Letter of Transmittal

Date: 3/10/2023

To: AIChE

From: Chemical Engineering Students

Subject: Closing Critical Gaps to Enable a Circular Economy of Plastics

Enclosed is a report evaluating the design, economics, and safety of a pyrolysis oil purification unit for the AIChE Design Project.

The attached report provides the design details for a two-distillation column purification process leading to three usable streams for a steam cracker, capital and operating costs, and details of safety precautions and potential hazards in the process.

CHE

Chemical Engineering Design II: AIChE Design Competition

Spring 2023

Team 12

Woroud Almutairi, Alex Bias, Elijah Ecklund, Riley McIntosh

Executive Summary

The purpose of this project is to have an efficient and cost-effective method of eliminating plastic waste. One of these methods is turning the plastic waste into ethylene. By pyrolyzing the plastic and purifying the resulting oil, the plastic waste can then become usable in an ethylene plant. The focus of this report was to create a purification unit for the pyrolysis oil and improving the quality, quantity, and affordability of the plastic waste supplied to the pyrolyzer [1].

The total capital costs for the process designed were found to be $186,000,000 with 97% of the cost coming from the tanks which were sized to hold a week's worth of each product stream. The variable operation costs are $8,906,000 and the fixed operating costs are $28,870,000. The process also spares pumps to ensure that minor maintenance and necessary repairs on these inexpensive parts does not necessitate a process shut down and the loss of productivity.

The feed of pyrolysis oil is supplied at 100 °F and 52,430 lb/hr. Along with this, the oil is considered to be complex with a wide boiling mixture of molecules. It is fractionated through two distillation columns into Pygas, Pyoil light cut, Pyoil medium cut, and Pyoil heavy cut. If water entered the oil, a process was implemented to eliminate it. Water was found to be all vaporized in the first column and combined with the Pygas, therefore, Pygas itself has to be dehydrated in an adsorber and then sent immediately as feed for the ethylene plant. It was found that the Pygas, after dehydration had a flow rate of 2,486 lb/hr. The Pyoil light cut and Pyoil medium cut have fears of contamination of salts and heavy metals, which a steam cracker can't handle. In order to ensure the safety of equipment, the light and medium cuts need to be sent through adsorbers to remove metals and chlorides and then stored in tanks for use as fuel in the steam cracker. It was found that the light cut's flow rate is 1,199 lb/hr and the medium's flow rate as 35,000 lb/hr. The Pyoil heavy cut is unsuitable for use in the steam cracker and is held in a tank to await waste treatment.

The process is designed with inherent safety in mind, using a two-column design to improve inherent safety and taking into account the effects of corrosion from water and chlorides in selecting materials of construction. Controls primarily focused on temperature, pressure, and level in the two distillation columns to ensure quality and safety of the process. To account for the possibility of an overpressure event, a pressure relief system using a rupture disk upstream of a pressure relief valve to vent gases to a flare is sized for the worst-case scenario of a fire in the column. Waste liquids collected by a knockout drum for this pressure relief system are sent to a process sewer assumed to be on site and compatible with the materials in this purification process.

The quality, quantity, and affordability of the plastic supply from the Bali Sorting Facility is suggested to be improved by focusing on opportunities for recycling afforded by the large tourist industry by pushing tourist buy-in. Additionally, by moving beyond households to small businesses and corporations, larger quantities of plastic could be obtained with less travel time, reducing overall costs.

Table of Contents

Table of Figures

Table of Tables

Brief Process Description

The contained process is designed to fractionate and purify a pyrolysis oil stream, the specifications of which are given by AIChE. The resulting components of interest are pyrolysis gas and pyrolysis oil light cut, medium cut, and heavy cut variants. The processes were created using Aspen HYSYS simulations [2].

The pyrolysis oil feed stream provided to the purification facility is initially stored in a tank. From there, it is pumped through a preheater into the 12th tray of the first distillation column. This 15-tray column fractionates the feed oil into 3 streams: Py Gas, pyrolysis oil Light cut, and a bottoms stream containing the Medium and Heavy cuts. The Py Gas exits the top of the column and moves through a condenser, partially being condensed into a reflux stream. The remaining Py Gas vapor leaves the condenser and flows through an adsorber unit, which uses a silica adsorbent to dehydrate the stream before it is sent off to the ethylene plant. The pyrolysis oil light cut stream is drawn from the 6th tray of the column and sent to a drum, from which it is pumped through two adsorbers, containing BASF Puricycle H and Puricycle HP adsorbents respectively, that remove all chlorides and heavy metals from the stream. After this, the light stream is cooled and pumped to a storage tank. The bottoms stream of the first column flows down into a reboiler, after which the boil-up is sent back into the column and the pyrolysis oil medium and heavy cut stream is pumped into the 9th tray of the second column as the feed stream.

The 15-tray second column separates the medium cut and heavy cut streams of the pyrolysis oil. The medium cut stream exits the top of the column as overhead vapor, then is completely condensed and sent to a drum along with the reflux for the column. From there, the reflux is pumped back into the column and the medium cut is pumped through two adsorbers, containing BASF Puricycle H and Puricycle HP adsorbents respectively, that remove all chlorides and heavy metals from the stream. The medium cut stream is then cooled and pumped to a storage tank. The heavy cut bottoms stream exits the column at the bottom, enters the reboiler, and is partially re-boiled, the vapor being sent as boil-up back into the column. From there, the liquid heavy cut stream is cooled and pumped to a storage tank that can hold a week’s worth of material, where it is held until it can be sent to a waste treatment plant.

Process Detail

Utilities for the process and conditions for the Raw Py Oil Feed are given in the tables below. Specifically, **Table 1** gives the Raw Py Oil information, **Table 2** gives the utilities available at battery limit, and **Table 3** states the costing of utilities used within the process.

Table 1: Initial Conditions of the Py Oil Feed

Mass flow rate (lb/hr)	52,430
Temperature (°F)	100.0
Pressure (Psig)	Defined by Vapor Pressure
Density (lb/ft^3)	49.10
Molecular Wight (g/mol)	182.0
Phase	Liquid

Table 2: Utilities Available at Battery Limit

Utility Stream	**Pressure (psig)**	**Temperature (°F)**
Cooling Water Supply	70.00	87.00
Cooling Water Return	40.00	107.0
Thermal Fluid Supply	230.0	750.0
Thermal Fluid Return	200.0	725.0

Table 3: Costing for Utilities

Utility Stream	**Cost (units specified per stream)**
Electricity ($/kW-hr)	0.2500
Fuel Gas ($/MBTU*HHV)	15.00
Cooling Water ($/MBTU)	0.5000
HP Steam, 600 psig, 750 °F ($/1000 kg)	51.90
MP Steam, 150 psig ($/1000 kg)	35.00
LP Steam, 50 psig ($/1000 kg)	22.90
Hot Nitrogen ($/L) [3]	2.000

Process Flow Diagram

T-101 Tower 1	TK-101 Feed Oil Tank	P-101 A/B Oil Feed Pump	E-101 Feed Heater	V-101 Reflux Drum	T-102 Tower 2	TK-102 Light Tank	P-102 A/B Reflux Pump	E-102 Condenser	V-102 A/B Water Adsorber	TK-103 Medium Tank	P-103 A/B Light Pump	E-103 Reboiler	V-103 Knockout Drum	TK-104 Heavy Tank
P-104 A/B Med/Hvy Pump	E-104 Condenser	V-104 A/B H Adsorber	P-105 A/B Medium Pump	E-105 Reboiler	V-105 A/B HP Adsorber	P-106 A/B Heavy Pump	V-106 Knockout Drum	E-106 Hvy Condenser	V-107 A/B H Adsorber	E-107 Med Condenser	P-107 A/B Reflux Pump	V-108 A/B HP Adsorber	E-108 Light Condenser	

