Letter of Transmittal

Date: March 10, 2023

To: Site Investment Team, Global Petrochemicals

Bali Project Team, Global Petrochemicals

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Subject: Completion of Process Design for Pyoil Purification Unit and Recommendations from “Cold Eyes” Review of Bali Sorting Facility

Enclosed is the successfully completed process design of a pyrolysis oil purification unit for the supply of recycled feedstock to an ethylene plant, also referred to as a steam cracker. As requested, this design purifies 52,432 pounds per hour of raw Pyoil derived from pyrolyzed plastic waste into on-specification feedstock to the Global Petrochemicals ethylene plant while effectively managing all high-consequence process safety risks and minimizing associated costs.

Additionally, enclosed are the recommendations from a “Cold Eyes” review of the Bali Sorting Facility in order to close the three circularity gaps: the quantity gap, the quality gap, and the affordability gap. By improving the facility’s operations to perform reliably and sustainably, the amount of recovered flexible plastics stream will be optimized which in return ensures the security of the feedstock for the Pyoil purification unit.

As per the given requirements, this report includes a cover page, table of contents, table of figures, table of tables, brief process description, economics, process safety, recommendations for improvement of the Bali Sorting Facility, conclusions, appendices, and references. This report will help analyze the various parameters affecting this process design and affecting the Bali Plastic Waste Collection and Sorting Facility.

If you have any questions and/or comments regarding the interpretation of this purification unit process design report or the recommendations for closing the three critical circularity gaps based on the “Cold Eyes” review, please feel free to contact our group.

2022-2023 AIChE Student Design Competition

Closing Critical Gaps to Enable a Circular Plastics Economy

Group 13

March 10, 2023

Executive Summary

Every year, millions of tons of plastics are burned, buried, or leaked to the environment impacting the ecosystems while a very little fraction of global plastic production is recycled. In order to increase the amounts recycled and eliminate waste, the global circular economy must be effectively established by achieving Chemical Recycling, a process to recover and upgrade plastic wastes. In fact, Global Petrochemicals, a leading manufacturer of basic chemicals and high-performance polymers, has committed to produce 10% of its virgin resin-quality plastics from recovered plastic waste using Chemical Recycling. In order to achieve this ambitious goal, Global Petrochemicals is investigating the use of Pyoil, derived from pyrolyzed plastic waste, as a feed source of a steam cracking facility.

In this report, a Pyoil purification unit is designed to purify 52,400 lb/hr of Pyoil feedstock into 165 lb/hr of Py Gas, 9,220 lb/hr of Light Cut, 14,200 Medium Cut, and 28,900 lb/hr of Heavy Cut. The Py Gas is fed directly to the Ethylene Plant while the Light and Medium cuts are sent to the plant as cracking furnace feedstock. The Heavy Cut is recommended to be purified further to be sold in order to increase profit margins. This process design includes distillation and adsorption processes to separate the cuts as well as remove contaminants from the Light and Medium Cuts before being sent to the steam cracker. In addition to the single distillation column and eight adsorption columns, this design consists of two phase separators, seven pumps, four heat exchangers, and four storage tanks which would require the addition of 14 operators. After optimizing the design, costing, and sizing, the capital cost of the unit totaled to be $44.9 million. The annual operating costs are estimated to be $7.5 million US dollars, with fixed annual costs amounting to $8.6 million. In summation, the annual cost of full operation would be estimated at $16.1 million.

To introduce inherent safety into the design, the process was operated at the lowest pressure deemed fit by the process requirements for each cut. The number of distillation columns, as well as operating costs, were minimized to simplify the system and thus further the inherent safety as described in the problem statement. For other safety precautions, a rupture disc in series with a safety relieve valve was added to the distillation column. Pressurized vessels, such as the distillation tower and storage tanks, were minimized when possible to ensure simplicity. The Hazard and Operability Study, Personnel Exposure Risk/Health Impact tables, and TNT Equivalency test were addressed to manage risks in the event of overpressure.

In addition to the design of the Pyoil purification plant, a “Cold Eyes Analysis” of operations in Bali, Indonesia was performed to develop ideas to improve the sorting and gathering of plastics and waste in communities. Innovative solutions were provided to improve the quantity gap, quality gap, affordability gap associated with the collection and sorting facility. Not only does a more efficient gathering and sorting system and infrastructure allow for more feed material for Pyoil, but also will positively influence a more circular economy where waste is used and not discarded. Reducing the amount of plastic waste in the environment will ultimately result in lessening damage to ecosystems and allow companies to be more environmentally conscious of their effects. The three gaps associated with the effectiveness and efficiency of the sorting facility can be improved with the following strategies: encouraging informal recycling sector, gaining revenue from contaminants, installing Near Infrared Optical Sorting Equipment, and implementing waste collection by incarcerated individuals. Total capital and variable costs of the proposed sorting facility innovations are $3.9 million and $354,000 respectively.

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Brief Process Description

Global Petrochemicals, a leading manufacturer of basic chemicals and high-performance polymers, would like to satisfy consumer demands to reduce plastic waste via optimized engineering process designs and societal innovations. In order to fulfill their commitment of producing 10% of their plastics from recovered plastic waste, Global Petrochemical must design and invest in new process technology for Pyoil purification and plastic waste recycling facilities in Bali, Indonesia. To fulfil the outlined goals of Global Petrochemicals, a preliminary design of a Pyoil purification system was created to purify oil from pyrolyzed plastic waste. Fractionated products from the pyrolyzed oil were designed to meet specification requirements and be transported to an ethylene plant in Southeast Asia. Process safety and economic performance were considered when creating and optimizing the preliminary design. In addition to the preliminary design of a Pyoil Purification Unit, strategies for improving the efficiency and effectiveness of a manual sorting and collection facility in Bali, Indonesia were included in final process economics [1].

The Pyoil Purification Unit was designed to produce four fractionated oil products of Py Gas, Pyoil Light Cut, Pyoil Medium Cut, and Pyoil Heavy Cut. As shown in Figure 1, the provided pyrolyzed feed was fractionated into the previously listed products using distillation. The Py Gas was delivered directly to the ethylene plant while the Light and Medium Cuts were processed further to remove contaminants that could potentially damage the steam cracking furnace in the ethylene plant. Free water was removed using knockout drums with a water boot, and other contaminants were removed using adsorption. The decontaminated products were stored in on-site tanks. Due to its high revenue potential, it was suggested that the Heavy Cut product be purified further to be sold by Global Petrochemical to increase profit margins of this design [2].

Safety was considered throughout all stages of the optimized preliminary design. Inherent safety was addressed during initial design when determining material of construction, operating conditions, physical location of units, etc. Other safety considerations included minimizing environmental impacts, designing process and instrumentation controls of the distillation column, pipe sizing, pressure relief design and sizing, process hazard identification, and analyzing atmospheric detonation of distillation inventory [2]. After the preliminary design was optimized considering safety and required specifications, an economic analysis of the capital and variable (operating) costs was generated. Strategies suggested to increase the amount of plastic waste collected and sorted in Bali were economically evaluated and included in a final capital and variable cost estimate of the project. Strategies for the plastic waste facility were suggested based on gaps in the quality of plastic sorted, quantity of plastic collected, and affordability of the facility required for successful operation of this project [3].

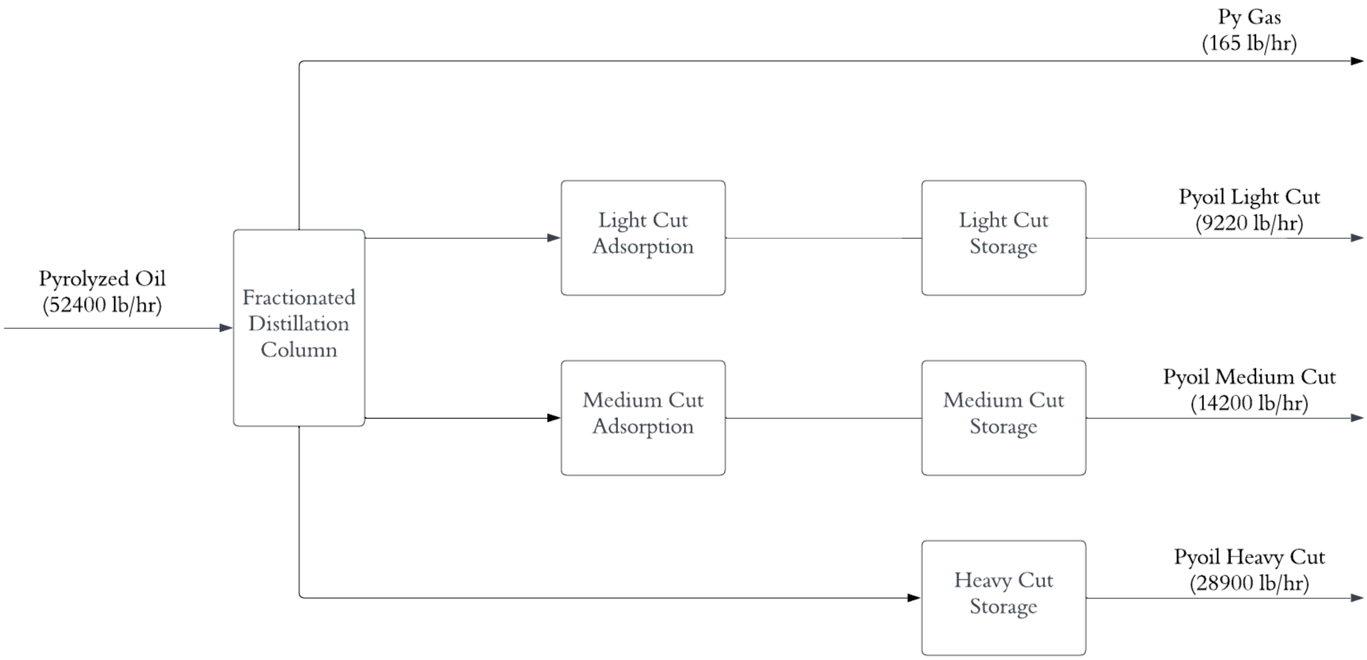

Figure 1. Pyrolyzed Oil Purification Block Diagram [2]

<u>Process Detail</u>

Process Flow Diagram for Process and Utilities areas

Preliminary design of the Pyoil purification unit began with the characterization of the provided raw Pyoil import, labeled Stream 1 in Figure 2. As described in the distillation section located in the Appendix of this report, Oil Manager in Aspen HYSYS Version 10 was used to characterize the feed using the true boiling points (TBP) and compositions located in Table 1. The distillation column, T-101, was designed and optimized to reduce capital costs, utility costs and risks of operation while meeting the required product specifications listed in Table 2. The detailed design considerations and calculations used to produce the converged distillation simulation are described in detail in the Appendix section of this document [2].

Operating temperatures of the condenser and reboiler were determined by provided utility temperatures, Table 3, and EBP of the Light and Medium Cuts respectively. The cooling water return temperature and required Py Gas temperature at 107°F was a consistent operating temperature for the condenser that provided separation of the Light Cut without other contaminants. Operating temperature for the reboiler was set to a max of 620°F in order to separate the Medium and Heavy Cuts without the Heavy Cut evaporating and contaminating the Medium Cut product [2].

