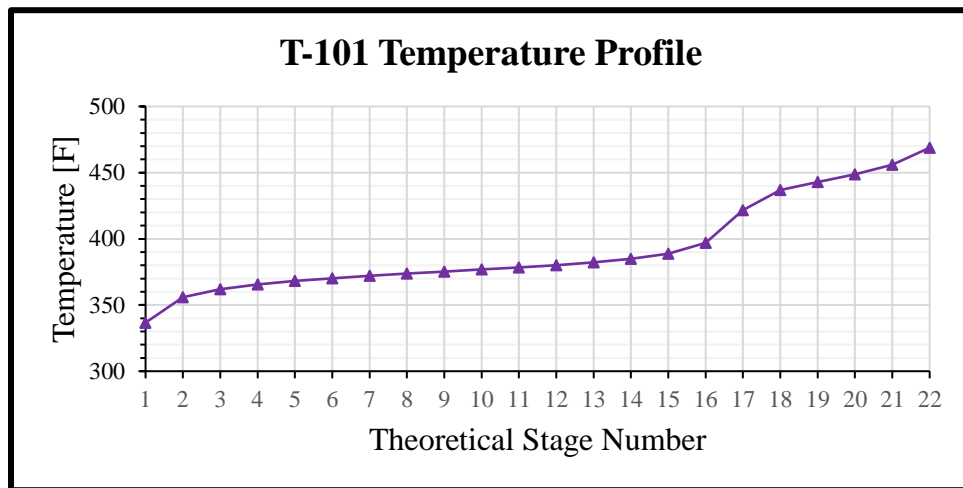

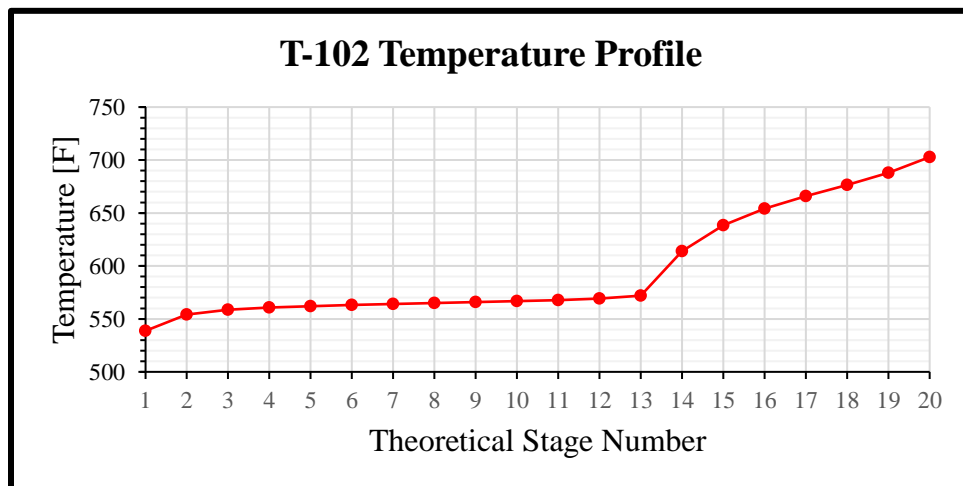*Figure 16. T-101 Temperature Profile*

Figure 17. T-102 Temperature Profile

REFERENCES

1. Yeo G., et al., "2022-2023 Student Design Competition Problem Statement and Rules," AIChE, Dec. 2022

2. "Products And Services," Singapore Refining Company, https://src.com.sg/business/#block2, Accessed Mar. 10, 2023

3. "Industrial Waste," NSL OilChem Waste Management, https://www.nsloilchem.com.sg/our-businesses/industrial-waste, Accessed Mar. 10, 2023

4. "Services," ECO Special Waste Management Pte Ltd, https://www.eco.com.sg/services, Accessed Mar. 10, 2023

5. Turton R., et al., *Analysis, Synthesis, and Design of Chemical Processes,* Pearson, 5th Ed., 2018

6. Maxwell C., "Cost Indices," Towering Skills, May 28 2020, https://www.toweringskills.com/financial-analysis/cost-indices/, Accessed Mar. 10, 2023

7. Nikraftar S., "How To Determine Location Factor," Nov. 15, 2020 http://nikraftar.com/how-to-determine-the-location-factor/#:~:text=A%20location%20factor%20is%20a,built%20in%20the%20base%20location, Accessed Mar. 10, 2023

8. "Chemical Plant Operator," Salary Expert, Mar. 9, 2023, https://www.salaryexpert.com/salary/job/chemical-plant-operator/indonesia, Accessed Mar. 10, 2023

