Letter of Transmittal

To: Global Petrochemical & Bali Project Team

From: Process Design Group $1.56 * 10^{28}$

Subject: Closing Critical Gaps to Enable a Circular Plastics Economy

The consumers of Global Petrochemicals demand action to reduce plastic waste, and the company has made a commitment to produce 10% of its virgin resin-quality plastics from recovered plastic waste. In order to accomplish this goal and continue to reduce their plastic-waste as part of their long-term goal, our team was tasked with design of a pyoil purification unit for supply of recycled feedstock to an on-site ethylene steam cracker. The unit that we have designed will purify a stream of pyrolysis oil, derived from plastic waste, into four separate streams to be sent to the ethylene plant or to be sent to other processing facilities. Additionally, our design takes process safety and economic performance into consideration.

The Bali Project Team has also asked for a cold eyes review of the Bali Manual Sorting Facility and recommendations for improved performance. This will help ensure sustainable operation of the community sorting facility and will ensure feedstock security for the development of a circular plastic recycling economy.

In this report, we have developed a purification unit that will primarily deliver feedstock to the adjacent Global Petrochemicals ethylene plant while minimizing the capital cost and variable operating cost. The design also includes treatment of contaminants, environmental concerns, and the health and safety of the personnel working this purification unit. We have also made recommendations for the Bali Project Team that will contribute to closing the recycling quantity, quality, and affordability issues that our design group was made aware of.

Technical and Economic Proposal

Closing Critical Gaps to Enable a Circular Plastics Economy

Group: $1.56 * 10^{28}$

March 10, 2023

Executive Summary

Each year, millions of tons of plastic waste are buried in landfills, burned, or released into the environment. This results in plastic waste making its way into our ecosystems and impacting natural life. To fix this problem of plastic pollution, we need to design a circular system to reuse the plastic that has already been produced and consumed. This will lower the amount of waste going to landfills, preventing excessive plastic waste from leaking into the environment. Global Petrochemicals is taking the first step in reducing plastic waste by replacing 10% of its virgin plastic production with recycled plastic.

The designed process looks at the purification of the pyrolysis oil. The pyrolysis oil has potential contaminants of chlorides, calcium, silica, and water. These contaminants can be present in concentrations up to 50 wppm each. Three adsorption columns were designed to remove the contaminants prior to distillation. To remove the contaminants, Amberlite, aluminum oxide, and activated carbon are utilized as adsorbents. For continuous operation, three adsorption columns are necessary for each type of adsorbent. This results in nine adsorption columns to be used for our process, with a total capital cost of $272,700.

After the contaminants are removed, the feed goes through two distillation columns in series. The first column separates the Py-Gas and Naphtha streams from the bottoms product. To save on energy costs, the bottoms product is then preheated before entering the second distillation column. The second column completes the final separation between the Gas Oil and the Pyoil Heavy Cut, which cannot be sent to the ethylene plant. Both towers use carbon steel sieve trays, resulting in the total capital cost for the distillation columns to be $1,106,400. The distillation columns produce Py-Gas, Naphtha, Gas Oil, and the Pyoil Heavy Cut streams. The Naphtha and Gas Oil streams were specified by Global Petrochemicals to have an End Boiling Point of 392 °F and 620 °F at atmospheric pressure, respectively.

With the addition of tanks, vessels, pumps, heat exchangers, and other processing equipment, the total capital cost for our design is $13,624,900. To operate our design, we have a variable cost of $2,252,100 and a fixed operating cost of $2,248,300 annually.

To further create a circular plastic economy, our team was asked to recommend improvements for a recycling program in Bali, Indonesia. The amount and quality of plastic collected are potential issues brought to us by the project team. To improve the quantity of plastic, we suggest that local companies adopt a sector of Bali to keep clean and host community clean up events. Since Bali is a popular tourist spot, installing vending machines around popular attractions can increase the recycling participation of tourists.

To increase the quality of the plastic collected, community members need to be informed. This can be done through flyers, efficient labelling, and random spot checks with participating families for improvement. Another way to increase quality is by including other facility equipment such as a shredder, magnet, and screens for easier separation of recycled goods.

Finally, a potential problem with the design of the Bali recycling proposal is the cost to run the program. By optimizing the pickup routes to run less often with larger quantity limits, we can lower the operating cost which will make the program more efficient to run.

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Process Description

The output from the pyrolyzer unit will be transported to the separation plant and enter the feed storage tank. This tank is designed to hold the feed to the pyrolysis purification unit for up to a week. From the feed tank, the stream is pumped to enter three adsorbers in series. The first adsorber contains the adsorbent Amberlite, which will remove the chloride contaminants. The second adsorber contains the adsorbent aluminum oxide to remove the calcium and silica present. The third and final adsorber contains activated carbon which will remove any water present in the stream. For each adsorbent type, three towers are required. During typical operation, the feed is directed through two adsorption towers of the same adsorbent. Once the contaminants have been removed from the stream, the clean feed enters the first distillation column on the eighth tray from the top. The first column contains twenty trays with the Py-Gas stream coming off as the overhead product. A side draw produces Naphtha, or the Pyoil light cut, which leaves the distillation column in vapor form on the eighteenth stage. The bottoms product from the first column is sent to a second distillation column for further separation. The overhead product of the first distillation column leaves as a vapor at 181.6 °F and 22 psia. The product stream then travels through a condenser where it is cooled to 107 °F and 20 psia. This stream is then split between the Py-Gas product stream and the reflux back into the column. The Naphtha side draw is sent to a fixed-tube heat exchanger that cools the stream from 392 °F to 100 °F and is then sent to a storage tank. From this storage tank, the Naphtha is pumped to reach a final pressure of 70 psig, where it is sent to a steam cracker for the ethylene plant.

The bottoms product exits the first distillation column at 470.6 °F and 24 psia. It is then heated by two fixed-tube heat exchangers in series that increase the temperature to 505.8 °F. The heated stream enters the second distillation tower at the twentieth stage from the top. The second distillation column contains a total of thirty trays. The Gas Oil, or Pyoil Medium Cut, comes off the second column as the overhead vapor. The bottoms product of the second distillation column comes off as the Pyoil Heavy Cut. The Gas Oil leaves the distillation column as a vapor at 534.9 °F and 22 psia where it then travels through two condensers in series. After the condensers, the stream is in liquid phase at 444.8 °F and 20 psia. The stream is split, sending reflux back into the column and producing the Gas Oil product. The Gas Oil product enters a second fixed-tube heat exchanger where the stream is condensed to 100 °F and is sent to a storage tank. From there, the Gas Oil is pumped to a steam cracker in the ethylene plant at 70 psig. The Pyoil Heavy Cut leaves the tower as a liquid at 730 °F and 24.5 psia where it is sent through two heat exchangers to bring the temperature down to 100 °F. The cooled stream is shipped to storage before pumping the liquid off-site at 50 psig to be used elsewhere. Table 1 shows the final product amounts from both the distillation columns.

An energy efficient aspect of our design includes heat integration between three heat exchangers. The bottoms product of the first distillation column is heated with the overhead product of the second distillation column. The overhead product of the second distillation column is condensed due to the temperature difference between the two streams. The second heat integration aspect is the heat exchanger before the inlet stream of the second distillation column. This stream is integrated with the condenser of the second distillation column, allowing the inlet stream to be pre-heated. This allows the overhead product of the second distillation column to be cooled to the necessary temperature for the distillation column to function. The third and final use of heat integration occurs between the Pyoil Heavy product stream of the second distillation column and the inlet stream entering the second distillation column. These heat integration strategies allow for an efficient transfer of heat while also reducing the operating cost of the facility.

The Process Flow Diagram for our design can be found in Figure 1. The Piping and Instrumentation Diagram in Figure 2 contains the recommended controls and pressure relief devices for the two distillation towers. Table 2 contains the information for the streams found in Figure 1.

Table 1: Distillation Columns Product Mass Flow Rate

T-104 & T-105	
Py-Gas ($\frac{lb}{hr}$)	1,331
Pyoil Light Cut (Naphtha) ($\frac{lb}{hr}$)	4,285
Pyoil Medium Cut (Gas Oil) ($\frac{lb}{hr}$)	39,980
Pyoil Heavy Cut ($\frac{lb}{hr}$)	6,831

Figure 1: Process Flow Diagram

T-101	Raw Pyoil Feed Tank
P-101 A/B	Raw Pyoil Feed Pump
T-101 A/B/C	Chloride Adsorbers
T-102 A/B/C	Metal Adsorbers
T-103 A/B/C	Water Adsorbers
T-104	Primary Dist. Tower
E-101	1st Reflux Condenser
V-102	1st Reflux Drum
P-102 A/B	1st Reflux Pump
E-102	Naphtha Condenser
P-103 A/B	Naphtha Prod. Pump
V-103	Naphtha Prod. Tank
E-103	1st Boilup Reboiler
P-104 A/B	1st Bottoms Pump
E-104	2nd Reflux Condenser
E-105	2nd Inlet Heat Exchanger
T-105	Secondary Distil. Tower
V-104	2nd Reflux Drum
P-105 A/B	2nd Reflux Pump
E-106 A/B	Gas Oil Condenser
V-105	Gas Oil Prod. Tank
E-107	2nd Boilup Reboiler
E-108	Heavy Cut Condenser
P-107 A/B	Heavy Cut Prod. Pump
V-106	Heavy Cut Prod. Tank
V-107	Knockout Drum
V-108	Liquid Seal Drum