Table 1. Raw Pyoil Import Streams from Pyrolyzer to Purification Unit

Mass flow rate, lb/hr	52,432
Temperature, °F	100
Pressure, psig	Defined by Vapor Pressure
Density, lb/ft^3	49.1
Molecular Weight	182.0
Phase	Liquid
Composition wt%	
Nitrogen	
Hydrogen	
Carbon Monoxide	
Carbon Dioxide	
Methane	0.01
Ethane	0.10
Ethylene	0.05
Propane	0.29
Propylene	0.26
Butane	0.36
C4 Olefins	0.62
1,3-Butadiene	0.07
Pentane	0.13
Hexane	0.40
C6+	97.69
TBP (True Boiling Point) at 760 MM HG (wt):	
IBP	-46 °F
5%	289 °F
10%	324 °F
30%	403 °F
50%	483 °F
70%	555 °F
90%	658 °F
95%	717 °F
EBP	844 °F

Table 2. Pyoil Purification Process Products and Stream Identification

Product Stream Name	**Py Gas**	**Pyoil Light Cut**	**Pyoil Medium Cut**	**Pyoil Heavy Cut**
Steam Cracker Feed Name	Olefin-rich Vapor fed directly to the Ethylene Plant	Naphtha (Cracking Furnace Feedstock)	Gas Oil (Cracking Furnace Feedstock)	Not suitable for Steam Cracker
PFD Stream Number	3	8	14	18
End boiling point (EBP °F)	N/A	392	620	--
Temperature (°F)	107	100	100	100
Pressure (psig)	Minimum 2.4	70	70	50

Table 3. Utilities Available

Utility	Pressure (psig)	Temperature (°F)
Cooling Water Supply	70	87
Cooling Water Return	40	107
Thermal Fluid Hot Oil Supply	230	750
Thermal Fluid Return	200	725

The distillation column was designed and optimized, specifications of each cut were met using various engineering assumptions and safety considerations. For instance, it was specified that free water must be removed from the light and medium cuts before they travel to the steam cracker; however, free water is also harmful for the adsorbent materials in T-102 to T-109, so free water was removed using V-101 and V-102 before adsorption. Free water was removed using a water boot attached to the reflux drum and knockout drum for the light and medium cuts respectively. Based on the separation properties of the free water and Pyoil, it was assumed that free water would enter the boot and then be processed as wastewater, labeled Utility Connection 1 in Figure 2. Specific assumptions regarding the design and sizing of adsorption columns are in the Adsorption section of the Appendix [1-4].

Material Balances

Material balances of the streams, Table 4, include the operating conditions and properties required to ensure safe operations and logical unit position. As shown in Table 3, mass flow of the Py Gas, Light, Medium, and Heavy Cut products are consistent with material feed mass flow [5].

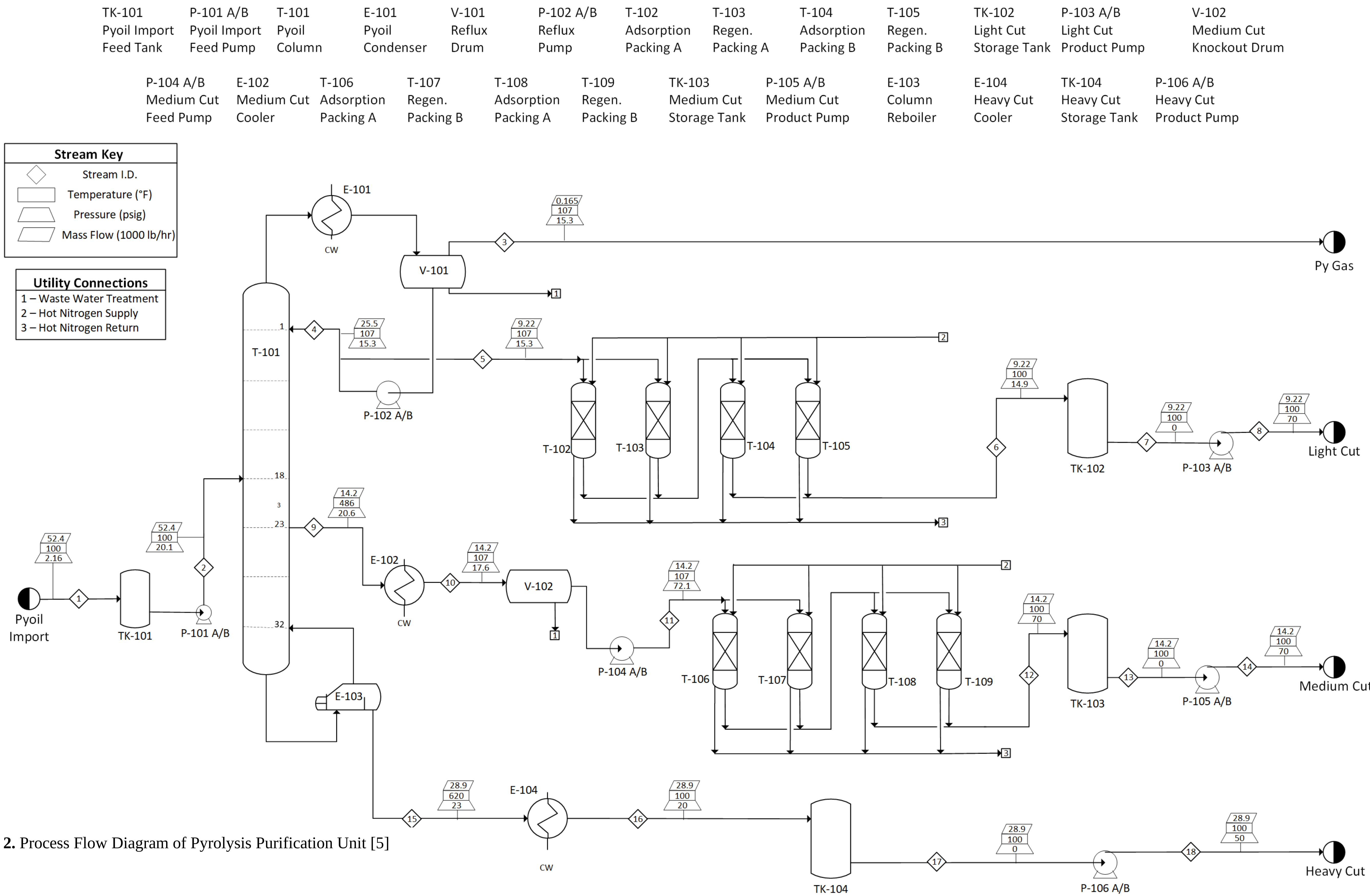

Figure 2. Process Flow Diagram of Pyrolysis Purification Unit [5]

Table 4. Flow Summary Table for Pyrolysis Oil Process Shown in Figure 2

Stream Number	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Temperature (°F)	100	100	107	107	107	100	100	100	486	107	107	100	100	100	620	100	100	100
Pressure (psig)	2.16	20.1	15.3	15.3	15.3	14.9	0	70	20.6	17.6	72.1	70.0	0	70.0	23.0	20	0	50
Vapor Fraction	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow (lb/hr)	52,400	52,400	165	25,500	9,220	9,220	9,220	9,220	14,200	14,200	14,200	14,200	14,200	14,200	28,900	28,900	28,900	28,900
Mole Flow (lbm/hr)	288	288	4.22	296	78.2	78.2	78.2	78.2	83.4	83.4	83.4	83.4	83.4	83.4	122.3	122.3	122.3	122.3
Enthalpy (Btu/lb)	-914	-914	-529	-915	-915	-915	-915	-915	-688	-688	-688	-688	-688	-688	-588	-588	-588	-588
Density (lbm/ft^3)	0.264	0.264	0.197	0.379	0.379	0.379	0.379	0.379	0.213	0.213	0.213	0.213	0.213	0.213	0.148	0.148	0.148	0.148
Mass Percent (%)																		
Methane	0.01	0.01	6.58	0.04	0.01	0.01	0.01	0.01	0	0	0	0	0	0	0	0	0	0
Ethane	0.1	0.1	22.2	1.18	0.26	0.26	0.26	0.26	0	0	0	0	0	0	0	0	0	0
Propane	0.29	0.29	21.2	3.89	1.22	1.22	1.22	1.22	0	0	0	0	0	0	0	0	0	0
Ethylene	0.05	0.05	13.9	0.48	0.11	0.11	0.11	0.11	0	0	0	0	0	0	0	0	0	0
Propylene	0.26	0.26	22.4	3.49	1.05	1.05	1.05	1.05	0	0	0	0	0	0	0	0	0	0
Butane	0.36	0.36	9.77	4.26	1.79	1.79	1.79	1.79	0	0	0	0	0	0	0	0	0	0
Hexane	0.40	0.40	0.65	3.40	2.25	2.25	2.25	2.25	0	0	0	0	0	0	0	0	0	0
Pentane	0.13	0.13	0.74	1.34	0.72	0.72	0.72	0.72	0	0	0	0	0	0	0	0	0	0
1,3-Butadiene	0.07	0.07	1.78	0.92	0.35	0.35	0.35	0.35	0	0	0	0	0	0	0	0	0	0
C4 Olefins	0.62	0.62	0.38	6.05	4.37	4.37	4.37	4.37	0	0	0	0	0	0	0	0	0	0
C6+	97.7	97.7	0.27	74.6	87.9	87.9	87.9	87.9	100	100	100	100	100	100	100	100	100	100

Sized Equipment List

As shown below, Table 5 is a detailed list of all sized major equipment utilized in the system as well as any essential design parameters for the equipment. Using the optimized Aspen HYSYS V10 distillation column simulation, the distillation tower was sized as discussed in detail in the *Distillation Section Detail* in the "Appendix" section. For sizing the heat exchangers, the heat duty was provided by the HYSYS simulation. The condenser and coolers are designed as counter current shell and tube heat exchangers. This design produces a higher driving force for effective heat transfer and minimal equipment area. A kettle reboiler was decided upon due to its connection to the distillation column and its cost-effective nature. For all heat exchangers, the LMTD method and the specifications of utilities were used to calculate sizing [2]. Pumps were sized by calculating discharge and upstream pressures while accounting for pressure drops across any relevant processes, such as heat exchangers, valves, and adsorbers. All pumps were designed to be electric and centrifugal single stage based on the capacity and total head of each [5]. The pumps were also oversized by 10% to account for unprecedented upsets and to maintain steady state operations. When designing the firewater pumps, the necessary size of the firewater pump was assumed to be approximately equal to the size of the largest pump in the system, the feed pump. All storage vessels were designed to accommodate a week hold up time, while the water knockout drum and reflux vessel were designed based on a five minute hold up time [4, 5]. The reflux drum was designed as a horizontal two-phase separator using Svrcek's and Monnery's methods in "Design Two-Phase Separators Within the Right Limits" [4]. The water knockout drum was sized with considerations of retention time of water droplets in medium cut oil to remove free water [6]. All equipment that encountered the Light Cut or the Py Gas from the column were designed using Stainless Steel or Clad Steel to prevent corrosion and fouling from the volatile components, while Carbon Steel was deemed sufficient for the Medium and Heavy cuts. Lastly, the sizing and design of the Adsorption systems are explained in further detail in the "Appendix" section.