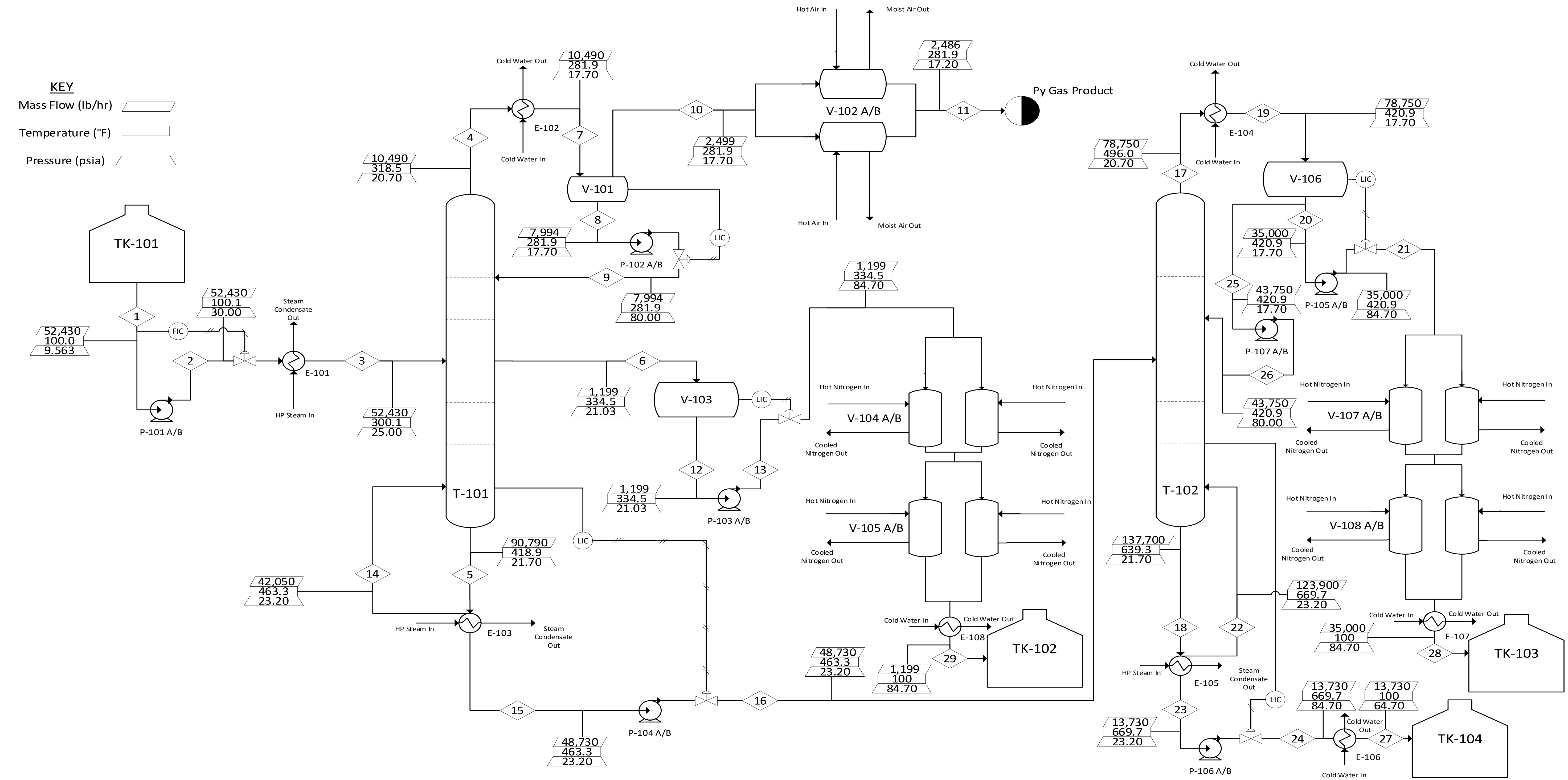


Figure 1: PFD

Material Balance

Table 4.1: Stream Summary

Stream #	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**
Temperature (°F)	100.0	100.0	300.0	318.5	418.9	334.5	281.9	281.9	281.9	281.9	281.9	334.5	334.5	463.3
Pressure (psia)	9.563	30.00	25.00	20.70	21.70	21.03	17.70	17.70	80.00	17.70	17.20	21.03	84.70	23.20
Vapor Fraction	0.000	0.000	1.380E-02	1.000	0.000	0.000	0.2801	0.000	0.000	1.000	1.000	0.000	0.000	1.000
Molar Enthalpy (Btu/lbmol)	-166600	-166600	-146700	-52290	-129400	-87740	-118000	-72850	-72820	-45160	-44020	-87740	-87730	-91570
Molar Flow (lbmole/hr)	288.1	288.1	288.1	135.8	513.2	11.19	135.8	97.75	97.75	38.04	37.31	11.19	11.19	274.3
Mass Flow (lbm/hr)	52430	52430	52430	10490	90790	1199	10490	7994	7994	2499	2486	1199	1199	42050
Density (lbm/ft^3)	48.00	48.03	20.94	0.2031	38.38	38.65	30.38	39.83	39.83	0.1522	0.1500	38.65	38.65	0.3851
Std. Ideal Liq. Vol. Flow (Barrel/day)	4565	4565	4565	995.9	7966	110.5	995.9	745.3	745.3	250.7	249.7	110.5	110.5	3763
Std. Ideal Liq. Vol. Flow (GPM)	133.2	133.2	133.2	29.05	232.4	3.222	29.05	21.74	21.74	7.311	7.285	3.222	3.222	109.8
Composition Flows (lbm/hr)	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Hydrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Methane	1.996	1.996	1.996	2.018	4.630E-06	8.570E-04	2.018	2.330E-02	2.330E-02	1.995	1.995	8.570E-04	8.570E-04	4.600E-06
Ethane	23.71	23.71	23.71	24.44	5.000E-04	2.430E-02	24.44	0.7479	0.7479	23.69	23.69	2.430E-02	2.430E-02	4.950E-04
Ethylene	12.77	12.77	12.77	13.08	1.560E-04	1.060E-02	13.08	0.3165	0.3165	12.76	12.76	1.060E-02	1.060E-02	1.550E-04
Propane	97.95	97.95	97.95	104.3	1.140E-02	0.1964	104.3	6.578	6.578	97.76	97.76	0.1964	0.1964	1.120E-02
Propene	90.30	90.30	90.30	95.30	8.080E-03	0.1639	95.57	5.436	5.436	90.13	90.13	0.1639	0.1639	7.960E-03
n-Butane	140.0	140.0	140.0	159.4	9.050E-02	0.5531	159.4	19.97	19.97	139.4	139.4	0.5531	0.5531	8.820E-02
1-Butene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1,3-Butadiene	28.92	28.92	28.92	32.51	1.420E-02	0.1029	32.51	3.693	3.693	28.82	28.82	0.1029	0.1029	1.380E-02
n-Pentane	54.57	54.57	54.57	70.00	0.1794	0.4147	70.00	15.84	15.84	54.15	54.15	0.4147	0.4147	0.1722
n-Hexane	176.7	176.7	176.7	275.9	2.894	2.615	275.9	102.0	102.0	173.9	173.9	2.615	2.615	2.715
H2O	13.31	13.31	13.31	13.58	1.290E-04	9.800E-03	13.58	0.2870	0.2870	13.30	0.000	9.800E-03	9.800E-03	1.280E-04
C6+ (Theoretical Components)	51790	51790	51790	9703	90790	1195	9703	7840	7840	1863	1863	1195	1195	42050

Table 4.2: Stream Summary Continued

Stream #	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
Temperature (°F)	463.3	463.3	496.0	639.3	420.9	420.9	420.9	669.7	669.7	669.7	420.9	420.9	100.0	100.0	100.0
Pressure (psia)	23.20	23.20	20.70	21.70	17.70	17.70	84.70	23.20	23.20	84.70	17.70	80.00	64.70	84.70	84.70
Vapor Fraction	0.000	0.000	1.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Molar Enthalpy (Btu/lbmol)	-142400	-142400	-104400	-160800	-263700	-131900	-131900	-130900	-164600	-164600	-131900	-131900	-272800	-165400	-102600
Molar Flow (lbmole/hr)	238.9	238.9	435.4	485.0	435.4	193.5	193.5	439.7	45.34	45.34	241.9	241.9	45.34	193.5	11.19
Mass Flow (lbm/hr)	48730	48730	78750	137700	78750	35000	35000	123900	13730	13730	43750	43750	13730	35000	1199
Density (lbm/ft^3)	38.02	38.02	0.3918	35.39	38.37	38.37	38.37	0.5974	35.08	35.08	38.37	38.37	50.91	47.55	45.86
Std. Ideal Liq. Vol. Flow	4204	4204	6919	11420	6919	3075	3075	10290	1129	1129	3844	3844	1129	3075	110.5
Std. Ideal Liq. Vol. Flow (GPM)	122.6	122.6	201.8	333.2	201.8	89.69	89.69	300.2	32.92	32.92	112.1	112.1	32.92	89.69	3.222
Composition Flows (lbm/hr)															
	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Hydrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Methane	2.400E-08	2.400E-08	5.400E-08	3.930E-19	5.402E-08	2.400E-08	2.400E-08	3.930E-19	2.960E-22	2.960E-22	3.000E-08	3.000E-08	2.960E-22	2.400E-08	4.770E-06
Ethane	4.860E-06	4.860E-06	1.090E-05	7.890E-16	1.093E-05	4.860E-06	4.860E-06	7.890E-16	8.850E-19	8.850E-19	6.070E-06	6.070E-06	8.850E-19	4.860E-06	7.230E-05
Ethylene	1.300E-06	1.300E-06	2.920E-06	1.180E-16	2.917E-06	1.300E-06	1.300E-06	1.180E-16	1.190E-19	1.190E-19	1.620E-06	1.620E-06	1.190E-19	1.300E-06	3.370E-05
Propane	1.790E-04	1.790E-04	4.030E-04	1.760E-13	4.034E-04	1.790E-04	1.790E-04	1.760E-13	2.700E-16	2.700E-16	2.240E-04	2.240E-04	2.700E-16	1.790E-04	3.980E-04
Propene	1.180E-04	1.180E-04	2.660E-04	8.690E-14	2.660E-04	1.180E-04	1.180E-04	8.680E-14	1.260E-16	1.260E-16	1.480E-04	1.480E-04	1.260E-16	1.180E-04	3.480E-04
n-Butane	2.320E-03	2.320E-03	5.210E-03	1.400E-11	5.201E-03	2.320E-03	2.320E-03	1.400E-11	2.950E-14	2.950E-14	2.890E-03	2.890E-03	2.950E-14	2.320E-03	8.510E-04
1-Butene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1,3-Butadiene	3.340E-04	3.340E-04	7.520E-04	1.470E-12	7.524E-04	3.340E-04	3.340E-04	1.470E-12	2.910E-15	2.910E-15	4.180E-04	4.180E-04	2.910E-15	3.340E-04	1.700E-04
n-Pentane	7.190E-03	7.190E-03	1.620E-02	2.390E-10	1.618E-02	7.190E-03	7.190E-03	2.380E-10	6.720E-13	6.720E-13	8.990E-03	8.990E-03	6.720E-13	7.190E-03	5.140E-04
n-Hexane	0.1789	0.1789	0.4025	3.190E-08	0.4025	0.1789	0.1789	3.180E-08	1.200E-10	1.200E-10	0.2236	0.2236	1.200E-10	0.1789	2.710E-03
H2O	1.020E-06	1.020E-06	2.290E-06	7.790E-17	2.287E-06	1.020E-06	1.020E-06	7.780E-17	7.710E-20	7.710E-20	1.270E-06	1.270E-06	7.710E-20	1.020E-06	4.860E-05
C6+ (Theoretical Components)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000

Sized Equipment List

The towers sizes could be calculated by taking the heuristics given in Turton that there is four feet above the first tray and six feet below the last tray in the tower. Along with the number of trays, assuming a 2 foot tray gap and a tray efficiency of 70%, the height can be calculated. The diameter for each tower was given in Aspen HYSYS. For the tanks and vessels, **Equation 1** was used. With this equation, for tanks the holdup time was given as 1 week, but for the vessels it was assumed to be 180 seconds for holdup and 120 seconds for surge [4]. Vessels were also assumed to have a D/L of 5 and be half filled. To find the diameter, the volumetric flow rate (Q_l) has to be multiplied by the hold up (t_{holdup}) and surge (t_{surge}) times. This is then all divided by pi over 4, multiplied by one minus f_a, multiplied by L over D. All of this is then taken to the power of 1 over 3.

$$D = \left(\frac{Q_L(t_{holdup}+t_{surge})}{\left(\frac{\pi}{4}\right)(1-f_a)\left(\frac{L}{D}\right)}\right)^{\frac{1}{3}} \qquad \textbf{Equation 1}$$

Given this Equation and assumptions, **Table 5** compiles all sizing information for the towers, tanks, and vessels.

Table 5: Towers, Tanks, and Vessels Sizing

Equipment	**PFD Name**	**Height (ft)**	**Diameter (ft)**	**MOC**	**Internals**
Towers	T-101	40.00	4.235	Ni	15 Sieve trays
	T-102	40.00	6.925	Ni	15 Sieve trays
Tanks	TK-101	225.2	45.04	Ni	
	TK-102	65.15	13.03	CS	
	TK-103	197.4	39.48	CS	
	TK-104	141.3	28.27	Ni	
Vessels	V-101	10.73	2.147	SS	
	V-102 A/B	3.75	1.25	SS	Silica
	V-103	5.158	1.031	Ni	
	V-104 A/B	6.750	2.250	CS	Puri H
	V-105 A/B	5.750	2.250	CS	Puri HP
	V-106	20.48	4.096	Ni	
	V-107 A/B	20.25	6.750	CS	Puri H
	V-108 A/B	20.25	6.750	CS	Puri HP

In order to find sizing for heat exchangers, area (A) had to be found. This was done using **Equation 2**. Where the heat duty (Q) would be divided by a correction factor (F), heat transfer coefficient (U), and the difference in temperatures using the LMTD method. It was assumed that the correction factor was one and the transfer coefficient was found as 120 in the GPSA [5]. This can be seen in **Table 6** below.

$$Q = A * U * F * LMTD$$ **Equation 2**

Table 6: Heat Exchanger Sizing

Equipment	PFD Name	Area (ft^2)	Duty (BTU/hr)	MOC	Type
Heat Exchangers	E-101	87.78	$5.730*10^6$	Ni	FH S/T
	E-102	71.30	$1.740*10^6$	SS	FH S/T
	E-103	197.2	$7.300*10^6$	Ni	FH S/T
	E-104	276.2	$1.200*10^7$	Ni	FH S/T
	E-105	1149	$1.300*10^7$	Ni	FH S/T
	E-106	280.3	$4.910*10^6$	Ni	FH S/T
	E-107	572.3	$6.500*10^6$	CS	FH S/T
	E-108	18.50	$1.670*10^5$	Ni	FH S/T

For pumps, a similar process is used in sizing, but Hydraulic Horsepower is required instead. This is found by dividing the multiplication of volumetric flow rate (Q) and change in pressure (ΔP) by 1715, as seen in **Equation 3**.

$$Hydralic\ HP = \frac{Q*\Delta P}{1715}$$ **Equation 3**

However, Purchased Horsepower is required. To receive this, the Hydraulic HP has to be divided by an assumed pump efficiency of 70%. This gives Brake Horsepower, which is then divided by an assumed motor efficiency of 80% to give Purchased Horsepower. This can be seen in **Table 7** below.