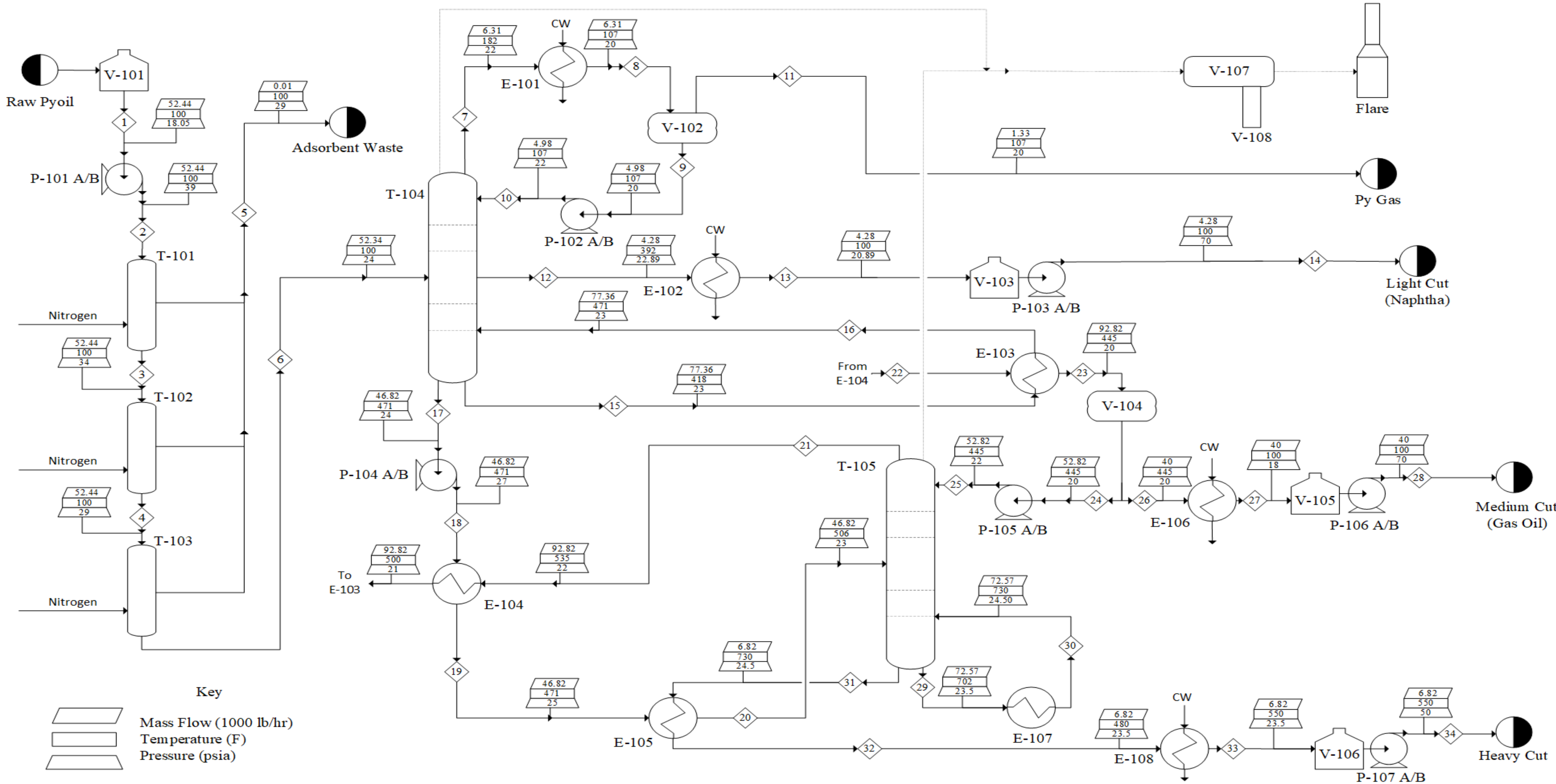


Figure 2: Piping & Instrumentation Diagram

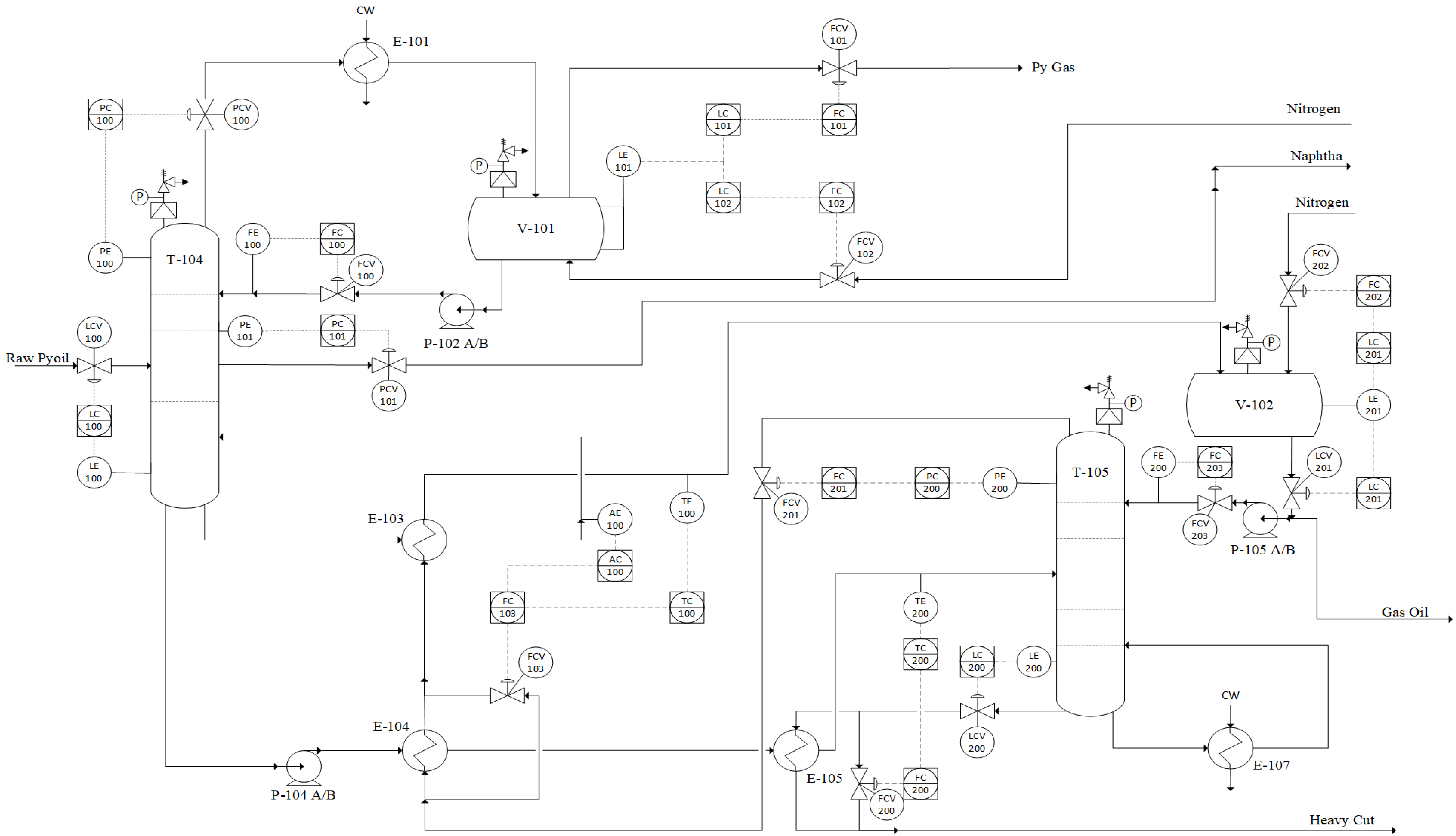


Table 2: Stream Tables (1 of 2)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Vapor Fraction	0	0	0	0	0	0	1	1	0	0	1	1	0	0	0	0	0	1
Temperature (F)	100	100	100	100	100	100	182	107	107	107	107	392	100	100	418	471	471	471
Pressure (psia)	18.05	39.00	34.00	29.00	24.00	24.00	22.00	20.00	20.00	22.00	20.00	22.89	20.89	70.00	23.00	23.00	24.00	27.00
Molar Flow ($\frac{lbmol}{hr}$)	288	288	288	288	-	288	85	85	61	61	24	32	32	32	232	232	232	232
Mass Flow ($\frac{lb}{hr}$)	52432	52432	52432	52432	10.486	52432	6306	6306	4975	4975	1331	4284	4284	4284	77360	77360	46816	46816
Std Ideal Liq Vol Flow (barrel/day)	4565	4565	4565	4565	-	4565	670	670	510	510	160	391	391	391	4013	4013	4013	4013
Molar Enthalpy Btu/lbmole)	-165085	-165085	-165085	-165085	-	-165085	-49571	-49571	-74485	-74485	-27518	-86281	-124741	-124741	-126276	-126276	-140379	-140379
Mass Density (lb/ft3)	48.03	48.03	48.03	48.03	-	48.03	0.25	0.25	40.56	40.56	0.19	0.36	45.78	45.78	38.46	38.46	38.28	38.28
Actual Volume Flow (barrel/day)	4666	4666	4666	4666	-	4666	108392	108392	524	524	30569	50666	400	400	13803	13803	5228	5228
Methane	0.0001	0.0001	0.0001	0.0001	0	0.0001	0.0039	0.0039	0.0001	0.0001	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ethane	0.0010	0.0010	0.0010	0.0010	0	0.0010	0.0223	0.0223	0.0024	0.0024	0.0024	0.0024	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ethylene	0.0005	0.0005	0.0005	0.0005	0	0.0005	0.0116	0.0116	0.0009	0.0009	0.0009	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Propane	0.0029	0.0029	0.0029	0.0029	0	0.0029	0.0520	0.0520	0.0159	0.0159	0.0159	0.0159	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Propene	0.0026	0.0026	0.0026	0.0026	0	0.0026	0.0473	0.0473	0.0127	0.0127	0.0127	0.0127	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
n-Butane	0.0036	0.0036	0.0036	0.0036	0	0.0036	0.0732	0.0732	0.0488	0.0488	0.0488	0.0488	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
1-Butene	0.0031	0.0031	0.0031	0.0031	0	0.0031	0.0600	0.0600	0.0363	0.0363	0.0363	0.0363	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
i-Butene	0.0031	0.0031	0.0031	0.0031	0	0.0031	0.0595	0.0595	0.0354	0.0354	0.0354	0.0354	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
13-Butadiene	0.0007	0.0007	0.0007	0.0007	0	0.0007	0.0143	0.0143	0.0088	0.0088	0.0088	0.0088	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
n-Pentane	0.0013	0.0013	0.0013	0.0013	0	0.0013	0.0424	0.0424	0.0438	0.0438	0.0438	0.0438	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
n-Hexane	0.0040	0.0040	0.0040	0.0040	0	0.0040	0.2646	0.2646	0.3307	0.3307	0.3307	0.3307	0.0008	0.0008	0.0000	0.0000	0.0000	0.0000
NBP[0]200*	0.0059	0.0059	0.0059	0.0059	0	0.0059	0.3486	0.3486	0.4638	0.4638	0.4638	0.4638	0.0594	0.0594	0.0022	0.0022	0.0003	0.0003
NBP[0]281*	0.0226	0.0226	0.0226	0.0226	0	0.0226	0.0003	0.0003	0.0004	0.0004	0.0004	0.0004	0.2310	0.2310	0.0430	0.0430	0.0127	0.0127
NBP[0]339*	0.1437	0.1437	0.1437	0.1437	0	0.1437	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.5416	0.5416	0.3355	0.3355	0.1616	0.1616
NBP[0]405*	0.2016	0.2016	0.2016	0.2016	0	0.2016	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1421	0.1421	0.3178	0.3178	0.2600	0.2600
NBP[0]480*	0.1807	0.1807	0.1807	0.1807	0	0.1807	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0192	0.0192	0.1450	0.1450	0.2046	0.2046
NBP[0]548*	0.2004	0.2004	0.2004	0.2004	0	0.2004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0051	0.0051	0.0946	0.0946	0.1930	0.1930
NBP[0]616*	0.1196	0.1196	0.1196	0.1196	0	0.1196	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0386	0.0386	0.0990	0.0990
NBP[0]684*	0.0576	0.0576	0.0576	0.0576	0	0.0576	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0143	0.0143	0.0413	0.0413
NBP[0]760*	0.0314	0.0314	0.0314	0.0314	0	0.0314	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0065	0.0065	0.0199	0.0199
NBP[0]821*	0.0136	0.0136	0.0136	0.0136	0	0.0136	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0025	0.0025	0.0076	0.0076
Chloride	0.00005	0.00005	0	0	0.25	0	0	0	0	0	0	0	0	0	0	0	0	0
Calcium	0.00005	0.00005	0.00005	0	0.25	0	0	0	0	0	0	0	0	0	0	0	0	0
Silica	0.00005	0.00005	0.00005	0	0.25	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	0.00005	0.00005	0.00005	0.00005	0.25	0	0	0	0	0	0	0	0	0	0	0	0	0

Table 2: Stream Tables (2 of 2)

