Date: March 06, 2015
To: Philips 66 Project Management Team
Project: Butamer Unit

Enclosed is the completed preliminary design and economic analysis for the installation of a Butamer Unit. The purpose of the project is to investigate whether or not it is economically feasible to install and run an isobutane conversion system at the Ponca City refinery. At present, the isobutane is purchased from a third party production plant in Kansas. The unit has two beneficial products: the isobutane that is split from the butane mix, the isobutane that is converted from n-butane, and the natural gas that is a by-product of the process.

The proposed Butamer Unit has a capacity of 7,800 BPD. The economic analysis is performed based on construction in 2016 and startup occurring mid-year 2017. The analysis is successfully completed for the process design, economics, safety, and environmental impacts of the endeavor. All the components are bound to the report and calculations are included in the Appendix.

Philips 66 Butamer Unit
CHE 4224
Senior Design Project
March 6, 2015

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Executive Summary

The purpose of this design is to establish a functioning Butamer Unit for the Ponca City refinery. The unit uses a mixed hydrocarbon stream and produces a 92.0 vol.% isobutane rich stream. A by-product of the process is a natural gas stream with pentane and other heavier carbon components and 1.0 vol.% n-butane.

The Butamer Unit design will cost $23MM. It will have estimated annual operating expenses of $5.5MM. With the estimated potential sales taken into account, this design is projected to have an NPV of $19MM, DCFROR of 42%, and a discounted payback period of 2.69 years. A total of 14 operators will be needed to operate this entire unit around the clock. The two greatest factors that affect these economics are the sales of our products and the raw material costs of the feed. The best case scenario for DCFROR is 125% and a worst case scenario is -35%. The best case scenario for NPV is $97MM and a worst case scenario is -$59MM. Therefore, it is recommended that care be taken to establish as many clients as possible to maximize demand and sales opportunities. In addition, a feed storage system should be developed so that large quantities of feed can be purchased at lower costs.

The overall process consists of two distillation columns and a reactor with necessary supporting pumps and heat exchangers between the main units. The raw material feed is pumped into the first column, de-isobutanizer tower, which produces the isobutane product stream in the tops, and an intermediate stream in the bottom. The intermediate stream is pumped to the second column, the de-butanizer tower. This column produces the natural gas by-product in the bottoms while the tops product enters the reactor. The reactor uses a platinum catalyst to convert the n-butane to isobutane and then feeds the product into the first column. The de-isobutanizer tower is optimized to have 24 trays inside the column and the de-butanizer tower is optimized to have 29 trays inside the column.

This design operates within the range of normal temperatures and pressures. There are no abnormal safety risks being added by the Butamer Unit. All fluids being handled within the unit are extremely flammable; therefore, the same measures should be taken to prevent potential fires or explosions that are taken in all parts of the refinery. There are no anticipated environmental hazards other than ensuring that the cooling water is returned to the environment at safe temperatures.

It is highly recommended that this design proceed onward to the detailed design stage. The initial economics show a potential for profit of $19MM within 5 years of startup. Also, there are no abnormal risks that make the process unsafe.

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1. Introduction

Phillips 66 refineries have a high demand of isobutane to increase the octane rating of the diesel fuels in the winter. Therefore, Phillips 66 wants to install a new Butamer Unit at the Ponca City refinery to use and sell isobutane. There is currently no isobutane being produced at the Ponca City refinery. Therefore, this Butamer Unit will be able to decrease the purchase and transportation costs of the isobutane. The refinery currently purchases the butane mixture at a reduced cost during the summer and stores it in salt caverns for later processing. The profit from the sale of the isobutane in Mt. Belleview, TX will outweigh the butane mix feed cost purchased from Conway, KS. The natural gasoline by-product of the process can also be sold to improve the economics of the project. Our group is responsible for performing the preliminary design, providing a basic control scheme, and the economic analysis to determine the feasibility of this project.

The main objective of the project is to meet the specification of 92.0 vol.% purity of isobutane in the top of the de-isobutanizer tower and 1 vol.% n-butane in the natural gasoline product at the bottom of the de-butanizer tower. The isobutane must be delivered at 20 psig above the bubble point temperature of 128 °F. The butane feed mix arrives at 95 °F and 100 psig with a volumetric flowrate of 7,800 BPD.

The block flow diagram of the isobutane production system (**Figure 1**), composition of the mixed butane feed (**Table 1**), component and transportation prices (**Table 2**), utility costs and specifications (**Tables 3, 4**), and the steam specifications (**Table 5**) were provided by Dr. Ramsey and Dr. Aichele and are presented in the Design Basis. The physical properties experts recommended using the Peng-Robinson equation of state in the Aspen simulation. The economic analysis is performed at a hurdle rate of 15% using a 5 year project evaluation life with 10-year MACRS depreciation and an effective tax rate of 40%.

The project will advance the goals of the organization by generating profit from isobutane production and eliminating payment to the third party. The project will also produce natural gas that can provide additional revenue in the future. This report includes the economic analysis, optimization and sensitivity analysis of the columns, safety and hazard analysis, and PFDs including one with an effective control scheme. This report includes all the sizing calculations performed in Microsoft Excel and also the Aspen simulation output in the Appendix. Isobutane is being delivered at 92.0 vol.% purity and 1.0 vol.% n-butane in the natural gasoline stream with a net present value of $19MM and annual revenue of $27MM. These specifications were achieved by sizing the de-isobutanizer tower, de-butanizer tower, and other additional equipment.

2. Design Basis

This project involves a preliminary design of the Phillips 66 Butamer Unit. The specifications are provided for temperature, pressure, and composition of the mixed butane feed stream. The feed is available at 95 °F, 100 psig, and a total volumetric flowrate of 7,800 BPD. The composition of the feed is shown below in **Table 1**.

Feed Stream Composition	
Component	**Feed (BPD)**
Water (H_2O)	1.1
Propane (C_3)	159.9
1-Butene	7.8
Iso-butene	15.6
Trans-2-butene	7.8
Cis-2-butene	7.8
Isobutane (i-C_4)	2496
n-Butane (n-C_4)	4784.3
2,2-dimethyl propane	7.8
Iso-pentane (i-C_5)	300.3
n-Pentane (n-C_5)	7.8
n-Hexane (n-C_6)	3.9

Table 1: *Mixed Butane Feed Composition from Conway, KS*

The butane mix feed will first enter the de-isobutanizer tower which separates the isobutane from n-butane. The isobutane product in the tops must meet the specification of 92.0 vol.% purity and must be delivered at 20 psig above the bubble point at 128 °F. The bottoms product of the de-isobutanizer tower enters a second fractionation column called the de-butanizer tower which separates n-butane from natural gasoline (C5+). The distillate of this tower enters a horizontal packed-bed Butamer reactor that partially converts the n-butane to isobutane. UOP has already designed the Butamer reactor to be 900 ft^3 with a pressure drop of 20 psi. The reaction stoichiometry is 1 mole of n-butane to 1 mole of isobutane. The reactor will be using a platinum catalyst of 44,200 lb priced at $13.25/lb and will be replaced every five years. The reactor feed temperature must enter in the range of 300-400 °F and 450 psig. The following Equation 1 is provided by UOP to determine the single pass conversion as a function of reactor feed temperature and composition:

$$\%Conversion = 100[F(T) - \left(\frac{i-C_4}{n-C_4}\right) * (1 - F(T))] \tag{1}$$

Where:

$F(T) = 0.867 - 0.00067T$

T = F at the reactor inlet

$\left(\frac{i-C_4}{n-C_4}\right)$ = molar ratio at the reactor inlet

The effluent of the reactor is recycled to the de-isobutanizer tower to further separate the unreacted n-butane from the isobutane and to also increase the purity of isobutane. The natural gasoline by-product in the bottoms of the de-butanizer tower must meet the specification of 1 vol.% n-butane. The process is summarized in the block flow diagram in **Figure 1**.

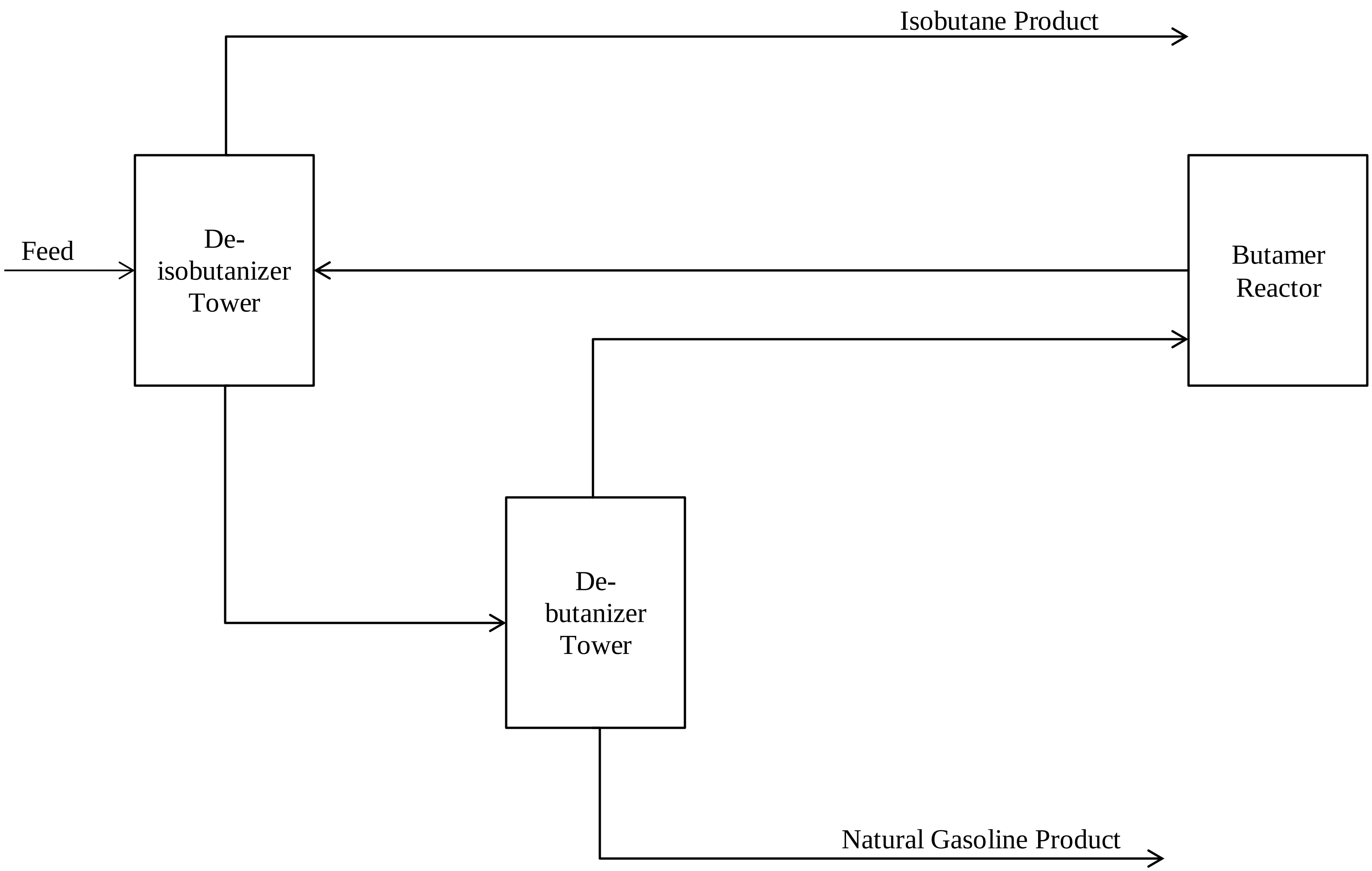

Figure 1: *Block Flow Diagram of the Process*

The transportation and component prices are specified below in **Table 2**. The mixed butane feed is purchased from Conway, KS, and is supplied to Ponca City, OK. The isobutane product will be sold in Mt. Belleview, TX. Natural Gasoline by-product will be sold in Oklahoma.

Component	**Purchase/Sale Location**	**Price**	**Unit**
Mixed Butanes	Conway	\$27.50	\$/BBL
Normal Butane	Mt. Belleview	\$31.78	\$/BBL
Isobutane (All C4's ex N-C4) (Sales Spec = 92 vol.% IC4s)	Mt. Belleview	\$37.67	\$/BBL
Natural Gasoline (C5+ Material and some N-C4)	Oklahoma	\$55.85	\$/BBL
Transportation (Conway-Oklahoma)		\$1.10	\$/BBL
Transportation (Oklahoma – Mt. Belleview)		\$1.35	\$/BBL

Table 2: *Component and Transportation Costs*

Utility costs are provided in **Table 3**. Utility and Steam specifications are provided in **Table 4** and **Table 5**, respectively. The utilities are readily available on site.

Utility Costs	
Electricity	\$0.07/kilowatt hours
160 psig steam (saturated)	\$3.30/thousand pounds
30 psig steam (saturated)	\$1.60/thousand pounds
Boiler feed water	\$0.12/thousand pounds (at 180 psig and 275 F)
Fuel gas	\$3.80/million BTU (HHV)
Cooling water	\$120/annual gallons per minute

Table 3: *Utility Costs*

Utility Specifications		
Extreme Maximum Temperature	117	ºF
Extreme Minimum Temperature	-22	ºF
Process Design Dry Bulb (Cooling Towers)	97	ºF
Process Design Dry Bulb (Air Coolers)	105	ºF
Process Design Wet Bulb (Cooling Towers)	76	ºF
Equipment Minimum Design Temperature	-11	ºF
Absolute Atmospheric Pressure	14.13	PSIA
Cooling Water Supply Temperature	87	ºF
Cooling Water Return Temperature	120	ºF
Cooling Water Supply Pressure	50	PSIG
Cooling Water Return Pressure	35	PSIG
Cooling Water Fouling Resistance	0.002	Hr- ºF-Ft^2
Instrument Air Pressure	85	PSIG
Instrument Air Dew Point	-10	ºF
Elevation Above Sea Level	992.33	Ft

Table 4: *Utility Specifications*

Steam Specifications								
Steam Systems	**Pressure**				**Temperature**			
	Min	Normal	Max	Design	Min	Normal	Max	Design
600 PSIG	595	600	615	660	700	750	775	800
175 PSIG	170	175	190	205	475	500	550	600
40 PSIG	35	40	45	50	Saturated	287	Saturated	350

Table 5: *Steam Specifications*

The Peng-Robinson equation of state was recommended by the physical properties experts. The economic evaluation criteria involved a hurdle rate of 15% using a 5 year project evaluation.

3. Technical Discussion

3.1 Design Philosophy

3.1.1 De-isobutanizer Column (T-1)

There are many factors that weighed in on specifically how this column is designed. The primary concern is ensuring that spec was met for the isobutane product in the distillate. To achieve this spec, the feed pressure is changed with the help of a feed pump (P-1) to a value that allows the separation to occur. Pressure of the column is set to 150 psia, significantly above the P_{min} to achieve the 92.0 vol.% isobutane purity in the overhead stream. Reflux and boil-up ratios are varied in order to find the optimal variation for separation.

The second consideration is cost. A trial and error method is utilized to find the minimum number of trays in the column along with the optimal feed location for both the feed stream and the recycle stream. The cheapest design option includes 26 trays inside the column. Although the column could technically operate with fewer numbers of trays, it is not an optimal set-up due to the difficulty in getting a good feed location. A larger diameter is needed for this column to accommodate the combination of feed, recycle, and resulting high vapor flowrates.

A column diameter of over 13 feet is targeted with maintenance in mind. The large diameter would provide enough space for a worker to easily enter with any equipment that might be necessary to perform routine maintenance. Secondly, ensuring a column diameter of that size would also prevent constructional failures due to extreme weather common in Oklahoma.

An air-cooled heat exchanger (E-1) is chosen for the condenser. The reason for this being the reflux only needed to be cooled 1°F and air is a free utility. For the reboiler, it is found that multimillion-dollar savings could be obtained by switching out the boiler feed water for medium pressure steam. See Appendix Reboiler (E-2) Optimization.

In order to meet the environmentally specified isobutane product stream temperature and pressure specification, a shell-and-tube heat exchanger along with a control valve were designed. These would deliver the product safely to storage at the desired conditions of 20 psig above the bubble point at 128 °F.

3.1.2 De-butanizer Column (T-2)

This column is very similar to the de-isobutanizer tower (T-1). To accomplish the 1.0 vol.% purity of n-butane in the bottoms, a column pressure of 151 psia is specified. Similar to the first column (T-1), the second one (T-2) is optimized by varying the reflux and boil-up ratios in order to accomplish the desired separation.

This column is more difficult when trying to find the optimum number of trays. It requires more trays to accomplish the separation than T-1. According to the sensitivity analysis in Aspen simulation (Appendix T-2 Optimization Data), the column can only be run with 30 to 31 computational stages due to the same maintenance reasons as T-1. The computational stage number of 31 is the only one to provide a good feed location. The column diameter is found to be 14 feet. This is above the assumed minimum column diameter threshold that is set. This column does not have a high vapor flow; therefore, a smaller diameter is expected.

The reboiler (E-6) and condenser (E-4) were selected for the same reasons as the de-isobutanizer column's reboiler (E-2) and condenser (E-1). Steam instead of boiler feed water is used because the high duty. If boiler feed water is to be used, four more reboilers in series are required to reduce cost. If medium pressure steam is used, there is less equipment, less operating and overhead costs, and less required maintenance.

A shell-and-tube heat exchanger cooler along with a control valve are necessary to reduce the temperature and pressure of the product so they can be safely stored at 100 psia and 100 °F. These specifications are environmentally specified.

3.1.3 Reactor

Since the reactor pressure is specified to be 430 psig (450 psig with a 20 psig pressure drop), the temperature is the only variable that can be changed to maximize percent conversion (Equation 1). By trying different temperatures within the given range of 300 °F – 400 °F, it is found that 300 °F is the best because it maximized the % conversion (see Appendix Reactor Optimization).

Due to the exothermic nature of the reaction, in order to eliminate the slight risk of runaway reaction, a jacketed reactor is recommended.

3.1.4 Service Factor

A service factor of 0.96 was implemented in the finding the operating costs. This is based on the assumption of 8409.6 operating hours per year. This is because there will be a few days of downtime year round in the worst case scenario.

3.2 Description of the Process

The purpose of the Butamer Unit is to convert butane mixed feed into a usable isobutane product with a by-product of natural gas. The purity of the top isobutane product stream being removed from (T-1) is 92.0 vol.%. The bottoms natural gas product stream coming out of (T-2) has a purity of 1.0 vol.%.

The pump (P-1) pumps 7,800 BPD of the hydrocarbon mixture into the de-isobutanizer tower (T-1) at 95 °F and 100 psig. The first fractionation column (T-1) separates the isobutane from the rest of the hydrocarbons. The hot vapors from T-1 are condensed using an air-cooled heat exchanger (E-1) to 153 °F. The condensed product is collected in the condensate receiver, D-1, that is designed for a holdup time of 5 minutes. Some of the condensed product is pumped using a reflux pump (P-2) back into T-1. The distillate stream with the isobutane product is cooled to 128 °F using a cooling water shell-and-tube heat exchanger (E-3). A control valve (CV-1) is added to decrease the pressure to 116 psig for safe product storage.

The bottoms of the T-1 utilizes a kettle reboiler (E-2) to heat the liquid at the bottom of the column. While some of the heated vapors are put back into T-1, the remaining of the bottoms stream is pumped using P-3 into the de-butanizer tower (T-2). The hot vapors of T-2 are condensed using an air-cooled heat exchanger (E-4). The condensed product is collected in the condensate receiver, D-2. Some of the condensed product is pumped using a reflux pump (P-4) back into T-2. The distillate stream is pumped using P-5 and heated using E-5 before it enters the packed bed reactor (R-1). The reactor partially converts the n-butane to isobutane and recycles it back to T-1. The bottoms product of T-2 utilizes a kettle reboiler (E-6) to heat the liquid at the bottom of the column. While some heated vapors are put back into T-2, the reminaing of the bottoms stream with natural gasoline by-product is cooled to 98 °F using a cooling water shell-and-tube heat exchanger (E-7). A control valve (CV-2) is added to decrease the pressure to 100 psig for safe natural gas product storage.

3.2.1 PFD with Temperatures and Pressures

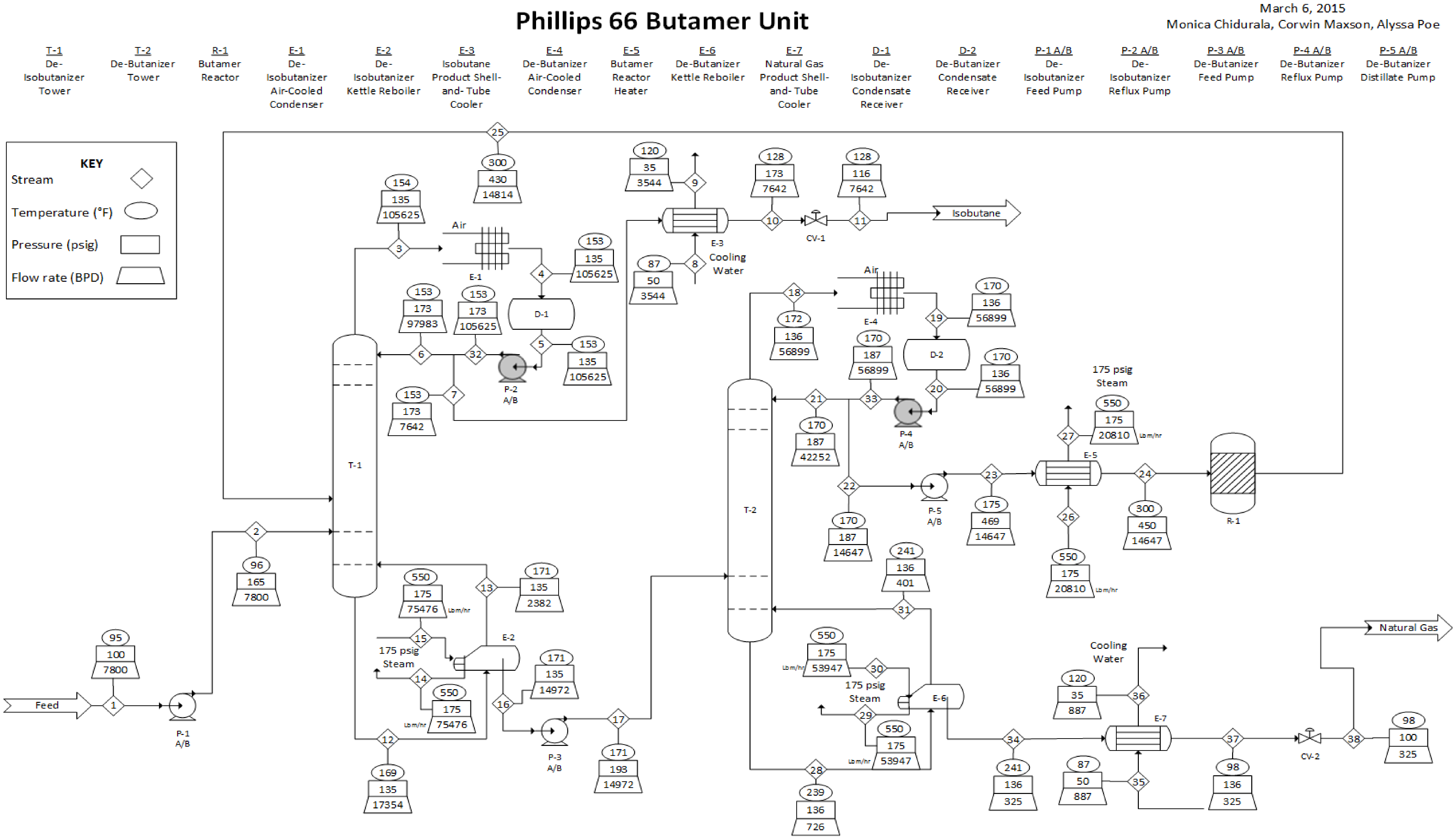

Figure 2: *PFD with Temperatures, Pressures, and Flowrates*

3.2.2 Stream Summary Table

Stream Number	1	2	3	4	5	32	6	7	8	9	10
Stream Label	Feed to De-isobutanizer feed pump, P-1	Feed Pump, P-1, Discharge	De-isobutanizer Condenser Feed	De-isobutanizer Condensate Receiver Feed	De-isobutanizer Reflux Pump Feed	De-isobutanizer Reflux Pump Discharge	De-isobutanizer Reflux	De-isobutanizer Tops Isobutane	Cooling water into Isobutane Cooler E-3	Cooling water out of Isobutane Cooler E-4	Isobutane Product HEX Discharge (Cooled)
Phase	Liquid	Liquid	Gas	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Mass Flow (lb/hr)											
Water ; H_2O (lb_m/hr)	0.000	0.000						0.000	31807.773	31807.773	0.000
Propane ; C_3H_8 (lb_m/hr)	0.020	0.020						0.021	0.000	0.000	0.021
1-butene ; C_4H_8 (lb_m/hr)	0.001	0.001						0.001	0.000	0.000	0.001
i-butene ; C_4H_8 (lb_m/hr)	0.002	0.002						0.002	0.000	0.000	0.002
Tr2-Butene ; C_4H_8 (lb_m/hr)	0.001	0.001						0.001	0.000	0.000	0.001
Cis2-Butene ; C_4H_8 (lb_m/hr)	0.001	0.001						0.001	0.000	0.000	0.001
I-Butane ; C_4H_{10} (lb_m/hr)	0.320	0.320						0.920	0.000	0.000	0.920
N-Butane ; C_4H_{10} (lb_m/hr)	0.613	0.613						0.054	0.000	0.000	0.054
2,2 Dimethyl Propane ; C_5H_{12} (lb_m/hr)	0.001	0.001						0.000	0.000	0.000	0.000
I-Pentane ; C_5H_{12} (lb_m/hr)	0.038	0.038						0.000	0.000	0.000	0.000
N-Pentane ; C_5H_{12} (lb_m/hr)	0.001	0.001						0.000	0.000	0.000	0.000
N-Hexane ; C_6H_{14} (lb_m/hr)	0.000	0.000						0.000	0.000	0.000	0.000
Total Flow lbmol/hr	1126.909	1126.909						1085.574	1767.098	1767.098	1085.574
Total Flow lb/hr	65646.310	65646.310						62692.500	31807.773	31807.773	62692.500
Total Flow cuft/hr	1824.740	1824.740	24710.000	24710.000	24710.000	24710.000	22922.000	1787.877	510.927	510.927	1787.877
Component Standard Volume Flow (BBL/DAY)											
Water ; H_2O (BBL/DAY)	1.100	1.100						0.517	2184.000	2184.000	0.517
Propane ; C_3H_8 (BBL/DAY)	159.898	159.898						159.898	0.000	0.000	159.898
1-butene ; C_4H_8 (BBL/DAY)	7.800	7.800						7.800	0.000	0.000	7.800
i-butene ; C_4H_8 (BBL/DAY)	15.600	15.600						15.600	0.000	0.000	15.600
Tr2-Butene ; C_4H_8 (BBL/DAY)	7.800	7.800						7.746	0.000	0.000	7.746
Cis2-Butene ; C_4H_8 (BBL/DAY)	7.800	7.800						6.905	0.000	0.000	6.905
I-Butane ; C_4H_{10} (BBL/DAY)	2495.968	2495.968						7031.034	0.000	0.000	7031.034
N-Butane ; C_4H_{10} (BBL/DAY)	4784.237	4784.237						412.920	0.000	0.000	412.920
2,2 Dimethyl Propane ; C_5H_{12} (BBL/DAY)	7.800	7.800						0.009	0.000	0.000	0.009
I-Pentane ; C_5H_{12} (BBL/DAY)	300.296	300.296						0.000	0.000	0.000	0.000