Table 7: Pump Sizing

Equipment	PFD Name	Flow (GPM)	Purchased Shaft Power (HP)	MOC	Type
Pumps	P-101	133.1	2.833	Ni	Centrifugal
	P-102	21.75	1.411	SS	Centrifugal
	P-103	3.225	0.2138	Ni	Centrifugal
	P-104	122.6	0.2553	Ni	Centrifugal
	P-105	89.68	6.256	Ni	Centrifugal
	P-106	32.93	2.109	Ni	Centrifugal
	P-107	112.2	7.277	Ni	Centrifugal

For piping, **Equations 4-7** were used. These equations were heuristics found in the Turton textbook [6]. **Equation 4** is used for liquid discharge, **Equation 5** for liquid suction, and **Equation 6** for vapor flow. This gives velocity (u). With the simulation giving the flow rate (Q), diameter can be calculated in **Equation 7**.

$$u = 5 + \frac{D}{3}$$ **Equation 4**

$$u = 1.3 + \frac{D}{6}$$ **Equation 5**

$$u = 20D$$ **Equation 6**

$$Q = \frac{\pi D^2}{4}(u)$$ **Equation 7**

Piping is also set to use Schedule 40 since the pressure of the liquid does not exceed the limit amount. [7]

Table 8: Piping Diameter size

Stream	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**
MOC	Ni	Ni	Ni	SS	Ni	Ni	SS	SS	SS
Dia (in)	6.500	3.500	3.500	2.000	8.500	1.500	3.000	3.000	1.500
Stream	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
MOC	SS	CS	Ni	Ni	Ni	Ni	Ni	Ni	Ni
Dia (in)	1.500	1.500	1.500	1.000	3.000	6.500	3.500	4.000	10.00
Stream	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**
MOC	Ni	Ni	Ni	Ni	Ni	Ni	Ni	Ni	Ni
Dia (in)	8.000	5.500	3.00	4.500	3.500	2.000	6.000	3.500	2.000
Stream	28	29	E-101	E-102	E-103	E-104	E-105	E-106	E-107
MOC	Ni	Ni	CS	CS	CS	CS	CS	CS	CS
Dia (in)	3.000	1.000	1.500	4.000	1.500	10.00	2.000	6.500	7.500
Stream	E-108	V-102	V-104	V-105	V-107	V-108			
MOC	CS	CS	CS	CS	CS	CS			
Dia (in)	1.500	2.000	2.000	2.000	2.000	2.000			

The flare system is sized using values from **Table 9**. The flare tip diameter (D_{min}) is calculated using the maximum velocity (V_{max}) and flow rate (Q) for the worst-case scenario overpressure event, in this case a fire using **Equation 8** [8].

$$D_{min} = 1.95\sqrt{\frac{Q}{V_{max}}} \qquad \textbf{Equation 8}$$

Using a fraction of heat intensity (f) value of 1 and a fraction of heat radiation (τ) equal to 0.2, the height of the flare (L) is calculated with **Equation 9**.

$$L = \sqrt{\frac{\tau * f * R}{4\pi * K}}$$ **Equation 9**

The purge gas requirement for the flare is calculated using a rounded diameter of 6 inches in **Equation 10**.

$$F_{pu} = 6.88 * D^2$$ **Equation 10**

The number of pilot burners is assumed to be based on flare diameter N=1. The pilot gas requirement is determined using **Equation 11**.

$$F_{pi} = 613N$$ **Equation 11**

A knock-out drum is also used as a part of the pressure relief system. The maximum design vapor velocity is determined with **Equation 12** using a vapor velocity factor (G) of value 0.2 and values from **Table 9**.

$$U = G\sqrt{\frac{\rho_l - \rho_v}{\rho_v}}$$ **Equation 12**

The required area (A) for the knock-out drum is determined using **Equation 13** below.

$$A = \frac{Q_a}{\left(60\,\frac{sec}{min}\right) * (U)}$$ **Equation 13**

The vessel diameter (d_{min}) is calculated using **Equation 14**.

$$d_{min} = 13.5\sqrt{A}$$ **Equation 14**

The vessel height (h) is calculated using **Equation 15**.

$$h = 3 * d$$ **Equation 15**

Table 9.1: Flare System and Knock-out Drum

Flare System	
Flare Tip	
Flare Tip Diameter (in)	6
Q (scfm)	37.5
R (Btu/hr)	52848
K (btu/hr*ft^2)	500
Length (ft)	41
Thickness (in)	0.2

Table 9.2: Flare System and Knock-out Drum

Knock-out Drum	
Velocity (ft/s)	1.77
Area (ft^2)	1.2
Min Diameter (in)	12
Height (in)	36
Thickness (in)	0.25
Purge Gas (Mscf/yr)	27.52
Pilot Gas (Mscf/yr)	613200
Amount stream (lbs/yr)	3.54E+08

Economics

Capital Cost

The capital cost includes every equipment listed. Each induvial cost is listed in **Table 10**. For the overall equipment cost, **Figure 2** shows the cost excluding the cost of tanks and **Figure 3** shows the total capital cost. It should be noted that the piping cost was based on length of the pipe. For the purpose of this process, it was assumed 20 feet of pipe was used per stream. Along with that, literature stated that Controls/Alarms were estimated to be around $500 for each instrument [9]. For calculating the capital cost, equations from Turton were used [6].

Table 10: Capital Costing of Equipment

Equipment	Cost ($)		Equipment	Cost ($)		Equipment	Cost ($)
T-101	419,700		TK-102	1,765,000		P-101A/B	104,400
T-102	1,081,000		TK-103	39,650,000		P-102A/B	60,790
V-101	33,700		TK-104	46,830,000		P-103A/B	109,700
V-102A/B	43,130		E-101	813,900		P-104A/B	106,300
V-103	24,060		E-102	358,100		P-105A/B	122,900
V-104A/B	68,550		E-103	1,037,000		P-106A/B	100,400
V-105A/B	68,550		E-104	2,464,000		P-107A/B	128,100
V-106	163,100		E-105	1,845,00		Piping	29,510
V-107A/B	513,800		E-106	1,012,000		PVR's	5,040
V-108A/B	513,800		E-107	1,339,000		Flare	38,420
TK-101	91,370,000		E-108	34,330		Instruments	41,000
Total Capital Cost ($)				**186,000,000**			

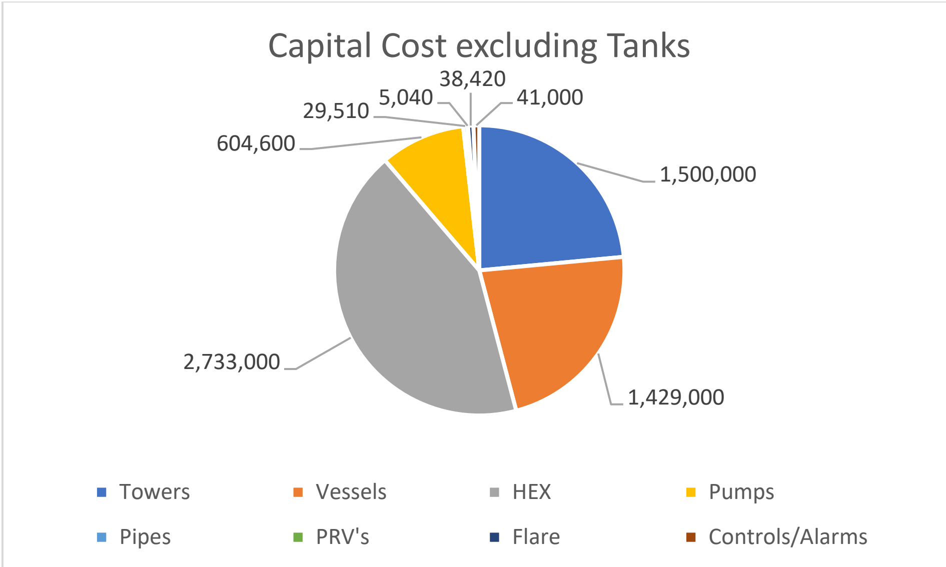


Figure 2: Capital Cost (Excluding Tanks)

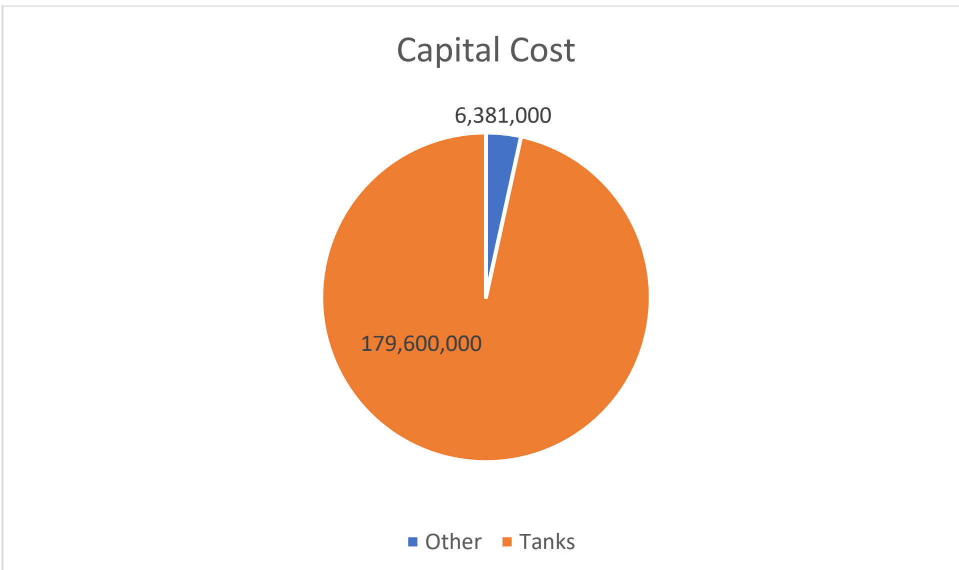


Figure 3: Capital Cost (Including Tanks)

Variable Cost

Variable Operating Cost was calculated using **Table 3** found above. With eight Heat Exchangers and fourteen pumps—spares included—**Table 11** below gives the prices for each individual Unit. It was assumed that the service factor for the year is 97%.

Table 11: HEX and Pump Operating Cost

Unit	Cost ($/yr)		Unit	Cost ($/yr)
E-101	832,000		P-101A/B	374.0
E-102	358,100		P-102A/B	186.0
E-103	1,037,000		P-103A/B	28.00
E-104	2,464,000		P-104A/B	34.00
E-105	1,845,000		P-105A/B	826.0
E-106	1,012,000		P-106A/B	278.0
E-107	1,339,000		P-107A/B	961.0
E-108	34,330			
Total HEX:	8,903,000		Total Pump:	2,687
Total Variable Cost ($/yr)		8,906,000		

Fixed Cost

The fixed costs estimate for this process includes labor costs, maintenance & repairs costs, and a catalysts & chemicals allowance. The annual values for the fixed costs are in **Table 12**.

Table 12: Fixed Costs Summary

Fixed Costs Summary	
Labor Costs ($)	223,700
Maintenance & Repairs ($)	11,160,000
Catalysts & Chemicals Allowance ($)	17,490,000
Total ($)	**28,870,000**

The labor costs in **Table 13** were calculated using a method found in [6], which calculates the number of operating laborers per shift using **Equation 16**.

$$N_{OL} = (6.29 + 31.7P^2 + 0.23N_{np})^{0.5} \quad \textbf{Equation 16}$$

This N_{OL} value can then be multiplied by 4.5, as it is typical to hire 4.5 operators for every required operator shift. This number is rounded up to the nearest whole number and multiplied by the typical annual average base salary for a process operator in Bali, Indonesia [10] to obtain an estimate for the cost of hiring operators per year. The cost of supervisory and clerical labor is estimated by multiplying this cost of operating labor by 0.18 [6]. The resulting costs for both operating labor and supervisory labor are added together to provide an estimate for the total labor cost per year for the process. These calculation values are all shown below in **Table 13**.