19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34
1	1	1	1	1	0	0	0	0	0	0	1	0	0	0	0
471	506	535	500	445	445	445	445	100	100	702	730	730	480	550	550
25.00	23.00	22.00	21.00	20.00	20.00	22.00	20.00	18.00	70.00	23.50	24.50	24.50	23.50	23.50	50.00
232	232	489	489	489	278	278	211	211	211	261	240	21	21	21	21
46816	46816	92818	92818	92818	52820	52820	39998	39998	39998	72560	72569	6818	6818	6818	6818
4013	4013	8039	8039	8039	4575	4575	3464	3464	3464	6445	5896	549	549	549	549
-140379	-140379	-105159	-105159	-105159	-135493	-135493	-135493	-173414	-173414	-159510	-128473	-161114	-161114	-204836	-204836
38.28	38.28	0.42	0.42	0.42	38.54	38.54	38.54	48.28	48.28	35.09	476947.71	34.76	34.76	722.74	722.74
5228	5228	943170	943170	943170	5858	5858	4436	3541	3541	9671	-424	838	838	-638	-638
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0003	0.0003	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0002	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.0127	0.0127	0.0140	0.0140	0.0140	0.0140	0.0140	0.0140	0.0085	0.0085	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.1616	0.1616	0.1779	0.1779	0.1779	0.1779	0.1779	0.1779	0.1291	0.1291	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.2600	0.2600	0.2862	0.2862	0.2862	0.2862	0.2862	0.2862	0.2459	0.2459	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
0.2046	0.2046	0.2252	0.2252	0.2252	0.2252	0.2252	0.2252	0.2339	0.2339	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000
0.1930	0.1930	0.2123	0.2123	0.2123	0.2123	0.2123	0.2123	0.2615	0.2615	0.0050	0.0053	0.0018	0.0018	0.0018	0.0018
0.0990	0.0990	0.0841	0.0841	0.0841	0.0841	0.0841	0.0841	0.1208	0.1208	0.4123	0.4270	0.2468	0.2468	0.2468	0.2468
0.0413	0.0413	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4419	0.4411	0.4509	0.4509	0.4509	0.4509
0.0199	0.0199	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1140	0.1049	0.2171	0.2171	0.2171	0.2171
0.0076	0.0076	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0267	0.0217	0.0833	0.0833	0.0833	0.0833
0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Process Detail

Table 3 shows the mass balance associated with our process including each piece of equipment. The mass fraction composition of each distillation column, T-104 and T-105 are shown in Tables 4 and 5, respectively.

Table 3: Process Mass Balance

Equipment	In (lb_m/hr)	Out (lb_m /hr)	Difference
T-104	52432	52432	0
T-105	42031	42031	0
P-101	52432	52432	0
P-102	4975	4975	0
P-103	2657	2657	0
P-104	42031	42031	0
P-105	54906	54906	0
P-106	41458	41458	0
P-107	6986	6986	0
E-101	6306	6306	0
E-102	2657	2657	0
E-103	50131	50131	0
E-104	48444	48444	0
E-105 (Shell)	48444	48444	0
E-105 (Tube)	6986	6986	0
E-106	41458	41458	0
E-107	70357	70357	0
E-108	6986	6986	0

Table 4: Mass Fraction Composition of T-104

	Feed Composition	PyGas Composition	Naphtha Composition	Bottoms Composition
Methane	0.0001	0.0039	0	0
Ethane	0.001	0.0394	0	0
Ethylene	0.0005	0.0197	0	0
Propane	0.0029	0.1142	0	0
Propene	0.0026	0.1024	0	0
n-Butane	0.0036	0.1418	0	0
1-Butene	0.0031	0.1221	0	0
i-Butene	0.0031	0.1221	0	0
13-Butadiene	0.0007	0.0276	0	0
n-Pentane	0.0013	0.0512	0	0
n-Hexane	0.004	0.156	0.0005	0

NBP[0]200*	0.0059	0.0995	0.0396	0.0002
NBP[0]281*	0.0226	0	0.1972	0.0073
NBP[0]339*	0.1437	0	0.5524	0.1103
NBP[0]405*	0.2016	0	0.1715	0.2101
NBP[0]480*	0.1807	0	0.028	0.1999
NBP[0]548*	0.2004	0	0.0089	0.2236
NBP[0]616*	0.1196	0	0.0017	0.1338
NBP[0]684*	0.0576	0	0.0002	0.0645
NBP[0]760*	0.0314	0	0	0.0352
NBP[0]821*	0.0136	0	0	0.0152

Table 5: Mass Fraction Composition of T-105

	Inlet Composition	Gas Oil Composition	Heavy Composition
Methane	0	0	0
Ethane	0	0	0
Ethylene	0	0	0
Propane	0	0	0
Propene	0	0	0
n-Butane	0	0	0
1-Butene	0	0	0
i-Butene	0	0	0
13-Butadiene	0	0	0
n-Pentane	0	0	0
n-Hexane	0	0	0
NBP[0]200*	0.0002	0.0002	0
NBP[0]281*	0.0073	0.0085	0
NBP[0]339*	0.1103	0.1291	0
NBP[0]405*	0.2101	0.2459	0
NBP[0]480*	0.1999	0.2339	0
NBP[0]548*	0.2236	0.2615	0.0013
NBP[0]616*	0.1338	0.1208	0.2097
NBP[0]684*	0.0645	0	0.4428
NBP[0]760*	0.0352	0	0.2418
NBP[0]821*	0.0152	0	0.1044

The parameters used for the equipment design and sizing are shown in the following tables. Table 6 outlines the pump parameters; Table 7 contains the heat exchanger parameters, and Table 8 contains the adsorber parameters. Both Tables 9 and 10 contain the parameters used for the distillation columns, T-104 and T-105, respectively. The vertical and horizontal vessel parameters used are shown in Tables 11 and 12, respectively. Lastly, the tankage parameters are given in Table 13.

Table 6: Pump Parameters (1 of 2)

	P-101	P-102	P-103
Flow ($\frac{lb}{hr}$)	52432	4975	38.93
Fluid Density ($\frac{lb}{ft^3}$)	44.35	44.35	62.37
Shaft Power (kw)	1.93	0.25	0.25
Type/Drive	Centrifugal	Centrifugal	Centrifugal
Material of Construction	Carbon Steel	Carbon Steel	Carbon Steel

Table 6: Pump Parameters (2 of 2)

	P-104	P-105	P-106	P-107
Flow ($\frac{lb}{hr}$)	37.92	38.65	38.65	34.75
Fluid Density ($\frac{lb}{ft^3}$)	62.37	62.37	62.37	62.37
Shaft Power (kw)	1.65	2.54	3.98	0.37
Type/Drive	Centrifugal	Centrifugal	Centrifugal	Centrifugal
Material of Construction	Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel

Table 7: Heat Exchanger Parameters & Sizing (1 of 2)

	E-101	E-102	E-103	E-104
Type	Fixed Tube	Fixed Tube	Fixed Tube	Fixed Tube
Area (ft^2)	118	75	3751	3067
Duty ($\frac{btu}{hr}$)	975000	1221000	8456000	8802440
Shell				
Temperature In	87	87	415.1	534.9
Temperature Out	107	107	462.3	500
Phase	Liquid	Liquid	Vapor	Liquid
MOC	Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel
Tube				
Temperature In	181.7	392	500	470.6
Temperature Out	107	100	444.8	470.6
Phase	Liquid	Liquid	Liquid	Liquid
MOC	Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel

Table 7: Heat Exchanger Parameters & Sizing (2 of 2)

E-105	E-106	E-107	E-108
Fixed Tube	Fixed Tube	Fixed Tube	Fixed Tube
213	1453	1258	334
930400	7993000	7402000	1823600
Shell			
470.6	87	750	87
505.8	107	750	107
Liquid	Liquid	Steam	Liquid
Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel
Tube			
701.4	444.8	701.1	534.9
480	100	730	444.8
Liquid	Liquid	Liquid	Liquid
Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel

Table 8: Adsorber Parameters & Sizing

	T-101	T-102	T-103
Adsorbent	Amberlite	Aluminum Oxide	Activated Carbon
Particle Size (in)	0.0220	0.0041	0.0603
Adsorption Capacity	0.54	0.2	0.4
Adsorbent Bulk Density ($\frac{lb}{in^3}$)	0.0245	0.0208	0.0188
Adsorbate Percentage of Stream (%)	0.005	0.01	0.005
Adsorbent Needed (in^3)	9305	206799	57262
Packed Height (ft)	4.17	11.72	7.64
Volume (ft^3)	5.40	120	34

Table 9: Distillation Column T-104 Parameters & Sizing

T-104	
Real Stages	20
Feed Stage	8
Maximum Pressure (psia)	24
Rectifying Section Diameter (ft)	3.60
Rectifying Tray Type	Sieve
Rectifying Tray Spacing (ft)	2
Stripping Section Diameter (ft)	5.62
Stripping Tray Type	Sieve
Stripping Tray Spacing (ft)	2
Total Volume (ft^3)	882
Material of Construction	Carbon Steel

Table 10: Distillation Column T-105 Parameters & Sizing

T-105	
Real Stages	30
Inlet Stage	20
Maximum Pressure (psia)	25
Column Diameter (ft)	7
Tray Type	Sieve
Tray Spacing (ft)	2
Total Volume (ft^3)	77
Material of Construction	Carbon Steel

Table 11: Vertical Vessel Design

	V-102	V-108
Operating Pressure (psia)	20	20
Terminal Velocity (Ut)	2.06	2.06
Holdup Time (min)	5	5
Surge Time (min)	5	5
Material of Construction	Carbon Steel	Carbon Steel
Total Height (ft)	30.40	19.71
Area (ft^2)	1.30	97
Volume of vessel (ft^3)	40	1,911

Table 12: Horizontal Vessel Design

Vessel Design - Horizontal		
	V-104	V-107
Diameter (ft)	6.3	11.1
Length (ft)	67	19.7
Material of Construction	Carbon Steel	Carbon Steel
Pressure (psia)	22	22
L/D Ratio	2.0	2.0
Volume (ft^3)	2144	1911
Holdup time (min)	5	5
Surge time (min)	5	5
Terminal Velocity ($\frac{ft}{s}$)	1.48	1.25
Cross Sectional Area (ft^2)	31	97

Table 13: Tankage Design

	V-101	V-103	V-105	V-106
Hourly Mass Flow ($\frac{lb}{hr}$)	52,432	4,285	39,977	6,839
Weekly Mass Volume Flow ($\frac{ft^3}{week}$)	179,401	14,662	136,785	23,400
Weekly Volume (ft^3)	179,401	14,662	136,785	23,400

Economics

Capital Cost

To find the capital costs for each piece of equipment, an approach found in the Turton et al. text was used. The cost of equipment at the base conditions of carbon steel and near-ambient pressure at 14.7 psia is determined by Equation 1. The specific K values of various equipment is found in Appendix A.1 in Turton et al [53] . The A values are based on our specific size calculations.

[1] $$\log(C_{p0}) = K_1 + K_2 \log(A) + K_3 \log(A)^2$$

The actual capital cost for the equipment used in this project is calculated using a material of construction factor (Fm) and a pressure factor (Fp). The material of construction factor (Fm) defines a value found in Turton et al. The other variable that plays a role in the actual capital cost is the pressure factor. Since capital costs increase with increasing pressure, it is important to take this into consideration. Since the gauge pressure of the vessels and columns are under 0.5 barg and the thickness of the wall is less than 0.0063, the pressure factor will be equal to 1. These factors are multiplied by the purchased cost as seen in Equation 2.