Table 5. Equipment Design List

Heat Exchangers	**E-101**	**E-102**		**E-103**		**E-104**
Type	Shell and Tube	Shell and Tube		Kettle Reboiler		Shell and Tube
Area (ft^2)	740	154		975		402
MOC	SS	SS		CS		CS
Process Stream Side	Shell	Shell		Shell		Shell
Heat Duty (MMBtu/Hr)	12.3	3.23		24.9		9.36
Design Pressure (psia)	80	85.3		87.7		87.7
Pump	**P-101 A/B**	**P-102 A/B**	**P-103 A/B**	**P-104 A/B**	**P-105 A/B**	**P-106 A/B**
Type	Centrf./Electric	Centrf./Electric	Centrif./Electric	Centrf./Electric	Centrif./Electric	Centrf./Electric
MOC	SS	SS	SS	CS	CS	CS
Shaft Power (kW)	2.24	3.73	1.5	1.5	2.24	2.24
Flow rate (lb/hr)	52,400	25,500	9,220	14,200	14,200	28,900
Inlet Pressure (psig)	2.16	17.5	0	17.6	0	0
Outlet Pressure (psig)	20.1	45.3	70	72.1	70	50
Efficiency (%)	75	75	75	75	75	75
Towers			**T-101**			
Theoretical Stages			22			
Feed Tray			11			
Medium Cut			16			
Tray Type			Sieve			
Tray Efficiency (%)			70%			
Number of Trays			32			
Tray Area (ft^2)			46.2			
Diameter (ft)			8			
Height (ft)			74			
L/D			9.25			
Tower Area (ft^2)			1,860			
Tower Volume (ft^3)			3,720			
Reflux Ratio			3.75			
Design Pressure (psia)			83			
MOC			SS Clad			
Vessels/Tanks	**TK-101**	**TK-102**	**TK-103**	**TK-104**	**V-101**	**V-102**
Type	Storage Tank	Storage Tank	Storage Tank	Storage Tanks	Reflux Drum	Knockout Drum
Orientation	Vertical	Vertical	Vertical	Vertical	Horizontal	Horizontal
MOC	SS Clad	SS Clad	CS	CS	SS Clad	CS
Liquid Holdup Time	1 Week	1 Week	1 Week	1 Week	5 mins	5 mins
Length (ft)	39.4	25.7	29.5	41.5	23.8	12.5
Diameter (ft)	3	7	9	11	4.75	5
L/D	3.03	3.67	3.27	3.78	5	3.33
Volume (ft^3)	5,230	989	1,870	3,950	422	245
Adsorption		**T-102 - 105**			**T-106 - T-109**	
Length (ft)		16.9			13.7	
Volume (ft^3)		295			454	

Economics

Capital Cost Estimate

To estimate the capital cost, the Chemical Engineering Plant Cost Index (CEPCI) from 2021 was used, and its value is 708 [7]. This was used in relation to the value 397 from 2001 found in Turton's *Analysis, Synthesis, and Design of Chemical Processes* [5]. To cost the equipment, the strategies described by Turton in the CAPCOST program included calculating the purchased cost (C_P^o), pressure factor (F_P), material factor (F_M), bare module cost (C_{BM}), and cost of materials required for installation (C_M). Then, the CEPCI was applied. For the material factor, most equipment was assumed to be Clad Steel or Stainless Steel, depending on the availability of costing information, to account for the corrosive material and high temperature nature of the process. When available, Clad Steel was used instead of Stainless Steel to provide protection against corrosion while still choosing the most economically friendly option.

The cost of materials required for installation "includes all piping, insulation and fireproofing, foundations and structural supports, instrumentation and electrical, and painting associated with the equipment" [5]. This was assumed to include all pressure relief valves and controls under instrumentation. The adsorption system for both the Light and Medium Cuts includes the costs of two beds of packing per stream, as well as two additional beds for regeneration. Due to lack of costing information from the referenced BASF PuriCycle® H and PuriCycle® adsorbents, an activated carbon adsorbent retailer was contacted, and the capital costs for packing were calculated based on that market price. Additionally, all pumps were spared in case of maintenance and mechanical issues. An electrical generator was also included in case of loss of electricity to the plant. The generator was sized and costed from a similar example to be able to support all pumps as well as the fire water pump in case of emergencies [8]. The total capital cost estimate for each piece of equipment is listed below in Table 6 and in total, amounts to $44.9 million.

Table 6. Equipment Capital Cost

Equipment	Description	Capital Costs (2021 $)
T-101	Distillation Tower	$688,000.00
TK-101	Feed Storage Tank	$2,320,000.00
TK-104	Heavy Cut Storage Tank	$1,930,000.00
TK-103	Medium Cut Storage Tank	$1,240,000.00
TK-102	Light Cut Storage Tank	$893,000.00
V-101	Reflux Drum	$4,320,000.00
V-102	Knock Out Drum	$230,000.00
T-102 – T-105	Adsorption System Light Cut	$3,000,000.00
T-106 – T-109	Adsorption System Medium Cut	$4,620,000.00
E-103	Reboiler	$10,100,000.00
E-101	Condenser	$11,400,000.00
E-102	Medium Cut Cooler	$41,800.00
E-104	Heavy Cut Cooler	$103,000.00
P-102	Reflux Pump and Spare	$51,200.00
P-101	Feed Pump and Spare	$46,300.00
P-104	Medium Cut Pump and Spare	$43,900.00
P-105	Medium Cut Storage Pump and Spare	$23,100.00
P-103	Light Cut Storage Pump and Spare	$21,900.00
P-106	Heavy Cut Storage Pump and Spare	$46,300.00
N/A	Distillation Trays	$280,000.00
N/A	Fire Water Pump and Spare	$46,300.00
N/A	Electric Generator	$11,000.00
N/A	Piping, Controls, Instrumentation	$3,420,000.000
Total Capital Cost:		**$44,900,000.00**

Variable Cost Estimate

The operating cost per year was calculated using the utilities' costs provided at the location of the plant. The main expenses are due to cooling water consumption, high pressure steam consumption, and pump usage. It should be noted that at this location, energy costs are higher because the most common source is from imported LNG [2]. The annual operating costs to run the process are listed in Table 7 in 2021 dollars and in total, amount to $7.55 million.

Table 7. Variable Operating Costs

Variable Operating Costs		
Utilities	**Yearly Operating Cost (2021$)**	**Consumption**
Distillation Condenser Cooling Water	$333,000.00	6.67e5 MBTU
Medium Cut Cooler Cooling Water	$87,800.00	1.76e5 MBTU
Heavy Cut Cooler Cooling Water	$254,000.00	5.09e5 MBTU
Total Cooling Water	**$675,000.00**	**13.5e5 MBTU**
Distillation Reboiler High Steam	$6,850,000.00	1.32e8 Kg
Total Steam	**$6,850,000.00**	**1.32e8 Kg**
Reflux Pump	$6,780.00	2.71e4 Kw-hr
Feed Pump	$4,220.00	1.69e4 Kw-hr
Medium Cut Pump	$3,150.00	1.26e4 Kw-hr
Medium Cut Storage Pump	$4,110.00	1.64e4 Kw-hr
Light Cut Storage Pump	$2,830.00	1.13e4 Kw-hr
Heavy Cut Storage Pump	$5,740.00	2.29e4 Kw-hr
Total Electricity	**$26,800.00**	**1.07e5 Kw-hr**
Total	**$7,550,000.00**	

Fixed Cost Estimate

The fixed costs of the plant were calculated using Turton's *Analysis, Synthesis, and Design of Chemical Processes* methods [5]. The labor cost was calculated with a correlation that takes into consideration all essential equipment, not including pumps, drums, or storage tanks. This correlation determined that fourteen operators would need to be hired. Annually, these operators would be paid $69,200 based on the average operators' salaries [9]. Lastly, the working capital is approximately 10% to 25% of the capital cost to account for overhead expenses, inventory, and additional costs to operate the plant [5]. Therefore, the working capital is $8.6 MM using 17% of the capital cost. In summation, the fixed costs total was estimated to be $8.6 million.

Table 8. Breakdown of Fixed Costs Estimate

Fixed Costs	
Description	**Yearly Cost (2021$)**
Labor	$966,000.00
Working Capital	$7,630,000.00
Total	**$8,600,000.00**

Process Safety

Minimizing Environmental Impacts

To minimize the environmental impact the system would impose during its operations, the project was inherently designed to reduce the amount of electricity and utilities needed while still meeting specifications for further downstream operation. The reflux ratio was minimized to not only reduce operating costs, but to reduce the amount of steam and cooling water needed for heat transfer, which would lead to less wastewater production and less energy and CO_2 emissions needed to produce steam. Designing the process with inherent safety in mind also results in lower risk of loss of containment, which would have a high possibility of disturbing the surrounding environment and could cause pollution to ground water with heavy hydrocarbons. Electric pumps were chosen over the usage of gas or oil to reduce CO_2 emissions. Due to the high volume that can be held in the storage tanks, it is recommended that the tanks are stored further away from the process equipment in case of an emergency explosion. This would ensure less damage and less risk of environmental pollution in case of an explosion. It would also protect personnel onsite. It was also assumed and suggested that these storage tanks only be used during maintenance and/or shutdown so that a week's worth of product is not consistently stored in them. For future detailed design, separating these tanks to reduce inventory per vessel could be an inherently safer design in case of over pressure or emergencies. These tanks were also designed thicker than common heuristics to prevent loss of containment in case of a tsunami or tropical storm because of the plant's location in the ring of fire, an area prone to natural disasters.

Process & Instrumentation Diagram (P&ID)

A detailed process and instrumentation diagram (P&ID), Figure 3, was created to understand the controllers and control valves needed for safe operation of the fractionated distillation column. Alarms for specified controllers were included for control loops where high or low operation would create dangerous conditions for operation. Table 9 displays the control valves used for the distillation column with associated manipulated and control variables needed for safe operation [10]. While T-101 is the only unit with a pressure relief valve and rupture disc, in reality, pressure relief valves would also be present on the reflux drum and reboiler; however, these were not sized nor included on the P&ID per required specifications. Pipe sizing was also included in the P&ID as shown in Table 10. Analysis of pipe sizing is required for safe operation and is dependent on the mass flow rate, density, fluid phase, and pressure drop across the pipe. Sizes recorded on the P&ID were designed to maximize safety and minimize cost based on the velocity of the vapor or liquid flowing through each pipe using Aspen HYSYS V.10.