Table 13: Labor Costs Calculations

Labor Costs Calculations	
Number of Particulate Solids Process Steps, P	0
Number of Non-Particulate Solids Process Steps, Nnp	15
Number of Operating Laborers per Shift, N_{OL}	3.121
Total Number of Operating Laborers (Calculated)	14.04
Total Number of Operating Laborers (Rounded)	15
Operator Salary ($, converted from IDR) [10]	12,640
Cost per Year for Operator Labor ($)	189,600
Cost per Year for Supervisory Labor ($)	34,120
Total Labor Cost per Year ($)	**223,700**

Based on a heuristic found in the Turton textbook, the maintenance costs for the process can be estimated by multiplying the fixed capital investment (FCI) by 0.06 [6]. The total cost of maintenance per year is shown in **Table 14**.

Table 14: Maintenance Costs Calculations

Maintenance Costs Calculations	
Fixed Capital Investment (FCI) ($)	186,000,000
Maintenance & Repairs ($/yr)	**11,160,000**

The catalysts and chemicals allowance includes the annual cost of BASF Puricycle H & Puricycle HP adsorbents, silica adsorbent for dehydration, and nitrogen gas for regenerating the Puricycle H and HP adsorbents. These catalysts and chemicals are needed for the adsorbers in the design. As the team was unable to obtain actual information regarding the Puricycle adsorbents from BASF, an estimated cost of $100/kg was assumed for both the H and HP adsorbents in these calculations, along with a density of 700 kg/m^3. The volumes of adsorbent needed were calculated using the LHSV values provided in the AIChE Problem Statement [1]. The details for the adsorbent quantities and costs is shown in **Table 15**.

Table 15.1: Adsorbent Costing

Puricycle H Costing		
Vessel	Light Adsorber (V-104 A/B)	Medium Adsorber (V-107 A/B)
Volume of Adsorbent (ft^3)	25.83	719.3
Volume of Adsorbent (m^3)	0.7313	20.37
Density of Adsorbent (kg/m^3)	700	700
Cost of Adsorbent ($/kg)	100	100
Number of Vessels	2	2
Adsorbent Cost ($)	102,400	2,852,000

Table 15.2: Adsorbent Costing

Puricycle HP Costing		
Vessel	Light Adsorber (V-105 A/B)	Medium Adsorber (V-108 A/B)
Volume of Adsorbent (ft^3)	25.83	719.3
Volume of Adsorbent (m^3)	0.7313	20.37
Density of Adsorbent (kg/m^3)	700	700
Cost of Adsorbent ($/kg)	100	100
Number of Vessels	2	2
Adsorbent Cost ($)	102,400	2,852,000
Silica Costing		
Vessel	Py Gas Dehydration Adsorber (V-102 A/B)	
Mass of Adsorbent (kg)	78.81	
Cost of Adsorbent ($/kg)	0.80	
Number of Vessels	2	
Adsorbent Cost ($)	126.09	

Nitrogen gas costing was done based on cost-per-liter values [3]. Given the cost per liter and the number of liters per year, an estimate for the yearly cost of nitrogen gas for the regeneration of the Puricycle H and HP adsorbents can be obtained as shown below in **Table 16**.

Table 16: Nitrogen Gas Costing

Nitrogen Gas Costing		
Vessel	Light Adsorbers (V-104 A/B, V-105 A/B)	Medium Adsorbers (V-107 A/B, V-108 A/B)
Volume Flow of Nitrogen Gas (L/yr)	965,000	1,930,000
Cost of Nitrogen Gas ($/L)	2	2
Number of Vessels Actively Using	2	2
Cost of Nitrogen Gas per Year ($)	3,860,000	7,720,000

Process Safety

Minimized Environmental Impact

Being a project with the purpose of reducing environmental impact through recycling plastics, minimizing environmental impact is of maximum interest in this process. To make this process as safe for the environment as possible, a focus was placed on inherent safety to minimize risk. Tanks should be placed as physically far as possible from the outside of the plant where there may be more human or environmental consequences to a loss of containment. In selecting between a single-column and two-column design, the DCISI Index **(Figure 4)** was used to determine that a two-column design was inherently safer as shown in **Table 17** [11].

Score	Chemical			Process		Distillation	
	Auto-ignition temperature	Flammability	Explosiveness (%)	Operating Pressure (bar)	Temperature (°C)	Reflux Ratio	Relative volatility
1	Column temperature < Auto ignition temperature	Non-flammable	Non explosive	0.5-5	0-70	$R \leq 1.5$	$\alpha > 1.1$
2	-	Combustible (flash point > 55°C)	0-20	6-25	< 0 & 71-150	$R \leq 2.5$	-
3	-	Flammable (flash point < 55°C)	21-45	26-50	151-300	$R \leq 3.5$	-
4	-	Easily flammable (flash point < 21°C)	46-70	51-200	301-600	$R < 5$	-
5	Column temperature > Auto ignition temperature	Very flammable (flash point < 0°C & boiling point < 35°C)	71-100	201-1000	> 600	$R \geq 5$	$\alpha \leq 1.1$

Figure 4: DCISI Index Reference Table [11]

Table 17: DCISI Index for Single and Two-Column Design

Column		Single-Column	T-101	T-102
Score	Auto-ignition temperature	5	1	1
	Flammability	2	2	2
	Explosiveness	4	4	4
	Operating Pressure	1	1	1
	Temperature	4	3	4
	Reflux Ratio	5	3	1
	Relative Volatility	1	1	1
Total Score		**1.833**	**1.267**	**1.167**

In the case of process failure, particularly due to overpressure, final disposition systems were chosen to minimize the environmental impact. Where possible without risking further overpressure, the fluid passing through the pressure relief system was put back into an earlier stage of the process. For all other materials, liquid and gas streams were separated with a knockout drum. The final disposition for the liquid relieved is a sewer system assumed to already be at the facility with the capability of handling the toxicity and volume of liquid that may come from a disaster in this process. The toxic gas is flammable, and therefore sent to a flare system before being released.

The most likely factors to cause process failure are liquid level, temperature, and pressure in the distillation column, so these variables are carefully controlled and monitored with alarms for the case of dangerous values. The full P&ID is shown in **Figure 5**. It is also assumed that all firefighting equipment and required PPE are already available at the facility.

P&ID with Controls and Alarm

For the process to flow at optimal efficiency, controls were installed in key locations. Along with this, alarms were also rigged in order to notify operators in the event of containment breach and/or disastrous levels. All of this can be seen in **Figure 5** of the P&ID below.

To sum up the controls, the system needs a flow controller for the feed rate, temperature, level, and pressure indicators and controllers for each tower, along with other pressure and temperature indicators throughout the process.

Along with this, Indicators are installed all throughout the process, most notably on the tanks and vessels. These do not need any controllers; however, the additional information ensures that the operators can make informed decisions regarding the processes of the procedures for the fractionation of pyrolysis oil. For these indicators and transmitter, the following naming scheme was used as seen in **Table 18**.

Table 19 shows every controller loop used in the process and **Table 20** shows all alarms used in the process.

Table 18: Indicators and Transmitters Code

Code	Name	Purpose
TT	Temperature Transmitter	Transmits Temperature
TI	Temperature Indicator	Receives signal from transmitter and indicates the temperature to the operator
PT	Pressure Transmitter	Transmits Pressure Level
PI	Pressure Indicator	Receives signal from transmitter and indicates the pressure level to the operator

Table 19.1: Controller Loops

Controller Loop	**Control Variable**	**Manipulate Variable**
FT 100 FIC 100	Feed flow rate into E-101	Feed flow rate leaving P-101
TT 101 TIC 101	Temperature of Feed entering T-101	Steam flow rate entering E-101
TT 102 TIC 102 (Slave to AIC 100)	Temperature of T-101	Steam flow rate entering E-103
AT 100 AIC 100 (Master of TIC 102)	Composition of bottoms product leaving T-101	Steam flow rate entering E-103
LT 100 LIC 100	Liquid level in T-101	Liquid flow rate leaving P-104
TT 103 TIC 103	Temperature entering V-101	Cooling water flow rate entering E-102
LT 101 LIC 101	Liquid level in V-101	Liquid flow rate leaving P-102
LT 102 LIC 102	Liquid level in V-103	Liquid flow rate leaving P-103
TT 106 TIC 106	Temperature of fluid in TK-102	Steam flow rate entering E-108
TT 107 TIC 107 (Slave to AIC 101)	Temperature of T-102	Steam flow rate entering E-105
AT 101 AIC 101 (Master of TIC 107)	Composition of bottoms product leaving T-102	Steam flow rate entering E-105

Table 19.2: Controller Loops

TT 108 TIC 108	Temperature entering V-106	Cooling water flow rate entering E-104
LT 103 LIC 103	Liquid level in T-102	Liquid flow rate leaving P-106
LT 104 LIC 104	Liquid level in V-106	Liquid flow rate leaving P-105
TT 110 TIC 110	Temperature of fluid in TK-103	Steam flow rate entering E-107
TT 111 TIC 111	Temperature of fluid in TK-104	Steam flow rate entering E-106

Table 20.1: Alarms and Systems in P&ID

Alarm	Name	Function	Location
PSH 101	Pressure Switch High	Alerts facility of overpressure event	T-101
TAH 102	Temperature Alarm High	Alerts facility of high temperature event	T-101
LA 100	Level Alarm	Alerts control room of abnormal fluid level in tower	T-101
LAH 100	Level Alarm High	Alerts control room of high fluid level in tower	T-101
LAL 100	Level Alarm Low	Alerts control room of high low level in tower	T-101
PSH 107	Pressure Switch High	Alerts facility of overpressure event	T-102
TAH 107	Temperature Alarm High	Alerts facility of high temperature event	T-102

Table 20.2: Alarm and Systems in P&ID

LA 101	Level Alarm	Alerts control room of abnormal fluid level in tower	T-102
LAH 101	Level Alarm High	Alerts control room of high fluid level in tower	T-102
LAL 101	Level Alarm Low	Alerts control room of high low level in tower	T-102