9. Sardar A., " Determination of Best Chloride Adsorbent To Use In Chloride Guards," LinkedIn, May 31, 2016, https://www.linkedin.com/pulse/determination-best-chloride-adsorbent-use-guards-ashish-sardar/, Accessed Mar. 10, 2023

10. Kumar S., et al., "Sorption of HCl from an Aromatic Hydrocarbon Mixture Using Modified Molecular Sieve Zeolite 13X," *American Chemical Society Omega*, Oct. 19, 2021

11. "High Adsorption 3A, 4A, 5A, 13X Synthetic Zeolite, Zeolite Molecular Sieves Desiccant," Xi'an Lyneng Purification Technology Co., Alibaba, https://www.alibaba.com/product-detail/High-Adsorption-3A-4A-5A-13X_1919092754.html?spm=a2700.details.0.0.1ade5903AziHG0, Accessed Mar. 10, 2023

12. "Products: Adsorbent," Jalon, https://www.jalonzeolite.com/products/, Accessed Mar. 10, 2023

13. Hawkins, D. "Guidance for Determining BACT Under PSD" United States Environmental Protection Agency, Jan. 4, 1979

14. *Memorandum of Understanding Between the United States of America and Indonesia*, Department of State: United States of America, Jun. 27, 2011 https://www.state.gov/wp-content/uploads/2019/02/11-627-Indonesia-Environmetal-Coop.pdf

15. "Liquification Plant Best Available Control Technology (BACT) Analysis," Alaska LNG, Apr. 13, 2022, https://dec.alaska.gov/media/25489/liquefaction-plant-bact-4-13-22-rev-3-final-signed.pdf

16. Alanazi N., et al., "Influence of Organic Chloride Concentration on Carbon Steel," Materials Performance, Jan. 6, 2021, https://materialsperformance.com/articles/corrosion-basics/2021/01/influence-of-organic-chloride-concentration-on-carbon-steel, Accessed Mar. 10, 2023

17. "Stainless Steel Chemical Compatibility," CP Lab Safety, https://www.calpaclab.com/stainless-steel-chemical-compatibility-chart/, Accessed Mar. 10, 2023

18. "Carbon Steel Chemical Compatibility," CP Lab Safety, https://www.calpaclab.com/carbon-steel-chemical-compatibility-chart/, Accessed Mar. 10, 2023

19. "Sizing, Selection, and Installation of Pressure-reliving Devices," *API Standard 520*, 9th ed., Jul. 2014

20. Mathias, M. "PSVs Discharging to Atmosphere: What to Consider,", Process Ecology, Aug. 30, 2021, https://processecology.com/articles/psvs-discharging-to-atmosphere-what-to-consider, Accessed Mar. 10, 2023

21. Denson W., et al., "Nonelectronic Parts Reliability Data," IIT Research Institute, 1991

22. "Hazard Communication", Occupational Safety and Health Administration, https://www.osha.gov/hazcom, Accessed Mar. 10, 2023

23. Safety Data Sheets, Air Gas, https://www.airgas.com/, Accessed Mar. 10, 2023

24. Crowl D., et al., "Fires and Explosions," *Chemical Process Safety Fundamentals with Applications*, Second Edition, Prentice Hall PTR, 2002

25. "Tray Components," Separations Processes, http://www.separationprocesses.com/Operations/POT_Chp02e.htm#:~:text=Weirs,used%20for%20low%20liquid%20loads, Accessed Mar. 10, 2023

26. "Nukemap," Nuclear Secrecy, https://nuclearsecrecy.com/nukemap/, Accessed Mar. 10, 2023

27. "Process Safety Management," Occupational Safety and Health Administration, https://www.osha.gov/process-safety-management, Accessed Mar. 10, 2023

28. "Hazards of Oil Refining Distillation Units," British Petroleum, Institution of Chemical Engineers, 2008, http://ndl.ethernet.edu.et/bitstream/123456789/42244/1/13.pdf

29. Johannes H., et al., "Applying the Extended Producer Responsibility towards Plastic Waste in Asian Developing Countries for Reducing Marine Plastic Debris," *Waste Management & Research*, Vol. 39, Issue 5, 2021, pg. 690-702

30. Gabelman A., "Adsorption Basics: Part 1," *Chemical Engineering Progress*, AIChE, July 2017

31. "PuriCycle: Enabling Plastics Circularity," BASF, https://catalysts.basf.com/industries/chemical/adsorbents/puricycle, Accessed Mar. 10, 2023

32. Sircar S., Conrad R., Koch W., *Closed-loop Thermal Regeneration of Adsorbents Containing Reactive Adsorbates,* United States Patent US4971606A, Air Product and Chemical Inc., 1989, https://patents.google.com/patent/US4971606A/en