Table 9. Control Loops and Alarms Needed for Distillation Column in Pyrolyzed Oil Purification System

Control Device	**Controlled Variable**	**Manipulated Variable**	**Failure Position**	**Alarms**
Level Controller LIC-100	Level of Py Oil in bottom of T-101	Outlet flowrate of bottoms liquid to E-102	Closed	High/Low
Temperature Controller TIC-105	Temperature (composition) of Medium Cut leaving T-101 at Tray 23	Flowrate of steam into E-102	Closed	High/Low
Pressure Controller PIC-101	Pressure of E-102	Flowrate of boil up vapors from E-102	Open	None
Level Controller LIC-101	Level of liquid in E-102	Flowrate of heavy Pyoil from E-102	Closed	High/Low
Temperature Controller TIC-107	Temperature (composition) of Light Cut leaving V-101	Flowrate of Cooling Water into E-101	Open	High/Low
Pressure Controller PIC-105	Pressure of V-101	Flowrate of Py-Gas out of V-101	Open	High/Low
Level Controller LIC-102	Level of liquid in V-101	Flowrate of Light Cut from V-101	Closed	High/Low

Table 10. Pipe Sizing of Fractionated Distillation Column Process for Pyrolyzed Oil Purification

Stream	**T-101 Feed**	**E-101 Feed**	**V-101 Py Gas**	**V-101 Liquid**	**Distillate**	**Reflux**	**Medium Cut**	**E-103 Feed**	**E-103 Vapor**	**E-103 Bottoms**
Vapor Phase	0	1	1	0	0	0	0	0	1	0
Pressure Drop (psi/100ft)	2	0.5	0.5	0.4	2	2	2	0.4	0.5	0.4
Calculated Diamter (in)	3.07	5.05	1.61	4.03	1.61	3.07	3.07	4.03	3.07	4.03
Nominal Pipe Size (in)	3	5	1 1/2	4	1 1/2	3	3	4	3	4

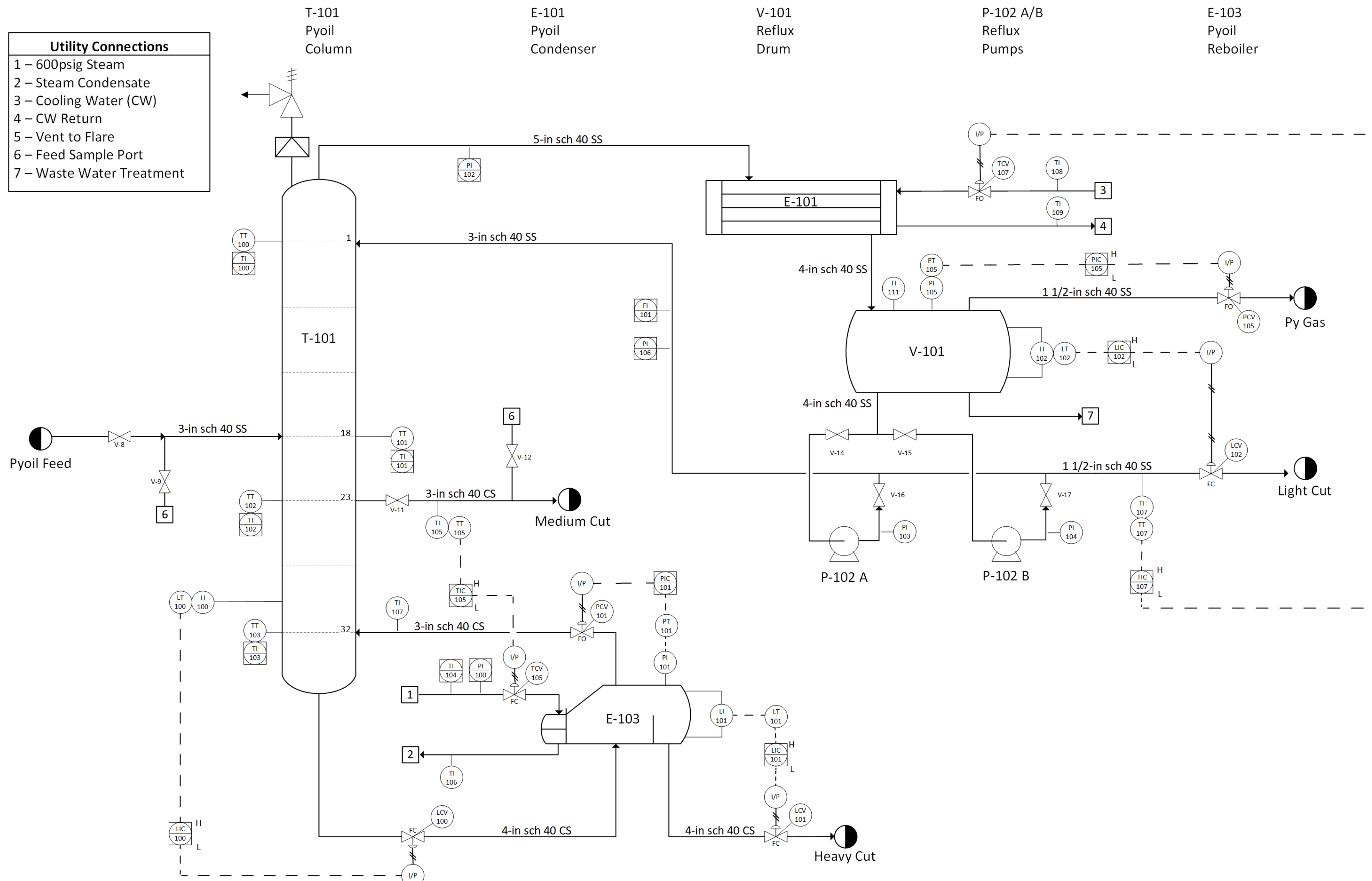

Figure 3. Process and Instrumentation Diagram of Fractionated Distillation System for Pyrolyzed Oil Purification

Pressure Relief Valve Sizing

As part of the design's intrinsic safety, a safety system was included in the distillation tower operations to prevent any loss of containment. This pressure relief system consists of a rupture disc, conventional pressure relief valve, and flare to relieve pressure. Due to the corrosive nature of Py Gas, a rupture disc was designed upstream of the pressure relief valve which will burst at the set pressure. In response to any over-pressure events or upsets, the system will vent vapor out of the top of the distillation tower through the relief valve and send the vapor to the available, onsite flare for safe and complete combustion of the vented material. Since the flare is assumed to already be onsite and safely perform combustion as stated on page 8 of "Part 1 Background and Technical Information," it is also assumed that a knock-out drum or liquid seal drum is already designed and in-place to ensure no liquid is sent to the flare [2]. Additionally, when sizing this system's equipment, the worst-case scenario was used in the event of overpressure which we assumed to be a fire.

To size the pressure relief system, the relief flow rate was first calculated using Equation (1) [11]. To find Q_{fire}, the amount of liquid exposed to the fire was estimated to include the liquid in the bottom of the tower and the liquid on the trays that will fall to the bottom since the operations will have stopped. As shown in Table 11, the height of this liquid was estimated to be 7.35 ft which makes the total wetted surface of the vessel 185 ft^2. Then, the environmental factor, F, was assumed to be 1 for no insulation in the worst-case scenario. With these variables, the total heat input through the surface of the vessel was calculated using Equation (2) to be 2,490,000 Btu/hr [11]. The other variables to calculate $\dot{m}_{relief}$, such as the densities and heat of vaporization under relieving conditions, were found through creating a stream in Aspen HYSYS with the estimated compositions of the heated liquid and running it through a heater with Q_{fire} added to it. With all these variables now known, $\dot{m}_{relief}$ was calculated to be 24,400 lb/hr.

$$\dot{m}_{relief} = \frac{Q_{fire}}{\lambda}\left(1 - \frac{\rho_v}{\rho_l}\right) \tag{1}$$

$$Q_{fire} = 34{,}500 F A_w^{0.82} \tag{2}$$

Table 11. Solving for Relief Flow Rate

Variable	Value	Unit
$\dot{m}_{relief}$	24,400	lb/hr
A_{Wetted}	185	ft^2
Q_{Fire}	2,490,000	Btu/hr
H_{Liquid}	7.35	ft
F (environmental factor)	1	--
ρ_{vapor}	2.17	lb/ft^3
ρ_{liquid}	30.5	lb/ft^3
λ (H_{vap})	94.6	Btu/lb

After solving for the relief rate of the critical flow, the areas of the rupture disc and relief valve were calculated using Equations (3) and (4) as well as Eqn. (9-10) from *Chemical Process Safety* [11]. For these calculations, the Z factor and C_p/C_v were found using the additional HYSYS stream mentioned above. In this case for a fire scenario, the inlet pressure, P_o, is 121% of the design pressure, and the set pressure is assumed to be 10% lower than the design pressure. Lastly, the γ and back pressure were low enough to assume the back pressure factor as 1. Using the variables shown in Table 12, the area for the relief valve vent and the area for the rupture disc were calculated to both be 1.57 in^2. Using this orifice area for the relief valve with Figure 9 of the *Supporting Documents* section, the standard valve body size is 3in x 4in which is an area of 1.84 in^2 [12]. Based on this standardized area, the relief valve's vent capacity was calculated to be 28,700 lb/hr. Because the relief flow rate is greater than 25% of the flow capacity, no chattering is expected to occur in the valve.

$$A_{rupture\ disc} = \frac{\dot{m}_{\text{relief}}}{XP_o}\sqrt{\frac{T_o Z}{MW}} \tag{3}$$

$$A_{relief\ vent} = \frac{\dot{m}_{\text{relief}}}{C_o K_b X P_o}\sqrt{\frac{T_o Z}{MW}} \tag{4}$$

Table 12. Solving for Areas of Rupture Disc and Relief Valve

Variable	Value	Unit
$A_{\text{relief vent}}$	1.57	in^2
$P_{\text{Max. Allowable Working}} = P_{\text{Design}}$	83	psi
$P_{\text{Max. Allowable Accumulated}} = P_o$	100	psi
P_{Set}	74.7	psi
$P_{\text{Final Disposition}}$	14.7	psi
$T_{\text{Design}} = T_o$	429	°F
P_{Back}	7.47	psi
K_b	1	--
Z factor	0.753	--
MW	213	lb/mol
C_o	0.975	--
$\gamma = C_p/C_v$	1.05	--
$A_{\text{Standard Sized Orifice for relief valve}}$	1.84	in^2
D_{Inlet} x D_{Outlet} (for relief valve)	3 x 4	in
$\dot{m}_{\text{capacity}}$	28,700	lb/hr
$A_{\text{Rupture Disc}}$	1.57	in^2
$D_{\text{Rupture Disc}}$	1.41	in

As requested, only a pressure relief system for the distillation tower was designed. However, more safety devices, such as a pressure relief valve on the kettle reboiler and many more, should be considered in the detailed design of the purification unit to ensure the unit is intrinsically safer especially in the event of overpressure.

Failure Rate Analysis

Below are rates reported from previous literature that documents the failure rates of each kind of control and safety equipment. Acknowledging these rates from previous studies can help ensure these metrics used to control operations and keep the system running at the correct specifications are a safe and reliable method. The rates are reported on an occurrence of failure per year basis, with the addition of the failure of a rupture disc in the event of an overpressure event [13, 14].

Table 13. Failure Rate Analysis

Failure Rate Analysis		
Equipment	**Failure Detail**	**Failure Rate (occurrence/time)**
Rupture Disc	Leakage	0.018/year
Rupture Disc	Fail	1E-04/demand
Controller	N/A	0.2/year
Control Valve	N/A	0.15/year
Flow Indicator	Liquid Reading	1.14/year
Hand Valve	N/A	0.13/year
Level Indicator	Liquid Reading	1.7/year
Pressure Indicator	N/A	1.41/year
Relief Valve	N/A	0.022/year

Personnel Exposure Risk

Table 14, below, is a list of the individual chemicals that comprise the Pyrolysis Oil as well as each cut. Due to the nature of the feed being variable, other chemicals that are common health risks, especially in crude refining, were also addressed. These limits, also known as the permissible exposure limits (PEL), are typically relayed in parts per million by volume (ppm) and are annually updated by ACGIH – the American Conference of Governmental and Industrial Hygienists [5]. The LD50 amount quantifies the lethal dose of 50% or higher of the animals tested [15, 16]. These numbers are necessary to know for the safety of the personnel involved in the distillation process. If exposed, they must know what actions to take based on the compound’s health impact.