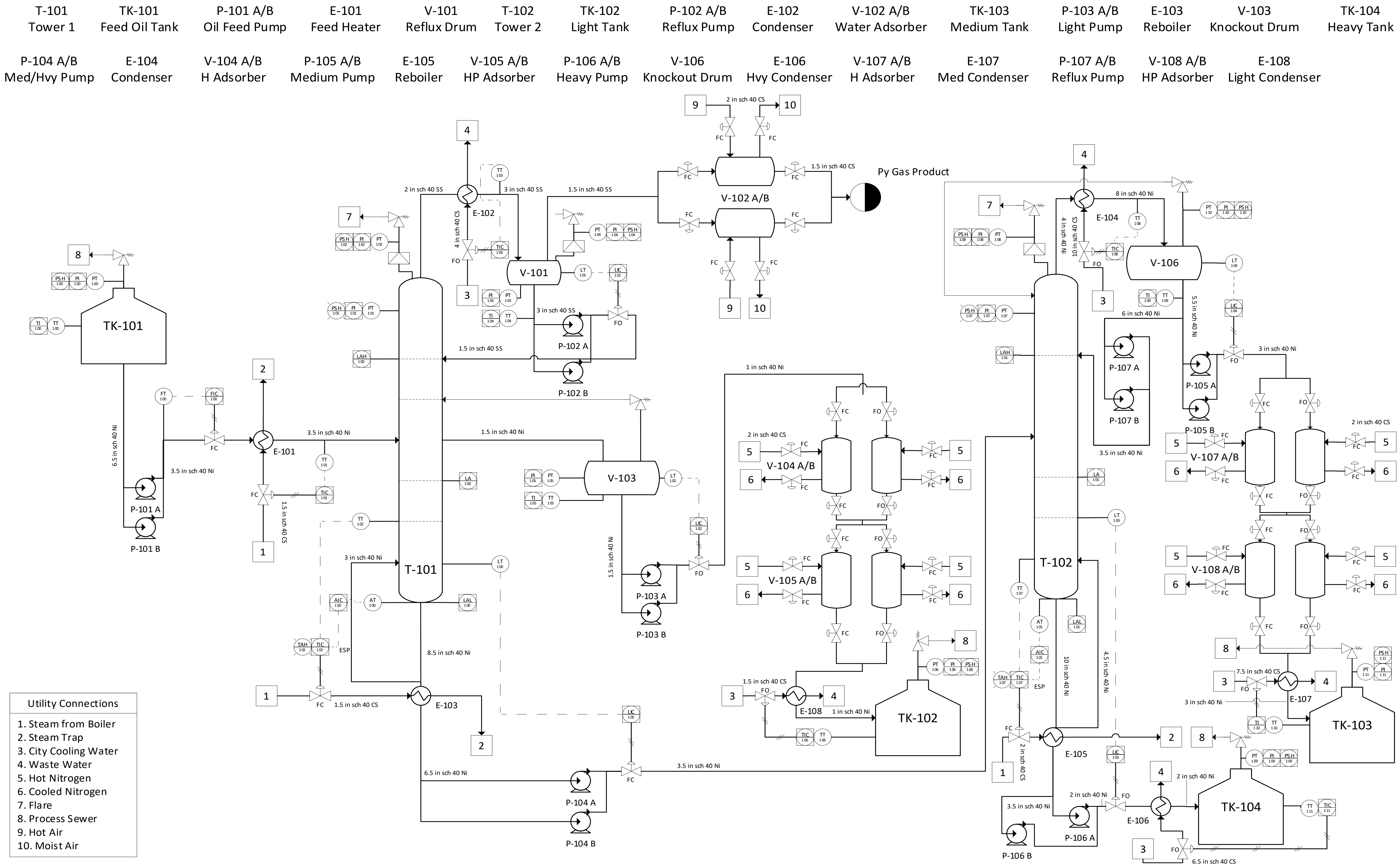

T-101 Tower 1
TK-101 Feed Oil Tank
P-101 A/B Oil Feed Pump
E-101 Feed Heater
V-101 Reflux Drum
T-102 Tower 2
TK-102 Light Tank
P-102 A/B Reflux Pump
E-102 Condenser
V-102 A/B Water Adsorber
TK-103 Medium Tank
P-103 A/B Light Pump
E-103 Reboiler
V-103 Knockout Drum
TK-104 Heavy Tank
P-104 A/B Med/Hvy Pump
E-104 Condenser
V-104 A/B H Adsorber
P-105 A/B Medium Pump
E-105 Reboiler
V-105 A/B HP Adsorber
P-106 A/B Heavy Pump
V-106 Knockout Drum
E-106 Hvy Condenser
V-107 A/B H Adsorber
E-107 Med Condenser
P-107 A/B Reflux Pump
V-108 A/B HP Adsorber
E-108 Light Condenser
Py Gas Product
Utility Connections
1. Steam from Boiler
2. Steam Trap
3. City Cooling Water
4. Waste Water
5. Hot Nitrogen
6. Cooled Nitrogen
7. Flare
8. Process Sewer
9. Hot Air
10. Moist Air

Figure 5: P&ID

Pressure Relief Valve Sizing

There are two types of pressure relief valves: liquid valves and vapor valves. All of the vapor valves are assuming critical flow, and this is determined by using the pressure and thermodynamic coefficient K from **Table 21** in **Equation 17** [12].

$$P_{cf} = P * \left(\frac{2}{(K+1)}\right)^{\frac{K}{(K-1)}} \qquad \textbf{Equation 17}$$

From this equation, the critical pressure was calculated to be 29.5 psia. This value, along with the coefficient of discharge, is a constant based on the ratio of specific heats k, set pressure plus overpressure allowance plus atmospheric pressure temperature, the compressibility factor, and the molecular weight for the valves was utilized in **Equation 18** to calculate the orifice area.

$$A = \frac{W}{K_D * C * P}\sqrt{\frac{T*Z}{M}} \qquad \textbf{Equation 18}$$

From this equation, a calculated area is obtained. Then, an orifice area that is larger than the calculated area is chosen. With this chosen area, the relief valves can be sized. For T-101, the size of the relief valve is 4 in by 6 in and for T-102, the size of the relief valve is 6 in by 8 in. The values for the calculations of the relief valve sizes are in **Table 21**.

Table 21: Pressure Relief Valve Calculation Values

Column	T-101	T-102
P(psia)	50	90
K	1.074	1.108
W(lb/h)	48.5	78473.95
MW	10499.74	180.9
Kd	0.62	0.62
C	323.5	326
T($R°$)	778.17	955.57
Z	1	1
Size (in x in)	4 x 6	6 x 8

The final disposition is assumed to have a pressure of 14.7 psia for the flare and process sewer, and pressure differences across discharge piping are equal to 0.5 psia per 100 ft of piping length [6]. These assumptions yield equal back pressure values for all tanks and vessels.

Failure Rate Analysis

For the process to run reliably, control systems must be operational at all times. Strict maintenance schedules can aid in this goal, but understanding how often the process control systems may fail helps in understanding where the process may be vulnerable to process disturbance or false response to disturbance. The failure rate of control systems as indicated by literature is shown in **Table 22**.

Table 22: Failure Rate of Control Systems [13]

Element	Failure Rate	Error Factor	Repair Time
Rupture Disk (Leakage)	2.000E-06/hour	10	N/A
Rupture Disk (Fail on Demand)	1.000E-04/demand	10	N/A
Temperature Sensors	1.900E-03/demand	5.8	3 hours
Temperature Sensors (Spurious)	7.000E-07/hour	2.6	3 hours
Pressure Sensors	7.000E-04/demand	3	2 hours
Pressure Sensors (Spurious)	8.700E-07/hour	2.5	2 hours
Flow Sensors	3.300E-04/demand	2.3	2 hours
Flow Sensors (Spurious)	4.300E-06/hour	5.3	2 hours
Level Sensors	2.100E-04/demand	3	3 hours
Level Sensors (Spurious)	8.200E-07/hour	5.6	3 hours
Reasonable Generic Failure Rate	1.000E-06/hour	3	N/A

The apparatus which fails most often are temperature sensors and spurious flow sensors. This should be kept in mind when creating maintenance schedules so that extra care may be taken to ensure these devices do not fail. For those apparatus which a specific failure rate could not be found in the literature, it was shown that a reasonable generic failure rate could be assumed. This is the rate that will be assumed for control systems such as AIC-100 and AIC-101.

Personal Exposure Risk

The pyrolysis oil purification process requires the use of hazardous, flammable, and toxic chemicals. All employees working in this process should be aware of the effects of contact with these chemicals and the safe exposure limits. A compilation of information about all chemicals used throughout the purification process is listed in **Table 23**.

Table 23.1: Personal Exposure Risk of Chemical Components

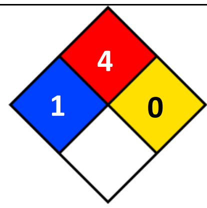
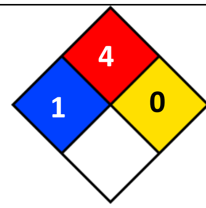
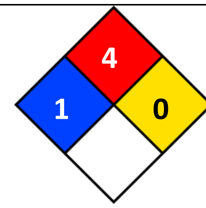
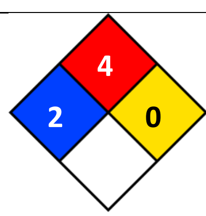
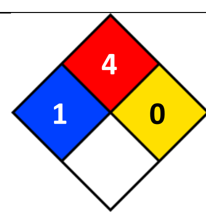
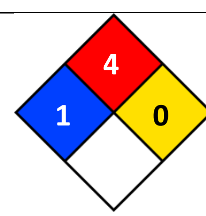
Chemical	**Methane [14]**	**Ethane [15]**	**Ethylene [16]**
OSHA Exposure	N/A	**ACGIH TLV** (United States, 3/2017). Oxygen Depletion [Asphyxiant].	**ACGIH TLV** (United States, 3/2019). TWA: 200 ppm 8 hours.
Hazard Diamond			
Fatal Exposure	N/A	N/A	N/A
Chemical	**Propane [17]**	**Propene [18]**	**n-Butane [19]**
OSHA Exposure	**NIOSH REL** (United States, 10/2016). TWA: 1800 mg/m³ 10 hours. TWA: 1000 ppm 10 hours. **OSHA PEL** (United States, 5/2018). TWA: 1800 mg/m³ 8 hours. TWA: 1000 ppm 8 hours. **ACGIH TLV** (United States, 3/2019). Oxygen Depletion [Asphyxiant]. Explosive potential.	**ACGIH TLV** (United States, 3/2019). TWA: 500 ppm 8 hours. **ACGIH TLV** (United States, 1/2005). TWA: 500 ppm 8 hours. Form: All forms	**NIOSH REL** (United States, 10/2016). TWA: 1900 mg/m³ 10 hours. TWA: 800 ppm 10 hours. **ACGIH TLV** (United States, 3/2017). STEL: 1000 ppm 15 minutes.
Hazard Diamond			
Fatal Exposure	>800,000 ppm	>65,000 ppm	658,000 mg/kg

Table 23.2: Personal Exposure Risk of Chemical Components

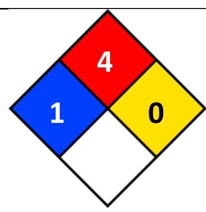
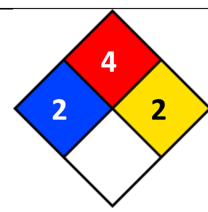
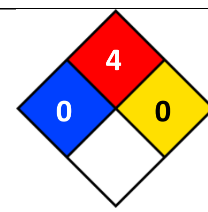
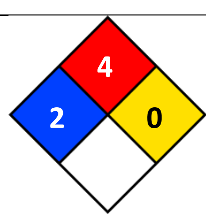
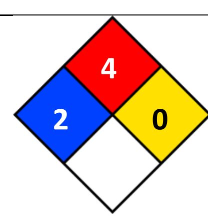
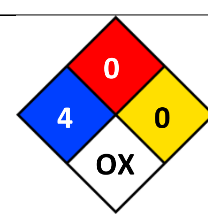
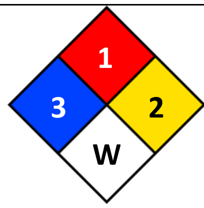
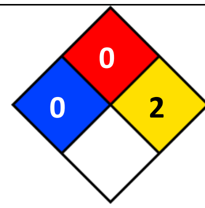
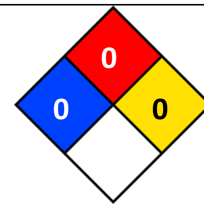
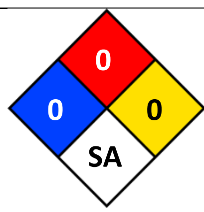
Chemical	1-Butene [20]	1,3-Butadiene [21]	n-Pentane [22]
OSHA Exposure	**ACGIH TLV** (United States, 3/2017). TWA: 250 ppm 8 hours.	**ACGIH TLV** (United States, 3/2017). TWA: 4.4 mg/m³ 8 hours. TWA: 2 ppm 8 hours. **OSHA PEL** (United States, 6/2016). STEL: 5 ppm 15 minutes. TWA: 1 ppm 8 hours.	**ACGIH TLV** (United States, 3/2019). TWA: 1000 ppm 8 hours. **NIOSH REL** (United States, 10/2016). CEIL: 1800 mg/m³ 15 minutes. CEIL: 610 ppm 15 minutes. TWA: 350 mg/m³ 10 hours. TWA: 120 ppm 10 hours. **OSHA PEL** (United States, 5/2018). TWA: 2950 mg/m³ 8 hours. TWA: 1000 ppm 8 hours.
Hazard Diamond			
Fatal Exposure	>200,000 ppm	>250,000 ppm	2,000 mg/kg
Chemical	**n-Hexane [23]**	**C6+ [24]**	**Chlorine [25]**
OSHA Exposure	**NIOSH REL** (United States, 10/2016). TWA: 180 mg/m³ 10 hours. TWA: 50 ppm 10 hours. **OSHA PEL** (United States, 5/2018). TWA: 1800 mg/m³ 8 hours. TWA: 500 ppm 8 hours.	N/A	**NIOSH REL** (United States, 10/2016). CEIL: 1.45 mg/m³ 15 minutes. CEIL: 0.5 ppm 15 minutes. **OSHA PEL** (United States, 5/2018). CEIL: 3 mg/m³ CEIL: 1 ppm
Hazard Diamond			
Fatal Exposure	25,000 mg/kg	N/A	>430 ppm