[2] $$C_p = (C_{p0})(F_m)(F_p)$$

Table 14 displays the breakdown of the total capital costs of the entire process along with the equipment associated with each cost. These amounts are compared in a pie graph shown in Figure 3.

Table 14: Capital Cost Breakdown

T-101A	\$13,800	P-101B	\$72,200	P-107B	\$68,500	V-103	\$546,200
T-101B	\$13,800	P-102A	\$71,500	E-101	\$132,600	V-104	\$692,900
T-101C	\$13,800	P-102B	\$71,500	E-102	\$132,300	V-105	\$1,666,500
T-102A	\$50,400	P-103A	\$71,400	E-103	\$352,000	V-106	\$655,100
T-102 B	\$50,400	P-103B	\$71,400	E-104	\$317,400	V-107	\$632,000
T-102C	\$50,400	P-104A	\$70,800	E-105	\$430,700	V-108	\$632,000
T-103A	\$26,700	P-104B	\$70,800	E-106	\$227,500	Trays T1	\$174,500
T-103B	\$26,700	P-105A	\$75,500	E-107	\$430,700	Trays T2	\$332,300
T-103C	\$26,700	P-105B	\$75,500	E-108	\$148,800	Flare	\$69,200
T-104	\$140,600	P-106A	\$82,800	V-101A	\$1,991,300	Control Valves	\$15,800
T-105	\$459,000	P-106B	\$82,800	V-101B	\$1,991,300	Alarms	\$16,800
P-101A	\$72,200	P-107A	\$68,500	V-102	\$55,300	Pressure Relief Devices	\$84,000
Total: \$13,624,900							

The bare module cost for the pumps, heat exchangers, and vessels are calculated using Equation 3. To find the bare module cost for trays inside of the distillation tower, the quantity was taken into consideration in Equation 6. The B values needed for this calculation were found in Appendix A from Turton et al. These values were found using variables from 2001. In order to bring these costs up to the current date, the Chemical Cost Index from 2001 was used along with the Chemical Cost Index from 2021 in Equation 7. The CEPCI from 2001 is 397 and the CEPCI from 2021 is 708.

[3] $$C_{BM} = (B_1)(C_{p0}) + (B_2)(C_p)$$

[4] $$C_{BM} = (F_{BM})(C_{p0})(N)(F_q)$$

[5] $$C_{BM1} = (C_{BM2})\frac{CEPCI1}{CEPCI2}$$

Tables 15-21 explain in further detail the calculations for each independent piece of equipment. This equipment includes pumps, heat exchangers, towers, and vertical and horizontal vessels.

Table 15: Constants Used in Capital Cost Equations

	Pumps	Heat Exchangers	Towers	Vertical Vessel	Horizontal Vessel
K_1	3.3892	4.3247	3.4974	3.4974	4.8509
K_2	0.0536	-0.303	0.4485	0.4485	-0.3973
K_3	0.1538	0.1634	0.1074	0.1074	0.1445
F_m	1.6	1	1	1	1
C_1	0	0	0	0	0
C_2	0	0	0	0	0

C_3	0	0	0	0	0
F_p	1	1	1	1	1
B_1	1.89	1.63	2.25	2.25	1.49
B_2	1.35	1.66	1.82	1.82	1.52

Table 16: Capital Cost Parameters of Pumps

	P-101	P-102	P-103	P-104	P-105	P-106	P-107
Type	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal
Shaft hp (KW)	1.93	0.25	0.25	1.65	2.54	3.98	0.37
C_p^0	2612	2589	2582	2561	2731	2998	2480
C_p	4179	4142	4131	4097	4369	4796	3968
C_{BM}	$37,800	$37,387	$37,294	$36,981	$39,436	$43,300	$35,823
C_{TM}	$44,520	$44,200	$44,100	$43,638	$46,600	$51,100	$42,300

Table 17: Capital Cost Parameters of Heat Exchangers (1 of 2)

	E-101	E-102	E-103	E-104
Type	Fixed Tube	Fixed Tube	Fixed Tube	Fixed Tube
Area (ft^2)	118	75	3571	3067
C_p^0	15,360	15,323	40,776	36,772
C_p	15,360	15,323	40,776	36,772
C_{BM}	$90,200	$90,000	$239,300	$215,800
C_{TM}	$132,600	$132,300	$352,000	$317,400

Table 18: Capital Cost Parameters of Heat Exchangers (2 of 2)

	E-105	E-106	E-107	E-108
Type	Fixed Tube	Fixed Tube	Fixed Tube	Fixed Tube
Area (ft^2)	213	1,453	1,258	334
C_p^0	16,093	26,350	24,944	17,229
C_p	16,093	26,350	24,944	17,229
C_{BM}	$188,900	$154,600	$292,800	$101,100
C_{TM}	$277,900	$227,500	$430,700	$148,800

Table 19: Capital Cost Parameters of Vessels and Towers (1 of 2)

	T-101	T-102	T-103	T-104	T-105	V-101
Type	Vertical	Vertical	Vertical	Vertical	Vertical	Horizontal
Volume (ft^2)	0.15	3.40	0.96	17	77	5,080

C_p^0	1,595	5,834	3,091	16,288	53,176	230,697
C_p	1,595	5,834	3,091	16,288	53,176	230,697
C_{BM}	$6,500.00	$23,800.00	$12,600.00	$118,300	$386,000	$1,238,400
C_{TM}	$13,800	$50,400	$26,700	$140,600	$459,000	$1,991,300

Table 20: Capital Cost Parameters of Vessels and Towers (2 of 2)

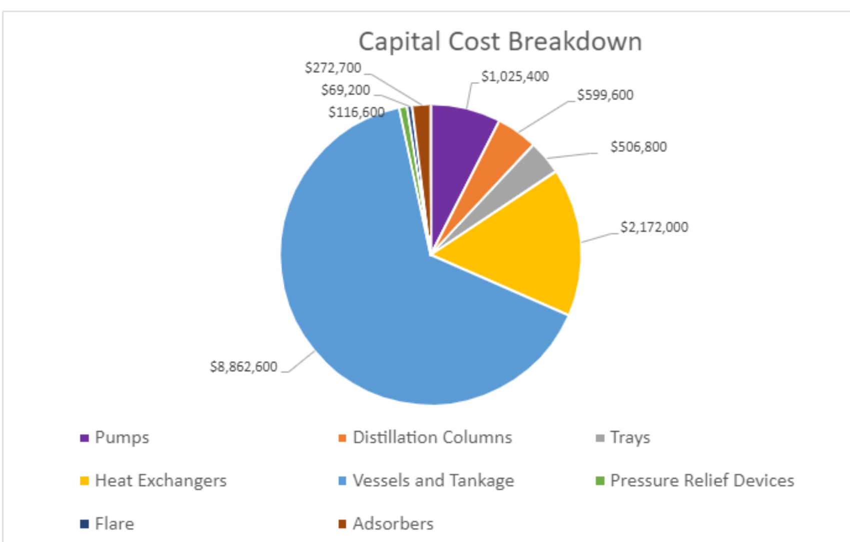
	V-102	V-103	V-104	V-105	V-106	V-107	V-108
Type	Vertical	Horizontal	Horizontal	Horizontal	Horizontal	Horizontal	Vertical
Volume (ft^2)	4	416	200	3,874	663	178	178
C_p^0	6,403	63,277	80,269	193,061	75,887	73,213	73,213
C_p	6,403	63,277	80,269	193,061	75,887	73,213	73,213
C_{BM}	$46,500	$339,700	$430,900	$1,036,400	$407,400	$393,100	$393,100
C_{TM}	$55,300	$546,200	$692,900	$1,666,500	$655,100	$632,000	$632,000

A Derrick flare was used in our process as the final disposition for any necessary components. The sizing of the flare is shown in Table 21, and the necessary equations were found in the Journal of the Air & Waste Management Association [49 & 50].

Table 21: Capital Cost of Onsite Flare

Flare	
Diameter (in)	48
Height (ft)	35
Cost	$69,200

Figure 3: Capital Cost Breakdown



Variable Operating Cost

The operating cost of all pumps is dependent on their yearly usage of electricity. The cost of electricity as a utility was given to us as $0.25 USD / kW-hr. The annual operating cost of all pumps is shown in Table 22.

Table 22: Operating Cost of Pumps

	P-101	P-102	P-103	P-104	P-105	P-106	P-107
Hydraulic Hp	1.89	0.17	0.25	1.16	1.78	3.90	0.37
Hydraulic Efficiency	0.84	0.62	0.84	0.62	0.62	0.84	0.84
Brake Hp	2.25	0.28	0.30	1.86	2.86	4.65	0.44
Motor Efficiency	0.87	0.84	0.87	0.84	0.84	0.87	0.87
Total Hp	2.58	0.33	0.34	2.22	3.41	5.34	0.50
Total Kw	1.93	0.25	0.25	1.65	2.54	3.98	0.37
Total Cost	$16,400	$2,200	$2,200	$14,100	$21,600	$33,900	$3,200

All condensers were costed as a function of the amount of cooling water used for each condenser. The cost of cooling water was given at $0.50 USD / MBTU. The parameters used to determine the MBTU needed, and the yearly operating cost of all condensers are shown in Table 23.

Table 23: Operating Cost of Condensers

	E-101	E-102	E-106	E-108
Duty ($\frac{Btu}{hr}$)	975,000	1,221,000	7,993,000	1,823,600
Mass Flow ($\frac{g}{hr}$)	13,658,959	17,105,220	111,975,444	25,547,157
Volume Flow ($\frac{gal}{min}$)	3,609	4,520	3,444	6,750
MBTU	8,284	10,375	7,905	15,495
Cost	$4,200	$5,200	$4,000	$7,800

The operating cost of the reboiler was a function of the volumetric flow rate and type of steam used. The type of steam used was high pressure steam, and the operating cost of the reboiler is displayed in Table 24.

Table 24: Operating Cost of Reboiler

E-107	
Volumetric ($\frac{kg}{hr}$)	4584
Mass (1000 kg)	38948
Cost	$2,021,500

Nitrogen was used in this purification process to regenerate the adsorbents, as well as for backflow into tankage when necessary. It was costed as a function of how much nitrogen would be used per year and is shown in Table 25. [10]

Table 25: Nitrogen Operating Cost

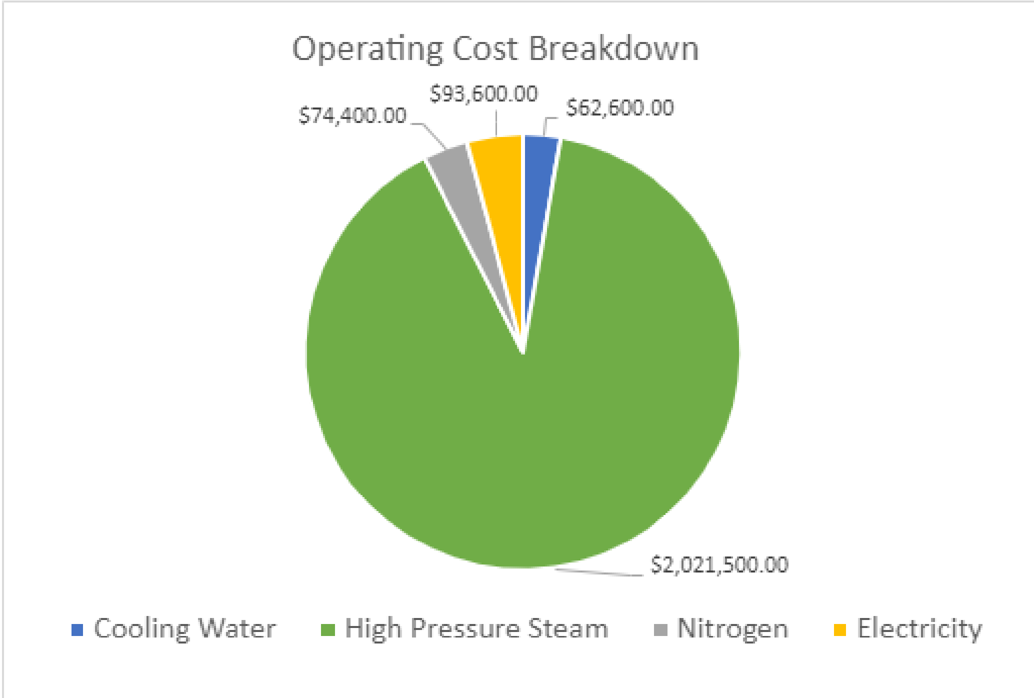
Nitrogen Operating Cost	
Mass Flow Rate ($\frac{m^3}{yr}$)	120000
Cost	$74,400

Table 26 outlines the total operating cost of the pumps which use electricity, the condensers which use cooling water, the reboiler using high pressure steam, and the nitrogen utility. A breakdown of cost per utility is shown in Figure 4.