N-Pentane ; C_5H_{12} (BBL/DAY)	7.800	7.800						0.000	0.000	0.000	0.000
N-Hexane ; C_6H_{14} (BBL/DAY)	3.900	3.900						0.000	0.000	0.000	0.000
Total Flow BBL/day	7799.998	7799.998	105625.000	105625.000	105625.000	105625.000	97983.000	7642.428	2184.000	2184.000	7642.428
Component Standard Volume Fraction											
Water ; H_2O	0.000	0.000						0.000	1.000	1.000	0.000
Propane ; C_3H_8	0.020	0.020						0.021	0.000	0.000	0.021
1-butene ; C_4H_8	0.001	0.001						0.001	0.000	0.000	0.001
i-butene ; C_4H_8	0.002	0.002						0.002	0.000	0.000	0.002
Tr2-Butene ; C_4H_8	0.001	0.001						0.001	0.000	0.000	0.001
Cis2-Butene ; C_4H_8	0.001	0.001						0.001	0.000	0.000	0.001
I-Butane ; C_4H_{10}	0.320	0.320						0.920	0.000	0.000	0.920
N-Butane ; C_4H_{10}	0.613	0.613						0.054	0.000	0.000	0.054
2,2 Dimethyl Propane ; C_5H_{12}	0.001	0.001						0.000	0.000	0.000	0.000
I-Pentane ; C_5H_{12}	0.038	0.038						0.000	0.000	0.000	0.000
N-Pentane ; C_5H_{12}	0.001	0.001						0.000	0.000	0.000	0.000
N-Hexane ; C_6H_{14}	0.000	0.000						0.000	0.000	0.000	0.000
Temperature F	95	95.56337	154	153	153	153	153	153.1823	87	120	128
Pressure psia	114.6959	160	150	150	150	150	150	150	50	50	150
Vapor Frac	0	0	1	0	0	0	0	0	0	0	0
Liquid Frac	1	1	0	1	1	1	1	1	1	1	1
Enthalpy Btu/lbmol	-63750.03	-63727.94						-63026.51	-123510	-122920	-63991.77
Enthalpy Btu/lb	-1094.356	-1093.977						-1091.358	-6855.854	-6822.938	-1108.072
Enthalpy Btu/hr	-71840000	-71816000						-68420000	-218250000	-217210000	-69468000
Density lb/cuft	34.642	34.61738						30.68793	61.71515	60.6799	32.02597
Average MW	58.25345	58.25345						57.75056	18.01528	18.01528	57.75056

Table 6: *Stream Summary Table*

Stream Summary Table

Stream Number	11	12	13	14	15	16	17	18	19	20	33
Stream Label	Isobutane Product to Storage	De-isobutanizer Reboiler Feed	De-isobutanizer Reboiler to Column	Steam into the Reboiler, E-2	Steam out of the Reboiler, E-3	De-isobutanizer Bottoms	De-butanizer Feed, P-3 Discharge	De-butanizer Condenser Feed	De-butanizer Condensate Receiver Feed	De-butanizer Reflux Pump Feed	De-butanizer Reflux Pump Discharge
Phase	Liquid	Liquid	Gas	Gas	Liquid	Liquid	Liquid	Gas	Liquid	Liquid	Liquid
Mass Flow (lb/hr)											
Water ; H_2O (lb_m/hr)	0.000			75475.577	75475.577	0.002	0.002				
Propane ; C_3H_8 (lb_m/hr)	0.021			0.000	0.000	0.001	0.001				
1-butene ; C_4H_8 (lb_m/hr)	0.001			0.000	0.000	0.001	0.001				
i-butene ; C_4H_8 (lb_m/hr)	0.002			0.000	0.000	0.002	0.002				
Tr2-Butene ; C_4H_8 (lb_m/hr)	0.001			0.000	0.000	0.008	0.008				
Cis2-Butene ; C_4H_8 (lb_m/hr)	0.001			0.000	0.000	0.024	0.024				
I-Butane ; C_4H_{10} (lb_m/hr)	0.920			0.000	0.000	0.335	0.335				
N-Butane ; C_4H_{10} (lb_m/hr)	0.054			0.000	0.000	0.601	0.601				
2,2 Dimethyl Propane ; C_5H_{12} (lb_m/hr)	0.000			0.000	0.000	0.003	0.003				
I-Pentane ; C_5H_{12} (lb_m/hr)	0.000			0.000	0.000	0.023	0.023				
N-Pentane ; C_5H_{12} (lb_m/hr)	0.000			0.000	0.000	0.001	0.001				
N-Hexane ; C_6H_{14} (lb_m/hr)	0.000			0.000	0.000	0.000	0.000				
Total Flow lbmol/hr	1085.574			0.000	0.000	2184.854	2184.854				
Total Flow lb/hr	62692.500			75475.577	75475.577	126540.000	126540.000				
Total Flow cuft/hr	1787.877	4060.000	557.000			3502.528	3502.528	13311.000	13311.000	13311.000	13311.000
Component Standard Volume Flow (BBL/DAY)											
Water ; H_2O (BBL/DAY)	0.517					29.155	29.155				
Propane ; C_3H_8 (BBL/DAY)	159.898					15.416	15.416				
1-butene ; C_4H_8 (BBL/DAY)	7.800					14.831	14.831				
i-butene ; C_4H_8 (BBL/DAY)	15.600					25.825	25.825				
Tr2-Butene ; C_4H_8 (BBL/DAY)	7.746					125.605	125.605				
Cis2-Butene ; C_4H_8 (BBL/DAY)	6.905					359.784	359.784				
I-Butane ; C_4H_{10} (BBL/DAY)	7031.034					5014.727	5014.727				
N-Butane ; C_4H_{10} (BBL/DAY)	412.920					8991.054	8991.054				
2,2 Dimethyl Propane ; C_5H_{12} (BBL/DAY)	0.009					41.658	41.658				
I-Pentane ; C_5H_{12} (BBL/DAY)	0.000					341.362	341.362				

N-Pentane ; C_5H_{12} (BBL/DAY)	0.000					8.513	8.513				
N-Hexane ; C_6H_{14} (BBL/DAY)	0.000					3.918	3.918				
Total Flow BBL/day	7642.428	17354.000	2382.000			14971.848	14971.848	56899.000	56899.000	56899.000	56899.000
Component Standard Volume Fraction											
Water ; H_2O	0.000			1.000	1.000	0.002	0.002				
Propane ; C_3H_8	0.021			0.000	0.000	0.001	0.001				
1-butene ; C_4H_8	0.001			0.000	0.000	0.001	0.001				
i-butene ; C_4H_8	0.002			0.000	0.000	0.002	0.002				
Tr2-Butene ; C_4H_8	0.001			0.000	0.000	0.008	0.008				
Cis2-Butene ; C_4H_8	0.001			0.000	0.000	0.024	0.024				
I-Butane ; C_4H_{10}	0.920			0.000	0.000	0.335	0.335				
N-Butane ; C_4H_{10}	0.054			0.000	0.000	0.601	0.601				
2,2 Dimethyl Propane ; C_5H_{12}	0.000			0.000	0.000	0.003	0.003				
I-Pentane ; C_5H_{12}	0.000			0.000	0.000	0.023	0.023				
N-Pentane ; C_5H_{12}	0.000			0.000	0.000	0.001	0.001				
N-Hexane ; C_6H_{14}	0.000			0.000	0.000	0.000	0.000				
Temperature F	127.9645	169	171	550	550	170.5685	170.9778	172	170	170	170
Pressure psia	131.122	150	150	190	190	150	175	151	151	151	202
Vapor Frac	0	0	1	1	0	0	0	1	0	0	0
Liquid Frac	1	1	0	0	1	1	1	0	1	1	1
Enthalpy Btu/lbmol	-63991.77					-59920.94	-59908.9				
Enthalpy Btu/lb	-1108.072					-1034.606	-1034.398				
Enthalpy Btu/hr	-69468000					-130920000	-130890000				
Density lb/cuft	32.02777					31.16907	31.14669				
Average MW	57.75056					57.91669	57.91669				

Table 7: *Stream Summary Table (Continued)*

Stream Summary Table

Stream Number	21	22	23	24	25	26	27	28	29	30	31
Stream Label	De-butanizer Reflux	De-butanizer Tops Product, P-5 Feed	Reactor Pre-Heater Feed	Feed to Reactor	Reactor Recycle	Steam into E-5, Reactor Feed Heater	Steam out of E-5, Reactor Feed Heater	De-butanizer Reboiler Feed	Steam out of E-6, Reboiler	Steam into E-6, Reboiler	De-butanizer Reboiler Column Feed
Phase	Liquid	Liquid	Liquid	Liquid	Liquid	Gas	Liquid	Liquid	Gas	Liquid	Gas
Mass Flow (lb/hr)											
Water ; H_2O (lb_m/hr)		0.002	0.002	0.002	0.002	20810.263	20810.263				
Propane ; C_3H_8 (lb_m/hr)		0.001	0.001	0.001	0.001	0.000	0.000				
1-butene ; C_4H_8 (lb_m/hr)		0.001	0.001	0.001	0.001	0.000	0.000				
i-butene ; C_4H_8 (lb_m/hr)		0.002	0.002	0.002	0.002	0.000	0.000				
Tr2-Butene ; C_4H_8 (lb_m/hr)		0.009	0.009	0.009	0.008	0.000	0.000				
Cis2-Butene ; C_4H_8 (lb_m/hr)		0.025	0.025	0.025	0.024	0.000	0.000				
I-Butane ; C_4H_{10} (lb_m/hr)		0.342	0.342	0.342	0.645	0.000	0.000				
N-Butane ; C_4H_{10} (lb_m/hr)		0.614	0.614	0.614	0.312	0.000	0.000				
2,2 Dimethyl Propane ; C_5H_{12} (lb_m/hr)		0.002	0.002	0.002	0.002	0.000	0.000				
I-Pentane ; C_5H_{12} (lb_m/hr)		0.003	0.003	0.003	0.003	0.000	0.000				
N-Pentane ; C_5H_{12} (lb_m/hr)		0.000	0.000	0.000	0.000	0.000	0.000				
N-Hexane ; C_6H_{14} (lb_m/hr)		0.000	0.000	0.000	0.000	0.000	0.000				
Total Flow lbmol/hr		2143.522	2143.522	2143.522	2143.519	0.000	0.000				
Total Flow lb/hr		123586.000	123586.000	123586.000	123586.000	20810.263	20810.263		53946.929	53946.929	
Total Flow cuft/hr	9884.000	3426.606	3426.606	3426.606	3465.666	334.099	334.099	170.000			94.000
Component Standard Volume Flow (BBL/DAY)											
Water ; H_2O (BBL/DAY)		28.573	28.573	28.573	28.573	1428.134	1428.134				
Propane ; C_3H_8 (BBL/DAY)		15.416	15.416	15.416	15.416	0.000	0.000				
1-butene ; C_4H_8 (BBL/DAY)		14.831	14.831	14.831	14.831	0.000	0.000				
i-butene ; C_4H_8 (BBL/DAY)		25.825	25.825	25.825	25.825	0.000	0.000				
Tr2-Butene ; C_4H_8 (BBL/DAY)		125.551	125.551	125.551	125.551	0.000	0.000				
Cis2-Butene ; C_4H_8 (BBL/DAY)		358.922	358.922	358.922	358.889	0.000	0.000				
I-Butane ; C_4H_{10} (BBL/DAY)		5014.722	5014.722	5014.722	9549.794	0.000	0.000				
N-Butane ; C_4H_{10} (BBL/DAY)		8987.810	8987.810	8987.810	4619.734	0.000	0.000				
2,2 Dimethyl Propane ; C_5H_{12} (BBL/DAY)		33.867	33.867	33.867	33.867	0.000	0.000				
I-Pentane ; C_5H_{12} (BBL/DAY)		41.065	41.065	41.065	41.066	0.000	0.000				
N-Pentane ; C_5H_{12} (BBL/DAY)		0.713	0.713	0.713	0.713	0.000	0.000				
N-Hexane ; C_6H_{14} (BBL/DAY)		0.018	0.018	0.018	0.018	0.000	0.000				

Total Flow BBL/day	42252.000	14647.314	14647.314	14647.314	14814.276	1428.134	1428.134	726.000			401.000
Component Standard Volume Fraction											
Water ; H_2O		0.002	0.002	0.002	0.002	1.000	1.000		1.000	1.000	
Propane ; C_3H_8		0.001	0.001	0.001	0.001	0.000	0.000		0.000	0.000	
1-butene ; C_4H_8		0.001	0.001	0.001	0.001	0.000	0.000		0.000	0.000	
i-butene ; C_4H_8		0.002	0.002	0.002	0.002	0.000	0.000		0.000	0.000	
Tr2-Butene ; C_4H_8		0.009	0.009	0.009	0.008	0.000	0.000		0.000	0.000	
Cis2-Butene ; C_4H_8		0.025	0.025	0.025	0.024	0.000	0.000		0.000	0.000	
I-Butane ; C_4H_{10}		0.342	0.342	0.342	0.645	0.000	0.000		0.000	0.000	
N-Butane ; C_4H_{10}		0.614	0.614	0.614	0.312	0.000	0.000		0.000	0.000	
2,2 Dimethyl Propane ; C_5H_{12}		0.002	0.002	0.002	0.002	0.000	0.000		0.000	0.000	
I-Pentane ; C_5H_{12}		0.003	0.003	0.003	0.003	0.000	0.000		0.000	0.000	
N-Pentane ; C_5H_{12}		0.000	0.000	0.000	0.000	0.000	0.000		0.000	0.000	
N-Hexane ; C_6H_{14}		0.000	0.000	0.000	0.000	0.000	0.000		0.000	0.000	
Temperature F	170	170.1224	175.1215	300	300	550	550	239	550	550	241
Pressure psia	202	151	464.6959	464.6959	444.6959	175	175	151	190	190	151
Vapor Frac	0	0	0	1	1	1	0	0	1	0	1
Liquid Frac	1	1	1	0	0	0	1	1	0	1	0
Enthalpy Btu/lbmol		-59682.31	-59531.24	-50110.21	-51057.42	-101460	-118940				
Enthalpy Btu/lb		-1035.153	-1032.533	-869.1312	-885.5601	-5631.86	-6601.942				
Enthalpy Btu/hr		-127930000	-127610000	-107410000	-109440000	-117240000	-137430000				
Density lb/cuft		31.10302	30.82582	5.586717	4.964065	1.444091	53.31192				
Average MW		57.65551	57.65551	57.65551	57.65552	18.01528	18.01528				

Table 8: *Stream Summary Table (Continued)*

Stream Summary Table

Stream Number	34	35	36	37	38
Stream Label	De-butanizer Natural Gas Product	Cooling Water into E-7	Cooling Water out of E-7	Natural Gas HEX Discharge	Natural Gas Feed to Storage
Phase	Liquid	Liquid	Liquid	Liquid	Liquid
Mass Flow (lb/hr)					
Water ; H_2O (lb_m/hr)	0.002	7963.120	7963.120	0.002	0.002
Propane ; C_3H_8 (lb_m/hr)	0.000	0.000	0.000	0.000	0.000
1-butene ; C_4H_8 (lb_m/hr)	0.000	0.000	0.000	0.000	0.000
i-butene ; C_4H_8 (lb_m/hr)	0.000	0.000	0.000	0.000	0.000
Tr2-Butene ; C_4H_8 (lb_m/hr)	0.000	0.000	0.000	0.000	0.000
Cis2-Butene ; C_4H_8 (lb_m/hr)	0.003	0.000	0.000	0.003	0.003
I-Butane ; C_4H_{10} (lb_m/hr)	0.000	0.000	0.000	0.000	0.000
N-Butane ; C_4H_{10} (lb_m/hr)	0.010	0.000	0.000	0.010	0.010
2,2 Dimethyl Propane ; C_5H_{12} (lb_m/hr)	0.024	0.000	0.000	0.024	0.024
I-Pentane ; C_5H_{12} (lb_m/hr)	0.925	0.000	0.000	0.925	0.925
N-Pentane ; C_5H_{12} (lb_m/hr)	0.024	0.000	0.000	0.024	0.024
N-Hexane ; C_6H_{14} (lb_m/hr)	0.012	0.000	0.000	0.012	0.012
Total Flow lbmol/hr	41.332	0.000	0.000	41.332	41.332
Total Flow lb/hr	2953.646	7963.120	7963.120	2953.646	2953.646
Total Flow cuft/hr	75.922	127.692	127.692	75.922	75.922
Component Standard Volume Flow (BBL/DAY)					
Water ; H_2O (BBL/DAY)	0.583	545.829	545.829	0.583	0.583
Propane ; C_3H_8 (BBL/DAY)	0.000	0.000	0.000	0.000	0.000
1-butene ; C_4H_8 (BBL/DAY)	0.000	0.000	0.000	0.000	0.000
i-butene ; C_4H_8 (BBL/DAY)	0.000	0.000	0.000	0.000	0.000
Tr2-Butene ; C_4H_8 (BBL/DAY)	0.054	0.000	0.000	0.054	0.054
Cis2-Butene ; C_4H_8 (BBL/DAY)	0.861	0.000	0.000	0.861	0.861
I-Butane ; C_4H_{10} (BBL/DAY)	0.005	0.000	0.000	0.005	0.005
N-Butane ; C_4H_{10} (BBL/DAY)	3.245	0.000	0.000	3.245	3.245
2,2 Dimethyl Propane ; C_5H_{12} (BBL/DAY)	7.791	0.000	0.000	7.791	7.791
I-Pentane ; C_5H_{12} (BBL/DAY)	300.296	0.000	0.000	300.296	300.296
N-Pentane ; C_5H_{12} (BBL/DAY)	7.800	0.000	0.000	7.800	7.800

N-Hexane ; C_6H_{14} (BBL/DAY)	3.900	0.000	0.000	3.900	3.900
Total Flow BBL/day	324.535	545.829	545.829	324.535	324.535
Component Standard Volume Fraction					
Water ; H_2O	0.002	1.000	1.000	0.002	0.002
Propane ; C_3H_8	0.000	0.000	0.000	0.000	0.000
1-butene ; C_4H_8	0.000	0.000	0.000	0.000	0.000
i-butene ; C_4H_8	0.000	0.000	0.000	0.000	0.000
Tr2-Butene ; C_4H_8	0.000	0.000	0.000	0.000	0.000
Cis2-Butene ; C_4H_8	0.003	0.000	0.000	0.003	0.003
I-Butane ; C_4H_{10}	0.000	0.000	0.000	0.000	0.000
N-Butane ; C_4H_{10}	0.010	0.000	0.000	0.010	0.010
2,2 Dimethyl Propane ; C_5H_{12}	0.024	0.000	0.000	0.024	0.024
I-Pentane ; C_5H_{12}	0.925	0.000	0.000	0.925	0.925
N-Pentane ; C_5H_{12}	0.024	0.000	0.000	0.024	0.024
N-Hexane ; C_6H_{14}	0.012	0.000	0.000	0.012	0.012
Temperature F	240.6726	87	120	98	98.13164
Pressure psia	151	65	50	151	114.7
Vapor Frac	0	0	0	0	0
Liquid Frac	1	1	1	1	1
Enthalpy Btu/lbmol	-69850.59	-123510	-122920	-76197.57	-76197.57
Enthalpy Btu/lb	-977.4524	-6855.854	-6822.882	-1066.269	-1066.269
Enthalpy Btu/hr	-2887000	-54547000	-54284000	-3.15E+06	-3.15E+06
Density lb/cuft	31.50163	61.71515	60.6781	37.80827	37.80329
Average MW	71.46188	18.01528	18.01528	71.46188	71.46188

Table 9: *Stream Summary Table (Continued)*

3.2.3 Utility Summary Table

Stream Number	8	9	15	14	26
Stream Label	Product HEX Cooling Water IN	Product HEX Cooling Water OUT	De-isobutanizer Reboiler Steam IN	De-isobutanizer Reboiler Steam OUT	Reactor HEX Steam IN
Temperature (°F)	87	120	550	550	550
Pressure (psig)	50	35	175	175	175
Air (lb/hr)	0	0	0	0	0
Steam (lb/hr)	0	0	75476	75476	20810
Cooling Water (lb/hr)	31808	31808	0	0	0

Stream Number	27	E-1		E-4	
Stream Label	Reactor HEX Steam OUT	De-isobutanizer Condenser Air IN	De-isobutanizer Condenser Air OUT	De-butanizer Condenser Air IN	De-butanizer Condenser Air OUT
Temperature (°F)	550	105	114	105	132
Pressure (psig)	175	0	0	0	0
Air (lb/hr)	0	13780675	13780675	5845632	5845632
Steam (lb/hr)	20810	0	0	0	0
Cooling Water (lb/hr)	0	0	0	0	0

Table 10: *Utility Summary Table*

Stream Number	32	33	35	36
Stream Label	De-butanizer Reboiler Steam IN	De-butanizer Reboiler Steam OUT	Natural Gas HEX Cooling Water IN	Natural Gas HEX Cooling Water OUT
Temperature (°F)	550	550	87	120
Pressure (psig)	175	175	151	151
Air (lb/hr)	0	0	0	0
Steam (lb/hr)	53947	53947	0	0
Cooling Water (lb/hr)	0	0	7963	7963

Table 11: *Utility Summary Table (Continued)*

Electricity consumption (kW)	
De-isobutanizer Feed Pump (P-1)	13
De-isobutanizer Reflux Pump (P-2)	62
De-butanizer Feed Pump (P-3)	21
De-butanizer Reflux Pump (P-4)	21
Reactor Pump (P-5)	101

Table 12: *Pump Electricity Consumption Summary*

3.2.4 PFD with Basic Control Strategy

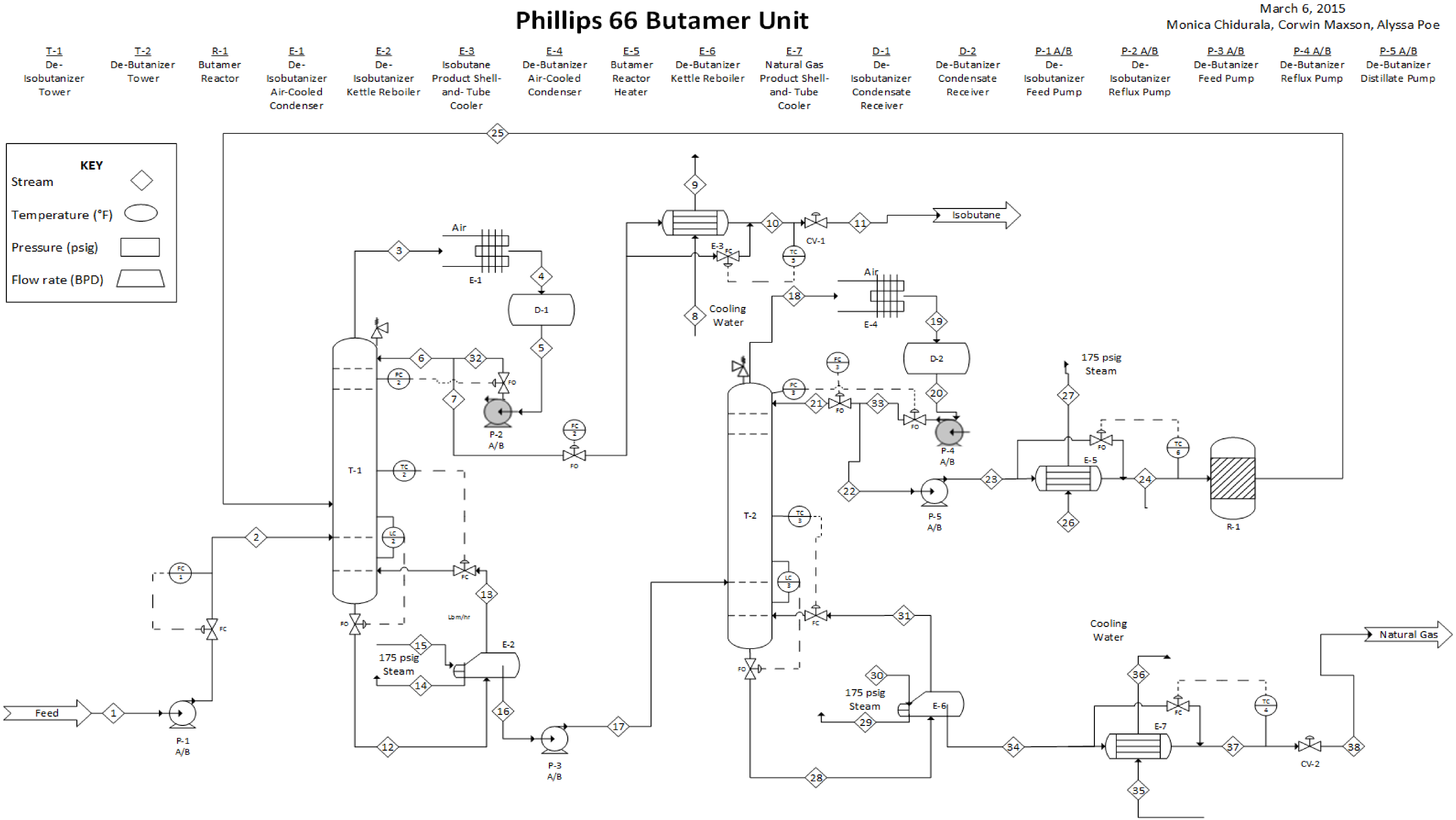

Figure 3: *PFD with Basic Control Scheme*

3.2.5 Control System Summary

	Number of Control Loops
Temperature	5
Level	2
Flow	3
Pressure	2
Other	0

Table 13: *Control Loop Summary*

Pairing of Controlled and Manipulated Variables

Loop	**Controlled Variable**	**Manipulated Variable**
Feed Flow	Flow of Feed to T-1 column	Flow of Feed to T-1 column
T-1 Liquid Level	Liquid level in bottom of column.	Flowrate of liquid out the bottom of the column.
T-1 Pressure	Pressure in the column	Flowrate of liquid leaving condensate receiver D-1
T-1 Temperature	Temperature in the column	Reflux ratio
Isobutane Flow	Flow of isobutane product to storage	Flow of isobutane product to storage
Isobutane Temperature	Temperature of isobutane product to storage	Flowrate of product stream through bypass around cooler
T-2 Liquid Level	Liquid level in bottom of column.	Flowrate of liquid out the bottom of the column.
T-2 Pressure	Pressure in the column	Flowrate of liquid leaving condensate receiver D-2
T-2 Temperature	Temperature in the column	Reflux ratio
T-2 Reflux	Reflux rate for T-2	Flowrate of reflux back into column
Natural Gas Temperature	Temperature of natural gas product to storage	Flowrate of product stream through bypass around cooler
Reactor Feed Temperature	Temperature of feed to reactor	Flowrate of reactor feed around bypass of heater

Table 14: *Manipulated vs. Controlled Variable Summary*

Control valve failure mode summary:

Loop Tag	**Service**	**Failure Position**	**Basis for specified failure position**
Feed Flow	Flow	CLOSED	This will stop the flow of feed to the entire unit.
T-1 Liquid Level	Level	OPEN	Drains liquid out of column to ensure the column does not fill up and overflow.
T-1 Pressure	Pressure	OPEN	Maximizes flow out of column which reduces the pressure as much as possible.
T-1 Temperature	Temperature	CLOSED	Stops flow from reboiler back into column thus reducing the temperature of the column.
Isobutane Flow	Flow	OPEN	Maximizes flow out of the column and towards storage.
Isobutane Temperature	Temperature	CLOSED	Forces all of the product stream through the cooler which causes product to be stored at lowest possible temperature.
T-2 Liquid Level	Level	OPEN	Drains liquid out of column to ensure the column does not fill up and overflow.
T-2 Pressure	Pressure	OPEN	Maximizes flow out of column which reduces the pressure as much as possible.
T-2 Temperature	Temperature	CLOSED	Stops flow from reboiler back into column thus reducing the temperature of the column.
T-2 Reflux	Flow	OPEN	Maximizes reflux flow which minimizes flow to reactor
Natural Gas Temperature	Temperature	CLOSED	Forces the entire product stream through the cooler which causes product to be stored at lowest possible temperature.
Reactor Feed Temperature	Temperature	OPEN	Maximizes bypass around reactor heater which minimizes temperature of the stream entering the reactor.