Table 23.3: Personal Exposure Risk of Chemical Components

Chemical	**Calcium [26]**	**Silica [27]**	**Water [28]**
OSHA Exposure	**OSHA PEL** TWA (Total Dust) 15 mg/m3 (50 mppcf*) **ACGIH TLV** TWA (inhalable particles) 10 mg/m3	**ACGIH TLV** (United States, 3/2019). TWA: 6.6 mg/m³ 8 hours. TWA: 5 ppm 8 hours. **NIOSH REL** (United States, 10/2016). TWA: 7 mg/m³ 10 hours. TWA: 5 ppm 10 hours.	N/A
Hazard Diamond			
Fatal Exposure	N/A	N/A	>90,000 mg/kg
Chemical	**Nitrogen [29]**		
OSHA Exposure	**ACGIH TLV** (United States, 3/2019). Oxygen Depletion [Asphyxiant].		
Hazard Diamond			
Fatal Exposure	174 ppm		

Atmospheric Detonation of Distillation Inventory

In the case of an overpressure or high temperature event, there is a possibility of column explosion. To prepare for the worst-case scenario, it was assumed that a detonation occurs when the column is filled completely with vapor. Detonation is calculated using **Equation 19** and values from **Table 24**.

$$W = \frac{P \cdot V_1}{k-1}\left[1 - \left(\frac{P_2}{P_1}\right)^{\frac{(1-k)}{k}}\right] \qquad \textbf{Equation 19}$$

Where TNT equivalency is calculated by dividing the work energy by the equivalent mass of TNT, or 4184 kJ/kg of TNT.

Table 24: TNT Equivalency Calculations for Distillation Columns

Column	**T-101**	**T-102**
Gauge Pressure, P (MPa)	0.4940	0.4940
Volume, V (L)	15960	42660
Initial Pressure, P_1 (MPa)	0.1013	0.1013
Final Pressure, P_2 (MPa)	0.5960	0.5960
Specific Heat Ratio, k	1.001	1.014
Stored Energy, W (kJ)	13950	36390
TNT Equivalent (kg)	**3.333**	**8.698**

As shown in the table, the TNT equivalent detonation of T-102 for a column filled with vapor is nearly three times as large as the TNT equivalence for T-101. Because T-102 is larger, it is logical that it would have a larger TNT equivalent detonation value.

Hazard and Operability Study

The largest column by volume in the pyrolysis oil purification process is T-102. For this column, a Hazard and Operability (HAZOP) Study was conducted to understand what variables detrimentally impact the process and ensure that the control system used is effective in managing the risk of these hazards. The results of the HAZOP Study are listed in **Table 25**.

Table 25.1: HAZOP for Largest Column, T-102

Guide Word	Deviation	Cause	Consequence	Action
NO	No flow into T-102	• Blockage in line 16 • Shutdown/failure upstream • P-104 failure • Valve failure	• Column dry out • Possible dangerous concentration • No operation • Pump cavitation/damage	• Install level alarm • Maintenance • Make bypass • Emergency plant shut down
	No cooling waters	• Pump failure • Source dried up	• Impure medium cut product/changes in product quality • High temperature in stream 19	• Maintenance • Emergency plant shut down
MORE OF	High level	• Blockage in line 23/24 • P-106 failure • Level control 103 failure • Valve failure	• Overpressure • Flooding • Changes in product quality	• Install high level alarm • Maintenance • Pressure relief system • Emergency plant shut down
	High temperature	• Fire • Loss of cooling water • Temperature control 107 failure • Analyzer control 101 failure • Valve failure	• Overpressure • Weeping • Changes in product quality	• Maintenance • Install high temperature alarm • Pressure relief system • Emergency plant shut down
	High pressure	• High level • High temperature • Valve failure	• Overpressure • Explosion • Changes in product quality	• Maintenance • Install high pressure alarm • Pressure relief system • Emergency plant shut down

Table 25.2: HAZOP for Largest Column, T-102

LESS OF	Low level	• Blockage in line 16 • Level control 103 failure • Pump 104 failure • Valve failure	• System failure • Pump cavitation/damage • Low pressure • Changes in product quality	• Maintenance • Install level alarm • Emergency plant shut down
	Low temperature	• Reboiler leak • Loss of high-pressure steam • Temperature control 107 failure • Analyzer control 101 failure • Valve failure	• Flooding • Loss of medium cut product in bottoms • Low pressure • Changes in product quality	• Maintenance • Install temperature alarm
	Low pressure	• Low temperature • Low level	• Changes in product quality	• Maintenance • Install pressure alarm
	Low atmospheric pressure	• Storm	• Plant damage	• Monitor severe weather
AS WELL AS	Water in column feed	• Failure in T-101	• Water in medium cut product • Poor product quality	• Maintenance • Emergency plant shut down

Improvement of Sorting Facility

Quantity Gap

While the proposed plan does well at educating and promoting participation in the local community, it seems to neglect to acknowledge the role of tourism in Bali's waste management problem. In a society whose primary industry is tourism, it is important to ensure that not only locals, but also tourists buy into recycling programs. This can be done with advertisements in the airport and at major tourist locations that advertise the proper way to recycle for Bali's facilities and providing well-labeled recycling receptacles in public areas for tourists and locals alike to use. Additionally, other locations with high levels of tourism have implemented recycling vending machines, or reverse vending machines, and have found incentives including free or reduced cost parking, free beach chairs, or a small amount of money back to be effective in encouraging recycling by tourists [30]. Bringing easy-to-use recycling programs to high-traffic tourist areas could be very effective in increasing the quantity of materials supplied to the Bali facility. Coca-Cola has found success with reverse vending machines in a collaboration with Merlin Entertainments group since 2022 [31]. For the same reasons, the quantity gap may be mitigated by working with commercial customers in addition to targeting households.

Quality Gap

To improve quality of materials within the sorting facility, flexible plastics could be included in the secondary plastic sorting to get a second set of eyes on these materials for which contamination is so adversarial. The second pass-through for all plastics would allow another opportunity for materials that are not fit for the steam cracker to be removed from the pyrolysis stream. Additionally, graphics provided at the site where recyclable materials are often acquired such as grocery stores which educate the population on proper waste management of their purchases may be effective in improving household segregation of waste and community buy-in to the program. Research shows the largest contributing factor to proper recycling as an individual choice is convenience. In fact, people may often throw non-recyclables into a recycling bin if that is closer to them than a normal trash bin [32]. From this information, it is recommended that the city put recycling and regular waste bins close together in public locations with easily understood information about how to recycle to make recycling as easy as possible and maintain quality.

Affordability Gap

With an expansion to high-traffic tourist sites, larger vehicles could be used to transport more waste at one time for these waste collection locations, reducing the overall cost of transportation. Additionally, if they are able, households could be encouraged to transport their own waste to the facility, also cutting down on transportation costs. The facility may also consider selling cleaned organic waste and allowing another location to compost the waste themselves to save space within the facility and devote the labor required for composting to sorting. The facility must consider the financial benefit of selling compost as opposed to organic waste for this purpose. The facility may also find financial benefit in working with more commercial customers, particularly from the tourism and manufacturing industries, where a larger fee can be charged for a higher volume of waste and less transportation costs are required.

Conclusions

The proposed design uses two distillation columns to fractionate Pygas, Pyoil light cut, Pyoil medium cut, and Pyoil heavy cut for use in a steam cracker. The Pygas is dehydrated in an adsorber, the Pyoil light cut and Pyoil medium cut streams have chlorides and heavy metals removed with two adsorbers in series, and the Pyoil heavy cut is stored for waste treatment, as it is unsuitable for use in the steam cracker. From the feed stream, 52,430 lb/hr of oil goes into the process, of that: 2486 lb/hr is Pygas, 1,199 lb/hr is Pyoil light cut, and 35,000 lb/hr is Pyoil medium cut are purified for use in the steam cracker while the rest is considered heavy cut and sent to a disposal site.

The project was found to cost $186,000,000 in capital costs, variable costs are $8,906,000 a year, and annual operating costs are $28,870,000. To minimize cost, a preheater was introduced. This allowed the feed to enter at a higher temperature, thus lowering the heat duty of the reboiler in T-101, allowing for a cheaper design. Among this, equipment material was carefully picked to allow the cheapest option while maintaining durability and longevity.

The design of this process emphasized inherent safety. A two-column design for the process was chosen because it was found to be inherently safer than a one column design, reducing hazards of this potentially dangerous process. Process controls and alarms, particularly focusing on level, temperature, and pressure, are included to increase process safety and ensure process efficacy. In the case of an overpressure event, a pressure relief system has been designed for the purification process. When necessary to remove material from a process, hazardous liquids are sent to a process sewer and hazardous gases are flared. Sizing for the pressure relief system was based on a fire.

Ideas for improvement in the effectiveness of the Bali Sorting Facility to improve the yield of pyrolysis oil to the purification unit include taking advantage of the high levels of tourism in the area by engaging tourists and engaging corporations or public facilities that would allow more plastic to be collected in one place than the current system which focuses on individual households does.

Appendices

Adsorption Section Detail

Adsorber for Water in Py Gas

As was mentioned in the problem statement from AIChE, the steam cracker of the ethylene plant cannot under any circumstance be fed free water. In this pyrolysis oil purification design, the first distillation column separates all water that could potentially be in the pyrolysis oil feed stream into the Py Gas stream. Therefore, the Py Gas stream requires dehydration to eliminate the possibility of free water being in its composition when it is fed directly to the ethylene plant.

The method of dehydration chosen for the Py Gas is an adsorber system filled with silica adsorbent. This adsorbent takes approximately 2-3 hours to regenerate using air at 293 F, and the calculated amount of silica in each of the two adsorber vessels (V-102 A/B) should adsorb any water within the Py Gas flow for about 4 hours [33]. After 4 hours, the vessels will switch roles, the adsorbing one swapping to regenerating and the regenerated one swapping to adsorbing.

To calculate the amount of silica adsorbent needed per vessel, the first metric needed was the estimated flow rate of water in the Py Gas stream. Because the composition given by the AIChE Problem Statement did not add up to 1, the assumption was made in this design that the remaining 0.03 wt% of the composition was water [1]. This led to 13.3 lb/hr of water within the feed stream and, because all the water left the distillation column through the Py Gas stream, 13.3 lb/hr of water in the Py Gas stream. This was converted into grams per hour of water to be adsorbed, then multiplied by 4 hours to obtain how many grams of water would need to be adsorbed in that time. It is estimated that after 4 hours, silica gel has adsorbed 30 grams of water per 100 grams of silica gel adsorbent [33]. Based on this, the mass of silica gel needed could be calculated. Costing of the silica adsorbent was done based on a $0.80/kg price found at an online vendor [34].

After the cost of the adsorbent was estimated, calculations were done to find the volume of adsorbent needed based on the silica gel’s density. Vessel sizing was done based on the optimal L/D ratio and choosing somewhat standard vessel lengths and diameters that give sufficient volume to hold the needed volume of adsorbent per vessel. All of the aforementioned calculation values can be found below in **Table 26**. Costing of the adsorber vessels was done with the other vessels in **Table 10** because the adsorbent costs are accounted for within the fixed costs catalysts & chemicals allowance section of this report.