Table 26: Total Operating Cost

Total Operating Cost	
P-101	$16,400
P-102	$2,200
P-103	$2,200
P-104	$14,100
P-105	$21,600
P-106	$33,900
P-107	$3,200
E-101	$4,200
E-102	$5,200
E-104	$37,400
E-105	$4,000
E-106	$4,000
E-107	$2,021,500
E-108	$7,800
Nitrogen	$74,400
Total	$2,252,100

Figure 4: Operating Cost Breakdown



Fixed Operating Cost

The fixed operating cost considers costs associated with the operators' salary, the need for absorbents to be purchased every year, as well as the maintenance associated with the plant's equipment. The fixed operating cost breakdown is shown in Table 27 and is further broken down in Figure 5.

Tables 27: Fixed Cost Breakdown

Fixed Operating Costs	
Labor	$1,127,800
Maintenance	$225,300
Adsorbents	$895,200
Total	$2,248,300

The equation to calculate the operating labor requirements for the chemical processing plant is shown in Equation 6. The equation considers the number of non-particle processing steps and processing steps in the plant. The annual salary from 2016 was brought to 2021 by using the Chemical Engineering Plant Cost Index ratio giving us our annual operating cost (AOC) for labor [53]. Table 28 outlines the variables used to calculate the labor cost.

[6] $$N_{OL} = \left(6.29 + 31.8P + 0.2N_{np}\right)^{0.5}$$

Tables 28: Labor Fixed Operating Costs

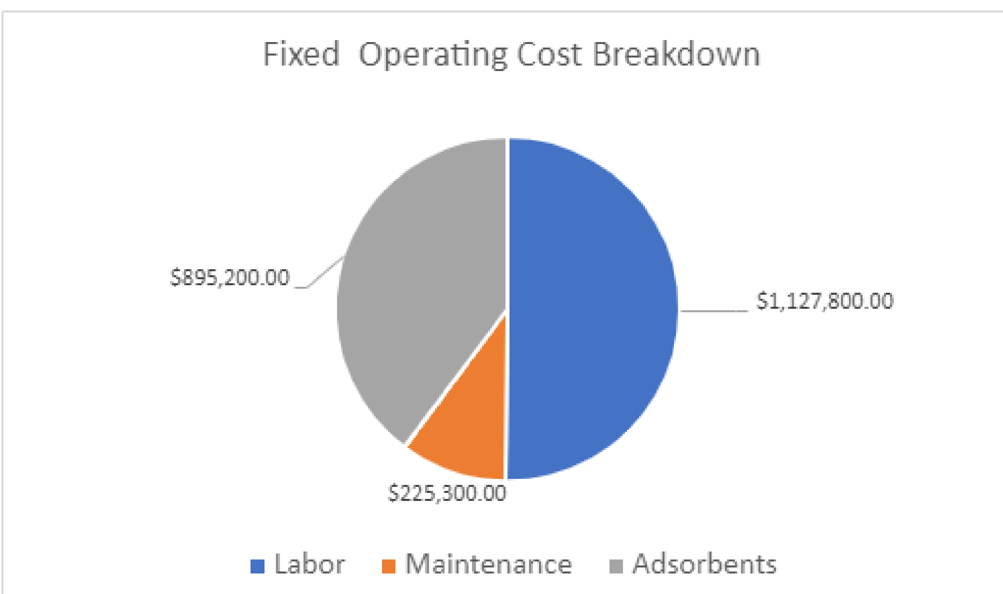
Labor	
P	0
Nnp	8
NOL	2.85
NOL*NO	$869,830
Yearly Salary (2016)	$66,910
AOC (2021)	$1,127,800
CEPCI (2016)	541.7
CEPCI (2021)	702.3
Maintenance	$225,300

Plant maintenance was taken as a percentage of the operating cost of the plant, it was assumed to be 10% of the total variable operating cost. The absorbents used and their fixed operating cost are shown in Table 29 [24,25,30]. Figure 5 combines the fixed operating cost, showing labor as the largest cost.

Table 29: Adsorbent Costing

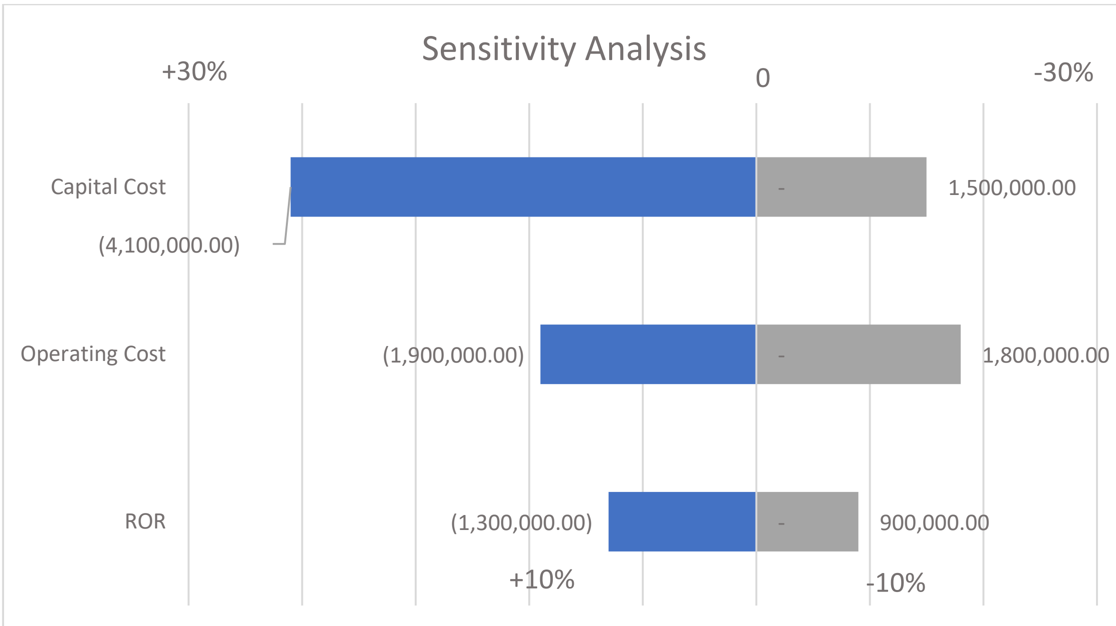
	T-101	T-102	T-103
Adsorbent	Amberlite	Aluminum Oxide	Activated Carbon
Volume (in^3)	9305	206799	57262
Mass (kg)	103	1954	489
Cost	$27,500	$244,300	$26,600

Figure 5: Pie Graph of Fixed Operating Cost



An economic analysis was conducted to compare variances in the capital cost, operating cost, rate of return (ROR), and how these would affect the net present value (NPV). Assuming a ROR of 15%, we varied the capital cost and operating cost of the process by 30%. Figure 6 shows the change in NPV from its original value. For the capital and operating cost, the NPV greatly increases when varied by +30% than its counterpart of –30%. Since the ROR was assumed, we wanted to see how a 10% change in the ROR would affect the NPV. A 10% difference in ROR would only have around a 1 million variation on the original NVP in both the positive and negative directions.

Figure 6: Sensitivity Analysis



Process Safety

Minimizing Environmental Impacts

In order to safely contain process materials and effectively treat possible harmful discharges to the environment, we chose to use tankage with a fixed roof on concrete pads. In addition to the necessary tankage, the feed tank is sized up in case of leakage elsewhere in the process or necessary removal of material from another tank. The tanks themselves also contain pressure relief devices in order to emit excess pressure when necessary. The Pyoil feed composition can vary which can result in variances in the water, chloride, and metals contaminants. One way to ensure the removal of water from the stream before entering the distillation column is with multiple adsorbers. The primary adsorber will remove most contaminants with the secondary adsorber being a guard adsorber in order to remove anything that passed through the first adsorber. A third adsorber will be used for regeneration as detailed below in the adsorption section. Once water is removed by these adsorbers, no water should enter the process unit as purification continues. The Py-Gas must maintain a minimum pressure above 2.4 psig in order to be fed to the ethylene plant. This will be ensured by the specification of the overhead product pressure of the first distillation column at 20 psia, and by using a pressure indicator at the top of the column to monitor the pressure of Py-Gas leaving the column.

Some of the main process concerns are the possibility of harmful discharges to the environment, pressure vessel bursts, a boiling liquid expanding vapor explosion, and a fire. In order to mitigate the possibility of harmful discharges to the environment, we have included rupture disks as well as pressure relief valves on both distillation columns. These relief devices feed to an onsite flare in order to combust vented material. Along with pressure indicators on the distillation columns, the tankage and vessels also have pressure indicators that way operators will be aware of any potential overpressure problems. The tanks also have minimal heat and pressure input into them, as all the material entering the tanks is kept at 100°F and below 30 psia. This will help reduce thermal hazards that are associated with pressurized vessel bursts. A boiling liquid expanding vapor explosion is overall the worst-case scenario due to the size and amount of heat being inputted into either of the distillation columns. The distillation columns are designed with a pressure indicator at the top of the column along with a rupture disk to notify operators if an overpressure event is occurring. Additionally, the columns should be designed, and tested in accordance with ASME standards. If an overpressure event occurs, the final disposition of the pressure needing to be relieved would be a flare. There are highly flammable materials present in the distillation columns, and exposure to oxygen would cause an overpressure event and subsequent fire. As stated above, prevention measures for this are the pressure indicator at the top of both columns as well as rupture disks to indicate if overpressure has occurred.

There are safety concerns along with the flammability of the components being used. Some of the most dangerous chemicals from the Personnel Exposure Risk outlined below are Benzene, 1,3-Butadiene, Hydrogen Sulfide, Xylene, Octane, and Hexane. These were chosen as dangerous chemicals based on their OSHA Chemical Exposure limits along with their LD50 values. Operators who are working should be made aware of these chemicals and wear proper protective equipment to combat lengthy exposure. Environmental concerns that the operators should be aware of include the loss of electric power, which would cause the pumps and control alarms to not work properly, and the possibility of natural disasters such as earthquakes and tsunamis based on the plant's location. The natural disasters could cause structural damage to the plant which could be detrimental.