33. "Salty About Coastal Construction? Use The Right Architectural Metal," Dahlstrom, Dec. 21, 2018, https://blog.dahlstromrollform.com/coastal-construction-correct-metal#:~:text=Stainless%20steel%20is%20the%20most,commonly%20used%20for%20coastal%20applications., Accessed, Mar. 10, 2023

34. Aisyah R., " Several Types of Three-Phase Separators," Chemical Engineering Portal, Dec. 2, 2021, https://missrifka.com/equipments/several-types-of-three-phase-separator.html, Accessed Mar. 10, 2023

35. Ondrey G., " Reboiler Circuits for Trayed Columns: These Vetted Recommendations Are Key To Efficient And Troublefree Separations," Gale Academic OneFile, Jan. 2011, https://go.gale.com/ps/i.do?p=AONE&u=googlescholar&id=GALE|A247530710&v=2.1&it=r&sid=AONE&asid=69f88153, Accessed Mar. 10, 2023

36. Heaslip B., "Heat Exchangers," Queens University, Fall 2008, https://1library.net/document/yr394ndj-heat-exchangers-prepared-by-bob-heaslip-kesco.html, Accessed Mar. 10, 2023

37. *Distillation Handbook*, Akron Varnish Works, 1998

38. "Indonesia Earthquakes and Tsunamis," ShelterBox, https://shelterbox.org/indonesia/#:~:text=Facts%20about%20the%20Indonesia%20Earthquakes,6.0%20magnitude%20almost%20every%20year., Accessed Mar. 10, 2023

39. "Cyclones in Indonesia," World Data, https://www.worlddata.info/asia/indonesia/cyclones.php, Accessed Mar. 10, 2023

40. "Earthquake Map," United States Geological Survey, https://earthquake.usgs.gov/earthquakes/map/?extent=-17.98196,-285.0293&extent=26.97781,-200.6543, Accessed Mar. 10, 2023

41. Carlson E., "Don't Gamble with Physical Properties for Simulations," *Chemical Engineering Progress*, AIChE, Oct. 1996

42. Kusenberg M., et al., "Hydrocarbons: Physical Data," Engineering Toolbox, 2017, https://www.engineeringtoolbox.com/hydrocarbon-boiling-melting-flash-autoignition-point-density-gravity-molweight-d_1966.html, Accessed Mar. 10, 2023

43. Taguchi, H. "Distillation Column Efficiency" PreFEED Corporation, Sep. 4, 2015, https://www.prefeed.com/common/pdf/1509E.pdf

44. "Distillation Tower Elevation and Support," The Piping Engineering World, https://www.pipingengineer.org/distillation-tower-elevation-and-support/, Accessed, Mar. 10, 2023

45. Svrcek W. Y., Monnery W. D., "Successfully Specify Three-Phase Separators," *Chemical Engineering Progress*, AIChE, Sep. 1994

46. Svrcek W. Y., Monnery W. D., "Design Two-Phase Separators Within the Right Limits," *Chemical Engineering Progress*, AIChE, Oct. 1993

47. "Typical Overall Heat Transfer Coefficients," Engineering Page, https://www.engineeringpage.com/technology/thermal/transfer.html, Accessed Mar. 10, 2023

48. "Guidelines for Allocation of Streams In Shell And Tube Exchanger," Equinox, Nov. 9, 2020, https://equinoxsoftwareservices.com/guidelines-for-allocation-of-streams-in-shell-and-tube-exchanger/, Accessed Mar. 10, 2023

49. Peters M., Timmerhaus K., *Plant Design and Economics for Chemical Engineers,* 4th Ed., 1991, Pg. 520-521

50. Katzen R., et al., "Ethanol Distillation: The Fundamentals," Katzen International Inc., 1999

51. Rossiter A., "Improve Energy Efficiency via Heat Integration," *Chemical Engineering Progress*, AIChE, Dec. 2010

52. Sloley A., "Effectively Control Column Pressure," *Chemical Engineering Progress*, AIChE, Jan. 2001

53. "Tray Column and Tray Types," Separation Processes, http://www.separationprocesses.com/Operations/POT_Chp02a.htm, Accessed Mar. 10, 2023

54. Hebert S., Sandford N.,"Consider Moving To Fixed Valves," *Chemical Engineering Progress*, AIChE, May. 2016, https://www.aiche.org/resources/publications/cep/2016/may/consider-moving-fixed-valves, Accessed Mar. 10, 2023

55. "Equipment Costs: Column Trays and Plates," Process Engineering, Jan, 15, 2000, https://processengineering.co.uk/article/1276022/equipment-costs-14-column-trays-and-plates, Accessed Mar. 10, 2023