Table 26: Silica Adsorption Calculations

Silica Adsorption Calculations	
Mass flow of H2O in Py Gas (lb/hr)	13.30
Mass flow of H2O Adsorbed (lb/hr)	13.30
Mass flow of H2O Adsorbed (g/hr)	6,031
Adsorbing Time (hrs)	4.000
Mass of H2O in 4 Hours of Flow (g)	24,120
Silica Needed to Adsorb 4 Hours of Flow (g)	80,410
Cost of Silica on $0.80/kg Basis [35] for 2 Vessels ($)	128.70
Density of Silica Gel [36] (kg/m^3)	720.8
Volume of Silica Needed per Vessel (ft^3)	3.944
L/D for Adsorber Vessels [6]	3.000
Vessel Diameter (ft)	1.25
Vessel Length (ft)	3.75
Vessel Volume (ft^3)	4.602

Adsorbers for Chlorides & Metals in Light and Medium Streams

The AIChE Problem Statement provides the insightful suggestion of using BASF Puricycle H and Puricycle HP adsorbents in series to “provide adequate protection” from chlorides and metals as contaminants within the Light and Medium streams. This design contains two adsorber systems in series for each stream that use the suggested adsorbents as prescribed. These are listed as V-104 A/B, V-105 A/B, V-107 A/B, and V-108 A/B in the PFD and in the costing calculations for the vessels. Adsorbent costing is accounted for within the fixed costs catalysts and chemicals allowance.

The BASF Puricycle H and Puricycle HP absorbents are assumed to be regenerable with by 500 °F nitrogen. Since adsorbers need zero flow of product flowing through them while they regenerate, the process would have to stop for regeneration if it was set up with a single adsorber bed per system. To counteract this, the design contains two adsorber beds per adsorption system to allow for one regenerating while the other adsorbs. This allows the process to be continuous. The adsorbent takes approximately 1 hour to saturate and approximately 1 hour to regenerate if using 500 °F nitrogen [33].

For vessel sizing, the LHSV value of 1 hr^{-1} from the AIChE problem statement was used in **Equation 20** [1, 35].

$$LHSV = \frac{Volumetric\ Flow\ of\ Liquid\ Feed\ per\ hr}{Volume\ of\ catalyst} \qquad \textbf{Equation 20}$$

This equation was used to find the volume of catalyst needed for the volumetric flow rate of feed given by the Aspen HYSYS simulation made for this purification process design. Because the Puricycle H and Puricycle HP adsorbents both have the same LHSV value, these volume calculations in **Table 27** below can count for both the H and HP adsorption systems on each stream. The volume of catalyst was used as a minimum value for potential vessel volumes, and along with an optimal L/D ratio of 3, was used to determine a feasible diameter and height for the adsorber vessels on each stream.

Table 27: Adsorbent and Vessel Volume Calculation Values

Adsorbent and Vessel Volume Calculation Values		
Stream Being Adsorbed	**Light**	**Medium**
LHSV (hr^{-1})	1	1
Volumetric Flow of Feed (ft^3/hr)	25.83	719.3
Volume of Adsorbent Needed (ft^3)	25.83	719.3
Vessel Volume (ft^3)	26.84	724.6
Vessel Diameter (ft)	2.25	6.75
Vessel Height (ft)	6.75	20.25

The design team did make an effort to contact BASF regarding their adsorbents but did not receive a response. With such limited information regarding their costs and densities, it was assumed by the team that both Puricycle H and Puricycle HP have costs of $100/kg and densities of 700 kg/m^3. These values were used to take the volumes of adsorbent needed (converted to m^3) and convert them first to masses in kilograms, then to costs in dollars. The results can be seen below in **Table 28**.

Table 28: Puricycle Adsorbent Costing

Puricycle Adsorbent Costing				
	H Light	**HP Light**	**H Medium**	**HP Medium**
Vol of Adsorbent per Vessel (m^3)	0.7313	0.7313	20.37	20.37
Number of Vessels per System	2	2	2	2
Total Cost of Adsorbent per Vessel ($)	102,400	102,400	2,852,000	2,852,000
Total H/HP Adsorbent Cost ($)	5,908,000			

Nitrogen gas for the regeneration of the Puricycle adsorbents is estimated to cost $2 per liter [3]. Based on flowrates of 0.5 gallons per minute for the Light stream adsorbers and one gallon per minute for the Medium stream adsorbers, the cost of nitrogen per year is $3,860,000 for the Light stream's adsorbers and $7,720,000 for the Medium stream's adsorbers.

During periods of unusually high levels of chlorides and metals, the design as it currently is should be able to sufficiently handle the feed. The volumes of adsorbent have been based on the total volumetric flowrate of the feed into the adsorbers, not of the metals or chlorides. Therefore, the adsorbent provided within the adsorbers as they are right now should be plenty sufficient to handle the extra contamination. Controls operation is also incorporated into the design, so if the adsorbers should become overwhelmed with contaminants, the level setpoints being sent to the LIC-104 and LIC-102 on the P&ID can be increased, causing the flowrates to the adsorbers to be decreased by the associated valves for a period of time.

Distillation Section Detail

The distillation design that was tested, optimized, and chosen is a two-tower configuration, designated as T-101 and T-102 for distillation towers one and two respectively.

For this design, pressure and temperature considerations for each individual tower were needed. The reboilers E-103 and E-105 utilize high pressure steam at 750 °F, while the condensers E-102 and E-104 use cooling water supplied at 87 °F. Reboiler pressures were calculated by adding 0.1 psi per stage, the condenser pressure drop, and the reboiler pressure drop to the condenser pressure.

For T-101, the top of the column had to have a minimum pressure of 2.4 psig. Because of this, a condenser pressure of 3 psig (17.7 psia) was chosen. This was to ensure that the Py gas vapor stream being sent to the ethylene plant would have sufficient pressure to reach the ethylene plant. Along with this, it doubled as a safety measure to allow the tower to operate above vacuum pressures, without the risk of air ingress into the process. The temperature in reboiler E-103 for T-101 had to be consistently above 392 °F because that is the Light cut's end boiling point. The Light stream and Py Gas are removed from tray 6 and through the overhead vapors of T-101, respectively.

For T-102, the pressure considerations were the same as in T-101, resulting in a condenser pressure of 3 psig. The temperature in reboiler E-105 needed to be above 620 °F because that is the end boiling point of the Medium cut. The Medium cut leaves T-102 through the overhead vapor stream, while the Heavy cut leaves as the bottoms product.

Modeling these towers in Aspen HYSYS, a few overall assumptions were made. The fluid packaging chosen for the model was Peng-Robison because it was found that engineers use it by default for atmospheric crude oil distillation, which is similar to this project [37]. It was also found that a "column with fewer than the minimum number of stages cannot achieve the desired separation", and for a binary distillation, this was seven [38]. Because this project is not a binary distillation and there are many more components involved, it was decided that ten stages would be the lowest feasible number for the tower configurations.

This is backed by an optimization calculation that involves multiplying the number of theoretical stages by the reflux ratio as the number of stages is varied. It is important to also adjust the feed stage location to minimize the reflux ratio with each variation of the total number of stages. The minimum value for the multiplied number of stages and reflux ratio is the most economically optimal column [6]. When the results are plotted, as shown in **Figure 6** and **Figure 7**, having a ten-tray column for both towers is optimal and efficiently allows minimization of overall energy consumption.

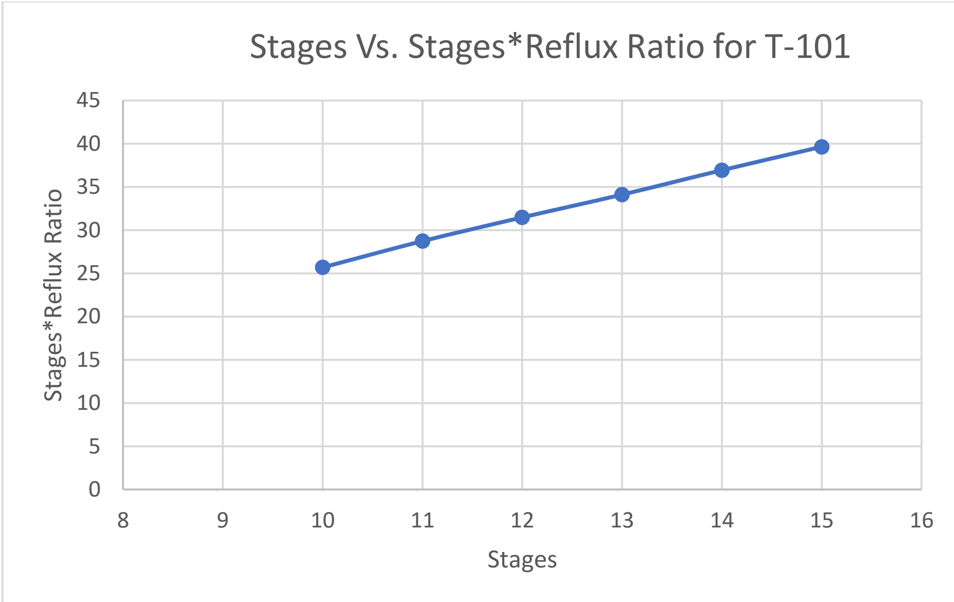


Figure 6: Stages Vs. Stages*Reflux for T-101

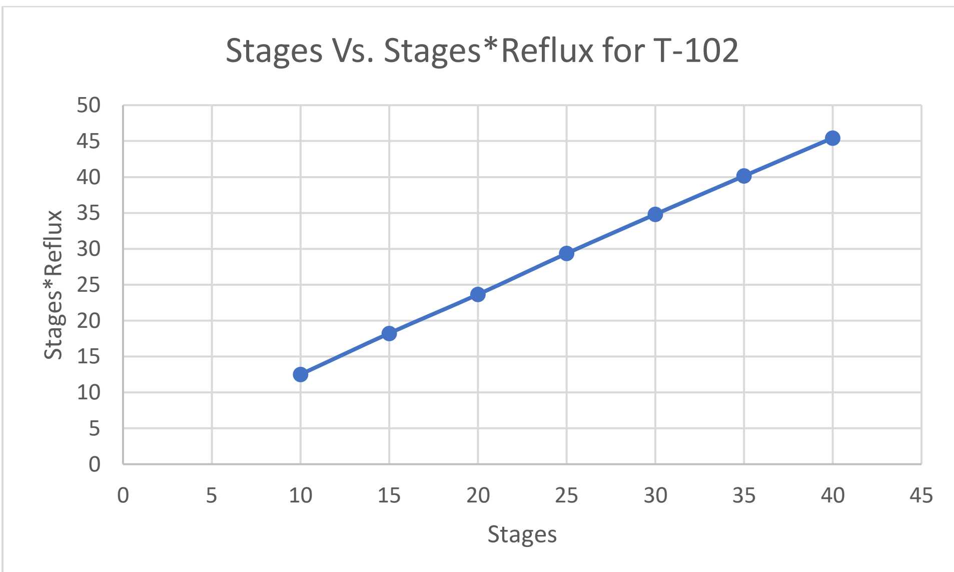


Figure 7: Stages Vs. Stages *Reflux for T-102

The distillation towers will be controlled to achieve their specifications in several ways. Each column has level alarms on both the top and bottom, as well as a level controller that manipulates the bottoms flowrate out of the column. This is to ensure that the levels inside remain safe and that they don't disturb the liquid/vapor flow profiles within the column. Each column also has a temperature controller, which manipulates the flowrate of high-pressure steam to the reboiler to allow the operator to control the temperature profile within the column. Additionally, the columns also have a pressure indicators and alarms to indicate when pressure is building above the desired amount within the towers. Finally, the temperatures of the reflux streams leaving the condensers are monitored, and the cooling water flowrate to the condenser is controlled. The reflux stream has an impact on the temperature profile within the tower. The temperature profile is essential to the separations within this design, so being able to control it is simply necessary to achieve specifications consistently.

Continuous removal of water occurs within the distillation column when it is neatly separated into the Py Gas stream (the overhead vapor of column T-101), which was shown to happen based on the composition of the Py Gas stream within the Aspen HYSYS model for this design. The Py Gas stream is then dehydrated using an adsorber system that is described within *Adsorber Section Detail*.