Table 14: Personnel Exposure Risk/Health Impact

Health Impact			
Compound	**OSHA Chemical Exposure Limit**	**NFPA Diamond Class**	**LD50**
1,3 Butadiene	1 ppm	242	N/A
Butane	N/A	140	N/A
Butene	N/A	140	5,000 mg/kg
Ethane	N/A	140	N/A
Ethylene	N/A	242	N/A
Methane	100 ppm	240	N/A
Pentane	1,000 ppm	140	5,000 mg/kg
Propane	1,000 ppm	240	N/A
Propylene	N/A	141	N/A
Benzene	1 ppm	230	50 mg/kg
Chlorine	1 ppm	400	5,800 mg/kg
Ammonia	50 ppm	310	350 mg/kg
Hydrogen Sulfide	10 ppm	440	49 mg/kg

Atmospheric Detonation of Distillation Inventory

When considering the risks in hydrocarbon processing, a TNT equivalency calculation was performed for the atmospheric detonation of all chemicals from the distillation tower. In this analysis, the worst-case scenario was considered by assuming that all material in the tower instantly and vent to atmosphere as a gas [3]. To calculate the equivalent mass of TNT to the mass of fractionator contents in the tower in the case of a detonation, the variables shown in Table 15 were used in Equation (5) [11]. The total volume of liquid and vapor in the tower was estimated to be 508 ft^3 using the liquid level in the bottom of the tower plus the amounts of liquid and vapor per tray which was provided by HYSYS. Using the volume of hydrocarbon present and the density, the mass of the hydrocarbon (m) was calculated to be 11,300 kg. Additionally, the explosion efficiency was assumed to be 2% while the heat of combustion of the hydrocarbon was estimated to be 7,000 kJ/mol [11, 4]. These variables determined the m_{TNT} to be 1,860 kg.

$$m_{TNT} = \frac{\eta m \Delta H_c}{E_{TNT}} \tag{5}$$

Table 15. Solving for TNT Equivalency

TNT Equivalency		
Variable	**Value**	**Units**
m_{TNT}	1,860	kg
η	2%	--
m	11,300	kg
ΔH_c	7,000	kJ/mol
E_{TNT}	4,690	kJ/kg
MW	182	g/mol
ρ	49.1	lb/ft^3
V	508	ft^3

The overpressure from such an explosion was estimated using the equivalent mass of TNT and the distance from the ground-zero point of the explosion, using the Eqns. (6-21) and (6-22) from *Chemical Process Safety* [10]. The ground-zero point, r, was estimated to be approximately 25.6 m, and the ambient pressure to be 101.3 kPa. With these variables, the scaled overpressure factor was calculated to be 3.73 using Eqn. (6-23) from *Chemical Process Safety* [11]. With the scaled overpressure, the peak side-on overpressure is 378 kPa. Based on Figure 10 of the *Supporting Documents* section and the peak overpressure, the damage from the explosion can be estimated as total destruction of buildings, heavy machine tools around 7000 lbs moved and badly damaged, and very heavy tools around 1,200 lbs possibly damaged but likely to survive overall [11].

Table 16. Peak Overpressure from Explosion Calculations

Damage Estimates Based on Overpressure		
Variable	**Value**	**Units**
z_e	2.08	$m/kg^{1/3}$
r	25.6	m
p_s	3.73	--
p_a	101	kPa
p_o	378	kPa

Hazard and Operability Study (HAZOP)

Table 17 demonstrates potential deviations that could occur in the process described below. The goal of a Hazard and Operability Study is to plan ahead and take action against what could be a very deadly, serious failure in the system [5]. A HAZOP contains crucial information about the deviation that could occur, its potential cause, and what can be done about it. The most common deviations include high pressure and high temperatures which could be addressed using indicators and alarms. Additionally, if there is a problem with the feed, a level alarm may be helpful to ensure the proper level is achieved within the distillation column.

Table 17: Hazard and Operability Study

HAZOP				
Guide Word	**Deviation**	**Cause**	**Consequence**	**Action**
lower/no	Lower feed flow rate or no feed flow	- Fouled pipes Valve failure - Pump failure - Feed piping leakage	- Column runs dry - No operation	- Consider low level alarms in tower - Plan for emergency shutdown
More	More feed flow	- Valve failure	- High liquid level in tower	- Consider high level alarm in tower
Higher	High pressure	- Cooling water valve failure - Cooling water no longer supplied - Fouling in condenser	- Pressure increases	- Insert pressure indicator in tower - Create emergency pressure relief
Lower	Low temperature	- Loss of steam to reboiler - Fouling in reboiler	- High liquid levels	- Consider high level alarm
No	Loss of electricity	- Storm damage Lightning strike	- Column runs dry - No operation	- Add generator to process
Higher	High pressure	- Fouling downstream - Valve failure	- Tube rupture/failure	- Add pressure relief valves - Insert pressure indicator in tower
Lower	Low pressure	- Feed pump failure - Failure in reboiler	- Bottoms level rises	- Consider level indicator

Recommendations for Improvement of the Bali Sorting Facility

Operation of the designed Pyrolysis Oil purification plant requires the production and supply of adequate amounts of pyrolyzed oil from plastics. A new sorting facility in Bali, Indonesia will supply the flexible plastic needed to make Pyrolysis Oil; however, the plastic must be sorted and reduced of contaminants before being brought to the plant. As a populous and low-income country, Bali produces large amounts of plastic waste, but does not have the waste management infrastructure to dispose of waste properly. It is estimated that 52% of waste generated in Bali is mismanaged with 33,000 tons of plastic waste entering the ocean each year [1,3].

To create an operational waste management infrastructure in Bali, a collection and sorting facility will be designed in Negara to serve as a model for future infrastructure to be created for the rest of the region. Negara, the largest sub-district of Jembrana, has high potential for plastic waste leakage reduction with the implementation of an improved waste management system. The goal of this design is to propose innovative strategies to the collecting and sorting facility in Negara in regard to closing quality, quantity, and affordability gaps associated with the new system. Table 18 displays proposed ideas, the methods required to implement them, and the effect they could have on the three gaps previously stated.

Table 18. Overall Ideas Regarding Plastic Waste Management in Bali, Indonesia Sorting Facility

Strategy	Method	Affected Gaps
Encouraging Informal Recycling Sector	- NGO dietary incentives - Government compensation for flexible plastic	Quantity Affordability
Revenue from Contaminants	- PVC, metals, secondarily paper/cardboard, and PET sold to industrial center of Surabaya, East Java	Affordability
Near Infrared Optical Sorting Equipment	- Government and Global Petrochemicals collaboration	Quality
Waste Collection by Incarcerated Individuals	- Government participation to provide optional community service to reduce prison time	Quantity Affordability

Possible recovery of plastic waste in Negara, shown in Table 19, will be used to determine the feasibility of proposed ideas. Plastic waste in Negara accounts for 24% of the total waste consumption of the sub-district, and 53.8% of this plastic waste are flexibles that can be used to create Pyrolysis Oil. Other contaminants should be disposed of in a fashion that benefits the economic and environmental systems of Negara. Provided operations of the waste management system in Negara, shown in Table 20, will also be considered in this "cold eyes analysis" of the collection and sorting facilities [3].

Table 19. Negara Recoverable Plastic Waste [3]

Recoverable Material	Amount (tons/yr)
PET	1,100
Other rigids	731
Plastic Bags	1,100
Other Flexibles	1,460

Table 20. Collection and Sorting Operating Specification

Variable	Value	Unit
Collection Vehicle Capacity	290	kg
Collection Operation Period	6	days/week
Collection Employees	20	drivers/day
Sorting Operation Period	6	days/week
Organic Waste Collected	32	ton/day
Inorganic Waste Collected	18	ton/day
Sorting Employees	48	workers/day
Average Employee Compensation	170	US dollars/month

Recommendations for Closing the Quantity Gap

Successful operation of the Pyrolyzer to create Pyrolysis Oil requires a sufficient quantity of plastic to be collected from the community of Negara. According to provided technical information, the only official collection is by the "formal sector" which includes household pickup performed by 20 drivers twice a week funded by the local government. This program has had minimal success in the past with more waste being collected by trash/waste pickers in the "informal sector" of waste collection. The issue arises when waste pickers only collect the valuable material like rigid plastics, glass, and metal. Most of the mishandled waste of Negara is dumped in ravines which transfers to the wa
e to nearby waterways [3].

Encouraging increased participation of the "informal sector" will generate more sorted plastic closing the quantity gap. Incentives based on the amount of waste collected should be provided with measurements of each type of waste (organic, plastic, metal, etc.) recorded at the local collection center. Incentives, paid for by NGOs or Global Petrochemical, should be provided for flexible plastics since they do not have return values similar to glass and metal. This would be an opportunity for any Negara resident to collect waste and deliver it in return for food or compensation. Of course, this idea would require the building of collection centers in villages every 5-10 miles with employees to run the collection centers. This idea would also require constant governmental or NGO funding to make a difference in plastic collection. When paired with the education program detailed in the technical information of this system, providing an incentive at collection agencies will boost the collection rates to meet the 18 ton/day goal of inorganic waste collection [17].

Another idea that would help close the quantity gap is trash collection by incarcerated individuals. In recent years, the overcrowding of prisons in Bali has posed a real challenge for Indonesian Government. The highest contributor to prison overcrowding is the amount and sentences associated with drug crimes. Indonesia has current alternative strategies to reduce prison time such as custodial services, community service, and fines, but these alternatives are not utilized in the justice system when providing verdicts [18]. If incarcerated individuals were able to reduce their prison time in exchange for community service in the form of collecting waste from ravines, there would be less pollution, less prison overcrowding, and more quantity of plastic for Pyrolysis Oil production. The largest benefit of this idea is the minimal cost of implementing it into the current justice system, and the amount of money it could save the Indonesian government. According to the University of Melbourne, the government would need to build and staff over 1,000 new prisons to fix the overcrowding issue; however, implementing or using community service will eliminate this cost [19]. Not only would this idea have little cost and save the government millions in new prison costs, but it would reduce the overcrowding of prisons which has become a major political issue in Bali over the last few years.

Increasing household collection participation will come with time after implementing ideas that create communication and community awareness for waste collection. The educational programs and marketing will be very beneficial; however, if residents notice people collecting waste on a daily basis, whether for compensation or community service, they will be reminded or be more inclined to recycle in their own home.

Recommendations for Closing the Quality Gap

The quality gap associated with this project is present due to the quality of flexible plastics required by the pyrolyzing plant. It would be impossible to design a system to create pyrolysis oil at reduced cost if rigid plastic and other contaminants needed to be removed from all imported plastics, so removing them at the sorting stage of production is more cost effect and practical. Increased and improved investments by the government and Global Petrochemicals when creating this facility will close the quality gap present when creating Pyrolysis Oil.

The main contaminants of flexible plastics of the inorganic waste are rigid plastics including polyethylene terephthalate (PET), high-density polyethylene (HDPE) and polypropylene (PP). Once separated from glass and metals, these rigid plastics can be bundled and sold; however, limiting the amount of PET, HDPE, and PP in flexible plastics product has proven difficult. Recent developments in technology have proven useful in increasing production rates and decreasing the amounts of contaminates in recycled plastic product. If Global Petrochemical were to purchase Near-Infrared (NIR) Optical Sorting Equipment for the sorting facility, the quantity of rigid contaminants in the flexible plastic product would decrease to only 10% [20, 17]. These contaminants, once melted, contribute to light and medium cuts with chlorine and sulfur that must be removed by adsorption. A lower amount of contaminants would require less activated catalyst in adsorption columns downstream which reduces capital cost of Pyrolysis Oil Purification. Also, a lower percent of rigid plastics would require less energy to break down plastics to Pyrolyzed Oil in the production Design. Average cost of one sorting unit is $75,000 and can handle 5 tons of plastic/day. To collect and sort the estimated 18 tons/day, four sorting units would be required at a capital cost of $400,000 [21].