Table 15: *Control Valve Fail Open/Closed Summary*

3.3 Technical Issues and Design Practices

3.3.1 De-Isobutanizer Column (T-1)

The de-isobutanizer splits the butane mix. It is the first column in the Butamer Unit accepting a feed and the recycle stream from the reactor in order to meet a top product spec of 92.0 vol.% isobutane. In order to meet the given specification the following design parameters were instituted.

Design

- Column Pressure is 150 psia – varied on Aspen Plus V8.2 simulation for convergence
- 26 trays and 25 theoretical stages – varied on Aspen Plus V8.2 simulation for convergence
- Diameter of 18 ft – Calculated on Aspen Fair tray sizing method
- Height of 78 ft (26 trays with 2 ft tray spacing, 8 ft skirt, 7 ft reboiler)
- Aspen Plus V8.2 simulated as total condenser and kettle reboiler
- Tray spacing of 2 ft due to column diameter – Turton Table 11.14 heuristic [1]
- 80% overall efficiency – calculated using O'Connell correlation CHE 4224 course notes [4]

Fenske's equation is used to determine the minimum number of stages (**Table 38**) which was used as a baseline to vary stages on Aspen Plus V8.2. A series of trial and error simulations were generated in order to determine the proper number of trays necessary to meet the product specifications along with minimizing cost. See Appendix *De-isobutanizer (T-1) Feed Stream Column Optimization Data*.

Tray diameter was found by the following assumptions:

Assumption

- 80% flooding
- Fair flooding calculation method
- Sieve trays

Pumps

At many points in the de-isobutanizer unit it was necessary to increase the pressure of the fluid in order to maintain or increase pressure to accomplish required separation. This is done so by utilizing two centrifugal pumps on the system plus two spares. Centrifugal pumps were chosen because they have a high capacity to pump fluid at high flow rates and work better under harsh conditions. The motor efficiencies for the pumps have been chosen using flow rates from Figure 12.3 in GPSA [5].

- P-1A De-isobutanizer (T-1) column feed pump. Discharge pressure was set at 160 psia in Aspen.

- P-1B Spare de-isobutanizer (T-1) column feed pump.
- P-2A De-isobutanizer (T-1) column reflux pump. Discharge pressure was set at 155 psia.
- P-2B Spare de-isobutanizer (T-1) column reflux pump.

Assumptions

- Both pumps and motor efficiencies were calculated using (p520-21 Peters and Timmerhaus, 1991) [3].
- All line loss pressure drops were assumed to be 5 psi.
- Pressure drop across the control valve was 10 psia according to Dr. Ramsey's and Dr. Aichele's last semester project statements [4].
- The pumps were sitting at ground level and were pumped to the desired heights with 2 ft tray spacing per stage + 7 ft for reboiler + 8 ft for skirt.
- P-2 reflux pump oversized by 10% using heuristics from Turton Table 11.11 [1].

Heat Exchangers

It is necessary for a condenser and a reboiler to be installed on the de-isobutanizer column (T-1). The hot vapors need to be condensed in order to isolate the product and the liquid at the bottom of the column needs to be heated to generate these hot vapors. There is a total count of three heat exchangers surrounding T-1. Final design measurements are indicated in **Table 16** Equipment Summary Table.

- E-1 Air-cooled heat exchanger utilized in the condensing process of the vapors.
- E-2 A kettle reboiler used to prepare the bottoms for re-entry back into the column.
- E-3 Shell-and-tube heat exchanger product cooler.

An air-cooled heat exchanger was chosen for the condenser because the vapors from the column needed to be cooled for condensation to occur. Since the condensate only needed to be cooled ~1 °F, an air-cooled heat exchanger can be used to save utility expenses since air is free. Aspen Plus V8.2 provided the necessary condenser duty (Appendix *Aspen Complete Simulation Report)* to calculate the area required for the heat exchanger (Appendix *Condenser (Air-cooled HEX) E-1 Design)*. It was designed to have an explosion proof motor.

A kettle reboiler (E-2) was installed for the bottoms. Steam was chosen as the utility because it brought down the overall utility cost of the system (Appendix *E-2 Reboiler Steam Temperature Optimization)*. Using steam instead of boiler feed water also allows a decrease in overall size of the reboiler from needing four heat exchangers in series to a single kettle reboiler. Kettle reboiler is chosen because it is the most commonly used in the industry for converting the bottoms product back to a vapor so it can re-enter the column. The temperatures were analyzed to help design this piece of equipment.

On the distillate product stream from the de-isobutanizer column, it was necessary to reduce the temperature and pressure of the isobutane product for safe storage. This reduction in temperature needed to be done by a shell-and-tube heat exchanger (E-3) because a large temperature drop is needed. The utility used is cooling water since it is the only suitable utility available for such large temperature difference.

Assumptions

- E-1
  - Condenser duty from Aspen Plus V8.2 Report.
  - Single pass OD 1", 5/8" fins at 10 fins/in, Pitch of 2 ¼" triangular – used from last semester's preliminary design requirements [4] for easier preliminary design calculations.
  - Air inlet temperature of 105 °F from **Table 4**.
  - Tube length of 249 ft from GPSA example [5].
  - Fan motor efficiency 92% from GPSA example [5].
  - Motor size (bhp) 800 hp using Design notes from last semester [4].
  - Material properties such as specific heat capacity obtained from Aspen Plus V8.2.
- E-2
  - Reboiler duty from Aspen Plus V8.2 Report.
  - $T_{steam,in} = T_{steam,out} = 550$ °F (medium pressure steam condenses at constant temperature) **Table 4**.
  - Approach temperature of 20 °F from heuristic for steam reboiler approach temperature, from Gas Purification by Cole & Nielsen.
  - Heat transfer coefficient of 140 (Btu/hr ft^2 °F) - heuristic given in Dr. Aichele's class notes [4] on estimating U; U chosen for steam reboilers; lowest for conservative purposes.
  - Correction factor is 0.90 – heuristics Turton Table 11.11 [1].
  - Material properties such as specific heat capacity obtained from Aspen Plus V8.2.
- E-3
  - Cooler duty obtained from Aspen Plus V8.2 Report.
  - Shell side is the isobutane and utility cooling water is tube side for easier maintenance, cleaning purposes, and to prevent fouling.
  - Cooling water temperatures in and out specified as 87 °F and 120 °F respectively as specified in **Table 4**.
  - U = 80 (Btu/hr ft^2 °F) heuristic given in Dr. Aichele's class notes [4] on estimating U; U chosen for steam reboilers; lowest for conservative purposes.
  - Correction factor = 0.9 – heuristics Turton Table 11.11 [1].
  - Material properties such as specific heat capacity obtained from Aspen Plus V8.2.

All three heat exchanger costs were calculated with the material of construction assumed to be carbon steel. See explanation at the end of this section.

Condensate Receiver

A drum is needed to act as a temporary storage for the condensate leaving the condenser E-1. This drum collects the liquid condensate at the inlet and is later pumped back into the column T-1 using reflux pump. Liquid flow rate is used to calculate the diameter and height.

Assumptions

- L/D ratio = 3 – Turton Table 11.6 heuristic 2.5 < L < 5 [1]
- T_{res} = 5min – Turton Table 11.6 heuristic # 5 [1]

By comparing the cost of a horizontal vs. vertical drum (Appendix *Condensate Receiver for De-Isobutanizer Tower D-1 Design)* it was determined that it was more economical to go with a horizontal condensate receiver.

Design

- Volume is 4,243 ft^3
- Diameter 12.25 ft
- Length 36 ft

3.3.2 De-butanizer Column (T-2)

The de-butanizer column (T-2) sends the isobutane mixture out of the distillate to the reactor. The natural gas bottoms product is sent to storage. The natural gas product is specified to be 1.0 vol.% n-butane purity. The following parameters are determined in the design:

Design

- Column Pressure is 151 psia – varied on Aspen Plus V8.2 simulation for convergence
- 29 trays and 30 theoretical stages – varied on Aspen Plus V8.2 simulation for convergence
- Diameter of 14 ft – Calculated on Aspen Fair tray sizing method
- Height of 86 ft (29 trays with 2 ft tray spacing, 8 ft skirt, 7 ft reboiler)
- Aspen Plus V8.2 simulated as total condenser and kettle reboiler
- Tray spacing of 2 ft due to column diameter – Turton Table 11.14 heuristic [1]
- 85% overall efficiency – calculated using O'Connel correlation CHE 4224 course notes [4]

Fenske's equation is used to determine the minimum number of stages (**Table 38**) which was used as a baseline to vary stages on Aspen Plus V8.2.

Tray diameter was determined by utilizing the following assumptions.

Assumptions

- 80% flooding
- fair flooding calculation method
- sieve trays

Pumps

At the feed to the de-butanizer column (T-2) it is necessary to increase the pressure of the flow by 25 psia for the fluid to be pumped to a certain height for optimal feed location. To do this, a centrifugal pump (P-3) is implemented. Similarly, the reflux going through the condenser needs to remain just above column pressure to ensure a good circulation back into the column. To do this a second centrifugal pump (P-4) is placed for the condensate. In total four pumps were added surrounding T-2, two pumps in service and two spares.

- P-3A De-butanizer column feed pump – discharge pressure is set at 175 psia on Aspen Plus V8.2.
- P-3B Spare de-butanizer column feed pump.
- P-4A De-butanizer column condenser reflux pump - discharge pressure set at 201 psia.
- P-4B Spare de-butanizer column condenser reflux pump.

Figure 12.3 in the GPSA [5] manual provided guidance on final pump selection design by utilizing the calculated pump head and flowrates. The following assumptions were made:

Assumptions

- Both pumps and motor efficiencies were calculated using (p520-21 Peters and Timmerhaus, 1991) [3].
- All line loss pressure drops were assumed to be 5 psi.
- Pressure drop across the control valve was 10 psia according to Dr. Ramsey's and Dr. Aichele's last semester project statements [4].
- The pumps were sitting at ground level and were pumped to the desired heights with 2 ft tray spacing per stage + 7 ft for reboiler + 8 ft for skirt.
- P-4 reflux pump oversized by 10% using heuristics from Turton Table 11.11 [1].

Heat Exchanger

Surrounding the de-butanizer column is a total of three heat exchangers. One is a total condenser, a kettle reboiler, and lastly a cooler for the natural gas product.

- E-4 Air-cooled heat exchanger used as a condenser for the column reflux
- E-6 Steam kettle reboiler used to prepare bottoms feed to circulate back into the column
- E-7 Natural gas product stream shell-and-tube cooler

An air-cooled heat exchanger (E-4) was chosen for the condenser because the vapors from the column needed to be cooled for condensation to occur. Since the condensate only needed to be cooled ~1 °F, an air-cooled heat exchanger can be used to save utility expenses since air is free. Aspen Plus V8.2 provided the necessary condenser duty (Appendix *Aspen Complete Simulation Report)* to calculate the area required for the heat exchanger (Appendix *Condenser (Air-cooled HEX) E-2 Design)*. It was designed to have an explosion proof motor.

A kettle reboiler (E-6) was installed for the bottoms. Steam was chosen as the utility because it brought down the overall utility cost of the system (Appendix *E-6 Reboiler Steam Temperature Optimization)*. Using steam instead of boiler feed water also allows a decrease in overall size of the reboiler from needing four heat exchangers in series to a single kettle reboiler. Kettle reboiler is chosen because it is the most commonly used in the industry for converting the bottoms product back to a vapor so it can re-enter the column. The temperatures were analyzed to help design this piece of equipment.

A shell in tube heat exchanger (E-7) was chosen for the natural gas product stream because the product needed to be cooled to a value of 100 °F for safe storage. Cooling water is used to accomplish this temperature drop. This reduction in temperature needed to be done by a shell-and-tube heat exchanger (E-7) because a large temperature drop is needed. The utility used is cooling water since it is the only suitable utility available for such large temperature difference.

Assumptions

- E-4
  - Condenser duty from Aspen Plus V8.2 Report.
  - Single pass OD 1”, 5/8”fins at 10fins/in, Pitch of 2 ¼” triangular – used from last semester’s preliminary design requirements [4] for easier preliminary design calculations.
  - From given specs air inlet temperature was 105 °F from **Table 4**.
  - Tube length of 119 ft from GPSA example [5].
  - Fan motor efficiency 92% from GPSA example [5].
  - Motor size (bhp) 350 hp using Design notes from last semester [4].
  - Material properties such as specific heat capacity obtained from Aspen Plus V8.2.
- E-6
  - Reboiler duty from Aspen Plus V8.2 Report.
  - $T_{steam,in} = T_{steam,out} = 550$ °F (medium pressure steam condenses at constant temperature) **Table 4**.
  - Approach temperature of 20 °F from heuristic for steam reboiler approach temperature, from Gas Purification by Cole & Nielsen.
  - Heat transfer coefficient of 140 (Btu/hr ft^2 °F) - heuristic given in Dr. Aichele's class notes [4] on estimating U; U chosen for steam reboilers; lowest for conservative purposes.
  - Correction factor is 0.90 – heuristics Turton Table 11.11 [1].
  - Material properties such as specific heat capacity obtained from Aspen Plus V8.2.
- E-7
  - Cooler duty obtained from Aspen Plus V8.2 Report.
  - Shell side is the natural gas and utility cooling water is tube side for easier maintenance, cleaning purposes, and to prevent fouling.
  - Cooling water temperatures in and out specified as 87 °F and 120 °F respectively as specified in **Table 4**.
  - U = 80 (Btu/hr ft^2 °F) heuristic given in Dr. Aichele's class notes [4] on estimating U; U chosen for steam reboilers; lowest for conservative purposes.
  - Correction factor = 0.9 – heuristics Turton Table 11.11 [1].

- Material properties such as specific heat capacity obtained from Aspen Plus V8.2.

All three heat exchangers were designed with carbon steel as the material of construction.

Condensate Receiver

A drum is needed on (T-2) to collect the reflux that has been condensed. It acts as temporary storage where the reflux is then pumped back into the column for farther separation. The flowrate of the liquid at a specified resonance time determines the size of the drum.

Assumptions

- L/D ratio = 3 – Turton Table 11.6 heuristic 2.5 < L < 5 [1]
- T_{res} = 5min – Turton Table 11.6 heuristic # 5 [1]

By comparing the cost of a horizontal vs. vertical drum (Appendix *Condensate Receiver for De-Isobutanizer Tower D-2 Design)* it was determined that a horizontal condensate receiver was more economical.

Design

- Volume is 2,356 ft^3
- Diameter 10 ft
- Length 30 ft

3.3.3 Reactor (R-1)

The reactor is specified in the memorandum as a horizontal packed bed reactor. Inside of the reactor, a platinum catalyst is installed to aid in the conversion of the n-butane to isobutane.

Design

- Conversion of n-butane to isobutane is 48.7% - Appendix *Reactor R-1 Design.*
- 29.5 ft height.
- 6.25 ft diameter.
- Operates at 300 ^oF and 450 psig with a 20 psi pressure drop as determined by UOP in the memorandum.
- Platinum mesh catalyst.
- Jacketed with cooling water – added for safety purposes

The effect of different temperatures on the conversion and overall design were evaluated. The design temperature of 300 ^oF was determined to be the best (Appendix *Reactor R-1 Optimization)*.

Assumptions

- Stream is well mixed in the reactor section
- Sized and costed as a regular process vessel – specified by Dr. Ramsey in class
- Mesh design of catalyst provides sufficient surface area to catalyze reaction.
- Catalyst is purchased/renewed once every 5 years

Pumps

The feed to the reactor unit comes from the top of the de-butanizer column. This stream is not at a high enough pressure for the reactor and must therefore go through a pump first. One pump (P-5) has been installed to handle this task with a spare having been specified.

- P-5A Reactor feed pump. Discharge pressure is set at 450 psig on Aspen Plus V8.2.
- P-5B The spare reactor pump

Figure 12.3 in the GPSA [5] manual provided guidance on final pump selection design by utilizing the flowrates. The following assumptions were made:

Assumptions

- Both pumps and motor efficiencies were calculated using (p520-21 Peters and Timmerhaus, 1991) [3].
- All line loss pressure drops were assumed to be 5 psi.
- The pumps were sitting at ground level and pumping to the desired heights.
- Pressure drop across the control valve was 10 psia according to Dr. Ramsey's and Dr. Aichele's last semester project statements [4].

Heat Exchanger

When the stream leaves the de-butanizer column, it is not hot enough for the conditions that were specified for the reactor. Therefore, a heat exchanger (E-5) has been installed to heat up the reactor feed stream. Medium pressure steam specified in **Table 4** is used to heat up the process stream.

- E-5 Process stream heater. Medium pressure steam is used as the hot stream. Outlet temperature on process stream set at 300 ^{o}F.

Assumptions

- E-5
  - Heater duty from Aspen Plus V8.2 Report.
  - Medium pressure steam 550 °F (medium pressure steam condenses at constant temperature) **Table 4**.
  - Shell side is the steam, and tube side is the process stream.

- Approach temperature of 20 °F from heuristic for steam reboiler approach temperature, from Gas Purification by Cole & Nielsen.
- Heat transfer coefficient of 140 (Btu/hr ft^2 °F) - heuristic given in Dr. Aichele's class notes [4] on estimating U; U chosen for steam reboilers; lowest for conservative purposes.
- Correction factor is 0.90 – heuristics Turton Table 11.11 [1].
- Material properties such as specific heat capacity obtained from Aspen Plus V8.2.

The material of construction for all the equipment in the Butamer Unit is carbon steel. This is because carbon steel is the cheapest material of construction and is suitable for hydrocarbons (non-corrosive).

All the motors used in the Butamer Unit are explosion proof motors. They are the safest of all listed in Turton [1].

Label	Equipment	Purpose	Function	Size / Capacity	Operation Conditions
De-isobutanizer Tower					
P-1	Feed Pump	Prepare feed for de-isobutanizer column	Increase pressure of feed to inlet of column	Q = 197 gpm Design Pressure = 230 psia Power =15 hp	Raise pressure of feed from 114.7 psia to 160 psia
T-1	Column	Separate isobutane from feed stream	Remove reacted isobutane to the product stream	Height = 70 ft Diameter = 18 ft # of trays = 26 Pressure = 150 psia	Remove isobutane at 150 psia
P-2	Distillate Reflux Pump	Prepare distillate for reflux back into column	Increase pressure from distillate for re-entry into the column	Q = 2,360 gpm Design Pressure = 237 psia Power = 75 hp	Raise pressure of distillate reflux to 155 psia
E-1	Condenser (Air-Cooled)	Prepare reflux for re-entry into the column	Causes a phase change from vapor to liquid	Duty = 9.14×10^7 BTU/hr Bare Surface Area = 21,497 ft^2	Cools reflux from 154.5°F to 153.2°F at P_{in} at 150psia
D-1	Condensate Receiver	Provide temporary storage for reflux	Collect Condensate from condenser E-1 at specific T_{res}	Volume = 4,243 ft^3 Type = Horizontal T_{res} = 5 min Length = 36 ft	Process vessel holds 2,060.5ft^3 for a resonance time of 5min and a flowrate of 6.87ft^3/s
E-2	Kettle Reboiler	Prepare feed to re-enter the column	Increase Temperature of bottoms reflux liquid	Duty =7.32×10^7 BTU/hr Area = 1,529 ft^2	Increase Temperature of the Bottoms product from 169 °F to 171 °F at 153 psia

Table 16: *Equipment Summary (De-Isobutanizer)*

Label	Equipment	Purpose	Function	Size / Capacity	Operation Conditions
			De-butanizer Tower		
P-3	Feed Pump	Prepare feed for debutanizer column	Increase Pressure of feed to optimum entry	Q = 422 gpm Design Pressure = 257 psia Power = 25 hp	Raise Pressure of feed to debutanizer column from 150 psia to 175 psia
T-2	Column	Separate heavy natural gas from rest of feed	Remove natural gas in product stream to storage	Height = 78 ft Diameter = 14 ft # of trays = 29 Pressure = 151 psia	Remove natural gas at 151 psia
P-4	Distillate Reflux Pump	Prepare distillate reflux for re-entry into column	Increase pressure from distillate for entry back into column	Q = 521 gpm Design Pressure = 251 psia Power = 25 hp	Raise pressure of distillate reflux to 155 psia from 151 psia
D-2	Condensate Receiver	Provide temporary storage for reflux	Collect Condensate from condenser at specific T_{res}	Volume = 2,356 ft^3 Type = Horizontal T_{res} = 5 min Length = 30 ft	Process vessel holds 1,109 ft^3 for a resonance time of 5min and a flowrate of 3.7 ft^3/s
E-4	Condenser (Air-Cooled)	Prepare reflux for re-entry into the column	Causes a phase change from vapor to liquid	Duty = $5.2x10^7$ BTU/hr Area = 10,273 ft^2	Cools reflux from 171.5°F to 170.1°F at 151 psia
E-6	Kettle Reboiler	Prepare feed to re-enter the column	Increase temperature of bottoms reflux liquid	Duty = $5.2x10^7$ BTU/hr Area = 1,339 ft^2	Increase Temperature of the Bottoms product from 238.7°F to 240.7°F at 154.4 psia

Table 17: *Equipment Summary (De-Butanizer)*

Label	Equipment	Purpose	Function	Size / Capacity	Operation Conditions
			Reactor		
P-5	Reactor Feed Pump	Prepare feed for entry into the reactor	Increase the pressure of the feed to the heat exchanger	Q = 413 gpm Design Pressure = 534 psia Power = 125 hp	Raise pressure of feed to the reactor from 151 psia to 464.7 psia
E-5	Reactor Feed Heat Exchanger (Heater)	Prepare feed to enter the reactor	Increase temperature of the feed stream entering the reactor	Duty = 2.02×10^7 BTU/hr Area = 520 ft^2	Increase Temperature of the reactor feed form 175°F to 300°F at 444.7 psia
R-1	Reactor (Packed Bed)	Prepare feed for recycle	React feed using a catalyst to send back to the de-isobutanizer column	Height = 29.5 ft Diameter = 6.25 ft Volume = 900 ft^3	Converts using catalyst 48.7% with a P_{drop} of 20 psi at 430 psia and temperature of 300°F
			Product Coolers		
E-7	Natural Gas Heat Exchanger (Cooler)	Prepare product to enter Natural Gas storage	Reduces the temperature of the product leaving the de-butanizer column	Duty = 2.62×10^5 BTU/hr Area = 80 ft^2	Cools product from the de-butanizer column from 240.7 °F to 98 °F at 151 psia
E-3	Isobutane Heat Exchanger (Cooler)	Prepare product to enter Isobutane storage	Reduces the temperature of the product leaving the de-isobutanizer column	Duty = 1.05×10^6 BTU/hr Area = 394 ft^2	Cools product from the de-isobutanizer column from 153.82 °F to 128 °F at 150 psia

Table 18: *Equipment Summary (Reactor & Product Coolers)*

3.4 Safety

3.4.1 Equipment and Process Safety

This process does not involve any extreme operating conditions. The highest temperature at any point of the process is 400 ^{o}F and the highest pressure is 450 psig. Both of these conditions occur within the reactor; therefore, it is assumed that the reactor needs the greatest amount of safety consideration. One of the larger concerns involved in this process is that all fluids in this process are flammable and could potentially pose an explosion risk if released as a vapor. The first step to analyze the existing inherent safety measures and any additional ones that are needed was to perform a potential consequence analysis. The potential consequence analysis was documented in **Table 22**.

Based on the results of the consequence analysis, it is determined that there were a few safeguards that needed to be included in the preliminary design. First, the reactor should be jacketed to help prevent a runaway reaction with the slightly exothermic chemistry. The reaction that converts n-butane to isobutane inside the reactor is exothermic and will cause the temperature of the fluid to increase as more material is reacted. It will be important to be able to keep the reactor temperature constant in the event that production is ever increased or there is an accidental increase in the feed rate to the reactor. It will also be necessary to have a thermometer near the exit from the reactor and also control valves which allow the available area of the reactor jacket to change. Since the reactor is a length of pipe for this system (indicated in the memorandum as a horizontal packed bed reactor), the jacket can be constructed as a series of segments each with their own line for the cooling water to leave. The segments can be connected by control valves which increase the available surface area for cooling when open and decrease the surface area when closed. In addition, a pressure gauge and pressure relief valve should be included as a second level of safety.

Both distillation columns involved in this process will need the same safety measures in place. It will be important to have pressure gauges and pressure relief valves at the top of each column. A flooded condenser system can be used with a pressure control system to help keep a constant pressure in the column during any variations in the feed rate to the columns.

The heat exchanger prior to the reactor involves a high pressure steam line that should be insulated to prevent operators from burning themselves. In addition, a temperature control system should be used with a control valve on the steam line to maintain a constant outlet temperature of the process line.

In more detailed versions of the design it is recommended that a flare system be considered in for use in the event of emergencies. Since all fluids involved are flammable, they can be easily combusted in a flare to prevent a hazardous release to the environment. Since petroleum based fluids are used in this process, it is important to minimize any releases to the environment as these fluids can be hazardous to plant and animal life.

None of the fluids involved are corrosive. The operating conditions are such that carbon steel can be used for the design of all process equipment and piping. The current design does not account for site storage of feed or products. It is recommended that future designs add feed storage for a three month supply. This would be 725,000 barrels of storage. In addition one month of product storage should be considered. This would be 237,000 barrels for isobutane and 10,000 barrels for natural gas.

The safety measurements mentioned above will make the overall process inherently safer. There are measures in place around all major pieces of equipment to help prevent releases or hazardous operating conditions. Since the entire process is operated at standard conditions an additional safety layer is provided.

This process is necessary in the winter months when addition of isobutane to diesel fuels is required. Therefore, it is assumed that this Butamer Unit will be shut down during summer months. It is recommended that the platinum catalyst in the reactor be replaced during such times. It is estimated that the catalyst can be used for five years without replacement; therefore, it may last through four successive winter cycles. The cost of replacing it during the summer is minimal compared to the cost that would be incurred if it failed during operation. In addition to this measure, it is recommended that a preventative maintenance schedule be developed for all seals and potential sources of leaks. All seals should be replaced during the summer months and the entire system should be checked for any signs of deterioration or weaknesses. Loss of containment which results in an explosive vapor cloud forming is the most significant hazard involved with this process. These steps will significantly reduce the risks and are highly recommended.