Pressure Relief Valve Sizing

To size the pressure relief, the maximum flow rate required to depressurize the distillation column is required. Commonly, a fire outside of the column causes the highest relief flow rate. As seen in Table 30, the relief flow rate required for an external fire situation is 34 lb/hr for the first distillation column. Since this value is low, our team also calculated the required flow rate for losing cooling water supply to the distillation column. As seen in Table 30, this created a minimum required relief flow rate of 4,975 lb/hr for the first distillation column. The calculations for the fire scenario and loss of cooling water were found in API Standard 521. Using the value from the loss of cooling water scenario and Equation 9, we calculated the required orifice area. This area was found to be 1.194 in^2. By rounding up to the next largest relief valve area, we get an area of 1.287 in^2 for the pressure relief on the first distillation column, T-104. [13]

Similarly, the second distillation column's relief valve can be calculated. Table 30, shows the calculations for the external fire situation, while Table 31 shows the calculations for the loss of cooling water scenario. Again, the external fire situation has a low required relief flow. Due to the large heat of vaporization value, more energy would need to be put into the system to increase the relief flow rate. This explains why the external fire situation does not create the worst-case scenario. For the loss of cooling water, we found a relief flow rate of 39,998 lb/hr, requiring an orifice area of 11.5 in^2. By sizing up to the next possible relief valve size, we get an actual relief valve area of 16 in^2 for the second distillation column, T-105.

[7]
$$\dot{m} = C_0 A P_0 \cdot \sqrt{\left[\frac{2 g_c mw}{RT_0}\frac{\gamma}{\gamma-1}\left[\left(\frac{P}{P_0}\right)^{\frac{2}{\gamma}} - \left(\frac{P}{P_0}\right)^{\frac{\gamma+1}{\gamma}}\right]\right]}$$

Table 30: Required Flow from Fire for T-104 and T-105

	Distillation Column 1	Distillation Column 2
Required Flow from Fire		
Heat Flux from fire ($\frac{BTU}{(hr)(ft^2)}$)	47,550	47,550
Insulation Factor	1	1
Wetted Area (ft^2)	113	113
Heat Duty from Fire	2,296,051	2,296,051
Heat of Vaporization ($\frac{BTU}{lbmol}$)	68,620	38,170
Required Relief Rate ($\frac{lb}{hr}$)	34	62

Table 31: Required Flow from Loss of Cooling Water

	T-104	T-105
P (psia)	24	25
MAWP (psia)	26.67	27.78

P (psia)	15.63	15.73
gamma	1.25	1.02
MW ($\frac{lb}{lbmol}$)	182	201.7
T_0	642	995
C_0	0.975	0.975
G_C	32.2	32.2
Gas Constant ($\frac{ft^3(psi)}{R°(lbmol)}$)	10.73	10.73
Required Relief Rate ($\frac{lb}{hr}$)	4,975	39,998
Area (in)	1.19	11.50
Orifice Actual Area (in)	1.28	16

Failure Rate Analysis

A fault is defined as an abnormal condition that may cause a reduction in, or loss of, the capability of a functional unit to perform a required function. Whereas a failure is defined as the termination of the ability of a functional unit to perform a required function.

Fault rates of the instrumentation in Figure 7 were taken from the book “Process Safety Calculations” [5]. The typical failure rate for common field devices shown in Figure 8 was taken from the “Plant Hazard Analysis and Safety Instrumentation Systems” book [8]. These will help with operator awareness of the possibility of control instruments failing.

Figure 7: Faults of Instruments per Year

Instrumentation (Anyakora et al., 1971)	
Item	**Faults/Year**
Controller	0.29
Control valve	0.60
Solenoid valve	0.42
I/P converter	0.49
Valve positioner	0.44
Magnetic flowmeter	2.18
Differential pressure flowmeter	1.73
Load cell solid flowmeter	3.75
Belt speed measurement	15.3
Liquid level measurement	1.70
Solid level measurement	6.86
Temperature measurement (excluding pyrometers)	0.35
Thermocouple	0.52
Resistance thermometer	0.41
Pressure switch	0.34
Flow switch	1.12
Optical pyrometer	9.70
PH-meter	5.88
Oxygen analyser	5.65
Carbon dioxide analyser	10.5
Hydrogen analyser	0.99
Controller settings	0.14

Figure 8: Failure Rates

Item Details	Failure Rate (Hr)	Safe Failure (%)	MTTF (Year)	Remarks
Pressure switch	4.0E-06	40	20–30	
Temperature switch	5.0E-06	40	20–30	
Flow switch	7.0E-06	60	20–30	
Level switch	5.0E-06	55	20–30	Float/displacer
Pressure transmitter	1.5E-06	>50	55	
DP transmitter	1.5E-06	50	40–60	
Flow meter	3.5E-06	25	50	Coriolis meter
Level instrument	7.0E-06	40–50	40–60	Displacer
RTD	5.0E-08	81.6	60–80	
Thermocouple	1.5E-06	95	60–80	
Temperature transmitter	5.0E-06	30	70	
I/P converter	4.0E-06	40		
Solenoid valve	2.0E-06	60	25–30	
Ball/Butterfly valve	3.0E-06	55	30	
Gate valve	2.0E-06	45	50	
Globe valve	2.5E-06	55	50	

Personnel Exposure Risk

Tables 32, 33, and 34 go into further detail on the chemicals present in our process and the exposure limits associated with each chemical. Table 33 also goes into detail on the flammability and reactivity of the included chemicals.

Table 32: OSHA Chemical Exposure Limits

Chemical	OSHA PEL: TWA		Chemical	OSHA PEL: TWA	
Nitrogen	-	-	Pentane	1000 ppm	2950 mg/m^3
Hydrogen	-	-	Hexane	500 ppm	1800 mg/m^3
Carbon Monoxide	35 ppm	40 mg/m^3	Styrene	100 ppm	-
Carbon Dioxide	5000 ppm	9000 mg/m^3	1,3-Butadiene	1 ppm	-
Methane	1000 ppm	0.1 mg/m^3	Clarified Oils (Petroleum), Catalytic Cracked	-	5 mg/m^3
Ethane	1000 ppm	-	Fuel Oil No. 6		5 mg/m^3
Ethylene	200 ppm	-	Naphthalene	10 ppm	5 mg/m^3
Propane	1000 ppm	1800 mg/m^3	Hydrogen Sulfide	5 ppm	-
Propylene	500 ppm	-	Sulfur	-	-
n-Butane	800 ppm	1900 mg/m^3	Fuel Oil No. 2	10 ppm	100 mg/m^3
n-Hexane	500 ppm	1800 mg/m^3	Kerosene Petroleum	-	200 mg/m^3
HCL	5 ppm	7 mg/m^3	Cumene	50 ppm	245 mg/m^3
Octane	500 ppm	2350 mg/m^3	Amberlite XAD-7	-	15 mg/m^3
n-Heptane	500 ppm	2000 mg/m^3	Aluminum Oxide	-	15 mg/m^3
m-Xylene	100 ppm	435 mg/m^3	Activated Carbon	-	15 mg/m^3
EthylBenzene	100 ppm	435 mg/m^3	Water	-	-
Benzene	1 ppm	-			
Toluene	200 ppm	-			

Table 33: NFPA Dimond Classification

Chemical	Health (Blue)	Flammability (Red)	Instability (Yellow)	Chemical	Health (Blue)	Flammability (Red)	Instability (Yellow)
Nitrogen	0	0	0	Pentane	0	4	0
Hydrogen	0	4	0	Hexane	2	3	0
Carbon Monoxide	3	4	0	Styrene	2	3	2
Carbon Dioxide	2	0	0	1,3-Butadiene	2	4	2
Methane	1	4	0	Clarified Oils (Petroleum), Catalytic Cracked	2	2	0
Ethane	1	4	0	Fuel Oil No. 6	2	2	0
Ethylene	1	4	0	Naphthalene	2	2	0
Propane	2	4	0	Hydrogen Sulfide	4	4	0
Propylene	1	4	3	Sulfur	2	2	2
n-Butane	1	4	3	Fuel Oil No. 2	0	2	0
n-Hexane	2	4	0	Kerosene Petroleum	1	2	0
HCL	3	0	1	Cumene	2	3	1
Octane	3	3	0	Amberlite XAD-7	0	4	0
n-Heptane	3	3	0	Aluminum Oxide	0	0	1
m-Xylene	3	3	0	Activated Carbon	1	2	0
EthylBenzene	3	3	0	Water	0	0	0
Benzene	2	3	0				
Toluene	2	3	0				

Table 34: Lethal Dose Limits (LD50)

Chemical	Inhalation	Oral	Chemical	Inhalation	Oral
Nitrogen	-	-	Pentane	364 g/m^3 (4 hours)	5000 g/kg
Hydrogen	-	-	Hexane	48000 ppm	25 g/kg
Carbon Monoxide	-	3760 ppm (1 hours)	Styrene	11.7 mg/L (4 hours)	1000 mg/kg
Carbon Dioxide	-	-	1,3-Butadiene	285 g/m^3 (4 hours)	5480 mg/kg

Methane	-	-	Clarified Oils (Petroleum), Catalytic Cracked	-	-
Ethane	-	-	Fuel Oil No. 6	-	-
Ethylene	-	-	Naphthalene	-	490 mg/kg
Propane	-	-	Hydrogen Sulfide	0.38 mg/L (16 hours)	-
Propylene	-	-	Sulfur	9.23 mg/L (4 hours)	2000 mg/kg
n-Butane	658000 mg/m^3 (4 hours)	-	Fuel Oil No. 2	4.6 mg/L (4 hours)	12 g/kg
n-Hexane	48000 ppm (4 hours)	15840 mg/kg	Kerosene Petroleum	-	15 g/kg
HCL	2810 ppm/hr	238 mg/kg	Cumene	3577 ppm (6 hours)	1400 mg/kg
Octane	24.88 mg/L (4 hours)	5 g/kg	Amberlite XAD-7	11 mg/L (4 hours)	2000 mg/kg
n-Heptane	73.5 mg/L (4 hours)	2000 mg/kg	Aluminum Oxide	2.3 mg/L (4 hours)	10000 mg/kg
m-Xylene	5267 ppm (6 hours)	4988 mg/kg	Activated Carbon	4.6 mg/L (4 hours)	8000 mg/kg
EthylBenzene	35500 mg/m^3 (2 hours)	3500 mg/kg	Water	-	90000 mg/kg
Benzene	10000 ppm (7 hours)	50 mg/kg			
Toluene	12500 mg/m^3 (4 hours)	5000 mg/kg			

Atmospheric Detonation of Distillation Inventory

Table 35 displays a TNT equivalency calculation for the atmospheric detonation of all chemicals from T-105 [12]. These values were found using Equation 8.

[8] $$m_{TNT} = \frac{nm\Delta H_C}{E_{TNT}}$$

Table 35: TNT Equivalency of Explosion

TNT Equivalency	
n	0.05
Mass (lbs)	48444
Energy of Explosion (BTU/lb)	13788
Energy of TNT (BTU/lb)	2016
Total (lbs)	16567
Tons	8

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

To do a thorough risk analysis, a Hazard and Operability Study was done on our two distillation columns. This is shown in Table 36.