The Light stream was chosen to be removed as a liquid on tray 6 of T-101 because this location is above the feed stage by at least 2 stages and because it is above the point in the column where 392 °F is reached. Being above the feed stage is important because the feed liquid contains heavier components that do not belong in the Light cut. Being not one, but two stages above the feed tray significantly reduced the heavy component compositions in the Light stream.

The table below **(Table 29)** contains the most important data from our distillation columns as they were modeled in Aspen HYSYS. This is the data that was used in our column-related calculations the most.

Table 29: Distillation Column Information

Column	T-101	T-102
Number of Theoretical Stages	10	10
Number of Actual Trays	15	15
Feed Stage Location	8	6
Feed Tray Location	12	9
Light Draw Tray Number	6	-
Condenser Pressure (psia)	17.70	17.70
Condenser Temperature (°F)	281.9	420.9
Condenser Pressure Drop (psi) [6]	3	3
Reboiler Pressure (psia)	23.20	23.20
Reboiler Temperature (°F)	463.3	670.0
Reboiler Pressure Drop (psi) [6]	1.5	1.5
Reflux Ratio	2.569	1.250

The tray type chosen for both distillation towers is the sieve tray because they are less expensive than bubble cap or valve trays, much easier to clean during a shutdown, and come highly recommended for services in which corrosion or fouling are anticipated [6]. Within the distillation towers, there are temperature and traffic profiles for both vapors and liquids. This is displayed below in **Table 30**.

Table 30: Temperature and Vapor/Liquid Traffic Profiles from Aspen HYSYS Simulation [2]

Column	T-101			T-102		
Stage Number	**Temperature (°F)**	**Net Liquid (lb/hr)**	**Net Vapor (lb/hr)**	**Temperature (°F)**	**Net Liquid (lb/hr)**	**Net Vapor (lb/hr)**
Condenser	281.9	7995	2499	420.9	43470	27940
1	318.5	9458	10490	495.9	51440	78470
2	325.1	10090	11960	526.2	55740	86790
3	329.8	10580	12590	540.1	56990	90740
4	334.5	9802	13080	548.5	55140	91250
5	339.8	10100	13500	556.4	120300	90140
6	346.3	10240	13800	560.8	120600	90300
7	356.1	9649	13930	594.2	139300	106900
8	382.7	82120	13350	604.3	146800	124800
9	400.9	88770	33380	624.4	148400	133400
10	418.9	90790	40040	640.1	137800	135600
Reboiler	463.3	48730	42060	670.0	13730	124100

References

[1] Miksiewicz, E., Yeo, G., & Machado, J. (2023). *AIChE 2022-2023 Student Design Competition Problem Statement and Rules.* American Institute of Chemical Engineers.

[2] Haydary, J. (2017) "Chemical Process Design and simulation: Aspen Plus and aspen hysys applications." Aspen.

[3] University of Arkansas. *Business Services Procurement*. Procurement. Retrieved March 9, 2023, from https://procurement.uark.edu/e-procurement/liquid-nitrogen-price-list.php

[4] Svrcek and Monnery. "Design Two-Phase Separators Within the Right Limits." 1 Nov. 2022, Stillwater, OK.

[5] GPSA. (2004). *Gpsa Engineering Data Book*. GPSA.

[6] Turton, R., Shaeiwitz, J. A., Bhattacharyya, D., & Whiting, W. B. (2018). *Analysis, synthesis and design of Chemical Processes*. Prentice Hall.

[7] Ligon, M. (2021, July 20). *Buy the right PVC pipe: Schedule 40 and schedule 80 PVC*. Commercial Industrial Supply Product Specs, Industry Knowledge & More. Retrieved March 9, 2023, from https://www.commercial-industrial-supply.com/resource-center/schedule-40-vs-schedule-80/#:~:text=Schedule%2080%20pipe%20has%20thicker,while%20schedule%2080%20has%20a%20.

[8] Stone, D., Lynch, S. Pandullo, R. Evans, L. & Vatavuk, W. (2012). *Flares. part II. capital and annual costs - tandfonline.com*. Retrieved March 8, 2023, from https://www.tandfonline.com/doi/pdf/10.1080/10473289.1992.10467008

[9] Process Puzzler. (2018, June 25). *Control Systems: Rule out a rule-of-thumb | chemical processing*. Control Systems: Rule Out A Rule-Of-Thumb. Retrieved March 10, 2023, from https://www.chemicalprocessing.com/automation/control-systems/article/11312700/control-systems-rule-out-a-rule-of-thumb

[10] ERI Institute. *Waste water treatment plant operator salary Bali, Indonesia*. Salary Expert. Retrieved March 9, 2023, from https://www.salaryexpert.com/salary/job/waste-water-treatment-plant-operator/indonesia/bali

[11] Mohd Shariff, A. (2018). *Distillation column inherent safety index at preliminary design stage*. IOP Conference Series: Materials Science and Engineering. Retrieved March 10, 2023, from https://iopscience.iop.org/article/10.1088/1757-899X/458/1/012047

[12] Fike. (2000). *Technical bulletin TB8102 rupture disc sizing*. Retrieved March 8, 2023, from http://www-eng.lbl.gov/~shuman/XENON/REFERENCES&OTHER_MISC/tb8102.pdf

[13] Cadwallader, L C. (1998, September 1). *Selected component failure rate values from fusion safety assessment tasks*. Idaho National Lab, USDOE Office of Energy Research. Retrieved March 10, 2023, from https://www.osti.gov/biblio/752571/

[14] *Methane;* SDS No. 001033 [Online]; Airgas: Radnor, PA, November 15, 2020. https://www.airgas.com/msds/001033.pdf (accessed 3/5/23).

[15] *Ethane;* SDS No. 001024 [Online]; Airgas: Radnor, PA, July 15, 2021. https://www.airgas.com/msds/001024.pdf (accessed 3/5/23).

[16] *Ethylene;* SDS No. 001022 [Online]; Airgas: Radnor, PA, June 3, 2021. https://www.airgas.com/msds/001022.pdf (accessed 3/5/23).

[17] *Propane;* SDS No. 001045 [Online]; Airgas: Radnor, PA, November 15, 2020. https://www.airgas.com/msds/001045.pdf (accessed 3/5/23).

[18] *Propylene;* SDS No. 001046 [Online]; Airgas: Radnor, PA, November 5, 2020. https://www.airgas.com/msds/001046.pdf (accessed 3/5/23).

[19] *n-Butane;* SDS No. 001007 [Online]; Airgas: Radnor, PA, January 6, 2020. https://www.airgas.com/msds/001007.pdf (accessed 3/5/23).

[20] *1-Butene;* SDS No. 001009 [Online]; Airgas: Radnor, PA, January 31, 2018. https://www.airgas.com/msds/001009.pdf (accessed 3/5/23).

[21] *1,3-Butadiene;* SDS No. 001008 [Online]; Airgas: Radnor, PA, February 1, 2018. https://www.airgas.com/msds/001008.pdf (accessed 3/5/23).

[22] *n-Pentane;* SDS No. 001133 [Online]; Airgas: Radnor, PA, February 22, 2021. https://www.airgas.com/msds/001133.pdf (accessed 3/5/23).

[23] *n-Hexane;* SDS No. 001060 [Online]; Airgas: Radnor, PA, February 22, 2021. https://www.airgas.com/msds/001060.pdf (accessed 3/5/23).

[24] *C6+;* SDS No. 0021127 [Online]; Airgas: Radnor, PA, July 20, 2020. https://www.airgas.com/msds/021127.pdf (accessed 3/5/23).

[25] *Chlorine Gas;* SDS No. 001015 [Online]; Airgas: Radnor, PA, February 11, 2021. https://www.airgas.com/msds/001015.pdf (accessed 3/5/23).

[26] *Calcium Metal;* SDS No. S25217 [Online]; Fisher Science Education: Rochester, NY, October 24, 2014. https://www.fishersci.com/content/dam/fishersci/en_US/documents/programs/education/regulatory-documents/sds/chemicals/chemicals-c/S25217.pdf (accessed 3/5/23).

[27] *Silicon Metal Powder;* SDS No. S255121 [Online]; Fisher Science Education: Rochester, NY, October 24, 2014.

https://www.fishersci.com/content/dam/fishersci/en_US/documents/programs/education/regulatory-documents/sds/chemicals/chemicals-s/S25521.pdf (accessed 3/5/23).

[28] *Water;* SDS No. LC26750 [Online]; LabChem: Zelienople, PA, June 26, 2020. https://www.labchem.com/tools/msds/msds/LC26750.pdf (accessed 3/5/23).

[29] *Nitrogen;* SDS No. 001040 [Online]; Airgas: Radnor, PA, August 31, 2021. https://www.airgas.com/msds/001040.pdf (accessed 3/5/23).

[30] Taylor, M. (2021, October 10). *Reduce, reuse, recycle with the Reverse Vending Machine*. USC Viterbi School of Engineering. Retrieved March 9, 2023, from https://illumin.usc.edu/reduce-reuse-recycle-with-the-reverse-vending-machine/

[31] Srivastava, T. (2022, July 14). *Coca-Cola debuts prize-giving recycling vending machine at theme parks*. The Drum. Retrieved March 9, 2023, from https://www.thedrum.com/news/2022/07/14/coca-cola-debuts-prize-giving-recycling-vending-machine-theme-parks

[32] Schumaker, E. (2016, August 3). *This is why you have so much trouble recycling*. HuffPost. Retrieved March 9, 2023, from https://www.huffpost.com/entry/psychology-of-why-people-dont-recycle_n_57697a7be4b087b70be605b3

[33] Microtonano: Micro to Nano. (n.d.). *TIN Indicating Silica Gel Desiccant*. Micro to Nano. Retrieved March 9, 2023, from https://www.microtonano.com/TIN-Indicating-Silica-Gel-Desiccant.php

[34] Made in China: Shao, N. (2021). *Factory price wholesale silica gel, desiccant Silica gel orange to green and silica gel blue to pink*. Made in China. Retrieved March 9, 2023, from https://1e79fdbf44cb6be5.en.made-in-china.com/product/uZLfwPgVgYRz/China-Factory-Price-Wholesale-Silica-Gel-Desiccant-Silicagel-Orange-to-Green-and-Silica-Gel-Blue-to-Pink.html

[35] WhatWhenHow: What-When-How. *The hydrotreating process part 3*. What-When-How RSS. Retrieved March 9, 2023, from http://what-when-how.com/petroleum-refining/the-hydrotreating-process-part-3/

[36] Researchgate: Bahrun, Mohd Hardyianto Vai & Bono, Awang & Dzilrazman, Nur & Kamin, Zykamilia. (2020). Recovery of Minor Palm Oil Compounds Using Packed Bed Adsorption Column. Jurnal Bahan Alam Terbarukan. 9. 21-29. 10.15294/jbat.v9i1.23461.

[37] Gutierrez: Gutierrez, J. P., Benitez, L. A., Martinez, J., Ruiz, L. A., & Erdmann, E. (2014, April). *Thermodynamic Properties for the Simulation of Crude Oil Primary Refining*. Juan Pablo Gutierrez et al. Int. Journal of Engineering Research and Applications . Retrieved March 10, 2023, from https://core.ac.uk/download/pdf/52480795.pdf

[38] Papers.ssrn: Digieneni, Y., Timipere Salome, F., & Joseph, A. (2020, August 24). *Determination of the minimum number of stages in a binary distillation column using*

Excel. SSRN. Retrieved March 9, 2023, from https://papers.ssrn.com/sol3/papers.cfm?abstract_id=3677964#:~:text=The%20study%20also%20looked%20at,stages%20obtained%20were%20approximately%20seven.

[39] Jenkins, S. (2015, March 19). *Economic indicators: CEPCI*. Chemical Engineering. Retrieved March 9, 2023, from https://www.chemengonline.com/economic-indicators-cepci/?printmode=1