3.4.2 Properties of Material Table

	Propane	1-Butene	Water	i-Butene	Tr2-Butene	Cis2-Butene	i-Butane
Chemical Formula	C_3H_8	C_4H_8	H_2O	C_4H_8	C_4H_8	C_4H_8	C_4H_{10}
Appearance	colorless	colorless	Colorless	Colorless	Colorless	colorless	colorless
Physical state	liquified gas	liquified gas	Liquid	liquified gas	gas	liquified gas	liquified gas
Molecular Weight (g/mole)	44.1	56.11	18.02	56.11	56.11	56.11	58.12
Odor	odorless	slightly aromatic	Odorless	characteristic	slight	slight	odorless
Auto Ignition Temp (°F)	1,004	824	1	869	615.2	617	860
Boiling Point (°F) at 1 atm	-43.8	43.3 °F)	212	19.6	33.8	38.7	-10
Melting Point (°F)	-306	-301 °F	32	-221.3	-157	-218	-256
Vapor pressure (psi)	190	63	0.334	24.3	15	27.3	72.2
Vapor density	1.52	1.94	0.62	1.94	1.9	1.94	2.01
% volatility (upper)	10%	10%	NA	10%	10%	10%	8%
% volatility (lower)	2%	2%	NA	2%	2%	2%	1%
Flash point (F)	-155	-112 °F	NA	-105	-98	10.42	-117
Classification	Flam Gas, Pressure Gas	Flam Gas, Pres Gas	NA	Flam Gas, Pres Gas	Flam Gas, Pres Gas	Flam Gas, Pres Gas	Flam Gas, Pres Gas
Salability in water (g/l)	0.04	2.21*E-6	NA	Insoluble	NA	NA	NA

Table 19: *Properties of Material Table*

	N-Butane	2,2 Dimethyl Propane	I-Pentane	N-Pentane	N-Hexane	platinum
Chemical Formula	C_4H_{10}	C_5H_{12}	C_5H_{12}	C_5H_{12}	C_6H_{14}	Pt
Appearance	colorless	colorless	Colorless	colorless	Colorless	metal
Physical state	gas	gas	liquid	liquid	liquid	wire
Molecular Weight (g/mole)	58.12	72.17	72.15	72.15	86.18	195.08
Odor	odorless		odorless	odorless	petrolic	NA
Auto Ignition Temp (°F)	550	842	788	500	453.2	
Boiling Point (°F) at 1 atm	31.1	192.2	81.9	95	56	6,921
Melting Point (°F)	-216		-258	-202	-139	3,222
Vapor pressure (psi)	35.2	21.7	11.2	8.4	4.95	NA
Vapor density	2.33	0.2001	2.49	0.626 (liquid)	0.659 (liquid)	NA
% volatiles (upper)	8%	8%	8%	8%	8%	NA
% volatilies (lower)	2%	1%	1%	1.4	1%	NA
Flash point (F)	-76	19.4	-60	-56.2	-14.78	NA
Classification	Flam Gas, Pres Gas	Flam Gas, Pres Gas	Flam Liquid	Flam Liquid, Toxic	Flam Gas, Pres Gas	catalyst
Salability in water (g/l)	0.061	NA	NA	NA	Insoluble	

Table 20: *Properties of Materials Table (Continued)*

The Properties of Materials used in the Butamer Unit are summarized in **Tables 19 & 20**. The information for these tables are obtained from SDS sheets in the Appendix (A-22 to A-235).

3.4.3 Reactivity Table

Chart Key	
No Reactivity	
Potential Reactivity	Self Polymerizable

	Propane	1-butene	Water	i-butene	Tr2-Butene	Cis2-Butene	I-Butane	N-Butane	2,3 Dimethyl Propane	I-Pentane	N-Pentane	N-Hexane
Propane		Self Polymerizable										
1-butene	Self Polymerizable			Self Polymerizable								
Water												
i-butene		Self Polymerizable			Self Polymerizable	Self Polymerizable	Self Polymerizable	Self Polymerizable	Self Polymerizable	Self Polymerizable	Self Polymerizable	Self Polymerizable
Tr2-Butene				Self Polymerizable								
Cis2-Butene				Self Polymerizable								
i-butane				Self Polymerizable								
N-Butane				Self Polymerizable								
2,3-Dimethylpropane				Self Polymerizable								
i-Pentane				Self Polymerizable								
N-Pentane				Self Polymerizable								
N-Hexane				Self Polymerizable								

Table 21: *Reactivity Matrix*

This reactivity table is generated with the help of CAMEO interactive matrix online. There are no potential hazards except that propane when mixed with 1-butene is self Polymerizable. 1-Butene mixed with i-butene is self-Polymerizable. I-Butene mixed with trans-2-butene and the rest of the chemicals is self-Polymerizable.

3.4.4 Potential Consequences Summary

A summary of potential consequences are given below.

	Hazard	Equipment Damage	Environmental Compliance	Loss of Life	Disruption of Other Business Units	Legal/PR	Community Impact
1	Loss of liquid feed containment	Low	High	Low	Low	High	High
2	Overpressure either column	Medium	Low	Low	Low	Low	Low
3	Vapor leak (explosion)	High	Low	High	High	High	High
4	Runaway reaction	High	Low	High	High	High	High

Table 22: *Potential Consequence Summary*

3.4.5 Existing Safeguards as Designed

The following are the existing safeguards as designed.

	Safeguard	**Purpose**
1	Jacketed Reactor	Prevent runaway reaction due to exothermic reaction.
2	Product Coolers	Product is cooled to temperatures that reduce risk during storage.
3	Normal operating pressures and temperatures.	Standard materials of construction may be used and there are no special risks created due to the usage of extreme pressures or temperatures.

Table 23: *Existing Safeguards as Designed*

3.4.6 Additional Safeguard Evaluations and Recommendations

- Consider addition of emergency shutdown control system. System should be focused on extremely high pressures or temperatures occurring in the columns or reactor.

3.4.7 Safety Assessment Summary

Potential for Project Termination:

- All of the fluids utilized in this process are extremely flammable and potentially explosive at the operating conditions of the system. If risk is excessive or cannot be managed, termination should be considered.

Major Concerns Requiring Significant Attention:

- Seals and potential sources of leaks should be inspected often and replaced or repaired as soon as signs of deterioration appear. Almost all parts of the process would release an explosive vapor if a leak occurred.
- When the platinum catalyst in the reactor needs replacing, the reactor needs to be purged of all fluids and detection equipment should be worn to ensure there are no lingering vapors.

Special Concerns – PSM related:

- Any leak would pose a significant risk of vapor cloud formation and subsequent explosion. This would lead to significant life loss and damage to the equipment. There is also a significant risk of fire if fluids are leaked.

Special Concerns – RMP related:

- The feed and the products are being transported off site. Any significant loss of containment during the transportation of these fluids can lead to a significant environmental damage and health risks from toxic vapors, explosions, or fires. Any loss of containment within the facility that led to an explosion can affect the nearby area and lead to damage, injuries, or possible deaths.

Overall Safety Characterization:

- While the effects of a loss of containment would likely be devastating, the risk of a loss of containment occurring is low. Relatively safe operating conditions have been designed, and systems are in place to prevent a runaway reaction. In addition, a control system (**Figure 3**) is established with the intent of preventing any abnormal temperatures or pressures. A preventative maintenance schedule would significantly reduce all other risks of leaks.

3.4.8 Inherently Safer Design Application Summary

Concept	How Incorporated in Preliminary Design
Minimize	Maximized conversion in reactor to reduce amount of recycle needed. Incorporated catalyst to reactor design to increase reaction rate. Reaction occurs continuously instead of in batches to reduce volume in the system.
Substitute	Lowest usable pressure steam is incorporated into all heating applications rather than highest available steam. No additional implementation is necessary. Fluids and catalyst type involved in system were specified prior to design and cannot be changed.
Moderate	All product streams are cooled and pressure is reduced to lead to safer storing conditions. All pressures and temperatures used in the process are as close to ambient as possible.
Simplify	Process involves three main sections. Each section is required for operation and no additional processing sections have been added. Some more complex designs led to slightly cheaper development however the added complexities led to significant additional hazards. Therefore the simplest design that achieved the goals is utilized.

Table 24: *Inherently Safer Design Application Summary*

3.4.9 Opportunities for Additional ISD in Detailed Design

- Consider ways to increase mixing capabilities inside the reactor. Different reactor designs may allow for increased conversion without changing the operating temperature and pressure. This would reduce the overall recycle rate and thus minimize the total volume of fluid in the system at any given time.
- Investigate addition of onsite storage of the process feed. This would reduce the distance the fluid is pumped to the system when in use. Less pumping distance would likely reduce incoming flow fluctuations which would help minimize risk of the system operating outside of designed temperatures and pressures.

3.5 Environmental Analysis

The coolers use cooling water as their utilities. It is important that the cooling water be returned at a safe temperature for the environment. Because of this limitation, the design is established in a manner that allows all the water to return at safe temperatures (120 °F). This is done by simply using more cooling water and increasing the heat exchanger surface area.

The feed and all the products are natural gas related fluids. Any loss of containment of liquids that occur off site would lead to significant environmental harm. Environmental consequences including harm to the surrounding wildlife can cause affected areas to be rendered uninhabitable. Oil spills have become a major focus of the general public and the EPA. Care should be taken to minimize any potential for spills.

There are no hazardous wastes or dangerous emissions associated with this process. No additional special permitting is required.

4. Economic Analysis

The economic analysis is performed using a five year project evaluation life and a hurdle rate of 15% as given in the memorandum. The equipment is depreciated using a 10-year MACRS depreciation method. The effective tax rate of 40% is also applied. The project has a net present value of $19MM and a future worth revenue of $128MM.

4.1 Capital Cost Estimates

The capital cost of equipment is estimated using the methods described in Turton [1]. The capital cost is calculated according to the following Equation (1):

$$\log_{10} C_p^0 = K_1 + K_2 \log_{10}(A) + K_3\left[\log_{10}(A)\right]^2 \tag{1}$$

Where:

- C_p^0 ≡ Equipment purchased cost at ambient operating pressure using carbon steel construction
- K_N ≡ Constant values found in Table A.1, Turton [1], classified by equipment type and description
- A ≡ Equipment size or capacity parameter

This correlation is valid only if the equipment capacities fall within the minimum and maximum capacity ranges tabulated in Table A.1, Turton [1]. The capacity and the units vary with the type of equipment. In the case where the equipment is not operating at ambient pressure, a pressure factor, $F_{P,vessel}$, is calculated using Equation (2):

$$F_{P,vessel} = \frac{\dfrac{(P+1)D}{D[850-0.6(P+1)]} + 0.00315}{0.0063} \tag{2}$$

Where:

- P ≡ Design pressure, barg
- D ≡ Diameter, m

This equation applies for t_{vessel} > 0.0063 m. If $F_{P,vessel}$ < 1, use $F_{P,vessel}$ = 1. This particular pressure factor only applies to process vessels. Process vessels include distillation towers, reactor, condensate receivers, and flash drums.

Equipment such as pumps, heat exchangers, and other equipment utilize another pressure factor correlation, F_P, shown in Equation (3) from Turton [1]:

$$\log_{10} F_P = C_1 + C_2 \log_{10} P + C_3(\log_{10} P)^2 \tag{3}$$

Where:

C_N ≡ Constant values found in Table A.2, Turton [1], classified by equipment type and description

P ≡ Design pressure, barg

The bare module cost, C_{BM}, is the original cost that includes correction factors for purchased price due to a variation in assumptions for different equipment. The bare module cost is applied for sieve trays, fans, and drives and can be found using Equation (4) from Turton [1]:

$$C_{BM} = C_P^0 F_{BM} \tag{4}$$

Where:

F_{BM} ≡ Bare module factor found in Figure A.19, Turton [1], using equipment type and identification number based on material of construction in Table A.6

The bare module cost for heat exchangers, process vessels, and pumps can be found using Equation (5) from Turton [1]:

$$C_{BM} = C_P^0 (B_1 + B_2 F_M F_P) \tag{5}$$

Where:

B_N ≡ Constant values found in Table A.4, Turton [1]

F_M ≡ Factor found in Figure A.18, Turton [1], using equipment type and identification number based on material of construction in Table A.3

F_P ≡ Calculated using Equations (2) or (3)

The Chemical Engineering Plant Cost Index (CEPCI) is used to justify changes that result from inflation. The CEPCI is based on equipment, machinery, supports, installation labor, buildings, materials, engineering, and supervision. Turton [1] utilizes 2001 CEPCI. A 2016 CEPCI value is necessary since it is the year of construction for the purposes of this project. The bare module cost needs to be adjusted to the future value of 2016 from 2001 using Equation (6), Turton [1]:

$$C_{BM,2016} = \frac{CEPCI_{2016}}{CEPCI_{2001}}(C_{BM}) \tag{6}$$

Where:

$CEPCI_{2016} \equiv 569.8$

$CEPCI_{2001} \equiv 397$

Since $CEPCI_{2016}$ is not published, the past CEPCI values can be utilized from Chemical Engineering Economic Indicators [2] journal to extrapolate the 2016 CEPCI value on Microsoft Excel. This method is included in the Appendix (*CEPCI)* and yields 569.8.

Carbon steel is the material of construction for all equipment in this project design. The reasons for this being the chemicals involved were hydrocarbons. It is also the cheapest option. The capital cost of this project includes spare pumps for replacement. The bare module costs of all the equipment are calculated for 2016 and are summarized in **Table 25**. A contingency rate of 18% [1] is applied to account for the cost as a protection against faulty information and oversights. It is added to the total bare module cost for a total module cost. The detailed costing calculations for each piece of equipment are included in the *Sample Calculations* section of the Appendix.

Summary of Equipment Capital Costs		
Butamer Reactor	R-1	($2,538,609)
De-Isobutanizer Tower	T-1	($7,952,777)
De-Butanizer Tower	T-2	($4,428,542)
Condensate Receiver 1	D-1	($607,449)
Condensate Receiver 2	D-2	($345,325)
De-Isobutanizer Reboiler	E-2	($696,397)
De-Butanizer Reboiler	E-6	($597,676)
Butamer Reactor Heater	E-5	($239,504)
Isobutane Cooler	E-3	($91,199)
De-Isobutanizer Feed Pump	P-1	($39,196)
De-Isobutanizer Reflux Pump	P-2	($85,007)
De-Butanizer Feed Pump	P-3	($48,927)
De-Butanizer Reflux Pump	P-4	($48,824)
Butamer Reactor Feed Pump	P-5	($138,899)
De-Isobutanizer Air-Cooler	E-1	($913,229)
De-Butanizer Air-Cooler	E-4	($609,237)
Natural Gas Cooler	E-7	($102,785)
Total		**($19,380,797)**
Total with 18% Contingency Rate		**($22,869,341)**

Table 25: *Summary of Equipment Capital Costs*

4.2 Revenue and Operating Expense Estimates

The production and sales of isobutane and natural gasoline products from the Butamer Unit is accounted for in the revenue. The costs associated with the project include the raw material costs, capital costs for each piece of equipment, a one-time purchase of the platinum catalyst for the given five year project evaluation life, and the operating costs. The transportation of the mixed butanes from Conway, KS to Ponca City and the transportation of the isobutane from Ponca City to the customer in Mt. Belleview are accounted as costs in the economic analysis calculations.

The yearly operating costs are calculated for equipment that require utilities such as electricity, air, steam, and cooling water. To estimate the electricity costs for a motor of a pump or an air-cooled heat exchanger, the head and the hydraulic horsepower are calculated using Equations (7) and (8) respectively.

$$Head(ft) = \frac{2.31 \cdot \Delta P(psi)}{spg} \tag{7}$$

Where:

ΔP ≡ Discharge pressure – suction pressure; account for frictional losses, psi
spg ≡ Specific gravity of the fluid

$$Hydraulichp = \frac{Q(gpm) \cdot H(ft) \cdot spg}{3960} = \frac{Q(gpm) \cdot \Delta P(psi)}{1715} \tag{8}$$

Where:

Q ≡ Flow rate pumped, gpm

The pump and motor efficiencies are estimated using Figures 14-37 and 14-38 from Plant Design and Economics [3] and are used to determine the actual brake horsepower needed and purchased. Brake horsepower needed is calculated using Equation (9):

$$Brakehp(bhp) = \frac{Hydraulichp}{\eta_{pump}} \tag{9}$$

Where:

η_{pump} ≡ Pump efficiency found using Q (gpm) in Figure 14-37 [3]

Purchased horsepower is calculated using Equation (10) after selecting a motor with a bhp greater than required bhp:

$$Purchasedhp = \frac{bhp}{\eta_{motor}} \tag{10}$$

Where:

η_{motor} ≡ Motor efficiency found using Q (gpm) in Figure 14-38 [3]

The purchased hp is converted to kilowatts and is manipulated according to Equation (11) to find the operational cost per year:

$$\frac{OperatingCost}{year} = Purchased(kW)X\frac{365days}{year}X\frac{24hours}{day}X\frac{\$0.07}{kWh}X0.96 \tag{11}$$

The service factor, 0.96, was estimated by Dr. Ramsey and Dr. Aichele in Fall 2014 for Design Project 1 [4]. The service factor assumes 8409.6 operating hours per year. The remaining number of hours account for required downtime for maintenance, cleaning equipment, and repairing instruments, etc. It also accounts for unexpected outages because of off-spec production, equipment trips, and other environmental constraints. The calculations are shown in the Appendix *Service Factor Calculations*. The pumps will be running continuously 365 days/year and 24 hours/day. The electricity utility cost is provided in Table 3. The operational costs to run the fans for the air-cooled heat exchangers (E-1 and E-4) are performed using the similar method as the pumps. However, the efficiency of fan motor is used from the GPSA [5] manual, instead of the pump efficiency, to determine the required bhp.

To determine the operating cost of the kettle reboilers (E-2 and E-6) and the shell-and-tube heat exchanger (E-5), the amount of medium pressure steam (550 °F at 175 psig from **Table 5**) needed is estimated using Equation (12):

$$\dot{m}(lb/hr) = \frac{Q(Btu/hr)}{\lambda_{steam}} \tag{12}$$

Where:

Q ≡ Kettle reboiler duty provided by Aspen (Appendix Report A-2), Btu/hr

λ_{steam} ≡ 970.4 Btu/lbm latent heat of vaporization from Engineering Toolbox [6]

The mass flow rate is converted to a yearly basis and multiplied by 365 days/year, 24 hours/day, and a service factor of 0.96 for maintenance since the equipment are expected to be operation year round. This value is multiplied by \$3.30/1000 lbm medium pressure steam cost as specified in **Table 3** to find the total operating cost per year.

The operating cost of shell-and-tube heat exchangers (E-3 and E-7) that utilize cooling water can be calculated by finding the mass flow rate of cooling water required from the heat exchanger design Equation (13):

$$\dot{m}(lb/hr) = \frac{Q(Btu/hr)}{C_p(Btu/lb\cdot°F)\cdot\Delta T(°F)} \tag{13}$$

Where:

Q ≡ Cooler duty provided by Aspen, Btu/hr

C_p ≡ Specific heat capacity of cooling water from Engineering ToolBox, Btu/lbm °F

ΔT ≡ Change in temperature of cooling water with supply and return temperatures specified in **Table 4**

The mass flow rate of water is converted to gpm and multiplied by 365 days/year, 24 hours/day, 60 minutes/hour, and a service factor of 0.96 for maintenance since the equipment are expected to be operation year round. This value is multiplied by $0.000228/gpm ($120/ annual gpm) cooling water cost as specified in **Table 3** to find the total operating cost per year. The utility consumption summary is provided below in **Table 26**.

Yearly Utility Consumption

Equipment		**Electricity (kW)/year**	**Steam (lb/year)**	**Cooling Water (lb/yr)**	**Air (lb/yr)**
De-Isobutanizer Air-Cooler	**E-1**	5619210			120718708813
De-Isobutanizer Reboiler	**E-2**		661166055		
Isobutane Cooler	**E-3**			278636091	
De-Butanizer Air-Cooler	**E-4**	2458405			51207736628
Butamer Reactor Heater	**E-5**		182297903		
De-Butanizer Reboiler	**E-6**		472575099		
Natural Gas Cooler	**E-7**			69756927	
De-Isobutanizer Feed Pump	**P-1**	112626			
De-Isobutanizer Reflux Pump	**P-2**	544361			
De-Butanizer Feed Pump	**P-3**	187711			
De-Butanizer Reflux Pump	**P-4**	186638			
Butamer Reactor Feed Pump	**P-5**	887545			

Table 26: *Utility Consumption on a Yearly Basis*

The utility consumption or operational costs for each piece of equipment are summarized in **Table 27** below. These calculations include the 0.96 service factor to account for maintenance for the year.

Yearly Utility Cost

Equipment		Electricity ($/year)	Steam ($/year)	Cooling Water ($/yr)	Air ($/yr)
De-Isobutanizer Air-Cooler	**E-1**	$393,345			Free
De-Isobutanizer Reboiler	**E-2**		$2,181,848		
Isobutane Cooler	**E-3**			$66,784,454	
De-Butanizer Air-Cooler	**E-4**	$172,088			Free
Butamer Reactor Heater	**E-5**		$601,583		
De-Butanizer Reboiler	**E-6**		$1,559,498		
Natural Gas Cooler	**E-7**			$16,719,579	
De-Isobutanizer Feed Pump	**P-1**	$7,884			
De-Isobutanizer Reflux Pump	**P-2**	$38,105			
De-Butanizer Feed Pump	**P-3**	$13,140			
De-Butanizer Reflux Pump	**P-4**	$13,065			
Butamer Reactor Feed Pump	**P-5**	$62,128			

Table 27: *Utility Costs on a Yearly Basis*

The operating labor costs are estimated using Turton Equation 8.3 [1]. The number of operators per shift can be calculated using Equation (14):

$$N_{OL} = (6.29 + 31.7P^2 + 0.23N_{np})^{0.5} \tag{14}$$

Where:

P ≡ 0; No number of particulate solids processing steps

N_{np} ≡ Sum of the equipment requiring operators

Operating Labor Requirements		
Equipment Type	**Number of Equipment**	**N_{np}**
Reactor	1	1
Distillation Columns	2	2
Condensate Receivers	2	-
Heat Exchangers	7	7
Pumps	5	-
Total		10
Number of Operators Required Per Shift, N_{OL}		2.93087
Operating Labor		14
Operator Salary [7]		$43,340
Labor Costs (2016)		**$606,760**

Table 28: *Estimation of Operating Labor Requirements*

Operating labor cost is 4.5 operators x N_{OL} as given in Turton [1]. A chemical plant operator median salary as of February 2015 is $43,340 per year [7]. The labor costs are operating labor multiplied by the operator salary.

There are no waste products in this process. Therefore, a waste treatment plan is not necessary. The only by-product being produced is natural gas and it will be sold in Oklahoma.

For safety purposes, the capital cost already involves spare pumps in case of emergency if a pump becomes dysfunctional. The estimated line losses while designing the pumps are much greater than the actual line losses due to the flow rates compared to the CRANE manual [8]. Therefore, the pumps have been oversized and the utility or operating costs are greater for estimation purposes than what they actually need to be. The reflux pumps (P-2 and P-4) have been oversized by 10% using a Turton [1] heuristic for towers. Due to the exothermic nature of the reaction in the horizontal packed bed reactor, it is safer to include a cooling water jacket. The operating cost for the cooling water jacket is not included in the economic analysis because it is not included in any of the class references. However, it must be included for detailed design. The manufacturing costs are out of scope of the analysis of this project. However, manufacturing estimates guided by Turton [1] such as direct supervisory and clerical labor, maintenance and repairs, operating supplies, laboratory charges, and total fixed and general manufacturing costs are included in **Table 43** of the Appendix to get a general estimate of what the NPV and DCFROR would be if all of these factors are taken into account. The fixed manufacturing costs take into account local taxes and insurance and plant overhead costs. The total general manufacturing costs take into account are: the administration costs, distribution and selling costs, and research and development.

The escalation of raw material purchase and transportation costs, sales and transportation costs of isobutane, and sales costs of natural gasoline are not accounted for in the economic analysis of this project. The operating costs may also escalate in the future years. The economic analysis is performed on the basis of washout assumption. A washout assumption assumes that the "escalation of the operating costs is offset by an equal increase in revenues producing uniform profit margins each year" [4].

4.3 DCFROR Analysis

As indicated in the memorandum, the economic analysis of this project involves a five year project evaluation life, 15% hurdle rate, and an effective tax rate of 40%. A 10-year MACRS depreciation method with a half year convention is chosen for this project because it is a worst case scenario. Using the MACRS method over a large period of time will make the overall NPV and DCFROR smaller because it slows the depreciation, and thus more taxes are paid. However, a 5-year and a 7-year MACRS depreciation methods are additionally performed for reference. These economic analyses are **Tables 44 & 45** of the Appendix.

The key parameters influencing the Net Present Value (NPV) and Discounted Cash Flow Rate of Return (DCFROR) are capital costs, raw material costs, sales costs, and operating costs. Sales costs of the isobutane and natural gasoline have the most impact among all of the parameters. This is because the products have a greater value and demand than the raw materials in the industry. Sales revenue is $56MM in 2017 and $117MM from 2018 to 2021. Increasing the production of isobutane and natural gasoline will significantly increase the NPV and DCFROR, making the project more economically favorable in the long run. Raw materials costs have the second most impact on the NPV and DCFROR because of their purchase price and transportation. Raw materials costs are $43MM in 2017 and $85MM from 2018 to 2021. Making the process more efficient will decrease the amount of raw materials required and will boost the NPV and DCFROR. A sensitivity analysis is run in the next section to see the impact of these parameters on the project estimation. A cash flow table with a 10-year MACRS depreciation method is presented below in **Table 29**. This is a revenue producing project. The undiscounted payback period for this project is 2.14 years as calculated below:

$$1+\frac{(\$22,869,341-\$6,644,308)}{\$14,276,961}=2.14yrs \tag{15}$$

The discounted payback period for this project is 2.69 years as calculated below:

$$2+\frac{(\$22,869,341-\$5,777,659-\$10,795,434)}{\$9,170,801}=2.69yrs \tag{16}$$

4.3.1 Cash Flow Table

Year	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**
End of Year	0	1	2	3	4	5
Production Natural Gas, "BBL/yr"	0	59227.63033	118455.2607	118455.2607	118455.2607	118455.2607
Production Iso-butane, "BBL/yr"	0	1,394,743	2,789,486	2,789,486	2,789,486	2,789,486
Natural Gas Sales Price, $/"BBL"	0	$55.85	$55.85	$55.85	$55.85	$55.85
Iso-butane Sales Price, $/"BBL"	0	$37.67	$37.67	$37.67	$37.67	$37.67
Sales Revenue	0	$55,847,838	$111,695,676	$111,695,676	$111,695,676	$111,695,676
Mixed Butanes, "BBL/yr"	0	1,423,500	2,847,000	2,847,000	2,847,000	2,847,000
Mixed Butanes Transportation, "$/BBL"	0	($1.10)	($1.10)	($1.10)	($1.10)	($1.10)
Isobutane Transportation, "$/BBL"	0	($1.35)	($1.35)	($1.35)	($1.35)	($1.35)
Mixed Butanes Purchased, "$/BBL"	0	($27.50)	($27.50)	($27.50)	($27.50)	($27.50)
— Mixed Butanes Purchased, "$/BBL"	0	($40,712,100)	($81,424,200)	($81,424,200)	($81,424,200)	($81,424,200)
— Isobutane Transportation, "$/BBL"	0	($1,882,903)	($3,765,807)	($3,765,807)	($3,765,807)	($3,765,807)
Net Revenue	0	$13,252,835	$26,505,670	$26,505,670	$26,505,670	$26,505,670
— Other Op Costs	0	($2,424,148)	($4,848,295)	($4,848,295)	($4,848,295)	($4,848,295)
— Depreciation		($2,286,934)	($4,116,481)	($3,293,185)	($2,634,548)	($2,108,553)
— Operators		($303,380)	($606,760)	($606,760)	($606,760)	($606,760)
— Write off						($8,429,639)
Taxable Income	0	$8,238,373	$16,934,133	$17,757,429	$18,416,066	$10,512,422
— Tax @ 40%	0	($3,295,349)	($6,773,653)	($7,102,972)	($7,366,427)	($4,204,969)
Net Income	0	$4,943,024	$10,160,480	$10,654,458	$11,049,640	$6,307,453
+ Depreciation		$2,286,934	$4,116,481	$3,293,185	$2,634,548	$2,108,553
+ Write off						$8,429,639
— Catalyst		($585,650)				
— Fixed Capital	($22,869,341)					
Cash Flow	($22,869,341)	$6,644,308	$14,276,961	$13,947,643	$13,684,188	$16,845,646
Discount Factor ($P/F_{i^*,n}$)	1.0000	0.8696	0.7561	0.6575	0.5718	0.4972
Discounted Cash Flow	($22,869,341)	$5,777,659	$10,795,434	$9,170,801	$7,823,979	$8,375,263

Discount Factor ($F/P_{i*,n}$)	3.059022863	2.66001988	2.313060766	2.011357188	1.74900625	1.520875
NPV @ i* =	**$19,073,796**	The project is economically attractive since NPV > 0				
DCFROR =	**42%**	The project is economically attractive since DCFROR > i*				
PWNR @ i*	**$41,943,136**					
PWNC @ i*	**($22,869,341)**					
BCR =	**1.834033468**	The project is economically attractive since BCR > 1				
FWR	**$128,305,013**					
GROR =	**41.19%**	7.99361E-15	This project is economically attractive since GROR > i*			

Table 29: *Cash Flow Table*

4.4 Sensitivity Analysis

A sensitivity analysis is performed on the sales, raw material costs, capital costs, and operating costs. These are the key parameters that have the most impact on the NPV and DCFROR as determined from the cash flow **Table 29**. They have been chosen based on the greatest prices involved in the evaluation of the project. These parameters are varied by ± 40% because the Turton [1] capital and operating costs methods are ± 40% off from the real estimates used in the industry. The following table summarizes the high and low values of these key parameters for DCFROR in **Table 30**. The best case scenario occurs at 125% DCFROR for sales and worst case scenario occurs at 35% DCFROR loss in sales.