Table 36: HAZOP (1 of 2)

What if:	Likelihood	Consequence	Ways to Prevent
Oxygen/air gets into pipe/distillation tower	High	Fire due to oxygen, heat, and combustible materials	No exposure to oxygen, no entry points for oxygen
Pressure introduced is too high (into distillation column)	Medium	Overpressure problems, increase in temperature and potential of fire	Pressure indicator, rupture disk, and pressure relief valve
Pressure introduced is too low (into distillation column)	Low	Proper separation will not occur (Product specifications will be incorrect)	Pump before entry into the distillation column
Flow into distillation tower is too high	Medium	Flooding will occur in tower	Flow indicator on the column & control valve
Flow into distillation tower is too low	Medium	Weeping will occur in tower	Flow indicator on the column & control valve
Loss of heating in an upstream column	Medium	Will send more liquid to distillation column and increase feed components and light components to the column	Temperature indicator on the reboiler and preheater for the column
Valve opening to an external pressure source	Medium	Overpressure problems or loss of pressure and potential of fire	Pressure indicator, rupture disk, and pressure relief valve no entry points for ambient air
Closed column outlets	Low	Material is trapped in the column which could cause overpressure problems	Pressure indicator and level indicator on the column and control valves
Condenser is not cool enough	Medium	Overhead product will be too hot (product specifications will be incorrect) and pressure will build in the column	Alarmed indicator and potentially backup cooling water storage
Condenser overfills	Medium	Spillage	Level indicator & control valve
Loss of cooling water	Medium	Overpressure problems, increase in temperature and potential of fire	Pressure indicator, rupture disk, and pressure relief valve for the column
Separation vessels have too much liquid	Medium	Proper separation will not occur (product specifications will be incorrect)	Level indicator & control valve and possible backup vessel
Separation vessels have too little liquid	Low	Proper separation will not occur (product specifications will be incorrect)	Level indicator & control valve and possible backup vessel

Reboiler temperature is beyond set point	Low	Too much bottoms product will re-enter the column	Temperature indicator on the reboiler, level indicator on the column, and bypass route
Reboiler does not have enough heat	Medium	Bottoms product will not re-enter the column	Alarmed indicator and potentially backup hot component
Reboiler overfills	Medium	Spillage	Level indicator & control valve
Loss of steam (reboilers, preheater)	Medium	Bottoms product will not re-enter the column and feed will not be correct temperature entering the column	Temperature indicator on the reboiler and flow indicator the column
Reflux is too high	High	Not enough product will form & too much liquid will reenter the column	Level indicator, control valve and backup vessel
Reflux is too low	Low	Not enough liquid reentering the column & product will have incorrect specifications	Level indicator, control valve and backup vessel
Fire occurs below column	Medium	Column is destroyed due to structural damage and flammable materials in the column	Fire alarm and fire retardant around column
Fire occurs inside column	Low	Column is destroyed due to flammable materials in the column	Fire alarm & fire retardant around column
Power is lost to column	Low	Column will be shut down (not operate) and tanks will continue to fill	Power alarm & possible backup generator
Process control is lost	Low	Indicators/alarms will not operate properly	Regularly testing of alarms and operator awareness
Failure of pressure controller	Medium	Overpressure problems, increase in temperature and potential of fire	Regularly testing of alarms and operator awareness
Failure of feed controller	Medium	Feed entering the column will not be the correct temperature	Regularly testing of control valves and operator awareness
Accumulation of non-condensables	Medium	Gunk up rupture disk, column gets dirty, tanks are dirty	Regular maintenance

Table 36: HAZOP (2 of 2)

HAZOP (continued)						
What if:	Equipment Damage	Environmental Compliance	Loss of Life	Disruption of other business units	Legal/PR	Community Impact
Oxygen/air gets into pipe/distillation tower	High	High	High	High	High	High

Pressure introduced is too high (into distillation column)	High	High	High	High	High	Medium
Pressure introduced is too low (into distillation column)	Medium	Medium	Medium	Medium	Medium	Medium
Flow into distillation tower is too high	Medium	High	Medium	High	Medium	Medium
Flow into distillation tower is too low	Medium	High	Medium	Medium	Medium	Medium
Loss of heating in an upstream column	Medium	Medium	Medium	High	Medium	Medium
Valve opening to an external pressure source	High	High	Medium	High	Medium	Medium
Closed column outlets	High	High	Medium	High	High	Medium
Condenser is not cool enough	High	High	Medium	High	Medium	High
Condenser overfills	Medium	High	Medium	High	Medium	Medium
Loss of cooling water	High	High	Medium	High	Medium	Medium
Separation vessels have too much liquid	Medium	Medium	Medium	Medium	Medium	Medium
Separation vessels have too little liquid	High	Medium	Medium	High	Medium	Medium
Reboiler temperature is beyond set point	Medium	Medium	Medium	Medium	Medium	Medium
Reboiler does not have enough heat	Medium	Medium	Medium	High	Medium	Medium
Reboiler overfills	Medium	High	Medium	High	Medium	Medium
Loss of steam (reboilers, preheater)	Medium	Medium	Medium	High	Medium	Medium
Reflux is too high	Medium	High	Medium	High	Medium	Medium
Reflux is too low	High	High	Medium	High	Medium	Medium
Fire occurs below column	High	High	High	High	High	High
Fire occurs inside column	High	High	High	High	High	High
Power is lost to column	Medium	High	Medium	High	Medium	Medium
Process control is lost	High	High	High	High	High	High
Failure of pressure controller	High	High	High	High	High	High
Failure of feed controller	Medium	Medium	Medium	High	Medium	Medium
Accumulation of non-condensables	Medium	High	Medium	Medium	Medium	Medium

<u>Recommendations for Improvement of the Bali Sorting Facility</u>:

The Quantity Gap

A large percentage of the waste that is mismanaged in Jembrana is eventually leaked into the ocean or dumped into the areas where the waste finds its way into local waterways. One thing that can be done about this is introducing trash traps in the rivers or local waterways to both collect and reduce the amount of waste that ends up in the ocean. Similarly how the “Adopt-A-Highway" system works, where a business or community group will adopt a section of a highway, local businesses in Bali could sponsor the trash traps. Additionally, signage would be placed near their sponsored/adopted trash trap to engage both the community and the tourists that visit. The economy of Bali is mainly supported by tourism which would increase their awareness by seeing the signage. This would help introduce a more circular business model where businesses are responsible and aware of the waste they contribute. Additionally, a waste vending machine can be implemented both for tourists and the local community. This system would accept specific kinds of plastic, (PET, HDPE, LDPE) and in return would supply either currency or coupons for local businesses. There would be a crusher in the machine as well so the machines would be able to separate the plastic types and crush them to conserve space. This would create a system of incentives as well as educate the community and tourists about their ability to recycle. Additionally, this would aid in the waste collection itself as the machine would serve a couple of communities and allow the household collection to potentially make fewer stops by visiting the vending machines instead of each household.

The Quality Gap

To help households segregate their waste more comprehensively, educating the households is necessary. One practical implementation would be handouts mailed to households explaining the separation of recyclables, which could decrease the contamination of organic/inorganic waste. Bins with signage posted stating what should and shouldn’t be placed inside of them could be delivered to each home. Random spot checks for participating homes would also help to decrease downstream contamination. Another helpful implementation would be educating the tourists as well by having posted signs near public trash cans. By including pictures of accepted items on the recycling containers, language barriers between tourists and community members will be minimized, leading to higher quality products. Getting hotels and resorts on board to educate tourists and implement the two-colored bag (organic/inorganic) system would be beneficial as well. Tracking the waste collection capacity as well as how effectively the waste was separated, and making this data public knowledge would educate the general public on the progress they are making towards the environment. This data could be brought up at community meetings, in a media campaign, or in building awareness in schools. Due to the limited landfill space in Bali and non-recyclable waste, implementation of tight regulation of landfill space is necessary to help with the overall amount of waste put into landfill. For the sorting facility, implementing a shredder, a magnet, and a screen to separate the incoming waste streams can help to separate types of recyclables. The shredder would be able to break down the large bags of waste that are incoming from communities. Since pickers are primarily looking for the most valuable items, a magnet would sort out most of the metals from the stream allowing the workers to sort for plastics primarily. Running the waste over a screen would allow the small particles like sand, bio waste, or ashes to be sorted out. These recovered particles could further be used in the composting process.

The Affordability Gap

There will always be some fractions of waste that cannot be recycled such as rubber or batteries. Introducing a "bulk-pickup" day to help remove this type of material would ease the separation from the recyclables. This could possibly generate a new source of revenue if the non-recyclable items are sold to companies that can recycle or repurpose those items. A designated "bulk-pickup" day would be set 1-2 times every month allowing for minimal storage of the bulk inside of homes. Since Bali is centered on an island in Indonesia, there is a limited amount of space. This would allow for more of the non-typical waste, such as electronics and furniture, to be potentially recycled, therefore creating more landfill space. To reduce the cost of household collection, installing a vending machine will somewhat centralize the location of the pickup which will help offset the traveling costs. To operate efficiently, changing the pickup system to reflect specified days of organic pickup and days of inorganic pickup would make the logistics of route traveling slightly faster. Currently, the routes are structured so that organic and inorganic waste is collected twice per week; however, the possibility of separate routes for organic waste and inorganic waste would maximize the input into each separate recycling facility. The ability to sell the compost made is very important. The possibility of selling it overseas could be another stream of revenue created.

Conclusions

To implement a circular plastic economy, some aspects of recycling need to be improved. Firstly, the process of collecting plastic recycling needs to be more efficient. To do that, instructing citizens on what can be recycled can improve the plastic quality entering the Bali Recycling Plant. This instruction can be through flyers, spot checks, or by including accepted items on labels for recycling containers. To improve the quantity of plastic waste, local businesses will be encouraged to participate in community cleanup efforts with a focus on separating typical waste from plastic waste. Recycling vending machines can also be installed in high traffic areas, giving an incentive for tourists to participate in the recycling efforts. Lastly, the costs associated with the recycling program could be lowered to make the project more feasible. To do this, having the pickup routes run less often and having specified days for recycling and compost waste can help to get more concentrated waste, with a lower operating cost.

Another gap that needs to be closed to create a more circular economy is the purification process for the oil coming out of the pyrolysis section. Our design separates the feed into four streams, as requested by Global Petrochemicals. The Py-Gas stream comes off the first distillation tower and is sent directly to the ethylene plant. The Naphtha stream also comes off the first distillation tower and is fed to a steam cracker in the ethylene plant for further processing. The Gas Oil stream comes off the overhead of the second distillation tower and is also sent to a steam cracker in the ethylene plant. Since the Naphtha and Gas Oil streams are fed to steam crackers, they cannot contain any water in the stream. To ensure that there is no water present, our design includes adsorption columns before the feed is separated. This allows for a minimal number of columns required for the design, while also removing the contaminants from the product streams. There are three adsorbers at the beginning of the process; the first is designed to remove the chlorides present in the feed stream. The second adsorber is designed to remove both Calcium and Silica. The final adsorber removes any free water in the pyoil stream.