Recommendations for Closing the Affordability Gap

The affordability gap associated with this design is created by the need to have a profitable system in a low-income country. One additional idea for this system is to sell PVC, metals, secondarily paper/cardboard, and PET to the industrial center of Surabaya, East Java close to Negara. Transportation costs would be minimal due to its close collection proximity, and revenue would be very beneficial to the affordability of this project [22]. As shown in Table 21 and Table 22, the current capital costs and costs and revenues per year associated with developing an efficient collecting and sorting facilities with the addition of the three ideas stated above. The capital cost includes the grassroots cost of building sorting facilities with a capacity of 17,700 tons of combined organic and inorganic waste each year in addition to the capital cost of suggested improvements [17, 5]. The costs and revenues per year of the waste management system includes worker compensation, yearly utility costs, transportation, and revenue for fertilizer, flexible plastics, and contaminating products [3, 17]. All calculations were based on a service factor of 0.97 to account for times when the sorting facility will need maintenance or service. The facility information provided the knowledge of 68 hired employees; however, incorporating 10 collection centers with 4 people working each center results in 108 total employees.

Table 21. Estimated Capital Cost of Developing Collection and Sorting Facilities

Variable	Amount	Unit	Cost ($)/Unit	Amount ($)/yr ($)
Grassroots Sorting Facility	14,000	ton/yr	198/ton	(2,770,000)
Collection Centers	3,700	ton/yr	198/ton	(733,00)
NIR Sorting Equipment	4	units	100,000/unit	(400,000)
Total Capital Cost	3,900,000			

Table 22. Estimated Cost and Revenue Breakdown of Collecting and Sorting Facility Per Year [22, 23]

Variable	Amount	Unit	Cost ($)/Unit	Amount($)/yr
Flexible Plastic Sales	2,480	ton/yr	51/ton	127,000
Organic Fertilizer Sales	1,170	ton/yr	10/ton	11,700
Rigid Plastic Sales	1,830	ton/yr	344/ton	630,000
Glass Sales	745	ton/yr	69/ton	51,400
Paper/Cardboard Sales	2,500	ton/yr	195/ton	490,000
Worker Compensation	108	employees	170/month	(214,000)
Plastic Collecting/Sorting	4,600	ton/yr	315/ton	(1,450,000)
Total Revenue	354,000			

Assuming Global Petrochemicals is paying for 50% of the total capital and variable costs with the other 50% being split between the government and NGO funding half of the total capital cost and revenue will be applied to the calculated costs of the Pyoil Purification design [2]. The adjusted capital and variable costs of the Pyoil Purification process design are located in Table 23.

Table 23. Adjusted Capital and Variable Costs of Pyoil Purification System Considering Bali Sorting and Collecting Facilities and Innovation

Total Costs ($)	Amount
Capital	46,900,000
Variable	7,730,000

Conclusions

It was determined one distillation column would be the most cost efficient and inherently safe design to properly separate the four cuts required from the Pyoil feed. The distillation column was optimized to ensure a reflux ratio that would minimize utility cost and column size. With a focus on the topic of safety, a pressure relief system was designed with a rupture disk in series with an emergency relief valve in the event of an overpressure event. Because the feed is variable and difficult to characterize based on compositions, EBP was utilized to successfully separate the cuts. For potential free water present in the feed stream, a water boot was added to the reflux drum for the Light Cut and a water knock out drum with a water boot was used for the Medium Cut. While the Py Gas is sent immediately to the cracking unit, the Medium and Light cut are sent to a series of adsorbers to remove contaminates, specifically chlorine and other heavy metals. Because the beds must be able to regenerate with hot nitrogen while processing continues, an additional set of adsorbers were designed in series for each stream. Finally, holding tanks were designed and sized assuming a week of holdup time was required. It was also assumed that storage tanks were only be used for maintenance, emergencies, and shutdowns to ensure that the cuts were being delivered at the appropriate temperatures. To design in an inherently safe matter, storage vessels were designed not to hold high pressures and additional pumps were added to deliver products at correct pressures.

For the “Cold Eyes Analysis” of the Bali waste collection and sorting facility, three gaps were addressed, and solutions were presented to close these gaps. The three gaps associated with the effectiveness and efficiency of the sorting facility can be improved with the following strategies: encouraging informal recycling sector, gaining revenue from contaminants, installing Near Infrared Optical Sorting Equipment, and implementing waste collection by incarcerated individuals.

Appendices

Adsorption Section Detail

Assuming packed beds can be regenerated with a hot nitrogen purge available on site, a second adsorber system in series for each Light Cut and Medium Cut was added to ensure processes can be continuous during regeneration. A valve system can be used to direct hot nitrogen for purging one set of adsorbers, while sending feed to the other set. As stated in [2], two beds in series are known to provide adequate protection of contaminants [2]. The main contaminates of interest are chlorine and other heavy metals. In the case of abnormally high contaminants, the analyzer controller connected to the adsorption system could detect the bed being full of contamination and thus switch over to the fresh regenerated system. Assuming regeneration can occur faster than one set of beds can become over saturated with containments, the frequency that the beds are regenerated and switched will be dependent on the contamination concentration in the variable feed stream.

Because little information could be gathered about the recommended adsorbents, trademarked BASF PuriCycle®H and PuriCycle®, information regarding the packing needed for sizing was taken as the same specifications from granular activated carbon for both adsorbents. This assumption was taken because activated carbon is a material also known to remove chlorine contaminates [24]. It was decided two separate adsorption systems for each cut would be most cost effective instead of combining the cuts together, to prevent another separation process after adsorption. Sizing of the adsorption was determined using the methods laid out in *Adsorption Basics* [25]. Using properties of granular activated carbon for modeling purposes, the following was calculated in Table 24 [26, 27].

Table 24. Adsorption Specifications [28]

	Adsorption Light cut	**Adsorption Medium Cut**
LHSV (1/hr)	1	1
Void Fraction	0.45	0.45
Packing Size (mm)	1	1
Superficial Velocity (ft/s)	0.00328	0.00328
Mass of Adsorbent (lb)	9220	14200
Length of 1 Bed (ft)	16.9	13.7
Volume of 1 Bed (ft^3)	295	454

The mass of adsorbent needed was calculated using the LHSV provided alongside the properties and flow rate of each cut taken from Aspen HYSY. A superficial velocity was determined from heuristics which was then used to calculate the cross sectional area of adsorption. The cross-sectional area, the volume of adsorbent required, and the bulk density of the packing was then utilized to determine the length and bed of the adsorbent needed. Although two different kinds of packing were recommended for protection from contamination from heavy metals, both beds were modeled as activated carbon for sizing purposes. Both of these systems for light and heavy cuts were then doubled to account for a regeneration period, with the system having a total of 8 adsorbers. It was assumed that there would be at least a 7°F temperature drop across the adsorber, as the Light Cut and Medium Cut needed to be supplied at 100°F but came out of the distillation process and cooler, respectively, at 107°F [2]. To account for the pressure drop across the adsorber, Equation (6) was used. It was determined pressure drop across the packing for light and medium cut was 0.410 and 0.101 psi, respectively [29].

$$-\frac{\Delta P}{\Delta L} = 150\left(\frac{\mu q}{d_p^2}\right)\left(\frac{(1-\epsilon)^2}{\varepsilon^3}\right) + 1.75\left(\frac{\rho q^2}{d_p}\right)\left(\frac{(1-\epsilon)}{\epsilon^3}\right) \tag{6}$$

Distillation Section Detail

The distillation tower is the major focus of the design for achieving separation of the four different cuts. The process design began using Aspen HYSYS V10 with Peng-Robinson as the equation of state shown in Figure 4. The first major problem encountered when simulating the purification unit was creating the feed stream based on the composition wt% of the chemicals, especially for C6+, as shown in Table 25 [2]. The feed stream was effectively simulated using Oil Manager in HYSYS utilizing provided True Boiling Points (TBP) at atmospheric pressure with feed compositions. As shown in Figure 5, the generated TBP curve demonstrated the same trend as provided values in Table 25. After the feed stream was specified, the distillation tower was configured using the End Boiling Points (EBP) of the Light and Medium Cuts as well as the condenser and reboiler temperatures. The condenser temperature was specified to be 107 °F at 30 psia for the Py Gas to remain a vapor and vent out of the reflux drum while the other overhead vapors condense into a liquid. The operating temperature and pressure of the condenser also provided Py Gas and Light Cut products at the required temperature and pressures needed to meet steam cracker and ethylene plant specifications. The Light Cut's EBP was specified as 392°F to ensure none of the heavier hydrocarbons were pulled into this cut. Next, the Medium Cut's EBP was specified as 620°F for the purpose of preventing the inclusion of Heavy Cut hydrocarbons. Lastly, the reboiler temperature was set to be 620°F since the Medium Cut and all lighter components in the Light Cut and Py Gas vaporize before this temperature. This temperature specification allows the Heavy Cut to remain a liquid pulled when provided as bottoms product from the reboiler. After setting the design specifications, a converged column was generated and adjusted for optimization.

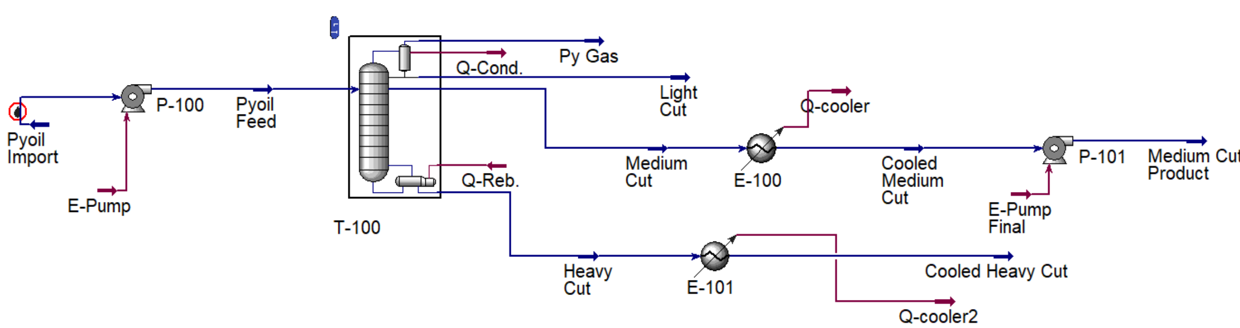

Figure 4. HYSYS Flowsheet of Distillation Tower Design

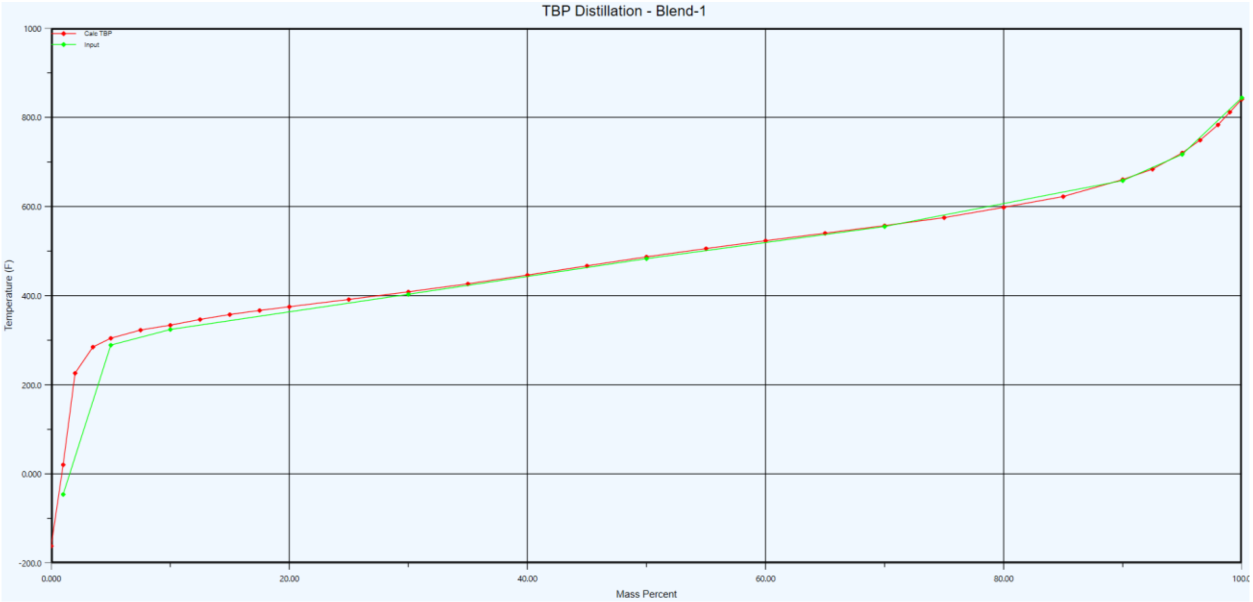

Figure 5. True Boiling Point Curve from HYSYS

Table 25. "Table 1: Raw Pyoil Import Streams from Pyrolyzer to Purification Unit" [2]