Sensitivity Analysis for DCFROR		
	Low	**High**
Sales	-35%	125%
Raw Material Costs	-10%	107%
Capital Costs	28%	70%
Operating Costs	37%	47%

Table 30: *Summary of Sensitivity Analysis for DCFROR*

These values are assembled into a tornado chart, **Figure 4**, below. It compares each of the key parameters and their impact on the DCFROR of the project. The base case for this analysis is a DCFROR value of 42%.

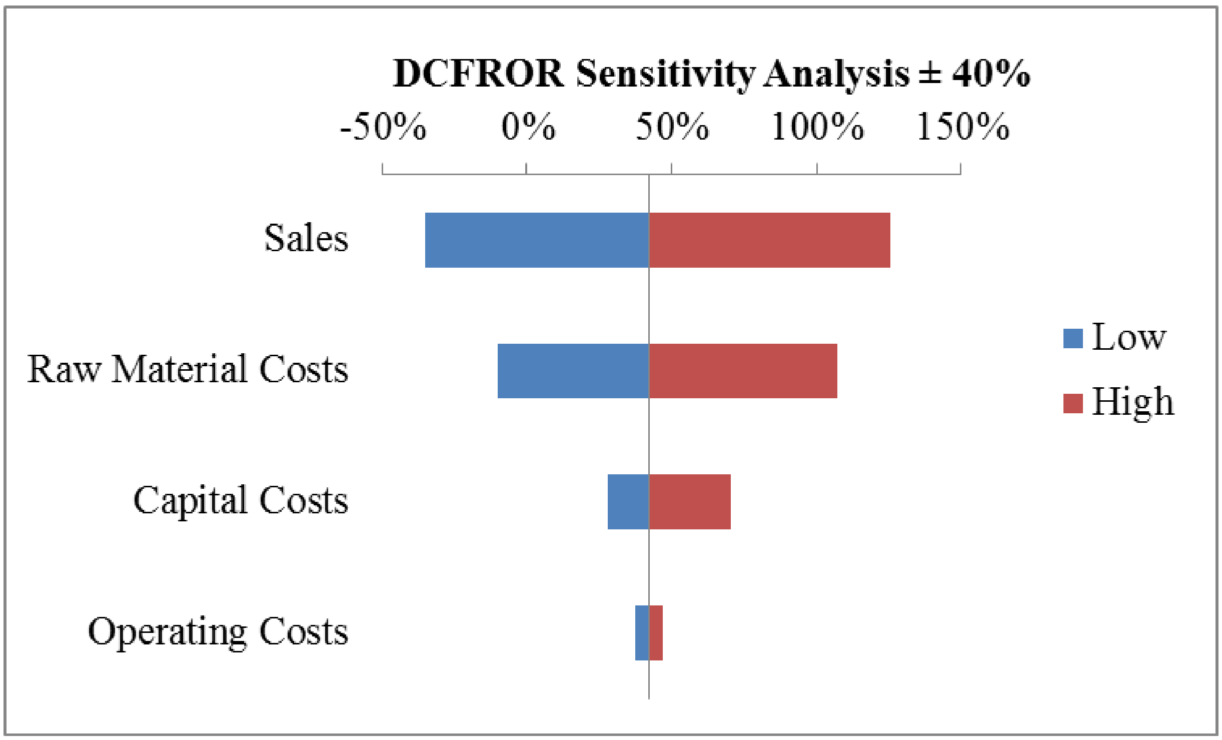

Figure 4: *DCFROR Sensitivity Analysis for ± 40%*

The following table summarizes the high and low values of the key parameters for NPV in **Table 31.** The best case scenario occurs when $97MM in sales and the worst case scenario occurs when ($59)MM in sales loss.

Sensitivity Analysis for NPV		
	Low	High
Sales	($59,132,099)	$97,279,690
Raw Material Costs	($40,573,640)	$78,721,231
Capital Costs	$12,168,022	$25,979,569
Operating Costs	$15,254,333	$22,893,258

Table 31: *Summary of Sensitivity Analysis for NPV ± 40%*

These values are assembled into a tornado chart, **Figure 5**, below. It compares each of the key parameters and their impact on the NPV of the project. The base case for this analysis is an NPV of $19MM.

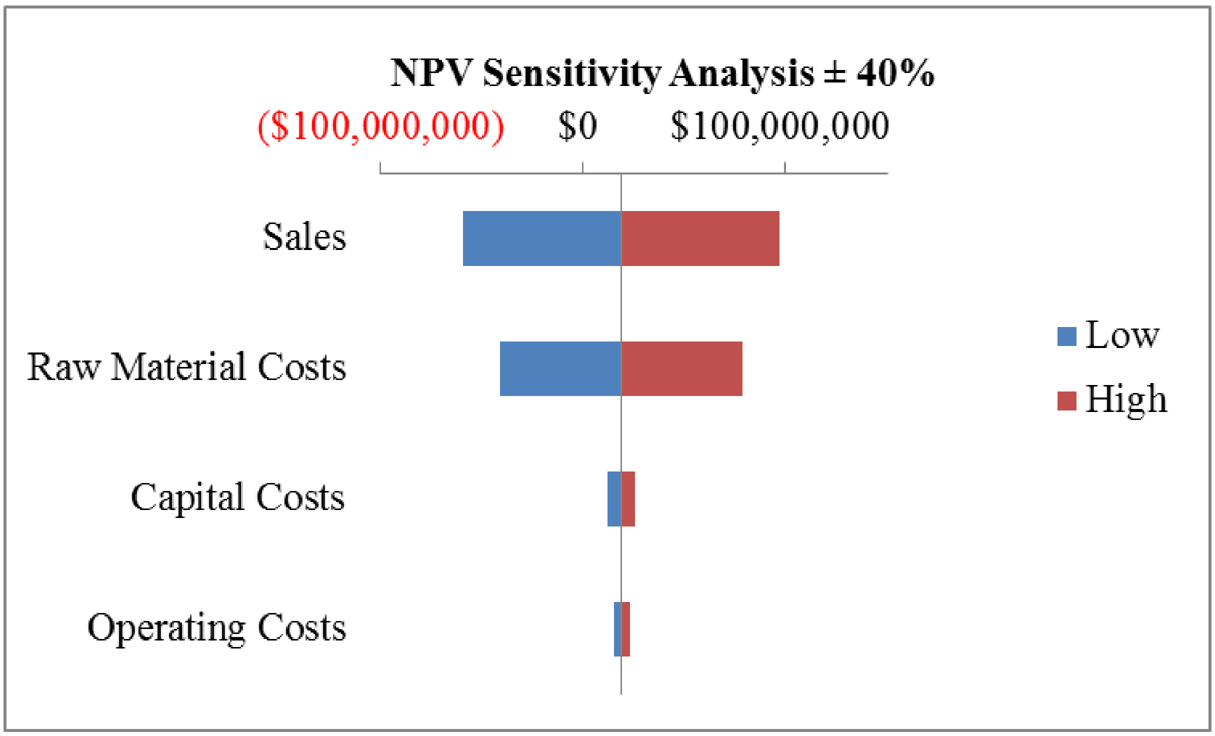

Figure 5: *NPV Sensitivity Analysis for ± 40%*

The sales are the most vulnerable or riskiest portion in the project because the demand for the product depends on the economy and other factors in the refinery. The sales prices for each individual product, isobutane and natural gasoline, also fluctuate unpredictably. Boosting the sales prices will increase the overall NPV and DCFROR for the project. If this is not possible, increasing the amount of product at the same sales price will also increase the NPV and

DCFROR. This can be done by making the unit more efficient by producing as much isobutane and natural gasoline as possible from least amount of raw materials possible.

The next risk factor in the project is the raw material costs. This depends immensely on the supplier and is out of our control. We may be able to find a cheaper supplier. The transportation costs may also fluctuate unpredictably from season to season, which are also out of our control. The effort to change this is the same as noted previously for sales. Increasing the efficiency of the unit by increasing the conversion in the horizontal packed bed reactor (R-1) will significantly increase the production of isobutane and decrease the amount of raw material needed for a certain amount of production per year. Selling more isobutane and natural gasoline will also offset the high raw material costs.

The capital and operating costs have very minor impacts on the sensitivities of this project. The design and operations of the equipment do not need any improvements.

5. Conclusion

It is highly recommended that this project be implemented into the Philips 66 Ponca City Refinery. It has a payback period of 2.69 years and an NPV of $19MM. The DCFROR is 42% with a future worth revenue of $128MM, making the Butamer Unit a very profitable investment.

6. Recommendations

When moving forward with a more detailed design of this Butamer Unit, there are a few specific opportunities for improvement that should be examined. The benefit of developing an on-site storage location for the feed to the process should be considered. Reducing the distance that the feed is pumped before it reaches the unit will help reduce variable flow rates leading into the first column. In addition to this, methods to ensure complete and thorough mixing in the reactor should be examined. At this time, the fluids simply flow through the reactor and it is assumed that everything is well mixed. For this level of the design, this assumption is sufficient since the addition of a hydrogen stream to the reactor section was not considered. In the future design steps it will likely be determined that a hydrogen stream is needed, and therefore mixing should be considered. A mesh catalyst is designed for the reactor; however, there may be alternative options that are more effective. Different configurations for the catalyst should be examined. It may be worthwhile to consider on-site storage of both the isobutane product and the natural gas product. While we intend to sell and transport both products to other sites, we may need to have some local storage for them as well.

One of the best ways to maximize the profitability of the Butamer Unit will be to find as many sales opportunities as possible. The total estimated sales of the products are the single greatest contributing factor to the overall economics. Therefore spend time and effort to establish connections with as many potential customers as possible. The cost of raw materials has the second largest impact on the economics. It is recommended that local storage be developed for the feed so that it can be purchased in excess when at a cheap price. This will reduce or possibly eliminate the need to purchase the feed at higher rates.

7. References

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8. Appendix

Report A-1: ***Minimum Column Pressure for Column T-1***

```
CHEMCAD 6.5.5                                                Page 1

Simulation: Minimum Column 1 Pressure    Date: 02/18/2015  Time: 18:25:37
STREAM PROPERTIES

Stream No.                      1
       Name
- - Overall - -
Molar flow lbmol/h       387.1051
Mass flow  lb/h        22074.3652
Temp F                    95.0000
Pres psia                 75.9406
Vapor mole fraction        0.0000
Enth MMBtu/h              -24.755
Tc F                     271.8969
Pc psia                  543.5579
Std. sp gr.  wtr = 1        0.561
Std. sp gr.  air = 1        1.969
Degree API               120.8904
Average mol wt            57.0242
Actual dens lb/ft3        33.5066
Actual vol ft3/hr        658.8067
Std liq  ft3/hr          630.7048
Std vap 60F scfh      146898.0938
- - Vapor only - -
Molar flow lbmol/h
Mass flow  lb/h
Average mol wt
Actual dens lb/ft3
Actual vol ft3/hr
Std liq  ft3/hr
Std vap 60F scfh
Cp Btu/lbmol-F
Z factor
Visc cP
Th cond Btu/hr-ft-F
- - Liquid only - -
Molar flow lbmol/h       387.1051
Mass flow  lb/h        22074.3652
Average mol wt            57.0242
Actual dens lb/ft3        33.5066
Actual vol ft3/hr        658.8067
Std liq  ft3/hr          630.7048
Std vap 60F scfh      146898.0938
Cp Btu/lbmol-F            35.0587
Z factor                   0.0235
Visc cP                    0.1475
Th cond Btu/hr-ft-F        0.0549
Surf. tens. dyne/cm        8.5578
```

Report A-1: ***Minimum Column Pressure for Column T-2***

```
CHEMCAD 6.5.5                                                Page 1

Simulation: Minimum Column 2 Pressure    Date: 02/18/2015  Time: 18:24:16
STREAM PROPERTIES

Stream No.                       1
       Name
- - Overall - -
Molar flow lbmol/h        702.5308
Mass flow  lb/h         40833.1953
Temp F                    177.4000
Pres psia                 151.4422
Vapor mole fraction         0.0000
Enth MMBtu/h               -41.887
Tc F                      305.6180
Pc psia                   551.0978
Std. sp gr.  wtr = 1         0.584
Std. sp gr.  air = 1         2.007
Degree API                110.6287
Average mol wt             58.1230
Actual dens lb/ft3         31.0813
Actual vol ft3/hr        1313.7545
Std liq  ft3/hr          1119.2437
Std vap 60F scfh       266595.4063
- - Vapor only - -
Molar flow lbmol/h
Mass flow  lb/h
Average mol wt
Actual dens lb/ft3
Actual vol ft3/hr
Std liq  ft3/hr
Std vap 60F scfh
Cp Btu/lbmol-F
Z factor
Visc cP
Th cond Btu/hr-ft-F
- - Liquid only - -
Molar flow lbmol/h        702.5308
Mass flow  lb/h         40833.1953
Average mol wt             58.1230
Actual dens lb/ft3         31.0813
Actual vol ft3/hr        1313.7545
Std liq  ft3/hr          1119.2437
Std vap 60F scfh       266595.4063
Cp Btu/lbmol-F             42.4417
Z factor                    0.0460
Visc cP                    0.09609
Th cond Btu/hr-ft-F         0.0452
Surf. tens. dyne/cm         5.8187
```

Report A-1: ***Bubble Point of Isobutane Product***

```
CHEMCAD 6.5.5                                                 Page 1

Simulation: Bubble Point of Product      Date: 02/19/2015  Time: 18:04:19
STREAM PROPERTIES

Stream No.                      1
       Name
- - Overall - -
Molar flow lbmol/h      1191.7235
Mass flow  lb/h        68839.0000
Temp F                   128.0000
Pres psia                111.1223
Vapor mole fraction        0.0000
Enth MMBtu/h              -75.833
Tc F                     275.4529
Pc psia                  534.3956
Std. sp gr.  wtr = 1        0.563
Std. sp gr.  air = 1        1.994
Degree API               119.6965
Average mol wt            57.7642
Actual dens lb/ft3        32.1182
Actual vol ft3/hr       2143.2991
Std liq  ft3/hr         1957.5500
Std vap 60F scfh      452233.5938
- - Vapor only - -
Molar flow lbmol/h
Mass flow  lb/h
Average mol wt
Actual dens lb/ft3
Actual vol ft3/hr
Std liq  ft3/hr
Std vap 60F scfh
Cp Btu/lbmol-F
Z factor
Visc cP
Th cond Btu/hr-ft-F
- - Liquid only - -
Molar flow lbmol/h      1191.7235
Mass flow  lb/h        68839.0000
Average mol wt            57.7642
Actual dens lb/ft3        32.1182
Actual vol ft3/hr       2143.2991
Std liq  ft3/hr         1957.5500
Std vap 60F scfh      452233.5938
Cp Btu/lbmol-F            38.0801
Z factor                   0.0347
Visc cP                    0.1278
Th cond Btu/hr-ft-F        0.0516
Surf. tens. dyne/cm        6.7617
```

De-isobutanizer (T-1) Feed Stream Column Optimization Data

Column 1 (T-1) Optimization Data Example calculations for 75 and 65 stages

# of Theoretical Stages	74	# of Theoretical Stages	64
# of Computational Stages	75	# of Computational Stages	65
# of Trays Inside the Column	73	# of Trays Inside the Column	63
Column Diameter (ft)	21	Column Diameter (ft)	21
Column Diameter rounded (ft)	21	Column Diameter rounded (ft)	21
Column Diameter (m)	6.4008	Column Diameter (m)	6.4008
Column Diameter rounded (m)	6.4	Column Diameter rounded (m)	6.4
Tray Spacing/tray (ft)	2	Tray Spacing/tray (ft)	2
$h_{Overhead}$ (ft)	3	$h_{Overhead}$ (ft)	3
h_{trays} (ft)	182	h_{trays} (ft)	158
h_{Btm} (ft)	7	h_{Btm} (ft)	7
h_{skirt} (ft)	8	h_{skirt} (ft)	8
h_{total} (ft)	200	h_{total} (ft)	176
Volume (f3)	69272.11801	Volume (f3)	60959.46385
Volume (m3)	1961.564711	Volume (m3)	1726.176946
Column Pressure (psig)	135.3	Column Pressure (psig)	135.3
	148.83		148.83
	185.3		185.3
Design Pressure (psig)	185.3	Design Pressure (psig)	185.3
Design Pressure (barg)	12.77598524	Design Pressure (barg)	12.77598524
Column Area (ft^2)	346.3605901	Column Area (ft^2)	346.3605901
Downcomer area/column area (ft2)	0.1	Downcomer area/column area (ft2)	0.1
Downcomer Area (ft2)	34.63605901	Downcomer Area (ft2)	34.63605901
Tray Area (ft2)	311.7245311	Tray Area (ft2)	311.7245311
Tray Area (m2)	28.95991655	Tray Area (m2)	28.95991655
Overall Efficiency	0.8	Overall Efficiency	0.8
Nactual	91	Nactual	79
Capital Cost: Tower		**Capital Cost: Tower**	
$F_{P,vessel}$	8.812985382	$F_{P,vessel}$	8.812985382
$C_p°$	1375563.414	$C_p°$	1187537.772
K_1	3.4974	K_1	3.4974
K_2	0.4485	K_2	0.4485
K_3	0.1074	K_3	0.1074
B_1	2.25	B_1	2.25
B_2	1.82	B_2	1.82
F_M	1	F_M	1
C_{BM} (2001)	25158550.55	C_{BM} (2001)	21719630.49
C_{BM} (2016)	**$36,384,117.02**	**C_{BM} (2016)**	**$31,410,775.26**
Capital Cost: Trays (Sieve)		**Capital Cost: Trays (Sieve)**	
$C_p°$	31188.84718	$C_p°$	31188.84718
K_1	2.9949	K_1	2.9949
K_2	0.4465	K_2	0.4465
K_3	0.3961	K_3	0.3961
N	73	N	63
F_Q	1	F_Q	1
F_{BM}	1	F_{BM}	1
C_{BM} (2001)	2276785.844	C_{BM} (2001)	1964897.373
C_{BM} (2016)	**$3,292,671.51**	**C_{BM} (2016)**	**$2,841,620.62**
TOTAL COST	**$39,676,788.53**	TOTAL COST	**$34,252,395.87**

De-isobutanizer (T-1) Reboiler (E-2) Optimization
Example calculations

Reboiler (Shell & Tube HEX)	
Reboiler Duty (Btu/hr)	99694994.7
$T_{H,in}$ (°F)	275
$T_{H,out}$ (°F)	275
$T_{C,in}$ (°F)	174.66
$T_{C,out}$ (°F)	176.15
Approach Temp (°F)	20
LMTD	99.59314237
U (Btu/hr ft^2 °F)	140
F (correction factor)	0.9
Area (ft^2)	**7945**
Area (m^2)	738.114335
Column Pressure (psia)	150
Computational Stages	75
Trays Inside the Column	73
Overall Efficiency	0.8
Nactual	91
Final Reboiler Pressure (psia)	159.1
Reboiler Pressure (barg)	9.935345255
ΔT_{steam}	0
$T_{avg, steam}$	275
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	102735.9797
Capital Cost	
$C_p°$	1601499.625
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$5,321,731.38
C_{BM} (2016)	**$7,696,250.10**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	863968494.9
$ per lb steam	$0.00012
Operating Cost Per Year	**$103,676.22**
Total Capital + Operating	**$7,799,926.32**

Reboiler (Shell & Tube HEX)	
Reboiler Duty (Btu/hr)	93639269.7
$T_{H,in}$ (°F)	275
$T_{H,out}$ (°F)	275
$T_{C,in}$ (°F)	174.66
$T_{C,out}$ (°F)	176.15
Approach Temp (°F)	20
LMTD	99.59314237
U (Btu/hr ft^2 °F)	140
F (correction factor)	0.9
Area (ft^2)	**7463**
Area (m^2)	693.335089
Column Pressure (psia)	150
Computational Stages	65
Trays Inside the Column	63
Overall Efficiency	0.8
Nactual	79
Final Reboiler Pressure (psia)	157.9
Reboiler Pressure (barg)	9.852608167
ΔT_{steam}	0
$T_{avg, steam}$	275
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	96495.53761
Capital Cost	
$C_p°$	1437137.206
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$4,775,560.38
C_{BM} (2016)	**$6,906,381.49**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	811488873.1
$ per lb steam	$0.00012
Operating Cost Per Year	**$97,378.66**
Total Capital + Operating	**$7,003,760.15**

De-isobutanizer (T-1) Condenser (E-1) Optimization

Example Calculations

Shell and Tube Cooler (Floating Head)		**Shell and Tube Cooler (Floating Head)**		**Shell and Tube Cooler (Floating Head)**	
Cooler Duty (Btu/hr)	1.28E+08	**Cooler Duty (Btu/hr)**	1.20E+08	**Cooler Duty (Btu/hr)**	1.14E+08
Mass flow rate (lbmol/hr)	1.20E+04	**Mass flow rate (lbmol/hr)**	1.20E+04	**Mass flow rate (lbmol/hr)**	1.20E+04
Cp (Btu/lbmol F)	35.4392	**Cp (Btu/lbmol F)**	35.4392	**Cp (Btu/lbmol F)**	35.4392
$\Delta T_{Bottoms}$ (°F)	1.31	**$\Delta T_{Bottoms}$ (°F)**	1.31	**$\Delta T_{Bottoms}$ (°F)**	1.31
$T_{H,in}$ (°F)	154.5	**$T_{H,in}$ (°F)**	154.5	**$T_{H,in}$ (°F)**	154.5
$T_{H,out}$ (°F)	153.19	**$T_{H,out}$ (°F)**	153.19	**$T_{H,out}$ (°F)**	153.19
$T_{C,in}$ (°F)	87	**$T_{C,in}$ (°F)**	87	**$T_{C,in}$ (°F)**	87
$T_{C,out}$ (°F)	120	**$T_{C,out}$ (°F)**	120	**$T_{C,out}$ (°F)**	120
LMTD	48.63636538	**LMTD**	48.63636538	**LMTD**	48.63636538
U (Btu/hr ft^2 °F)	80	**U (Btu/hr ft^2 °F)**	80	**U (Btu/hr ft^2 °F)**	80
F (correction factor)	0.9	**F (correction factor)**	0.9	**F (correction factor)**	0.9
Area (ft^2)	**36456**	**Area (ft^2)**	**34235**	**Area (ft^2)**	**32453**
Area (m^2)	3386.871768	**Area (m^2)**	3180.534205	**Area (m^2)**	3014.981059
Pressure at condenser (psia)	150	**Pressure at condenser (psia)**	150	**Pressure at condenser (psia)**	150
Number of stages	75	**Number of stages**	65	**Number of stages**	55
N_{actual}	93.75	**N_{actual}**	81.25	**N_{actual}**	68.75
Pressure drop for stages	9.375	**Pressure drop for stages**	8.125	**Pressure drop for stages**	6.875
Pressure (psia)	159.375	**Pressure (psia)**	158.125	**Pressure (psia)**	156.875
Pressure (barg)	9.954305837	**Pressure (barg)**	9.868121371	**Pressure (barg)**	9.781936905
$\dot{m}_{water}$ (lbmol/hr)	215288.7238	**$\dot{m}_{water}$ (lbmol/hr)**	202168.9102	**$\dot{m}_{water}$ (lbmol/hr)**	191644.6634
$\dot{m}_{water}$ (lb/hr)	3875197.028	**$\dot{m}_{water}$ (lb/hr)**	3639040.383	**$\dot{m}_{water}$ (lb/hr)**	3449603.941
ρ_{water} (lb/ft3)	62.42	**ρ_{water} (lb/ft3)**	62.42	**ρ_{water} (lb/ft3)**	62.42
$\dot{m}_{water}$ (gpm)	7740.170902	**$\dot{m}_{water}$ (gpm)**	7268.480617	**$\dot{m}_{water}$ (gpm)**	6890.107486
Cp (Btu/lbmol F)	17.9691	**Cp (Btu/lbmol F)**	17.9691	**Cp (Btu/lbmol F)**	17.9691
ΔT_{water} (°F)	33	**ΔT_{water} (°F)**	33	**ΔT_{water} (°F)**	33
Capital Cost (Shell and Tube - Floating Head		**Capital Cost (Shell and Tube - Floating Head)**		**Capital Cost (Shell and Tube - Floating Head)**	
$C_p°$	627957.5337	$C_p°$	575409.5815	$C_p°$	534656.0614
K_1	4.8306	K_1	4.8306	K_1	4.8306
K_2	-0.8509	K_2	-0.8509	K_2	-0.8509
K_3	0.3187	K_3	0.3187	K_3	0.3187
B_1	1.63	B_1	1.63	B_1	1.63
B_2	1.66	B_2	1.66	B_2	1.66
F_M	1	F_M	1	F_M	1
Pressure (psia)	159.375	Pressure (psia)	158.125	Pressure (psia)	156.875
Design Pressure (psia)	209.375	Design Pressure (psia)	208.125	Design Pressure (psia)	206.875
Pressure (barg)	13.40168448	Pressure (barg)	13.31550002	Pressure (barg)	13.22931555
$F_{P,vessel}$	1.016060847	$F_{P,vessel}$	1.015920419	$F_{P,vessel}$	1.015779555
C_1	-0.00164	C_1	-0.00164	C_1	-0.00164
C_2	-0.00627	C_2	-0.00627	C_2	-0.00627
C_3	0.0123	C_3	0.0123	C_3	0.0123
C_{BM} (2001)	$2,082,722.27	C_{BM} (2001)	$1,908,304.39	C_{BM} (2001)	$1,773,023.26
C_{BM} (2016)	**$3,012,018.14**	**C_{BM} (2016)**	**$2,759,776.24**	**C_{BM} (2016)**	**$2,564,133.63**
Operating Cost		**Operating Cost**		**Operating Cost**	
Converting to per min basis	504576	Converting to per min basis	504576	Converting to per min basis	504576
Service Factor	0.96	Service Factor	0.96	Service Factor	0.96
Cooling water (gpm)	7740.170902	Cooling water (gpm)	7268.480617	Cooling water (gpm)	6890.107486
$/ gpm year cooling water	120	$/ gpm year cooling water	120	$/ gpm year cooling water	120
$/ gpm	2.28E-04	$/ gpm	2.28E-04	$/ gpm	2.28E-04
Operating Cost Per Year	**$856,000.98**	**Operating Cost Per Year**	**$803,835.81**	**Operating Cost Per Year**	**$761,990.77**
Total Capital + Operating	**$3,868,019.12**	**Total Capital + Operating**	**$3,563,612.05**	**Total Capital + Operating**	**$3,326,124.40**