The fourth stream produced by the distillation columns is the Pyoil Heavy Cut that cannot be sent to a steam cracking furnace or be sent to the ethylene plant for other purification. To further our commitment to creating a circular economy, the uses for the Heavy Cut must be explored. Some possible dispositions our group found include microbial degradation of the organic compounds present, possible use as a diesel fuel or other transportation fuels, and road construction. Certain microbes can break down hydrocarbons and hydrocarbon pollutants. These microorganisms break down hydrocarbons as sources of energy. The result of microbial breakdown is compost that can potentially be used in soil. Some factors influencing the ability of microbes to breakdown hydrocarbons are the temperature and pH of the substrate being degraded. The heavy cut can also be refined and used as transportation fuel, such as diesel fuel or as an additive for motor gasoline, as these fuels have a heavier grade. Depending on the amount of bitumen present in the heavy cut, bituminous sand or tar sand can be produced which is used in road construction, roofing, and waterproofing. Rubber can be mixed with bitumen which can form latex, sheet rubber or rubber powder and can be used for tire making or road repairs based on the durability of the material.

Using cost approximations from Turton, we found that the capital cost is $13,624,900. The variable operating cost from the utility streams is $2,252,100. The fixed operating costs for labor, maintenance, and adsorbents were found to be $2,248,300.

Appendices

Adsorption Section Detail

In order to properly design our adsorbers, several assumptions about both their specifications and the specifications of the adsorbents themselves had to be made. For example, to determine the volume of each adsorber, the physical specifications of each adsorbate had to be found. This includes the adsorption bulk densities and their respective adsorption capacities. However, since commercially available adsorbents vary slightly in their specific physical qualities depending on the manufacturer, reputable sellers of the materials were referenced using their specifications along with their costs. These includes the Dow Chemical Company, Millipore, and Hapman Advantages. However, many of the values used for the particle sizes were as a "mesh." Thus, a mesh conversion chart had to be utilized for exact sizes as well [16, 4, 25, 23]. As for selecting which adsorbents we would use, Amberlite was chosen for the first adsorber due to its reported ability to remove chlorides from aqueous solutions, assuming that this would be appropriate for the raw pyoil feed [24, 26]. Likewise, activated aluminum oxide was chosen for the second adsorber due to its reported ability to remove silica and calcium contaminants [25]. Finally, activated carbon was chosen as the final adsorbent since it can purportedly remove water from a stream without affecting any hydrocarbons within it. Furthermore, it was assumed that activated charcoal, the material that were costed, was comparable enough in makeup to the activated carbon to be specified as identical [30]. The presence of chlorides in the stream was found to be 0.005% of the total as given by the wppm value in the project statement. Silica, calcium, and water however were assumed to be comparable in percentage to this value. Thus, the amount of adsorbent required for each was dependent on these assumptions as well as the values taken from the aforementioned sources. As for the adsorbers themselves, it was assumed when sizing them that the optimal height to diameter ratio was between 2.5 and 4 to 1, using 3.25 to 1 as an average spec [48].

During the process, for the adsorbent to be regenerated, we elected to employ three parallel adsorber towers for each of the three adsorbents in our design, meaning nine towers in total. The idea behind this is it allows each tower to serve a specific duty which alternates during breakthrough. The first tower serves as the primary adsorber, purifying the stream until it is fed into the second tower which theoretically shouldn't have to adsorb anything until breakthrough occurs, but it still functions as a safety should any contaminants slip past. When breakthrough does occur however, this second tower will allow the excess contaminants to be adsorbed before the valve to the first tower is shut and regeneration begins. In other words, this second tower allows us to operate the first tower all the way until breakthrough occurs rather than having to shut it off beforehand. As for the third tower, this one serves as a standby. Since we wish to operate without interruption, having this tower allows us to immediately begin operating with another pair of towers while the first is being regenerated with nitrogen gas. To summarize this process, the inlet stream is connected to three towers in parallel and each of the tower outlets are connected to the next tower in series. By opening and closing their respective valves, this allows us to control which tower serves as the primary before being fed into the secondary. This way, regeneration can occur without interruption to our continuous process.

This method also has the added benefit of allowing us to handle unusually high levels of contaminants. Since our process always operates with a secondary adsorber tower, any premature breakthrough that occurs will be handled by this secondary tower. Of course, depending on the level of excess, it would theoretically be possible to overwhelm both towers and cause a second breakthrough before initial regeneration had finished. This scenario would require a very high level of contaminants over the course of a long period, and is relatively unlikely. If such an event does occur, however, regeneration can be accelerated by feeding a higher volume of nitrogen gas into the necessary tower.

Distillation Section Detail

To get the most efficient separation, our design utilized two distillation columns. This allowed us to preheat the feed stream entering the second distillation column without having an excess of wasted heat. This also allowed us to design integrated heat exchangers to save on operating costs. The first distillation column was designed to separate the Py-Gas and the Naphtha from the heavier components. To do this, we assumed that the Py-Gas needed to be a gas at 107 °F with an ethylene recovery of 100%. We also used the constraints given by the problems statement for the Naphtha stream to have an End Boiling Point of 392 °F at atmospheric pressure. These constraints can be seen in Table 37.

Table 37: T-104 Constraints and Sizing

T-104	
Constraints	
Maximum Volume (m^3)	520
PyGas Component Recovery (Ethylene)	100%
PyGas Stream Temperature (°F)	107
Naphtha Stream Temperature (°F)	392
Column Sizing	
Theoretical Stages	10
Stage Efficiency	0.5
Real Stages	20
Feed Stage	8
Column Top Pressure (psia)	20
PyGas Flow Rate ($\frac{lb}{hr}$)	1331
Naphtha Draw Stage	18
Naphtha Draw Phase	Vapor
Naphtha Draw Pressure (psia)	22.89
Naphtha Flow Rate (lb/hr)	4285
Bottoms Temperature	470.6
Column Bottoms Pressure (psia)	24
Column Bottoms Flow Rate ($\frac{lb}{hr}$)	46,816
Rectifying Section Diameter (ft)	3.6
Rectifying Tray Type	Sieve
Rectifying Tray Spacing (ft)	2
Rectifying Section Volume (ft^3)	213
Stripping Section Diameter (ft)	5.616
Stripping Tray Type	Sieve
Stripping Tray Spacing (ft)	2
Stripping Section Volume (ft^3)	669

Total Volume (ft^3)	882
Material of Construction	Carbon Steel

With these constraints the column was optimized by changing the number of stages, the feed stage location, and the pressures at the top and bottom of the column. After the optimal configuration was found, different types of trays were used to find the most efficient separation. The sieve trays had the smallest required diameter and are also the cheapest option. The first distillation tower sizing results can also be found in Table 37. The column profile for T-104 can be found in Table 38.

Table 38: T-104 Column Profile

Theoretical Stages	Temp	Pres	Net Liq	Net Vap
	(°F)	(psia)	$(\frac{lbmol}{hr})$	$(\frac{lbmol}{hr})$
Condenser	107	20	60.59	---
1__Main	181.6	22	68.29	84.83
2__Main	197.7	22.11	68.96	92.53
3__Main	206.2	22.22	59.67	93.2
4__Main	245.5	22.33	618.2	83.91
5__Main	274.8	22.44	664.2	354.3
6__Main	302	22.56	676.7	400.3
7__Main	336.8	22.67	715.7	412.8
8__Main	367.7	22.78	761.3	451.8
9__Main	392	22.89	775.8	497.5
10__Main	418	23	721.2	543.7
Reboiler	470.6	24	---	489.1

This process was similarly repeated for the second distillation column. We found this second column to be necessary to not include the heaviest components in the gas oil stream. The second column allowed us to preheat the feed into this column, lowering the amount of high-pressure steam required for the reboiler. To separate the products into Gas Oil and the Pyoil Heavy Cut, we assumed that the Heavy Cut had to be able to vaporize at 730 °F. We also used the constraint given in the problem statement for the Gas Oil stream to have an End Boiling Point at 620 °F at atmospheric pressure. These constraints are shown in Table 39. We optimized the second distillation columns size using the same variables as before, giving the sizing values in Table 39. Table 40 shows the column profiles for T-105.

Table 39: T-105 Constraints and Sizing

T-105	
Constraints	
Maximum Volume (m^3)	520
Gas Oil D86 BP (°F)	620
Heavy Stream Temperature (°F)	730
Column Sizing	

Theoretical Stages	15
Stage Efficiency	0.5
Real Stages	30
Feed Stage	20
Column Top Pressure (psia)	20
Gas Oil Flow Rate ($\frac{lb}{hr}$)	39,998
Heavy Stream Temperature (°F)	730
Heavy Stream Pressure (psia)	24
Heavy Stream Flow Rate ($\frac{lb}{hr}$)	6,818
Column Diameter (ft)	7
Tray Type	Sieve
Tray Spacing (ft)	2
Total Volume (ft^3)	77
Material of Construction	Carbon Steel

Table 40: Column Profiles for T-105

Theoretical Stages	Temp	Pres	Net Liq	Net Vap
	(F°)	(psia)	($\frac{lbmol}{hr}$)	($\frac{lbmol}{hr}$)
Condenser	444.8	20	278.5	---
1__Main	535	22	273.8	489.3
2__Main	573.7	22.11	283.3	484.7
3__Main	590.4	22.21	286.6	494.2
4__Main	598.6	22.32	287.3	497.5
5__Main	603.1	22.43	287.1	498.2
6__Main	605.7	22.54	286.2	497.9
7__Main	607.6	22.64	284.2	497
8__Main	609.2	22.75	280.2	495.1
9__Main	611.4	22.86	270.3	491.1
10__Main	615.5	22.96	259.2	481.2
11__Main	649.5	23.07	296.2	237.9
12__Main	663.5	23.18	304.8	274.9
13__Main	673.6	23.29	300.2	283.6
14__Main	685.1	23.39	287.4	279
15__Main	701.5	23.5	260.8	266.2
Reboiler	730	24.5	---	239.6

With the configuration of the two distillation columns in series, we can save on energy consumption. This is because we are able to utilize heat integration between the columns. The first columns reboiler uses the excess heat from the second column’s condenser. Due to the temperature difference between the two, the first column gets all its necessary heat input from the Gas Oil reflux stream. The second column's condenser needs more heat removed before being fed back into the column. To do this, the condenser is combined with a preheat stream to the second distillation column. This allows the condenser to cool down to the desired temperature while also heating the secondary feed. Finally, the secondary feed is put through another heat exchanger, where the heavy product stream is used as a heating fluid. The design for these heat exchangers is displayed in Table 41 with detailed inlet and outlet stream specifications.

Due to the preheat for T-105 the reboiler only needs $7.5 * 10^6$ BTU/hr, which is the only use of high-pressure steam in the design.

Table 41: Heat Integration Calculations for E-103, E-104, and E-105

	E-103	E-104	E-105
Q ($\frac{BTU}{hr}$)	8,456,000	8,802,440	930,400
U ($\frac{BTU}{(hr)(ft^2)(F°)}$)	90	86	71
Correction Factor, (F°)	0.75	0.75	1
LMTD	34	45	61
A (ft^2)	3,751	3,067	213
Cooling Fluid			
	Boil-up - First Column	Preheat Stream	Preheat Stream
Temperature in (°F)	415.1	470.6	470.6
Temperature out (°F)	462.3	470.6	505.8
Pressure in (psia)	24	27	25
Pressure out (psia)	23	25	23
Tube or Shell?	Shell	Tube	Shell
Heating Fluid			
	Reflux - Second Column	Reflux - Second Column	Heavy Product
Temperature in (°F)	500	534.9	701.4
Temperature out (°F)	444.8	500	480
Pressure in (psia)	22	21	24.5
Pressure out (psia)	21	20	23.5
Tube or Shell?	Tube	Shell	Tube

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