Mass flow rate, lb/hr	52,432
Temperature, °F	100
Pressure, psig	Defined by Vapor Pressure
Density, lb/ft^3	49.1
Molecular Weight	182.0
Phase	Liquid
Composition wt%	
Nitrogen	
Hydrogen	
Carbon Monoxide	
Carbon Dioxide	
Methane	0.01
Ethane	0.10
Ethylene	0.05
Propane	0.29
Propylene	0.26
Butane	0.36
C4 Olefins	0.62
1,3-Butadiene	0.07
Pentane	0.13
Hexane	0.40
C6+	97.69
TBP (True Boiling Point) at 760 MM HG (wt):	
IBP	-46 °F
5%	289 °F
10%	324 °F
30%	403 °F
50%	483 °F
70%	555 °F
90%	658 °F
95%	717 °F
EBP	844 °F

When optimizing the distillation tower design, many different configurations were considered. For example, the three different types of condensers were evaluated. When using a full condenser, the only product leaving the condenser was the vapor Py Gas while the condensed liquid was sent back to the tower. However, this caused the reflux ratio to be too high causing flooding in the tower as well as higher costs for reboiler utility and overall capital cost. Then, the total condenser did not allow the Py Gas to vent as a vapor product, so the supplied cooling water was not cold enough to provide a liquid Py Gas outlet. In this event, safety also becomes a concern since pressure could build up in the conder as well as lead to chattering of pressure valves due to low boiling points of Py Gas. Lastly, the partial condenser was selected for this design because of its reasonable reflux ratio and capability to vent the Py Gas as a vapor product while also separating the Light Cut from the reflux stream.

In addition to the various condenser comparisons, a two-tower design was considered with the Py Gas and Light Cut pulled separately by a partial condenser while the Medium and Heavy cuts left as the bottoms product that fed into the second tower that would separate the two heavier cuts. However, when comparing this design to the design with one tower, the overall capital costs and utility consumption were lower for the single tower design. In addition to being the more economically attractive option, the design of one tower versus two towers also was inherently safer by simplifying the process through a decrease in equipment and process complexity.

Additional design configurations were considered before the feed enters the tower. For example, preheating the feed using heat integration before the tower was implemented to decrease the duty required of the reboiler, but the reflux ratio significantly increased due to the increase of vapor being sent to the condenser. For this reason, the preheater was not included in the design. Also, a flash drum before the tower was considered to proactively separate the Py Gas from the other cuts. However, a flash drum couldn't ensure none of the Light Cut would vaporize with the Py Gas based on the End Boiling Points. Lastly, when considering the kettle reboiler design, the hot oil was assumed to create high pressure steam for the reboiler rather than utilizing hot oil to operate the reboiler due to hot oil's contaminated nature and lack of purity. Therefore, high pressure steam is utilized for reboiler operations based on the supplied temperature of 750°F in order to effectively provide enough heat transfer to vaporize the boilup vapor [2].

Once the condenser type and number of towers were established, the number of stages, N, and the entering feed stage were manipulated in order to optimize the reflux ratio, RR, while minimizing the tower's capital cost. While adjusting N, the HYSYS would not converge at stages below 18. For this reason, the recorded data for this iteration started at stage 18. The reflux ratio directly affects the operating cost of the column while the number of stages, N, directly affects the capital cost of the trays and tower. Therefore, the most economically attractive tower design will have the lowest N*RR which is the product of number of stages and reflux ratio. As shown in Figure 6, the lowest N*RR was at 22 stages. After identifying the optimal number of tower stages, the reflux ratio was further optimized by iterating the feed stage and Medium Cut stage. As shown in Tables 26 and 27, the most optimal feed stage was 12 while the optimal Medium Cut stage was 16. The feed inlet location and Medium Cut outlet location produced internals where the hydraulic plots were satisfied with two passes. As a result of minimizing the reflux ratio in the most economically attractive way, the tower configuration design minimizes the overall energy consumption of the condenser and reboiler

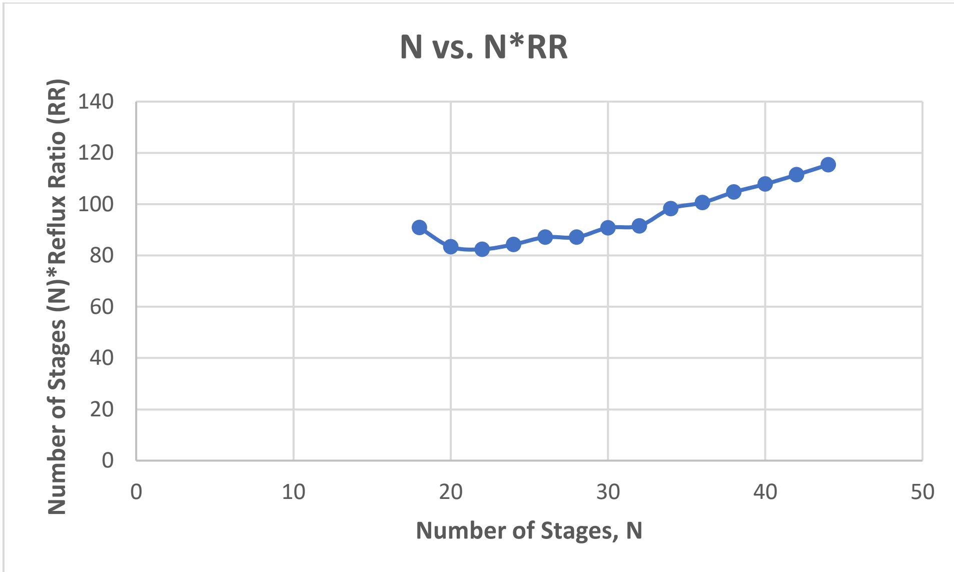

Figure 6. Number of Stages vs. Number of Stages*Reflux Ratio

Table 26. Optimizing Feed Stage

Feed Stage	Medium Cut Stage	RR
2	16	18.8
4	16	6.54
6	16	4.86
8	16	4.09
10	16	3.80
12	16	3.72
14	16	3.79
16	16	4.36
18	16	15.7

Table 27. Optimizing Side Cut Stage

Feed Stage	Medium Cut Stage	RR
12	12	4.91
12	14	3.96
12	16	3.72
12	18	4.33
12	20	15.7

The trays in the distillation column are sieve trays due to the high feed liquid flow rate and cheaper cost. The high feed flow rate allows the column to have constant vapor and liquid flow throughout the column. Therefore, sieve trays work effectively to avoid flooding and weeping in the column. The sieve trays also were the lowest cost of all trays considered which helped lower the capital cost of the tower. The actual number of trays in the column is 32 based on the assumption of 70% tray efficiency [30]. The column height was calculated based on the number of actual trays. The tray spacing was assumed to be 2 feet per tray [4]. Along with the tray spacing, heuristics for adding 4 feet for the top of the column and 6 feet for the bottom of the column were used to calculate the height [4]. Using the given assumptions, the height of the column is 74 feet. Aspen HYSYS determined the column diameter to be 8 feet. This gives an L/D ratio of 9.25 which is an acceptable range according to Turton, confirming a reasonable column height and diameter [4].

The column operating pressure is set to be above atmospheric as well as above the minimum distillation pressure of 2.4 psig [2]. The final operating pressure of the column 33 psia. This provides some cushion above atmospheric pressure to allow the control system to easily maintain the desired pressure. The design pressure of the tower is 83 psia. The condenser operating pressure is 30 psia using an assumption of 3 psi pressure drop from the column through the condenser [4]. The condenser pressure was used to find the pressure of the reboiler. The pressure drop through the reboiler was assumed to be 1.5 psi [4]. The heuristic of 0.1 psi pressure drop per tray was also assumed throughout the column [4]. Thus, the reboiler pressure is 37.7 psia. These pressures were specified in HYSYS in the converged simulation ensuring that the pressures are both feasible to control and effective for the separations.

Despite no water specification, trace levels of water can be present in the feed. However, free water cannot be fed to the steam cracker. Therefore, water boots were incorporated into the process design to remove any water from the Light Cut and Medium Cuts before being sent to the adsorption columns. The first water boot was added to the reflux drum and selected as a feature for this vessel in HYSYS to remove any water from the Light Cut. For the Medium Cut, a water knockout drum with a water boot was added to the stream leaving the distillation column. These additional design configurations allow for the continuous removal of free water from the tower, eliminating any free water from the products that feed the steam cracking furnaces.

The following plots in Figures 7 and 8 depict the temperature profile per stage of the tower and the vapor/liquid flow profile per stage throughout the tower. In Figure 7, the temperature in column increases from the top at the condenser to the bottom at the reboiler with a significant increase in temperature after the condenser. In Figure 8, the vapor and liquid molar flow rates throughout the column are as expected with the vapor flow rates being larger in the top of the tower while the liquid flow rates are larger in the bottom of the tower.