Air-Cooled Condenser is cheaper

Condenser (Air-cooled HEX)	
Distillate Flow Rate (lb/hr)	6.27E+04
Reflux Ratio	1.5
ṁ (lb/hr) (Vapor Flow Rate)	156730
P_{inlet} (psia)	200
U_x (Btu/hr ft^2 ^{o}F)	3.9
$T_{H,in}$ (^{o}F) (T_1 - GPSA notation)	179.83
$T_{H,out}$ (^{o}F) (T_2 - GPSA notation)	138.29
$T_{C,in}$ (^{o}F) (t_1 - GPSA notation)	117
$T_{C,out}$ (^{o}F) (t_2 - GPSA notation)	139.83
ΔT_1 (^{o}F)	21.29
ΔT_2 (^{o}F)	40
LMTD	29.66818058
Δt_a (oF)	20.6094
P	0.363361452
R	1.819535699
F	0.91
CMTD	26.99804433
Condenser Duty (Btu/hr)	1,776,250
Required Extended Surface Area, A_x (ft^2)	**16870**
Required Extended Surface Area, A_x (m^2)	1567.27361
APSF	118.8
Face Area, F_a (ft^2)	142.003367
Assumed Tube Length (ft)	9
Width (ft)	15.8
APF	5.58
Number of Tubes, N_t	336
Tube Area, A_t (in^2)	0.5945
Tube Area, A_t (ft^2)	0.004128472
Tube Side Mass Velocity, G_t (lb/ft^2*s)	31.3849173
Tube ID, D_i (in)	0.87
Tube ID, D_i (ft)	0.0725
$\dot{m}_{air}$ (lb/hr), W_a - GPSA notation	359110.0016
Air Face Mass Velocity, G_a (lb/ft^2*h)	2528.88371
Air-side film coefficient, h_a (Btu/h*ft2*°F)	7.8
AR	21.4
GPSA Method (bare tube area)	
U_b (Btu/hr ft^2 ^{o}F)	83.46
Required Bare Surface Area, A_b (ft^2)	**789**
Required Bare Surface Area, A_b (m^2)	73.300467
GPSA Method (fan sizing)	
Number of Fans, N_f	1
Minimum fan area, FAPF (ft^2)	56.8013468
Fan diameter (ft)	9
Pressure drop factor, Fp	0.083
T_a, avg (°F)	128.415
Number of rows	4
Air density ratio, Dr	0.92
Air static pressure drop, ΔP_a	0.360869565
Actual air volume (ACFM)	86857.35608
Fan total pressure, PF (in H_2O)	0.467786426
BHP, per fan	9.132134346
Efficiency of fan motor speed reducer	0.92
Actual fan motor needed (hp)	9.926232985
Motor size, rounded to discrete size, hp	7.5
Motor size, kW	5.59275

Capital Cost	
$C_p°$	43965.20061
K_1	4.0336
K_2	0.2341
K_3	0.0497
B_1	0.96
B_2	1.21
F_M	1
Pressure (psia)	150
Pressure (barg)	9.307922342
$F_{P,vessel}$	0.993047359
C_1	-0.125
C_2	0.15361
C_3	-0.02861
C_{BM} (2001)	$95,034.62
C_{BM} (2014)	**$138,692.66**
Motor Capital Cost	
$C_p°$	2272.186266
K_1	2.4604
K_2	1.4191
K_3	-0.1798
F_{BM}	1.5
Shaft Power, kW	4.92162
C_{BM} (2001)	3408.279399
C_{BM} (2014)	**$4,974.01**
Operating Cost	
Purchased (hp)	8.522727273
Purchased (kW)	6.355397727
Service Factor	0.96
$ per kWh	0.06
Operating Cost Per Year	**$3,206.78**

De-isobutanizer (T-1) Optimization Results

# of Computational Stages	Reboiler Duty (BTU/hr)	Condenser Duty (BTU/hr)	Optimum Feed Stage	Column PWC	Reboiler PWC	Condenser PWC	Total PWC	
22	5118080.912	2700664.333	20	$3,094,724	$5,118,081	$2,700,664	$10,913,469	NO
23	4993743.128	2647017.977	21	$2,580,741	$4,993,743	$2,647,018	$10,221,502	NO
24	4890901.658	2602543.862	22	$3,362,439	$4,890,902	$2,602,544	$10,855,884	NO
25	4808895.961	2566702.499	23	$2,794,713	$4,808,896	$2,566,702	**$10,170,311**	YES
26	76162294.7	9.43E+07	25	$2,869,331	$4,956,219	$2,628,046	$10,453,596	NO
35	83295339.5	1.07E+08	34	$15,712,836	$5,746,321	$3,070,353	$24,529,510	NO
45	85244099	1.09E+08	44	$20,247,265	$5,974,026	$3,158,079	$29,379,370	NO
55	88776531	1.14E+08	54	$24,608,848	$6,395,992	$3,326,124	$34,330,964	NO
65	93639269.7	1.20E+08	64	$34,252,396	$7,003,760	$3,563,612	$44,819,768	NO
75	99694994.7	1.28E+08	74	$39,676,789	$7,799,926	$3,868,019	$51,344,734	NO
25	8.95E+07	7.98E+07	24	$8,326,517	$6,484,980	$2,143,248	$16,954,744	NO
26			25	$8,814,984				NO
						Minimum	**$10,170,311**	

# of Computational Stages 25 (unhighlighted) - New Design with 25 stages col 1, 15 stages col 2 = more expensive
# of Computational Stages 26 (last one) - New Design with 95 F Recycle Feed with Heat Exchanger

	Design without purge in recycle	Design with 10% purge in recycle
Isobutane Flow (BBL/DAY)	8367.24	6920.82
$/BBL Isobutane 92% vol spec	$37.67	$37.67
Sales/day	$315,194	$260,707
Sales/year	$115,045,785	$95,158,161
Loss	$19,887,624	

Table 32: *Optimization of T-1 Tray Count/Location*

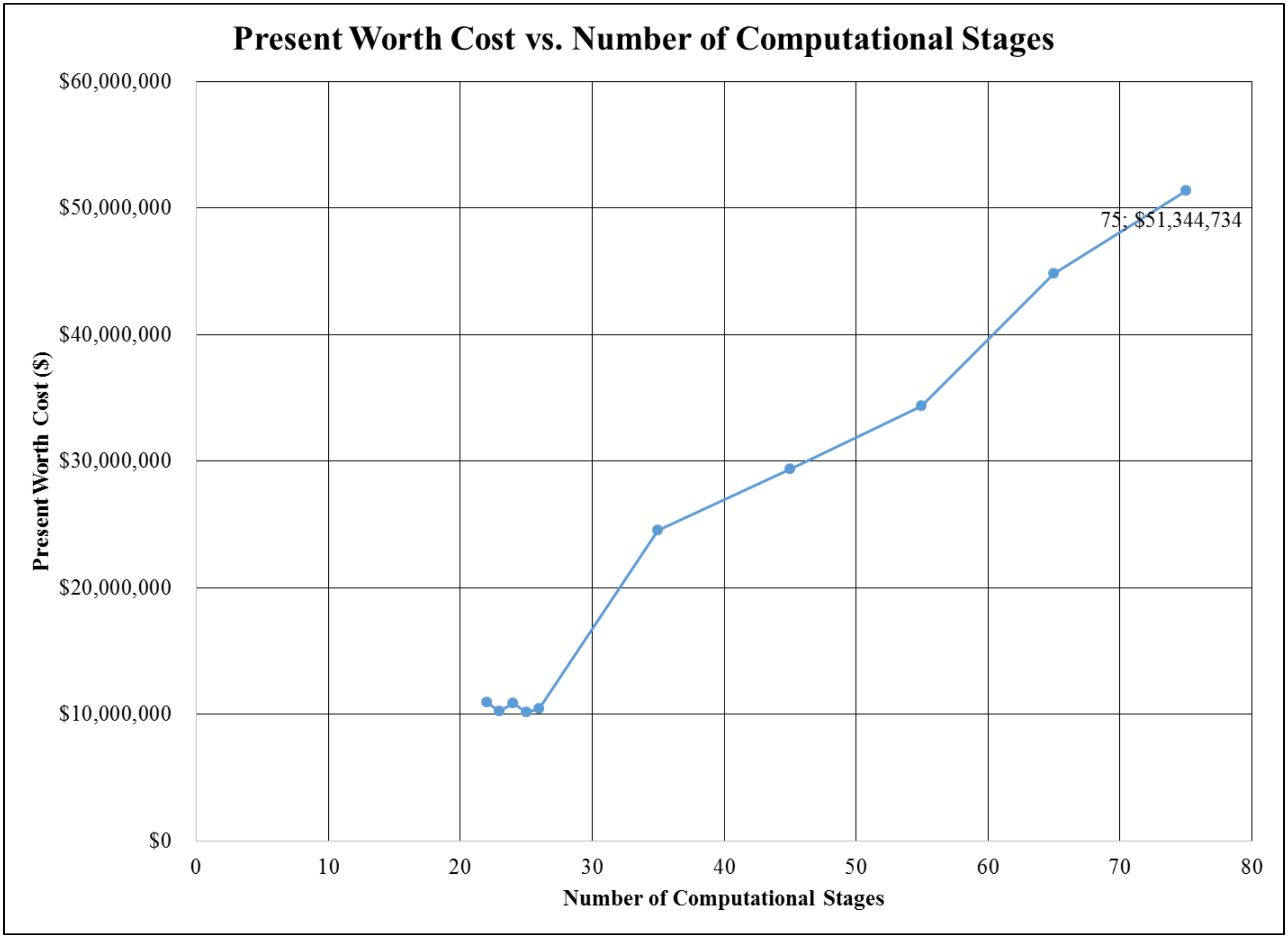


Table 33: *Present Worth Cost vs. Number of Computational Stages*

Optimum Feed Stage location and Diameters; T-1 Optimization Using Sensitivity Analysis on Aspen Plus

Row/Case	Status 24 stages	VARY 1 B1 S8 FEEDS STAGE	REBOILER BTU/HR	CONDENS BTU/HR		
21	OK	22	75546818.9	-93549874.4	YES	YES
22	OK	23	77541609.4	-96146616.6	NO	NO
23	OK	24	80233281.9	-99654851.4	NO	NO
			75546818.9	-93549874.4		

Section starting stage:	2	
Section ending stage:	29	
Stage with maximum diameter:	29	
Column diameter:	12.6238	ft
Downcomer area / Column area:	0.1	
Side downcomer velocity:	0.336068	ft/sec
Flow path length:	8.67316	ft
Side downcomer width:	1.97531	ft
Side weir length:	9.17258	ft

***Finding the tray sizing and optimum feed stage location – Example calculation**

De-isobutanizer (T-1) Recycle Optimization Data – Example Calculations

# of Theoretical Stages	74	# of Theoretical Stages	64
# of Computational Stages	75	# of Computational Stages	65
# of Trays Inside the Column	73	# of Trays Inside the Column	63
For stages 2-29:		For stages 2-29:	
Column Diameter (ft)	21	Column Diameter (ft)	21
Column Diameter rounded (ft)	21	Column Diameter rounded (ft)	21
Column Diameter (m)	6.4008	Column Diameter (m)	6.4008
Column Diameter rounded (m)	6.4	Column Diameter rounded (m)	6.4
Tray Spacing/tray (ft)	2	Tray Spacing/tray (ft)	2
$h_{Overhead}$ (ft)	3	$h_{Overhead}$ (ft)	3
h_{trays} (ft)	182	h_{trays} (ft)	158
h_{Btm} (ft)	7	h_{Btm} (ft)	7
h_{skirt} (ft)	8	h_{skirt} (ft)	8
h_{total} (ft)	200	h_{total} (ft)	176
Volume (f3)	69272.11801	Volume (f3)	60959.46385
Volume (m3)	1961.564711	Volume (m3)	1726.176946
Column Pressure (psig)	135.3	Column Pressure (psig)	135.3
	148.83		148.83
	185.3		185.3
Design Pressure (psig)	185.3	Design Pressure (psig)	185.3
Design Pressure (barg)	12.77598524	Design Pressure (barg)	12.77598524
Column Area (ft^2)	346.3605901	Column Area (ft^2)	346.3605901
Downcomer area/column area (ft2)	0.1	Downcomer area/column area (ft2)	0.1
Downcomer Area (ft2)	34.63605901	Downcomer Area (ft2)	34.63605901
Tray Area (ft2)	311.7245311	Tray Area (ft2)	311.7245311
Tray Area (m2)	28.95991655	Tray Area (m2)	28.95991655
Overall Efficiency	0.8	Overall Efficiency	0.8
Nactual	91	Nactual	79
Capital Cost: Tower		**Capital Cost: Tower**	
$F_{P,vessel}$	8.812985382	$F_{P,vessel}$	8.812985382
$C_p°$	1375563.414	$C_p°$	1187537.772
K_1	3.4974	K_1	3.4974
K_2	0.4485	K_2	0.4485
K_3	0.1074	K_3	0.1074
B_1	2.25	B_1	2.25
B_2	1.82	B_2	1.82
F_M	1	F_M	1
C_{BM} (2001)	25158550.55	C_{BM} (2001)	21719630.49
C_{BM} (2016)	**$36,384,117.02**	**C_{BM} (2016)**	**$31,410,775.26**
Capital Cost: Trays (Sieve)		**Capital Cost: Trays (Sieve)**	
$C_p°$	31188.84718	$C_p°$	31188.84718
K_1	2.9949	K_1	2.9949
K_2	0.4465	K_2	0.4465
K_3	0.3961	K_3	0.3961
N	73	N	63
F_Q	1	F_Q	1
F_{BM}	1	F_{BM}	1
C_{BM} (2001)	2276785.844	C_{BM} (2001)	1964897.373
C_{BM} (2016)	**$3,292,671.51**	**C_{BM} (2016)**	**$2,841,620.62**
TOTAL COST	**$39,676,788.53**	TOTAL COST	**$34,252,395.87**

De-isobutanizer (T-1) Recycle Optimization of Reboiler (E-2) – Example Calculations

Reboiler (Shell & Tube HEX)	
Reboiler Duty (Btu/hr)	69106956.8
$T_{H,in}$ (°F)	275
$T_{H,out}$ (°F)	275
$T_{C,in}$ (°F)	174.66
$T_{C,out}$ (°F)	176.15
Approach Temp (°F)	20
LMTD	99.59314237
U (Btu/hr ft^2 °F)	140
F (correction factor)	0.9
Area (ft^2)	**5508**
Area (m^2)	511.709724
Column Pressure (psia)	150
Computational Stages	45
Trays Inside the Column	43
Overall Efficiency	0.8
Nactual	54
Final Reboiler Pressure (psia)	155.4
Reboiler Pressure (barg)	9.680239235
ΔT_{steam}	0
$T_{avg, steam}$	275
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	71214.91838
Capital Cost	
$C_p°$	866058.5738
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$2,877,884.59
C_{BM} (2016)	**$4,161,976.24**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	598888977.6
$ per lb steam	$0.00012
Operating Cost Per Year	**$71,866.68**
Total Capital + Operating	**$4,233,842.92**

De-isobutanizer (T-1) Recycle Condenser (E-1) Optimization – Example Calculation

Shell and Tube Cooler (Floating Head)		**Shell and Tube Cooler (Floating Head)**		**Shell and Tube Cooler (Floating Head)**	
Cooler Duty (Btu/hr)	8.41E+07	**Cooler Duty (Btu/hr)**	8.46E+07	**Cooler Duty (Btu/hr)**	8.55E+07
Mass flow rate (lbmol/hr)	1.20E+04	**Mass flow rate (lbmol/hr)**	1.20E+04	**Mass flow rate (lbmol/hr)**	1.20E+04
Cp (Btu/lbmol F)	35.4392	**Cp (Btu/lbmol F)**	35.4392	**Cp (Btu/lbmol F)**	35.4392
$\Delta T_{Bottoms}$ (°F)	1.31	**$\Delta T_{Bottoms}$ (°F)**	1.31	**$\Delta T_{Bottoms}$ (°F)**	1.31
$T_{H,in}$ (°F)	154.5	**$T_{H,in}$ (°F)**	154.5	**$T_{H,in}$ (°F)**	154.5
$T_{H,out}$ (°F)	153.19	**$T_{H,out}$ (°F)**	153.19	**$T_{H,out}$ (°F)**	153.19
$T_{C,in}$ (°F)	87	**$T_{C,in}$ (°F)**	87	**$T_{C,in}$ (°F)**	87
$T_{C,out}$ (°F)	120	**$T_{C,out}$ (°F)**	120	**$T_{C,out}$ (°F)**	120
LMTD	48.63636538	**LMTD**	48.63636538	**LMTD**	48.63636538
U (Btu/hr ft^2 °F)	80	**U (Btu/hr ft^2 °F)**	80	**U (Btu/hr ft^2 °F)**	80
F (correction factor)	0.9	**F (correction factor)**	0.9	**F (correction factor)**	0.9
Area (ft^2)	**24020**	**Area (ft^2)**	**24169**	**Area (ft^2)**	**24422**
Area (m^2)	2231.53006	**Area (m^2)**	2245.372607	**Area (m^2)**	2268.877066
Pressure at condenser (psia)	150	**Pressure at condenser (psia)**	150	**Pressure at condenser (psia)**	150
Number of stages	75	**Number of stages**	65	**Number of stages**	55
N_{actual}	93.75	**N_{actual}**	81.25	**N_{actual}**	68.75
Pressure drop for stages	9.375	**Pressure drop for stages**	8.125	**Pressure drop for stages**	6.875
Pressure (psia)	159.375	**Pressure (psia)**	158.125	**Pressure (psia)**	156.875
Pressure (barg)	9.954305837	**Pressure (barg)**	9.868121371	**Pressure (barg)**	9.781936905
$\dot{m}_{water}$ (lbmol/hr)	141846.0323	**$\dot{m}_{water}$ (lbmol/hr)**	142728.5561	**$\dot{m}_{water}$ (lbmol/hr)**	144222.2765
$\dot{m}_{water}$ (lb/hr)	2553228.582	**$\dot{m}_{water}$ (lb/hr)**	2569114.009	**$\dot{m}_{water}$ (lb/hr)**	2596000.977
ρ_{water} (lb/ft3)	62.42	**ρ_{water} (lb/ft3)**	62.42	**ρ_{water} (lb/ft3)**	62.42
$\dot{m}_{water}$ (gpm)	5099.721494	**$\dot{m}_{water}$ (gpm)**	5131.450441	**$\dot{m}_{water}$ (gpm)**	5185.153447
Cp (Btu/lbmol F)	17.9691	**Cp (Btu/lbmol F)**	17.9691	**Cp (Btu/lbmol F)**	17.9691
ΔT_{water} (°F)	33	**ΔT_{water} (°F)**	33	**ΔT_{water} (°F)**	33
Capital Cost (Shell and Tube - Floating Head		**Capital Cost (Shell and Tube - Floating Head**		**Capital Cost (Shell and Tube - Floating Head**	
C_p°	358840.743	C_p°	361702.1688	C_p°	366581.0892
K_1	4.8306	K_1	4.8306	K_1	4.8306
K_2	-0.8509	K_2	-0.8509	K_2	-0.8509
K_3	0.3187	K_3	0.3187	K_3	0.3187
B_1	1.63	B_1	1.63	B_1	1.63
B_2	1.66	B_2	1.66	B_2	1.66
F_M	1	F_M	1	F_M	1
Pressure (psia)	159.375	Pressure (psia)	158.125	Pressure (psia)	156.875
Design Pressure (psia)	209.375	Design Pressure (psia)	208.125	Design Pressure (psia)	206.875
Pressure (barg)	13.40168448	Pressure (barg)	13.31550002	Pressure (barg)	13.22931555
$F_{P,vessel}$	1.016060847	$F_{P,vessel}$	1.015920419	$F_{P,vessel}$	1.015779555
C_1	-0.00164	C_1	-0.00164	C_1	-0.00164
C_2	-0.00627	C_2	-0.00627	C_2	-0.00627
C_3	0.0123	C_3	0.0123	C_3	0.0123
C_{BM} (2001)	$1,190,153.10	C_{BM} (2001)	$1,199,559.16	C_{BM} (2001)	$1,215,654.03
C_{BM} (2016)	**$1,721,190.95**	**C_{BM} (2016)**	**$1,734,793.94**	**C_{BM} (2016)**	**$1,758,070.22**
Operating Cost		**Operating Cost**		**Operating Cost**	
Converting to per min basis	504576	Converting to per min basis	504576	Converting to per min basis	504576
Service Factor	0.96	Service Factor	0.96	Service Factor	0.96
Cooling water (gpm)	5099.721494	Cooling water (gpm)	5131.450441	Cooling water (gpm)	5185.153447
$/ gpm year cooling water	120	$/ gpm year cooling water	120	$/ gpm year cooling water	120
$/ gpm	2.28E-04	$/ gpm	2.28E-04	$/ gpm	2.28E-04
Operating Cost Per Year	**$563,988.40**	**Operating Cost Per Year**	**$567,497.37**	**Operating Cost Per Year**	**$573,436.49**
Total Capital + Operating	**$2,285,179.35**	**Total Capital + Operating**	**$2,302,291.30**	**Total Capital + Operating**	**$2,331,506.71**

De-isobutanizer (T-1) Recycle Optimization

# of Computational Stages	Reboiler Duty (BTU/hr)	Condenser Duty (BTU/hr)	Optimum Feed Stage	Column PWC	Reboiler PWC	Condenser PWC	Total PWC	
22	81566842.1	1.02E+08	21	$3,094,724	$5,550,159	$2,911,227	$11,556,110	NO
23	80233491.2	1.01E+08	22	$2,580,741	$5,399,447	$2,847,943	$10,828,131	NO
24	79140876.8	9.91E+07	23	$3,362,439	$5,278,493	$2,796,264	$11,437,196	NO
25	78263245.6	9.79E+07	24	$10,020,202	$5,182,211	$2,754,851	$17,957,264	NO
26	77552742.7	9.70E+07	25	$2,869,331	$5,105,840	$2,721,151	**$10,696,321**	YES
35	71646629.6	9.06E+07	34	$15,712,836	$4,486,763	$2,502,214	$22,701,813	NO
45	69106956.8	8.72E+07	44	$20,247,265	$4,233,843	$2,385,956	$26,867,063	NO
55	67894729.3	8.55E+07	54	$24,608,848	$4,115,062	$2,331,507	$31,055,416	NO
65	67236752.5	8.46E+07	64	$34,252,396	$4,052,086	$2,302,291	$40,606,773	NO
75	66848363.8	8.41E+07	74	$39,676,789	$4,014,781	$2,285,179	$45,976,749	NO
						Minimum	**$10,696,321**	

Table 34: *De-isobutanizer Optimization (Recycle)*

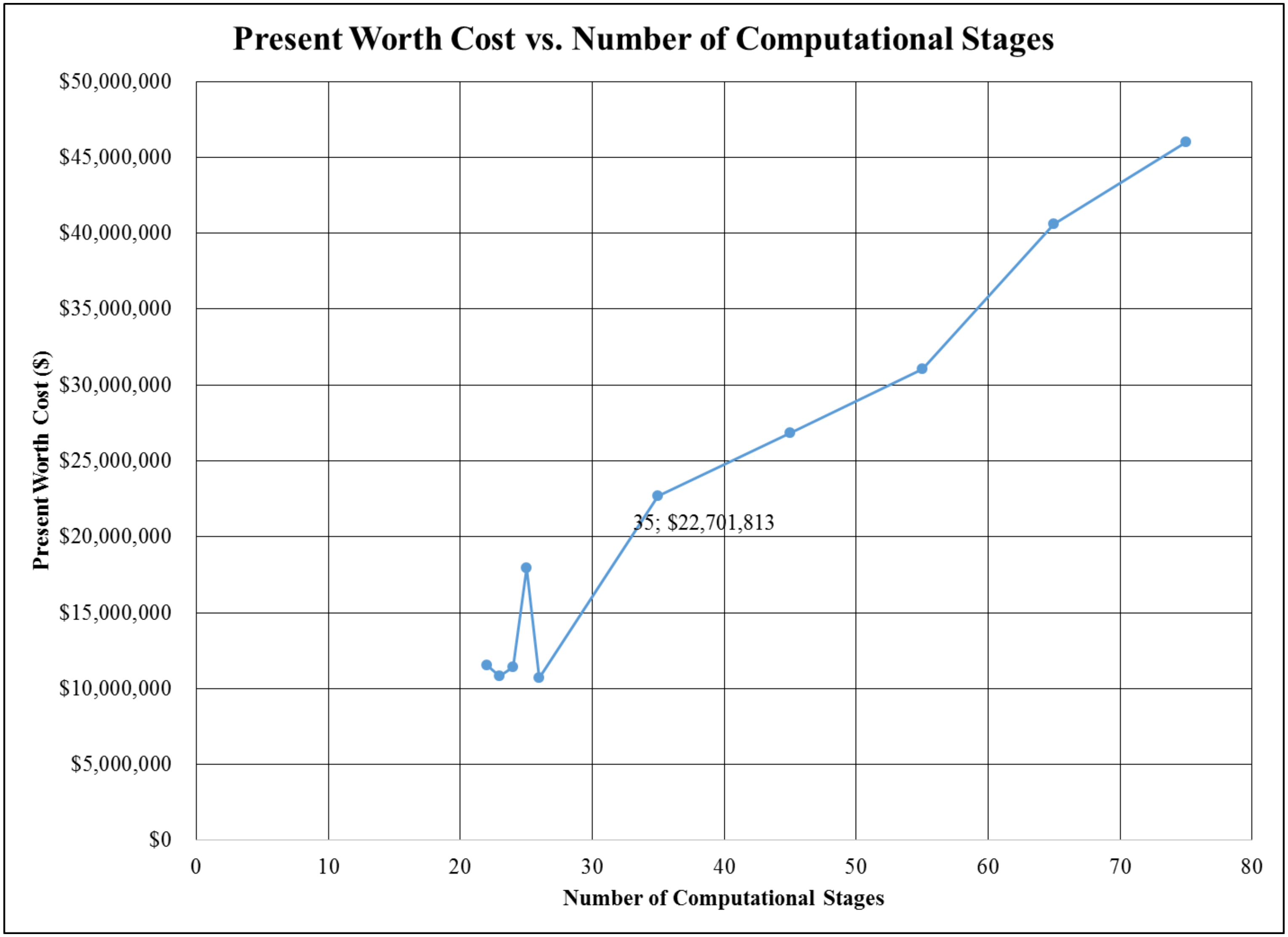


Table 35: *Present Worth Cost vs. Number of Computational Stages*

Optimum Feed Stage location and Diameters; T-1 (Recycle) Optimization Using Sensitivity Analysis on Aspen Plus

75 stages – Example

Row/Case	Status					
		VARY 1	REBOILER	CONDENS		
		B1				
		S8				
		FEEDS				
		STAGE				
			BTU/HR	BTU/HR		
8	OK	9	95950666.2	-123674422	NO	NO
9	OK	10	89432880.3	-114843842	NO	NO
10	OK	11	84321252.9	-107915464	NO	NO
11	OK	12	80294241.8	-102454110	NO	NO
12	OK	13	77097695.2	-98116062.4	NO	NO
13	OK	14	74560093	-94669379.2	NO	NO
14	OK	15	72527667.3	-91907038.4	NO	NO
15	OK	16	70904380.4	-89697454	NO	NO
16	OK	17	69617034.3	-87941738.2	NO	NO
17	OK	18	68604185.9	-86557048	NO	NO
18	OK	19	67832590.8	-85497919.8	NO	NO
19	OK	20	67283340.1	-84738934.6	NO	NO
20	OK	21	66964155.6	-84288832.7	NO	NO
21	OK	22	66848363.8	-84111902.8	YES	YES
22	OK	23	66870956.5	-84120754.3	NO	NO
23	OK	24	67009978.4	-84283927.4	NO	NO
24	OK	25	67217459.4	-84537134	NO	NO
25	OK	26	67475684.4	-84856250.1	NO	NO
26	OK	27	67778046.9	-85232376.4	NO	NO
27	OK	28	68124649.3	-85665062.7	NO	NO
28	OK	29	68507516.4	-86143970.1	NO	NO
29	OK	30	68929840.2	-86672850.7	NO	NO
30	OK	31	69424229.4	-87293448.6	NO	NO
31	OK	32	69922458.1	-87919391.6	NO	NO
32	OK	33	70413074.4	-88536533.9	NO	NO
33	OK	34	70908950.3	-89162042	NO	NO
34	OK	35	71355026.2	-89726400.8	NO	NO
35	OK	36	71811264.6	-90304757	NO	NO
36	OK	37	72190775.3	-90788158.2	NO	NO
37	OK	38	72566567.4	-91267593.6	NO	NO
38	OK	39	72912902.6	-91710562.5	NO	NO
39	OK	40	73219668.2	-92103385.5	NO	NO
40	OK	41	73547596.2	-92524719.7	NO	NO
41	OK	42	73868972.4	-92938172.2	NO	NO
42	OK	43	74194128	-93357099.3	NO	NO
43	OK	44	74509291.6	-93763523.7	NO	NO
44	OK	45	74824282.3	-94170268.7	NO	NO
45	OK	46	75150141.6	-94591276.5	NO	NO
46	OK	47	75480836	-95018961	NO	NO
47	OK	48	75814436.8	-95450898.2	NO	NO
48	OK	49	76160233.6	-95898892.6	NO	NO
49	OK	50	76514472.2	-96358207.8	NO	NO
50	OK	51	76883320.9	-96836639.8	NO	NO
51	OK	52	77268028.9	-97336068.4	NO	NO
52	OK	53	77664372.7	-97850855.5	NO	NO
53	OK	54	78086068.2	-98398835.5	NO	NO
54	OK	55	78518399.9	-98961014.8	NO	NO
55	OK	56	78969610.3	-99548099.1	NO	NO
56	OK	57	79449036	-100172184	NO	NO
57	OK	58	79954361.6	-100830301	NO	NO
58	OK	59	80481004.4	-101516820	NO	NO
59	OK	60	81043317.2	-102250051	NO	NO
60	OK	61	81651862.7	-103044001	NO	NO
61	OK	62	82287427.5	-103874069	NO	NO
62	OK	63	82962596.8	-104756454	NO	NO
63	OK	64	83690053.2	-105707712	NO	NO
64	OK	65	84474068.5	-106733880	NO	NO
65	OK	66	85323760.4	-107847311	NO	NO
66	OK	67	86271239.9	-109089274	NO	NO
67	OK	68	87281945.5	-110415951	NO	NO
68	OK	69	88409617.4	-111898206	NO	NO
69	OK	70	89676243.3	-113564407	NO	NO
70	OK	71	91130522.2	-115480203	NO	NO
71	OK	72	92774808.8	-117649581	NO	NO
72	OK	73	94663239.4	-120147062	NO	NO
73	OK	74	96976925.5	-123211640	NO	NO
74	OK	75	99798393.6	-126957946	NO	NO
			66848363.8	-84111902.8		

Section starting stage:	2	
Section ending stage:	74	
Stage with maximum diameter:	3	
Column diameter:	20.6826	ft
Downcomer area / Column area:	0.1	
Side downcomer velocity:	0.268303	ft/sec
Flow path length:	14.2099	ft
Side downcomer width:	3.23632	ft
Side weir length:	15.0282	ft

De-butanizer (T-2) Optimization Data – Example Calculations

# of Theoretical Stages	19
# of Computational Stages	20
# of Trays Inside the Column	18
For stages 2-29:	
Column Diameter (ft)	19
Column Diameter rounded (ft)	19
Column Diameter (m)	5.7912
Column Diameter rounded (m)	5.79
Tray Spacing/tray (ft)	2
$h_{Overhead}$ (ft)	3
h_{trays} (ft)	46
h_{Btm} (ft)	7
h_{skirt} (ft)	8
h_{total} (ft)	64
Volume (f3)	18145.83917
Volume (m3)	513.8320985
Column Pressure (psig)	110.3
	121.33
	160.3
Design Pressure (psig)	160.3
Design Pressure (barg)	11.05229592
Column Area (ft^2)	283.528737
Downcomer area/column area (ft2)	0.1
Downcomer Area (ft2)	28.3528737
Tray Area (ft2)	255.1758633
Tray Area (m2)	23.70641695
Overall Efficiency	0.8
Nactual	23
Capital Cost: Tower	
$F_{P,vessel}$	7.071574673
$C_p°$	318007.8051
K_1	3.4974
K_2	0.4485
K_3	0.1074
B_1	2.25
B_2	1.82
F_M	1
C_{BM} (2001)	4808362.573
C_{BM} (2016)	**$6,953,819.78**
Capital Cost: Trays (Sieve)	
$C_p°$	22777.60989
K_1	2.9949
K_2	0.4465
K_3	0.3961
N	18
F_Q	1
F_{BM}	1
C_{BM} (2001)	409996.9781
C_{BM} (2016)	**$592,934.72**
TOTAL COST	**$7,546,754.50**

# of Theoretical Stages	29
# of Computational Stages	30
# of Trays Inside the Column	28
For stages 2-29:	
Column Diameter (ft)	14
Column Diameter rounded (ft)	14
Column Diameter (m)	4.2672
Column Diameter rounded (m)	4.27
Tray Spacing/tray (ft)	2
$h_{Overhead}$ (ft)	3
h_{trays} (ft)	70
h_{Btm} (ft)	7
h_{skirt} (ft)	8
h_{total} (ft)	88
Volume (f3)	13546.54752
Volume (m3)	383.5948769
Column Pressure (psig)	110.3
	121.33
	160.3
Design Pressure (psig)	160.3
Design Pressure (barg)	11.05229592
Column Area (ft^2)	153.93804
Downcomer area/column area (ft2)	0.1
Downcomer Area (ft2)	15.393804
Tray Area (ft2)	138.544236
Tray Area (m2)	12.87107402
Overall Efficiency	0.8
Nactual	35
Capital Cost: Tower	
$F_{P,vessel}$	5.346394448
$C_p°$	236217.5516
K_1	3.4974
K_2	0.4485
K_3	0.1074
B_1	2.25
B_2	1.82
F_M	1
C_{BM} (2001)	2829989.707
C_{BM} (2016)	**$4,092,711.00**
Capital Cost: Trays (Sieve)	
$C_p°$	9506.904657
K_1	2.9949
K_2	0.4465
K_3	0.3961
N	28
F_Q	1
F_{BM}	1
C_{BM} (2001)	266193.3304
C_{BM} (2016)	**$384,966.90**
TOTAL COST	**$4,477,677.91**

De-butanizer (T-2) Reboiler (E-6) Optimization – Example Calculations

Reboiler (Shell & Tube HEX)	
Reboiler Duty (Btu/hr)	59747241.5
$T_{H,in}$ (°F)	275
$T_{H,out}$ (°F)	275
$T_{C,in}$ (°F)	219.88
$T_{C,out}$ (°F)	222.62
Approach Temp (°F)	20
LMTD	53.73835829
U (Btu/hr ft² °F)	140
F (correction factor)	0.9
Area (ft²)	**8824**
Area (m²)	819.776072
Column Pressure (psia)	110.3
Computational Stages	20
Trays Inside the Column	18
Overall Efficiency	0.85
Nactual	21
Final Reboiler Pressure (psia)	112.4
Reboiler Pressure (barg)	6.7154936
ΔT_{steam}	0
$T_{avg, steam}$	275
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	61569.70476
Capital Cost	
$C_p°$	1926139.372
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$6,400,498.74
C_{BM} (2016)	**$9,256,355.80**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	517776589.2
$ per lb steam	$0.00012
Operating Cost Per Year	**$62,133.19**
Total Capital + Operating	**$9,318,488.99**

Reboiler (Shell & Tube HEX)	
Reboiler Duty (Btu/hr)	57904840.4
$T_{H,in}$ (°F)	275
$T_{H,out}$ (°F)	275
$T_{C,in}$ (°F)	219.88
$T_{C,out}$ (°F)	222.62
Approach Temp (°F)	20
LMTD	53.73835829
U (Btu/hr ft² °F)	140
F (correction factor)	0.9
Area (ft²)	**8552**
Area (m²)	794.506456
Column Pressure (psia)	110.3
Computational Stages	30
Trays Inside the Column	28
Overall Efficiency	0.85
Nactual	33
Final Reboiler Pressure (psia)	113.6
Reboiler Pressure (barg)	6.798230688
ΔT_{steam}	0
$T_{avg, steam}$	275
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	59671.10511
Capital Cost	
$C_p°$	1822204.377
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$6,055,126.12
C_{BM} (2016)	**$8,756,880.37**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	501810125.5
$ per lb steam	$0.00012
Operating Cost Per Year	**$60,217.22**
Total Capital + Operating	**$8,817,097.58**

De-butanizer (T-2) Condenser (E-4) Optimization – Example Calculation

Shell and Tube Cooler (Floating Head)		**Shell and Tube Cooler (Floating Head)**		**Shell and Tube Cooler (Floating Head)**	
Cooler Duty (Btu/hr)	1.28E+08	**Cooler Duty (Btu/hr)**	1.20E+08	**Cooler Duty (Btu/hr)**	1.14E+08
Mass flow rate (lbmol/hr)	2.15E+03	**Mass flow rate (lbmol/hr)**	2.15E+03	**Mass flow rate (lbmol/hr)**	2.15E+03
Cp (Btu/lbmol F)	35.4392	**Cp (Btu/lbmol F)**	35.4392	**Cp (Btu/lbmol F)**	35.4392
$\Delta T_{Bottoms}$ (^{o}F)	1.88	**$\Delta T_{Bottoms}$ (^{o}F)**	1.88	**$\Delta T_{Bottoms}$ (^{o}F)**	1.88
$T_{H,in}$ (^{o}F)	160.63	**$T_{H,in}$ (^{o}F)**	160.63	**$T_{H,in}$ (^{o}F)**	160.63
$T_{H,out}$ (^{o}F)	158.75	**$T_{H,out}$ (^{o}F)**	158.75	**$T_{H,out}$ (^{o}F)**	158.75
$T_{C,in}$ (^{o}F)	87	**$T_{C,in}$ (^{o}F)**	87	**$T_{C,in}$ (^{o}F)**	87
$T_{C,out}$ (^{o}F)	120	**$T_{C,out}$ (^{o}F)**	120	**$T_{C,out}$ (^{o}F)**	120
LMTD	54.72310854	**LMTD**	54.72310854	**LMTD**	54.72310854
U (Btu/hr ft^{2} ^{o}F)	80	**U (Btu/hr ft^{2} ^{o}F)**	80	**U (Btu/hr ft^{2} ^{o}F)**	80
F (correction factor)	0.9	**F (correction factor)**	0.9	**F (correction factor)**	0.9
Area (ft^{2})	**32401**	**Area (ft^{2})**	**30427**	**Area (ft^{2})**	**28843**
Area (m^{2})	3010.150103	**Area (m^{2})**	2826.759581	**Area (m^{2})**	2679.601229
Pressure at condenser (psia)	125	**Pressure at condenser (psia)**	125	**Pressure at condenser (psia)**	125
Number of stages	20	**Number of stages**	30	**Number of stages**	40
N_{actual}	25	**N_{actual}**	37.5	**N_{actual}**	50
Pressure drop for stages	2.5	**Pressure drop for stages**	3.75	**Pressure drop for stages**	5
Pressure (psia)	127.5	**Pressure (psia)**	128.75	**Pressure (psia)**	130
Pressure (barg)	7.756601951	**Pressure (barg)**	7.842786417	**Pressure (barg)**	7.928970884
$\dot{m}_{water}$ (lbmol/hr)	215288.7238	**$\dot{m}_{water}$ (lbmol/hr)**	202168.9102	**$\dot{m}_{water}$ (lbmol/hr)**	191644.6634
$\dot{m}_{water}$ (lb/hr)	3875197.028	**$\dot{m}_{water}$ (lb/hr)**	3639040.383	**$\dot{m}_{water}$ (lb/hr)**	3449603.941
ρ_{water} (lb/ft3)	62.42	**ρ_{water} (lb/ft3)**	62.42	**ρ_{water} (lb/ft3)**	62.42
$\dot{m}_{water}$ (gpm)	7740.170902	**$\dot{m}_{water}$ (gpm)**	7268.480617	**$\dot{m}_{water}$ (gpm)**	6890.107486
Cp (Btu/lbmol F)	17.9691	**Cp (Btu/lbmol F)**	17.9691	**Cp (Btu/lbmol F)**	17.9691
ΔT_{water} (^{o}F)	33	**ΔT_{water} (^{o}F)**	33	**ΔT_{water} (^{o}F)**	33
Capital Cost (Shell and Tube - Floating Head)		**Capital Cost (Shell and Tube - Floating Head)**		**Capital Cost (Shell and Tube - Floating Head**	
C_p°	533485.6769	C_p°	489846.3176	C_p°	455943.212
K_1	4.8306	K_1	4.8306	K_1	4.8306
K_2	-0.8509	K_2	-0.8509	K_2	-0.8509
K_3	0.3187	K_3	0.3187	K_3	0.3187
B_1	1.63	B_1	1.63	B_1	1.63
B_2	1.66	B_2	1.66	B_2	1.66
F_M	1	F_M	1	F_M	1
Pressure (psia)	127.5	Pressure (psia)	128.75	Pressure (psia)	130
Design Pressure (psia)	177.5	Design Pressure (psia)	178.75	Design Pressure (psia)	180
Pressure (barg)	11.2039806	Pressure (barg)	11.29016506	Pressure (barg)	11.37634953
$F_{P,vessel}$	1.012336692	$F_{P,vessel}$	1.012488636	$F_{P,vessel}$	1.012640075
C_1	-0.00164	C_1	-0.00164	C_1	-0.00164
C_2	-0.00627	C_2	-0.00627	C_2	-0.00627
C_3	0.0123	C_3	0.0123	C_3	0.0123
C_{BM} (2001)	$1,766,093.08	C_{BM} (2001)	$1,621,749.46	C_{BM} (2001)	$1,509,620.01
C_{BM} (2016)	**$2,554,111.26**	**C_{BM} (2016)**	**$2,345,362.54**	**C_{BM} (2016)**	**$2,183,201.73**
Operating Cost		**Operating Cost**		**Operating Cost**	
Converting to per min basis	504576	Converting to per min basis	504576	Converting to per min basis	504576
Service Factor	0.96	Service Factor	0.96	Service Factor	0.96
Cooling water (gpm)	7740.170902	Cooling water (gpm)	7268.480617	Cooling water (gpm)	6890.107486
$/ gpm year cooling water	120	$/ gpm year cooling water	120	$/ gpm year cooling water	120
$/ gpm	2.28E-04	$/ gpm	2.28E-04	$/ gpm	2.28E-04
Operating Cost Per Year	**$891,667.69**	**Operating Cost Per Year**	**$803,835.81**	**Operating Cost Per Year**	**$761,990.77**
Total Capital + Operating	**$3,445,778.95**	**Total Capital + Operating**	**$3,149,198.35**	**Total Capital + Operating**	**$2,945,192.49**

De-butanizer (T-2) Optimization Data

# of Computational Stages	**Reboiler Duty (BTU/hr)**	**Condenser Duty (BTU/hr)**	**Optimum Feed Stage**	**Column PWC**	**Reboiler PWC**	**Condenser PWC**	**Total PWC**	
20	59747241.5	1.28E+08	3	$7,546,754	$9,318,489	$3,445,779	$20,311,022	NO
25	58567028	1.09E+08	4	$7,765,869	$8,996,005	$2,800,325	$19,562,199	NO
26	58396044.8	1.07E+08	4	$6,781,333	$8,950,177	$2,724,795	$18,456,305	NO
27	58247711.8	1.09E+08	4	$5,877,146	$8,909,957	$2,799,485	$17,586,587	NO
30	57904840.4	1.20E+08	4	$4,477,678	$8,817,098	$3,149,198	$16,443,974	NO
31	52891543.7	5.42E+07	4	$3,746,059	$7,528,711	$1,240,438	$12,515,208	YES
40	57286229.9	1.14E+08	5	$5,845,816	$8,652,713	$2,945,192	$17,443,721	NO
15	6.44E+07	6.44E+07	4	$3,042,626	$10,654,707	$1,501,670	$15,199,003	NO
31				$3,746,059				
						Minimum	**$12,515,208**	

15* - New Design with 25 stages col 1, 15 stages col 2 = more expensive
31* (last) - New Design with 95 ^{0}F Recycle Feed

Table 36: *Present Worth Cost of Computational Stage and Feed Location for T-2*

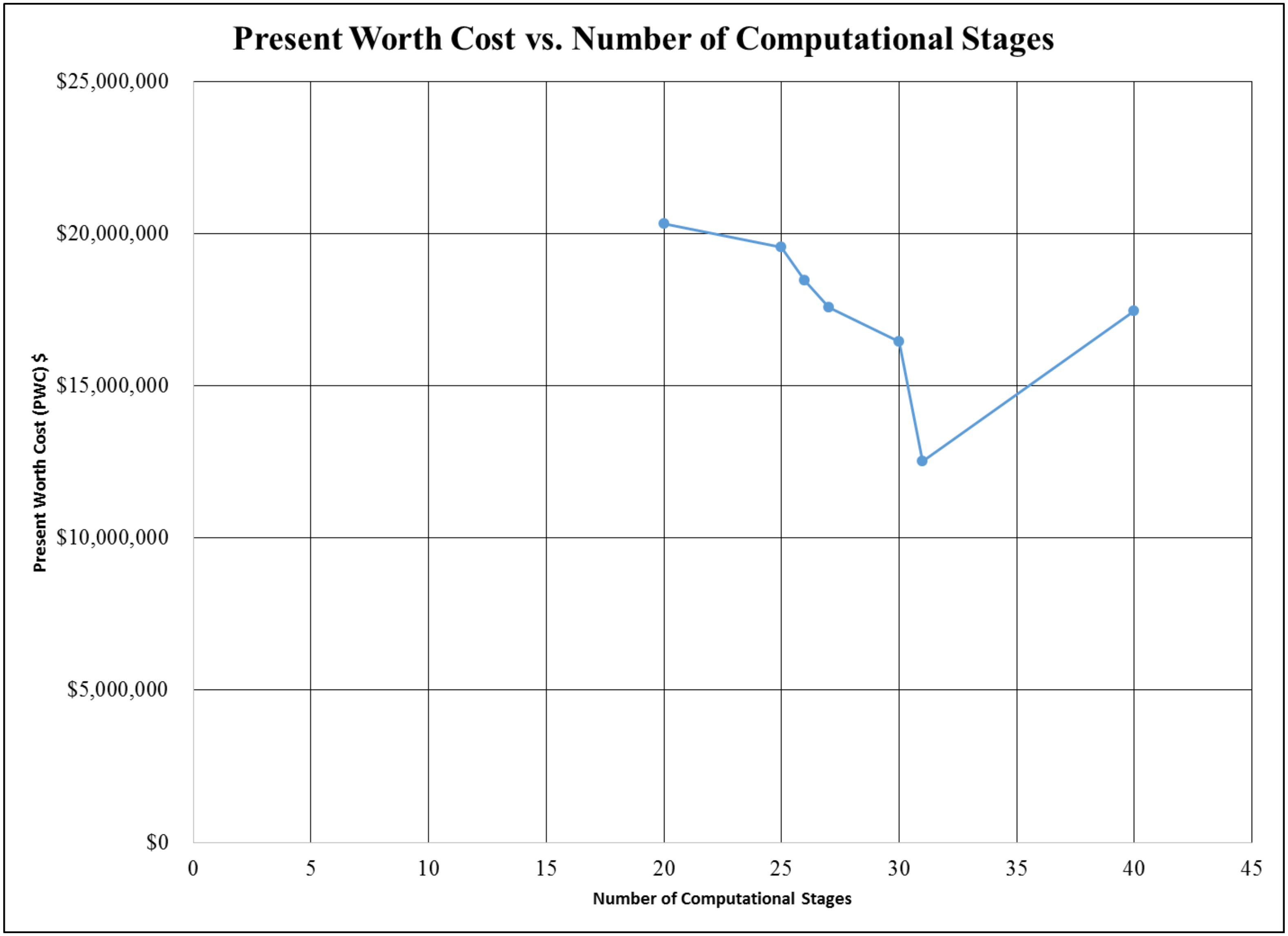


Figure 6: *Present Worth Cost vs. Number of Computational Stages for T-2*

Optimum Feed Stage location and Diameters for De-butanizer, T-2 Optimization Using Sensitivity Analysis on Aspen Plus

20 stages – Example Calculations

Row/Case	Status					
		VARY 1	REBOILER	CONDENS		
		B2				
		S10				
		FEEDS				
		STAGE				
			BTU/HR	BTU/HR		
1	OK	2	59828920.2	-61302086	NO	NO
2	OK	3	59747241.5	-61217793	YES	YES
3	OK	4	59803392.2	-61274496	NO	NO
4	OK	5	59932849.6	-61406375	NO	NO
5	OK	6	60119832.9	-61597420	NO	NO
6	OK	7	60359442.9	-61842795	NO	NO
7	OK	8	60662466.8	-62153737	NO	NO
8	OK	9	61048968.1	-62551070	NO	NO
9	OK	10	61567888.8	-63085353	NO	NO
10	OK	11	62335242.2	-63876315	NO	NO
11	OK	12	63658616.4	-65241085	NO	NO
12	OK	13	66743411.9	-68421058	NO	NO
13	OK	14	75893008.1	-77843293	NO	NO
			59747241.5	-61217793		

Section starting stage:	2	
Section ending stage:	19	
Stage with maximum diameter:	6	
Column diameter:	18.3073	ft
Downcomer area / Column area:	0.1	
Side downcomer velocity:	0.206955	ft/sec
Flow path length:	12.578	ft
Side downcomer width:	2.86466	ft
Side weir length:	13.3023	ft

	Fenske's Equation Nmin: Aspen Report K values (De-Isobutanizer Tower)											
	Water	**Propane**	**1-Butene**	**i-Butene**	**Trans-2-Butene**	**Cis-2-Butene**	**i-Butane**	**n-Butane**	**2,2-Dimethyl Propane**	**i-Pentane**	**n-Pentane**	**n-Hexane**
	0.77092	1.8135	0.91305	0.92601	0.80035	0.75739	0.99144	0.79488	0.61714	0.42556	0.35212	0.15544
	0.781	1.8317	0.92425	0.93726	0.81062	0.76716	1.0035	0.80486	0.62547	0.43178	0.35778	0.1583
	0.78692	1.8446	0.93154	0.94456	0.81723	0.77341	1.0114	0.81135	0.63091	0.43585	0.36179	0.16005
	0.79075	1.8553	0.93698	0.94997	0.82206	0.77796	1.0173	0.81619	0.63498	0.43891	0.36506	0.16124
	0.83387	2.0629	1.0274	1.0393	0.89998	0.85014	1.1158	0.8969	0.70394	0.49046	0.42818	0.17801
	0.83587	2.0734	1.032	1.0439	0.90392	0.85379	1.1209	0.90104	0.70747	0.49303	0.4311	0.17892
	0.8419	2.0964	1.0423	1.0541	0.91293	0.86219	1.1319	0.91034	0.71529	0.49905	0.43848	0.18099
	0.85014	2.1247	1.0552	1.0669	0.92436	0.87287	1.1458	0.92212	0.72509	0.50659	0.44731	0.18375
	0.85955	2.1575	1.0707	1.0823	0.93804	0.88565	1.1625	0.93626	0.73693	0.51544	0.45693	0.18719
K_{avg}	0.817	1.984	0.993	1.005	0.870	0.822	1.078	0.866	0.677	0.471	0.404	0.172

x_D	0.92	mole fraction of more volatile component (i-butane) in the overhead distillate
x_B	0.08	mole fraction of more volatile component (i-butane) in the bottoms
$\alpha_{i\text{-}but:n\text{-}but}$	1.245	Kavg of isobutane/Kavg of n-butane
N_{min}	**23**	

Table 37: *Fenske's Equation (De-isobutanizer T-1)*

	Aspen Report K values (De-Butanizer Tower)											
	Water	**Propane**	**1-Butene**	**i-Butene**	**Trans-2-Butene**	**Cis-2-Butene**	**i-Butane**	**n-Butane**	**2,2-Dimethyl Propane**	**i-Pentane**	**n-Pentane**	**n-Hexane**
	0.85241	2.1353	1.0614	1.0728	0.92983	0.87797	1.1522	0.92762	0.73051	0.51118	0.45389	0.18548
	0.85914	2.1652	1.0749	1.0862	0.94159	0.88887	1.167	0.93982	0.74096	0.51895	0.46304	0.18824
	0.86385	2.1859	1.0843	1.0956	0.94979	0.89647	1.1773	0.94835	0.74824	0.52433	0.4692	0.19021
	0.86762	2.2012	1.0913	1.1026	0.95595	0.90221	1.185	0.95478	0.75368	0.52833	0.47352	0.19175
	0.87159	2.2243	1.1014	1.1127	0.9647	0.91025	1.1963	0.96393	0.76167	0.53414	0.48052	0.19373
	1.5395	3.3752	1.7988	1.8238	1.6294	1.5445	1.9296	1.6416	1.3098	0.96867	0.83511	0.43899
	1.5667	3.4022	1.8209	1.8461	1.6513	1.5659	1.9514	1.663	1.3276	0.98455	0.84949	0.44906
K_{avg}	1.060	2.527	1.290	1.306	1.146	1.084	1.394	1.148	0.910	0.653	0.575	0.262

x_D	0.99	mole fraction of more volatile component (n-butane) in the overhead distillate
x_B	0.01	mole fraction of more volatile component (n-butane) in the bottoms
$\alpha_{n\text{-}but:2,2\text{-}Dim}$	1.262	Kavg of n-butane/Kavg of 2,2-dimethyl propane
N_{min}	**40**	

Reference: Fenske's Equation From Design of Equilibrium Stage Separation Processes: Single Stage Flash Continuous Distillation, Attachment C, Dr. Wagner & Dr. Whiteley

Table 38: *Fenske's Equation (De-butanizer T-2)*

E-2 Reboiler Steam Temperature Optimization – Example Calculations

Reboiler (Shell & Tube HEX) for De-Isobutaniz	
Reboiler Duty (Btu/hr)	7.32E+07
$T_{H,in}$ (^oF)	550
$T_{H,out}$ (^oF)	550
$T_{C,in}$ (^oF)	168.955
$T_{C,out}$ (^oF)	170.577
Approach Temp (^oF)	20
LMTD	380.2334234
U (Btu/hr ft^2 ^oF)	140
F (correction factor)	0.9
Area (ft^2)	**1529**
Area (m^2)	142.048687
Column Pressure (psia)	150
Computational Stages	26
Trays Inside the Column	24
Overall Efficiency	0.8
Nactual	30
Final Reboiler Pressure (psia)	153
Reboiler Pressure (barg)	9.51476506
ΔT_{steam}	0
$T_{avg, steam}$	550
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	75475.57708
Capital Cost	
$C_p°$	144912.1637
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$481,538.43
C_{BM} (2016)	**$696,397.45**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	634719413
$ per lb steam	$0.00330
Operating Cost Per Year	**$2,094,574.06**
Total	**$2,790,971.51**

Reboiler (Shell & Tube HEX) for De-Isobut	
Reboiler Duty (Btu/hr)	7.32E+07
$T_{H,in}$ (^oF)	287
$T_{H,out}$ (^oF)	287
$T_{C,in}$ (^oF)	168.955
$T_{C,out}$ (^oF)	170.577
Approach Temp (^oF)	20
LMTD	117.2321299
U (Btu/hr ft^2 ^oF)	140
F (correction factor)	0.9
Area (ft^2)	**4959**
Area (m^2)	460.705977
Column Pressure (psia)	150
Computational Stages	26
Trays Inside the Column	24
Overall Efficiency	0.8
Nactual	30
Final Reboiler Pressure (psi	153
Reboiler Pressure (barg)	9.51476506
ΔT_{steam}	0
$T_{avg, steam}$	287
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	75475.57708
Capital Cost	
$C_p°$	732351.6278
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$2,433,580.74
C_{BM} (2016)	**$3,519,427.17**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	634719413
$ per lb steam	$0.00330
Operating Cost Per Year	**$2,094,574.06**
Total	**$5,614,001.23**