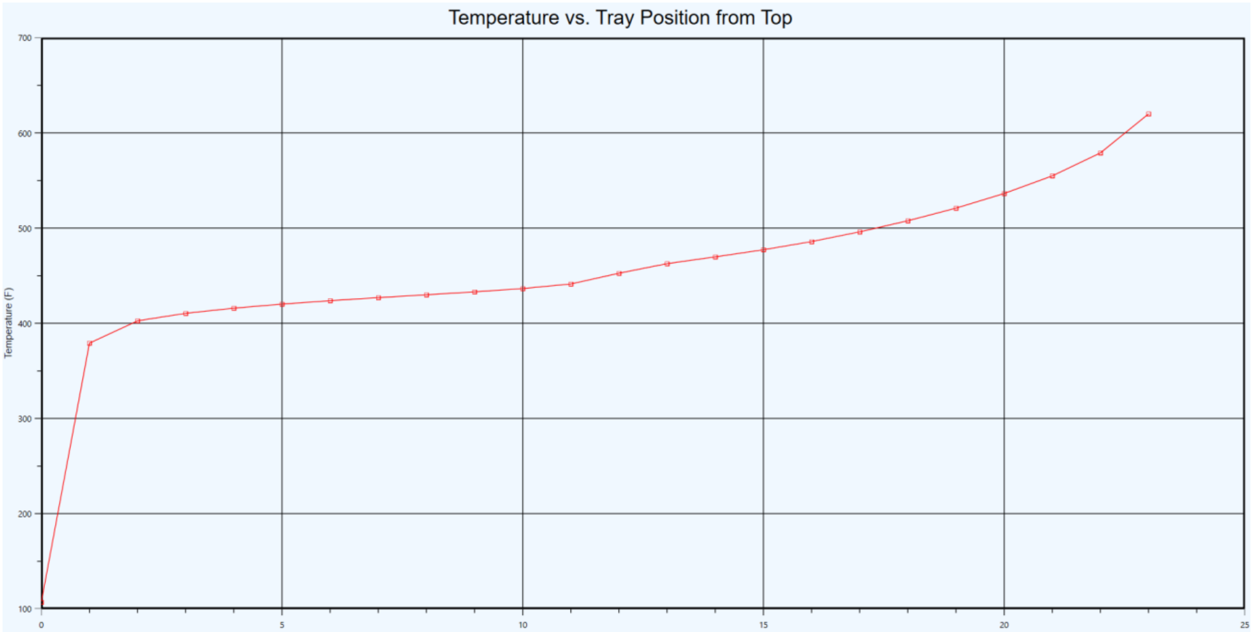

Figure 7. Tower Temperature Profile Per Stage from HYSYS

Figure 8. Vapor/Liquid Traffic Per Stage Profile from HYSYS

FIG. 5-7
API Pressure Relief Valve Designations

Standard Orifice Designation	Orifice Area cm²	Orifice Area (in.²)										
D	0.710	0.110	•	•	•							
E	1.265	0.196	•	•	•							
F	1.981	0.307	•	•	•							
G	3.245	0.503			•	•						
H	5.065	0.785			•	•						
J	8.303	1.287				•	•					
K	11.858	1.838					•	•				
L	18.406	2.853					•		•			
M	23.226	3.60							•			
N	28.000	4.34							•			
P	41.161	6.38							•			
Q	71.290	11.05								•		
R	105.226	16.0								•	•	
T	167.742	26.0										•
		in.	1 × 2	1.5 × 2	1.5 × 3	2 × 3	3 × 4	3 × 6	4 × 6	6 × 8	6 × 10	8 × 10
		mm	25 × 50	38 × 50	38 × 75	50 × 75	75 × 100	75 × 150	100 × 150	150 × 200	150 × 250	200 × 250
			Valve Body Size (Inlet Diameter times Outlet Diameter)									

Figure 9. Standard Pressure Relief Valve Sizes [12]

Table 6-9 Damage Estimates for Common Structures Based on Overpressure (these values are approximations)[1]

Pressure psig	Pressure kPa	Damage
0.02	0.14	Annoying noise (137 dB if of low frequency, 10–15 Hz)
0.03	0.21	Occasional breaking of large glass windows already under strain
0.04	0.28	Loud noise (143 dB), sonic boom, glass failure
0.1	0.69	Breakage of small windows under strain
0.15	1.03	Typical pressure for glass breakage
0.3	2.07	"Safe distance" (probability 0.95 of no serious damage below this value); projectile limit; some damage to house ceilings; 10% window glass broken
0.4	2.76	Limited minor structural damage
0.5–1.0	3.4–6.9	Large and small windows usually shatter; occasional damage to window frames
0.7	4.8	Minor damage to house structures
1.0	6.9	Partial demolition of houses, made uninhabitable
1–2	6.9–13.8	Corrugated asbestos shatters; corrugated steel or aluminum panels, fastenings fail, followed by buckling; wood panels (standard housing), fastenings fail, panels blow in
1.3	9.0	Steel frame of clad building slightly distorted
2	13.8	Partial collapse of walls and roofs of houses
2–3	13.8–20.7	Concrete or cinder block walls, not reinforced, shatter
2.3	15.8	Lower limit of serious structural damage
2.5	17.2	50% destruction of brickwork of houses
3	20.7	Heavy machines (3000 lb) in industrial buildings suffer little damage; steel frame buildings distort and pull away from foundations
3–4	20.7–27.6	Frameless, self-framing steel panel buildings demolished; rupture of oil storage tanks
4	27.6	Cladding of light industrial buildings ruptures
5	34.5	Wooden utility poles snap; tall hydraulic presses (40,000 lb) in buildings slightly damaged
5–7	34.5–48.2	Nearly complete destruction of houses
7	48.2	Loaded train wagons overturned
7–8	48.2–55.1	Brick panels, 8–12 in thick, not reinforced, fail by shearing or flexure
9	62.0	Loaded train boxcars completely demolished
10	68.9	Probable total destruction of buildings; heavy machine tools (7000 lb) moved and badly damaged, very heavy machine tools (12,000 lb) survive
300	2068	Limit of crater lip

[1] V. J. Clancey, "Diagnostic Features of Explosion Damage," paper presented at the *Sixth International Meeting of Forensic Sciences* (Edinburgh, 1972).

Figure 10. Damage Estimates based on Overpressure [11]

References

[1] Miksiewicz, E., Yeo, G., & Machado, J. (2023). AIChE 2022-2023 Student Design Competition Problem Statement & Rules. AIChE.

[2] Miksiewicz, E., Yeo, G., & Machado, J. (2023). AIChE 2022-2023 Student Design Competition Problem Statement & Rules. *Appendix 1.* AIChE.

[3] Miksiewicz, E., Yeo, G., & Machado, J. (2023). AIChE 2022-2023 Student Design Competition Problem Statement & Rules. *Appendix 2.* AIChE.

[4] Svrcek, William & Monnery, Wayne. (1993). Design two-phase separators within the right limits. 89. 53-60.

[5] Turton, R.; Shaeiwitz, J. A.; Bhattacharyya, D.; Whiting, W. B. *Analysis, synthesis, and design of Chemical Processes*; Prentice Hall: Boston, 2018.

[6] Stewart, M., & Arnold, K. (2008). *Gas-Liquid And Liquid-Liquid Separators* (pp. 131–174). Gulf Professional Publishing.

[7] Maxwell, C. (n.d.). *Cost Indices – Towering Skills*. https://www.toweringskills.com/financial-analysis/cost-indices/

[8] *Kubota GL14000 Diesel Generator 14KW Auto Start GL 14000-AUTO from KUBOTA - Acme Tools*. (n.d.). https://www.acmetools.com/kubota-gl14000-diesel-generator-14kw-auto-start-gl-14000-auto/407001000648.html?srsltid=Ad5pg_F6PXvN-6cAb3nz_5jTyTCHhTDIEw8lzaJ5TN3qaQWFF0UKDijN2Rc

[9] 51-8091 chemical plant and system operators. https://www.bls.gov/oes/current/oes518091.htm (accessed Nov 15, 2022).

[10] Seborg, D. E., Edgar, T. F., Mellichamp, D. A., & Doyle, F. J., III. (2016). *Process Dynamics and Control*. John Wiley & Sons.

[11] Crowl, D. A., & Louvar, J. F. (2001). *Chemical Process Safety: Fundamentals with Applications*. Pearson Education.

[12] *Engineering Data Book*; Gas Processors Suppliers Association: Tulsa, OK, 2018.

[13] Cadwallader, L. C. (1998, September 1). *Selected component failure rate values from fusion safety assessment tasks*. OSTI.gov. Retrieved March 9, 2023, from https://www.osti.gov/servlets/purl/752571

[14] Ahmad, A. (n.d.). *Failure Rate Data*. UTM. Retrieved March 9, 2023, from https://arshadahmad.files.wordpress.com/2016/09/raam-p5-3-fta-quantification.pdf

[15] *Permissible Exposure Limits – OSHA Annotated Table Z-1 | Occupational Safety and Health Administration*. (n.d.). https://www.osha.gov/annotated-pels/table-z-1

[16] Airgas. (n.d.). *SDS Search*. Airgas, an Air Liquide Company. Retrieved March 9, 2023, from https://www.airgas.com/sds-search

[17] University of Oxford, University of Leeds, Ellen Macarthur Foundation, Common Seas, PEW Charitable Trusts, & SYSTEM IQ. (2020). Breaking the Plastic Wave (pp. 68-70). *https://www.pewtrusts.org/-/media/assets/2020/07/breakingtheplasticwave_report.pdf*. PEW Charitable Trusts. Retrieved March 4, 2023, from https://www.pewtrusts.org/-/media/assets/2020/07/breakingtheplasticwave_report.pdf

[18] Siddharta, A. (2019, May 15). *Indonesia Government Looks for Alternatives to Overcrowded Prisons*. VOA. https://www.voanews.com/a/indonesia-government-looks-for-alternatives-to-overcrowded-prisons/4918101.html

[19] Kusadrianto, M. D. (2020, September 28). *Indonesia's prison overcrowding crisis: Criminal Code delay a chance to fix it?* Indonesia at Melbourne. https://indonesiaatmelbourne.unimelb.edu.au/indonesias-prison-overcrowding-crisis-criminal-code-delay-a-chance-to-fix-it/

[20] *TITUS MRF Services Waste Managment Consulting*. (n.d.). [Slide show]. http://www.titusmrfservices.net/. https://www.vinylinfo.org/wp-content/uploads/2020/06/Vinyl-Recycling-Summit_Titus.pdf

[21] STEINERT GmbH. (n.d.). *UniSort PR with camera-based NIR for an optimal sorting result • STEINERT*. https://steinertglobal.com/magnets-sensor-sorting-units/sensor-sorting/nir-sorting-systems/unisort-pr/

[22] *Recycling - Secondary Material Price Indicator*. (n.d.). Eurostat Statistics Explained. Retrieved March 4, 2023, from https://ec.europa.eu/eurostat/statistics-explained/index.php?title=Recycling_%E2%80%93_secondary_material_price_indicator#Average_prices_and_trade_volumes

[23] Bonoli, A., Zanni, S., & Awere, E. (2019). Organic waste composting and sustainability in low-income communities in Palestine: lessons from a pilot project in the village of Al Jalameh, Jenin. *International Journal of Recycling of Organic Waste in Agriculture*, *8*(3), 253–262. https://doi.org/10.1007/s40093-019-0264-8

[24] *ACT - Activated Carbon Basic Information*. (2021). https://www.activatedcarbon.com.au/carbonbasics.htm

[25] Gabelman, A., P. E. & American Institute of Chemical Engineers. (2017). Adsorption Basics: Part 1. *Back to Basics*.

[26] Flood, T. (2020, October 12). *General Carbon Corporation*. General Carbon Corporation. https://generalcarbon.com/

[27] Company, R. E. (n.d.). *VGAC12x40SpecSheet | Riley Equipment Company*. Riley Equipment Company. https://recofiltration.com/assets/files/pdf/VGAC12x40SpecSheet.pdf

[28] Naylor, W. F., & Rester, D. O. (1995, July 1). *Determining activated carbon performance*. OSTI.gov. Retrieved March 9, 2023, from https://www.osti.gov/biblio/99393#:~:text=In%20a%20bed%20of%20granular,treated%20and%20the%20GAC%20particles.

[29] *Reactors*. (n.d.). The University of Manchester. https://personalpages.manchester.ac.uk/staff/tom.rodgers/Interactive_graphs/reactors.html?reactors/ergun/index.html[]:~:text=A%20commonly%20used%20equation%20for,tells%20us%20a%20number%20of

[30] Big Chemical Encyclopedia. https://chempedia.info/info/tray_efficiency_factors/ (accessed Nov 15, 2022).