E-6 Reboiler Steam Temperature Optimization – Example Calculations

Reboiler (Shell & Tube HEX) for De-Butaniz	
Reboiler Duty (Btu/hr)	5.24E+07
$T_{H,in}$ (°F)	550
$T_{H,out}$ (°F)	550
$T_{C,in}$ (°F)	238.68
$T_{C,out}$ (°F)	240.671
Approach Temp (°F)	20
LMTD	310.3234355
U (Btu/hr ft^2 °F)	140
F (correction factor)	0.9
Area (ft^2)	**1339**
Area (m^2)	124.397117
Column Pressure (psia)	151
Computational Stages	31
Trays Inside the Column	29
Overall Efficiency	0.85
Nactual	34
Final Reboiler Pressure (psia	154.4
Reboiler Pressure (barg)	9.611291662
ΔT_{steam}	0
$T_{avg,\ steam}$	550
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	53946.9291
Capital Cost	
$C_p°$	124369.4356
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$413,275.61
C_{BM} (2016)	**$597,676.24**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	453672095
$ per lb steam	$0.00330
Operating Cost Per Year	**$1,497,117.91**
Total	**$2,094,794.16**

Reboiler (Shell & Tube HEX) for De-Butaniz	
Reboiler Duty (Btu/hr)	5.24E+07
$T_{H,in}$ (°F)	775
$T_{H,out}$ (°F)	775
$T_{C,in}$ (°F)	238.68
$T_{C,out}$ (°F)	240.671
Approach Temp (°F)	20
LMTD	535.3238829
U (Btu/hr ft^2 °F)	140
F (correction factor)	0.9
Area (ft^2)	**777**
Area (m^2)	72.185631
Column Pressure (psia)	151
Computational Stages	31
Trays Inside the Column	29
Overall Efficiency	0.85
Nactual	34
Final Reboiler Pressure (psia	154.4
Reboiler Pressure (barg)	9.611291662
ΔT_{steam}	0
$T_{avg,\ steam}$	775
λ_{steam} (Btu/lbm)	970.4
$\dot{m}_{steam}$ (lb/hr)	53946.9291
Capital Cost	
$C_p°$	70776.47773
K_1	4.4646
K_2	-0.5277
K_3	0.3955
B_1	1.63
B_2	1.66
F_M	1
Pressure (psia)	194.7
Design Pressure (psia)	244.7
Pressure (barg)	15.8372575
$F_{P,vessel}$	1.019860005
C_1	-0.00164
C_2	-0.00627
C_3	0.0123
C_{BM} (2001)	$235,187.94
C_{BM} (2016)	**$340,127.13**
Operating Cost	
Hours/yr	8409.6
Service Factor	0.96
Mass of steam/yr	453672095
$ per lb steam	$0.00330
Operating Cost Per Year	**$1,497,117.91**
Total	**$1,837,245.05**

Reactor (R-1) Optimization

T (°F)	300
F(T)	0.666
iC4	706.55
nC4	1316.4
iC4/nC4	0.536728958
% Conv	48.67325281
Pressure (psig)	450
Pdrop (psi)	20
Volume (ft3)	900
Volume (m3)	25.48512
Actual Height (ft)	23.5
Actual Diameter (ft)	6.95
Actual Diameter (m)	2.11836
Actual Volume (ft3)	891.5123275
Actual Volume (m3)	25.24477628

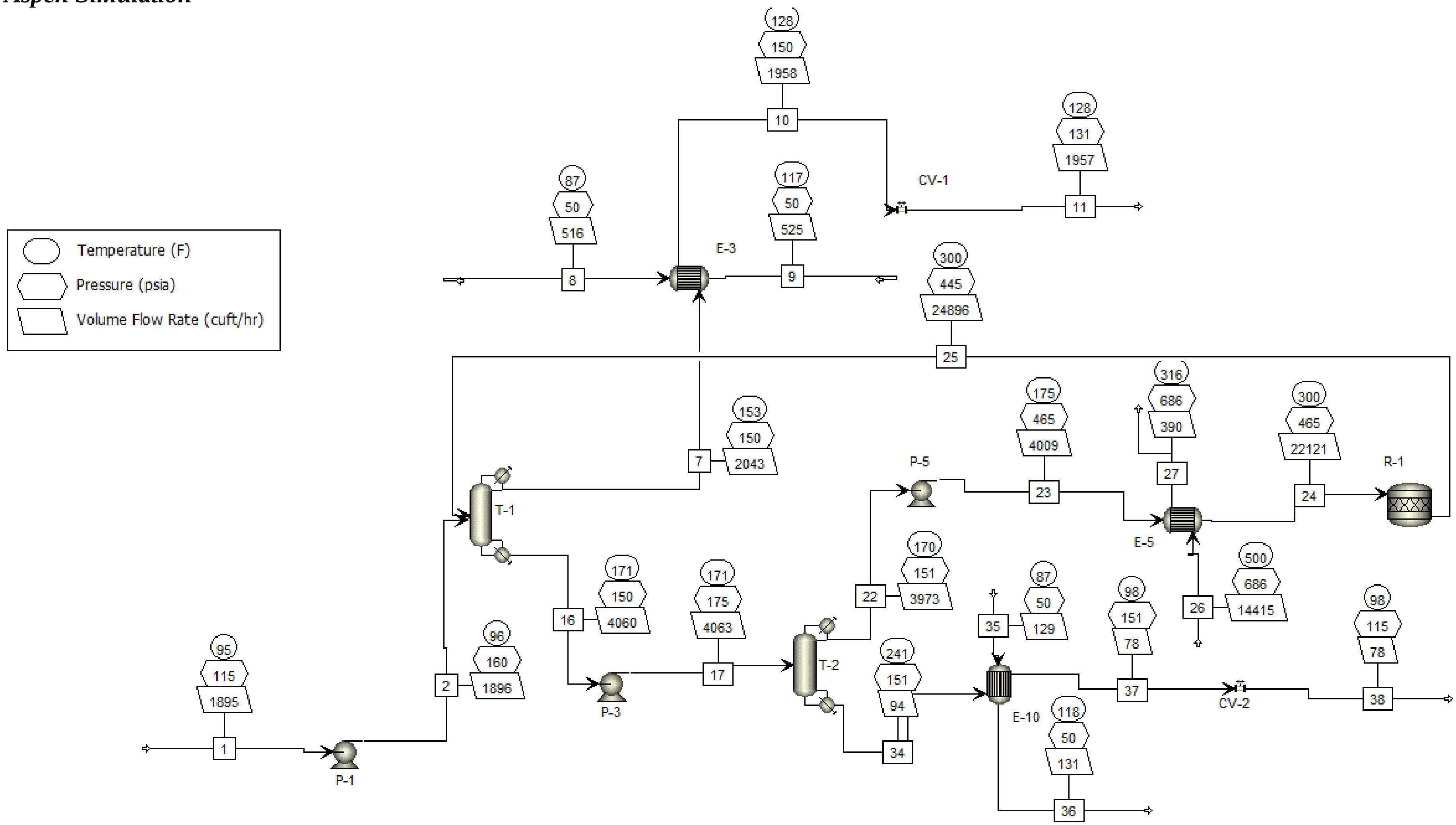
T (°F)	310
F(T)	0.6593
iC4	706.55
nC4	1316.4
iC4/nC4	0.536728958
% Conv	47.64364441
Pressure (psig)	450
Pdrop (psi)	20
Volume (ft3)	900
Volume (m3)	25.48512
Actual Height (ft)	23.5
Actual Diameter (ft)	6.95
Actual Diameter (m)	2.11836
Actual Volume (ft3)	891.5123275
Actual Volume (m3)	25.24477628

T (°F)	400
F(T)	0.599
iC4	706.55
nC4	1316.4
iC4/nC4	0.536728958
% Conv	38.37716879
Pressure (psig)	450
Pdrop (psi)	20
Volume (ft3)	900
Volume (m3)	25.48512
Actual Height (ft)	23.5
Actual Diameter (ft)	6.95
Actual Diameter (m)	2.11836
Actual Volume (ft3)	891.5123275
Actual Volume (m3)	25.24477628

NOTE: The highest conversion occurs at 300 °F and produces the lowest diameters for both columns

Computer Printout(s)

Aspen Simulation



Aspen Stream Summary Table

	Units	1	2	7	8	9	10	11	16	17	22	23	24	25	26	27	34	35	36	37	38
From			P-1	T-1		E-3	E-3	CV-1	T-1	P-3	T-2	P-5	E-5	R-1		E-5	T-2		E-10	E-10	CV-2
To		P-1	T-1	E-3	E-3		CV-1		P-3	T-2	P-5	E-5	R-1	T-1	E-5		E-10	E-10		CV-2	
Substream: MIXED																					
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Vapor	Vapor	Vapor	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Component Mole Flow																					
WATER	LBMOL/HR	0.8900109	0.8900109	0.4185526	1767.099	1767.099	0.418553	0.418553	23.58996	23.58996	23.11837	23.11837	23.11837	23.1185	1155.519	1155.519	0.471585	441.636	441.636	0.471585	0.471585
PROPA-01	LBMOL/HR	26.79723	26.79723	26.79721	0	0	26.79721	26.79721	2.583569	2.583569	2.583569	2.583569	2.583569	2.583557	0	0	5.61E-14	0	0	5.61E-14	5.61E-14
1-BUT-01	LBMOL/HR	1.21728	1.21728	1.217254	0	0	1.217254	1.217254	2.314637	2.314637	2.314611	2.314611	2.314611	2.31461	0	0	2.53E-05	0	0	2.53E-05	2.53E-05
ISOBU-01	LBMOL/HR	2.430656	2.430656	2.430625	0	0	2.430625	2.430625	4.023942	4.023942	4.023911	4.023911	4.023911	4.023911	0	0	3.11E-05	0	0	3.11E-05	3.11E-05
TRANS-01	LBMOL/HR	1.234841	1.234841	1.226315	0	0	1.226315	1.226315	19.88509	19.88509	19.87661	19.87661	19.87661	19.87656	0	0	0.008474	0	0	0.008474	0.008474
CIS-2-01	LBMOL/HR	1.269639	1.269639	1.124031	0	0	1.124031	1.124031	58.56429	58.56429	58.42409	58.42409	58.42409	58.41868	0	0	0.140202	0	0	0.140202	0.140202
ISOBU-02	LBMOL/HR	352.1107	352.1107	991.8807	0	0	991.8807	991.8807	707.4365	707.4365	707.4358	707.4358	707.4358	1347.207	0	0	0.00064	0	0	0.00064	0.00064
N-BUT-01	LBMOL/HR	700.7218	700.7218	60.4781	0	0	60.4781	60.4781	1316.871	1316.871	1316.396	1316.396	1316.396	676.6277	0	0	0.475329	0	0	0.475329	0.475329
2:2-D-01	LBMOL/HR	0.9392821	0.9392821	0.00106285	0	0	0.001063	0.001063	5.016523	5.016523	4.078327	4.078327	4.078327	4.078304	0	0	0.938196	0	0	0.938196	0.938196
2-MET-01	LBMOL/HR	37.86619	37.86619	8.60E-07	0	0	8.60E-07	8.60E-07	43.04441	43.04441	5.17819	5.17819	5.17819	5.178214	0	0	37.86622	0	0	37.86622	37.86622
N-PEN-01	LBMOL/HR	0.9934012	0.9934012	4.52E-10	0	0	4.52E-10	4.52E-10	1.084162	1.084162	0.090761	0.090761	0.090761	0.090761	0	0	0.993401	0	0	0.993401	0.993401
N-HEX-01	LBMOL/HR	0.4376635	0.4376635	1.40E-19	0	0	1.40E-19	1.40E-19	0.439724	0.439724	0.00206	0.00206	0.00206	0.00206	0	0	0.437664	0	0	0.437664	0.437664
HYDRO-01	LBMOL/HR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PLATI-01	LBMOL/HR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Component Std Volume Flow																					
WATER	BBL/DAY	1.099986	1.099986	0.5172993	2184	2184	0.517299	0.517299	29.1554	29.1554	28.57255	28.57255	28.57255	28.57271	1428.134	1428.134	0.582844	545.8287	545.8287	0.582844	0.582844
PROPA-01	BBL/DAY	159.8979	159.8979	159.8979	0	0	159.8979	159.8979	15.41605	15.41605	15.41605	15.41605	15.41605	15.41598	0	0	3.35E-13	0	0	3.35E-13	3.35E-13
1-BUT-01	BBL/DAY	7.799899	7.799899	7.799732	0	0	7.799732	7.799732	14.83137	14.83137	14.83121	14.83121	14.83121	14.8312	0	0	0.000162	0	0	0.000162	0.000162
ISOBU-01	BBL/DAY	15.5998	15.5998	15.5996	0	0	15.5996	15.5996	25.82541	25.82541	25.82521	25.82521	25.82521	25.82521	0	0	0.0002	0	0	0.0002	0.0002
TRANS-01	BBL/DAY	7.799899	7.799899	7.746044	0	0	7.746044	7.746044	125.6046	125.6046	125.5511	125.5511	125.5511	125.5508	0	0	0.053526	0	0	0.053526	0.053526
CIS-2-01	BBL/DAY	7.799899	7.799899	6.905374	0	0	6.905374	6.905374	359.7837	359.7837	358.9224	358.9224	358.9224	358.8892	0	0	0.861316	0	0	0.861316	0.861316
ISOBU-02	BBL/DAY	2495.968	2495.968	7031.034	0	0	7031.034	7031.034	5014.727	5014.727	5014.722	5014.722	5014.722	9549.794	0	0	0.004537	0	0	0.004537	0.004537
N-BUT-01	BBL/DAY	4784.237	4784.237	412.9195	0	0	412.9195	412.9195	8991.054	8991.054	8987.81	8987.81	8987.81	4619.734	0	0	3.24535	0	0	3.24535	3.24535
2:2-D-01	BBL/DAY	7.799899	7.799899	0.00882601	0	0	0.008826	0.008826	41.65775	41.65775	33.86687	33.86687	33.86687	33.86668	0	0	7.790876	0	0	7.790876	7.790876
2-MET-01	BBL/DAY	300.2961	300.2961	6.82E-06	0	0	6.82E-06	6.82E-06	341.3618	341.3618	41.06541	41.06541	41.06541	41.0656	0	0	300.2963	0	0	300.2963	300.2963
N-PEN-01	BBL/DAY	7.799899	7.799899	3.55E-09	0	0	3.55E-09	3.55E-09	8.512528	8.512528	0.71263	0.71263	0.71263	0.712628	0	0	7.799899	0	0	7.799899	7.799899
N-HEX-01	BBL/DAY	3.89995	3.89995	1.25E-18	0	0	1.25E-18	1.25E-18	3.918309	3.918309	0.018358	0.018358	0.018358	0.018359	0	0	3.899951	0	0	3.899951	3.899951
HYDRO-01	BBL/DAY	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PLATI-01	BBL/DAY																				
Volume Flow	BBL/DAY	7799.998231	7799.998231	7642.428282	2184	2184	7642.428	7642.428	14971.85	14971.85	14647.31	14647.31	14647.31	14814.28	1428.134	1428.134	324.535	545.8287	545.8287	324.535	324.535
Component Std. Vol. Fraction																					
WATER		0.000141024	0.000141024	6.77E-05	1	1	6.77E-05	6.77E-05	0.001947	0.001947	0.001951	0.001951	0.001951	0.001929	1	1	0.001796	1	1	0.001796	0.001796
PROPA-01		0.0204997	0.0204997	0.0209223	0	0	0.020922	0.020922	0.00103	0.00103	0.001052	0.001052	0.001052	0.001041	0	0	1.03E-15	0	0	1.03E-15	1.03E-15
1-BUT-01		0.000999987	0.000999987	0.00102058	0	0	0.001021	0.001021	0.000991	0.000991	0.001013	0.001013	0.001013	0.001001	0	0	4.99E-07	0	0	4.99E-07	4.99E-07
ISOBU-01		0.00199997	0.00199997	0.00204118	0	0	0.002041	0.002041	0.001725	0.001725	0.001763	0.001763	0.001763	0.001743	0	0	6.16E-07	0	0	6.16E-07	6.16E-07
TRANS-01		0.000999987	0.000999987	0.00101356	0	0	0.001014	0.001014	0.008389	0.008389	0.008572	0.008572	0.008572	0.008475	0	0	0.000165	0	0	0.000165	0.000165
CIS-2-01		0.000999987	0.000999987	0.000903557	0	0	0.000904	0.000904	0.024031	0.024031	0.024504	0.024504	0.024504	0.024226	0	0	0.002654	0	0	0.002654	0.002654
ISOBU-02		0.3199959	0.3199959	0.92	0	0	0.92	0.92	0.334944	0.334944	0.342365	0.342365	0.342365	0.644635	0	0	1.40E-05	0	0	1.40E-05	1.40E-05
N-BUT-01		0.6133639	0.6133639	0.0540298	0	0	0.05403	0.05403	0.600531	0.600531	0.613615	0.613615	0.613615	0.311843	0	0	0.01	0	0	0.01	0.01
2:2-D-01		0.000999987	0.000999987	1.15E-06	0	0	1.15E-06	1.15E-06	0.002782	0.002782	0.002312	0.002312	0.002312	0.002286	0	0	0.024006	0	0	0.024006	0.024006
2-MET-01		0.0384995	0.0384995	8.93E-10	0	0	8.93E-10	8.93E-10	0.0228	0.0228	0.002804	0.002804	0.002804	0.002772	0	0	0.925313	0	0	0.925313	0.925313
N-PEN-01		0.000999987	0.000999987	4.65E-13	0	0	4.65E-13	4.65E-13	0.000569	0.000569	4.87E-05	4.87E-05	4.87E-05	4.81E-05	0	0	0.024034	0	0	0.024034	0.024034
N-HEX-01		0.000499994	0.000499994	1.63E-22	0	0	1.63E-22	1.63E-22	0.000262	0.000262	1.25E-06	1.25E-06	1.25E-06	1.24E-06	0	0	0.012017	0	0	0.012017	0.012017
HYDRO-01		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Mole Flow	LBMOL/HR	1126.909	1126.909	1085.574	1767.099	1767.099	1085.574	1085.574	2184.854	2184.854	2143.522	2143.522	2143.522	2143.519	1155.519	1155.519	41.33176	441.636	441.636	41.33176	41.33176
Mass Flow	LB/HR	65646.31	65646.31	62692.5	31834.78	31834.78	62692.5	62692.5	126540	126540	123586	123586	123586	123586	20817	20817	2953.646	7956.197	7956.197	2953.646	2953.646
Temperature	F	95	95.56337	153.1823	87	117.4671	128	127.9645	170.5685	170.9778	170.1224	175.1215	300	300	500	316.306	240.6726	87	117.5194	98	98.13164
Pressure	PSIA	114.6959	160	150	50	50	150	131.122	150	175	151	464.6959	464.6959	444.6959	686.3496	686.3496	151	50	50	151	114.7
Vapor Fraction		0	0	0	0	0	0	0	0	0	0	0	1	1	1	0	0	0	0	0	0
Liquid Fraction		1	1	1	1	1	1	1	1	1	1	1	0	0	0	1	1	1	1	1	1
Solid Fraction		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Molar Enthalpy	BTU/LBMOL	-63750.03	-63727.94	-63026.51	-123510	-122920	-63991.8	-63991.8	-59920.9	-59908.9	-59682.3	-59531.2	-50110.2	-51057.4	-101460	-118940	-69850.6	-123510	-122920	-76197.6	-76197.6
Mass Enthalpy	BTU/LB	-1094.356	-1093.977	-1091.358	-6855.854	-6822.94	-1108.07	-1108.07	-1034.61	-1034.4	-1035.15	-1032.53	-869.131	-885.56	-5631.86	-6601.94	-977.452	-6855.85	-6822.88	-1066.27	-1066.27
Enthalpy Flow	BTU/HR	-71840000	-71816000	-68420000	-218250000	-2.2E+08	-6.9E+07	-6.9E+07	-1.3E+08	-1.3E+08	-1.3E+08	-1.3E+08	-1.1E+08	-1.1E+08	-1.2E+08	-1.4E+08	-2887000	-5.5E+07	-5.4E+07	-3149400	-3149400
Molar Entropy	BTU/LBMOL-R	-104.4147	-104.399	-102.653	-39.76135	-38.7058	-104.261	-104.251	-98.1727	-98.167	-97.8839	-97.81	-84.5836	-85.3351	-14.1812	-32.8493	-118.116	-39.7614	-38.704	-128.198	-128.177
Mass Entropy	BTU/LB-R	-1.79242	-1.792151	-1.777523	-2.20709	-2.1485	-1.80537	-1.80519	-1.69507	-1.69497	-1.69774	-1.69646	-1.46705	-1.48009	-0.78718	-1.82342	-1.65285	-2.20709	-2.1484	-1.79394	-1.79364
Molar Density	LBMOL/CUFT	0.5946773	0.5942546	0.5313876	3.425711	3.368246	0.554557	0.554588	0.538171	0.537784	0.539463	0.534655	0.096898	0.086099	0.080159	2.959261	0.440817	3.425711	3.368146	0.529069	0.528999
Mass Density	LB/CUFT	34.642	34.61738	30.68793	61.71515	60.6799	32.02597	32.02777	31.16907	31.14669	31.10302	30.82582	5.586717	4.964065	1.444091	53.31192	31.50163	61.71515	60.6781	37.80827	37.80329
Average Molecular Weight		58.25345	58.25345	57.75056	18.01528	18.01528	57.75056	57.75056	57.91669	57.91669	57.65551	57.65551	57.65551	57.65552	18.01528	18.01528	71.46188	18.01528	18.01528	71.46188	71.46188
Liq Vol 60F cuft/hr		1824.74	1824.74	1787.877	510.9271	510.9271	1787.877	1787.877	3502.528	3502.528	3426.606	3426.606	3426.606	3465.666	334.099	334.099	75.92203	127.6917	127.6917	75.92203	75.92203

Table 39: *Aspen Stream Summary Table*

Report A-2: Aspen Complete Simulation Report

```
BLOCK:  CV-1     MODEL: VALVE
 -----------------------------
   INLET STREAM:          10
   OUTLET STREAM:         11
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           1085.57         1085.57
0.00000
       MASS(LB/HR   )           62692.5         62692.5
0.00000
       ENTHALPY(BTU/HR  )      -0.694678E+08   -0.694678E+08
0.00000

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR

                          ***  INPUT DATA  ***

    VALVE OUTLET PRESSURE     PSIA                          131.122
    VALVE FLOW COEF CALC.                                   NO

                          FLASH SPECIFICATIONS:
    NPHASE                                                    2
    MAX NUMBER OF ITERATIONS                                 50
    CONVERGENCE TOLERANCE
0.000100000

                          ***  RESULTS  ***

    VALVE PRESSURE DROP       PSI                            18.8780

 BLOCK:  CV-2     MODEL: VALVE
 -----------------------------
   INLET STREAM:          37
   OUTLET STREAM:         38
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
```

```
      MOLE(LBMOL/HR)           41.3318         41.3318
0.00000
      MASS(LB/HR   )           2953.65         2953.65
0.00000
      ENTHALPY(BTU/HR  )      -0.314938E+07   -0.314938E+07
0.00000

                  ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR

                      ***  INPUT DATA  ***

    VALVE OUTLET PRESSURE     PSIA                       114.700
    VALVE FLOW COEF CALC.                                NO

                      FLASH SPECIFICATIONS:
    NPHASE                                                 2
    MAX NUMBER OF ITERATIONS                              50
    CONVERGENCE TOLERANCE
0.000100000

                      ***  RESULTS  ***

    VALVE PRESSURE DROP       PSI                         36.3000

 BLOCK:  E-3      MODEL: HEATX
 -----------------------------
   HOT SIDE:
   ---------
   INLET STREAM:           7
   OUTLET STREAM:          10
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE
   COLD SIDE:
   ----------
   INLET STREAM:           8
   OUTLET STREAM:          9
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                  ***  MASS AND ENERGY BALANCE  ***
                                     IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
      MOLE(LBMOL/HR)           2852.67         2852.67
0.00000
      MASS(LB/HR   )           94527.3         94527.3
0.00000
      ENTHALPY(BTU/HR  )      -0.286675E+09   -0.286675E+09   -
0.207917E-15
```

```
                  ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E              0.00000      LB/HR
    PRODUCT STREAMS CO2E           0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION    0.00000      LB/HR
    UTILITIES CO2E PRODUCTION      0.00000      LB/HR
    TOTAL CO2E PRODUCTION          0.00000      LB/HR

                      ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                    50
   CONVERGENCE TOLERANCE
0.000100000

   FLASH SPECS FOR COLD SIDE:
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                    50
   CONVERGENCE TOLERANCE
0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED HOT OUTLET TEMP
     SPECIFIED VALUE               F                   128.0000
     LMTD CORRECTION FACTOR                              1.00000

   PRESSURE SPECIFICATION:
     HOT  SIDE PRESSURE DROP       PSI                   0.0000
     COLD SIDE PRESSURE DROP       PSI                   0.0000

   HEAT TRANSFER COEFFICIENT SPECIFICATION:
     HOT LIQUID    COLD LIQUID     BTU/HR-SQFT-R       149.6937
     HOT 2-PHASE   COLD LIQUID     BTU/HR-SQFT-R       149.6937
     HOT VAPOR     COLD LIQUID     BTU/HR-SQFT-R       149.6937
     HOT LIQUID    COLD 2-PHASE    BTU/HR-SQFT-R       149.6937
     HOT 2-PHASE   COLD 2-PHASE    BTU/HR-SQFT-R       149.6937
     HOT VAPOR     COLD 2-PHASE    BTU/HR-SQFT-R       149.6937
     HOT LIQUID    COLD VAPOR      BTU/HR-SQFT-R       149.6937
     HOT 2-PHASE   COLD VAPOR      BTU/HR-SQFT-R       149.6937
     HOT VAPOR     COLD VAPOR      BTU/HR-SQFT-R       149.6937

                    ***  OVERALL RESULTS  ***

   STREAMS:
                   --------------------------------------
                   |                                    |
   7         ----->|               HOT                  |-----> 10
   T=  1.5318D+02  |                                    |       T=
1.2800D+02
```

```
   P=  1.5000D+02  |                                    |       P=
1.5000D+02
   V=  0.0000D+00  |                                    |       V=
0.0000D+00
                   |                                    |
   9        <-----|                COLD                |<----- 8
   T=  1.1747D+02  |                                    |       T=
8.7000D+01
   P=  5.0000D+01  |                                    |       P=
5.0000D+01
   V=  0.0000D+00  |                                    |       V=
0.0000D+00
                   --------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY            BTU/HR            1047856.6006
     CALCULATED (REQUIRED) AREA      SQFT                  182.7829
     ACTUAL EXCHANGER AREA           SQFT                  182.7829
     PER CENT OVER-DESIGN                                    0.0000

   HEAT TRANSFER COEFFICIENT:
     AVERAGE COEFFICIENT (DIRTY)     BTU/HR-SQFT-R         149.6937
     UA (DIRTY)                      BTU/HR-R            27361.4443

   LOG-MEAN TEMPERATURE DIFFERENCE:
     LMTD CORRECTION FACTOR                                  1.0000
     LMTD (CORRECTED)                F                      38.2968
     NUMBER OF SHELLS IN SERIES                             1

   PRESSURE DROP:
     HOTSIDE, TOTAL                  PSI                     0.0000
     COLDSIDE, TOTAL                 PSI                     0.0000

                        ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:

                                HOT
         -------------------------------------------------------------
         |                                                           |
  HOT IN |                      LIQ                                  |
HOT OUT
  ------> |
|------>
   153.2 |                                                           |
128.0
         |                                                           |
  COLDOUT |                     LIQ                                  |
COLDIN
  <------ |
|<------
```

```
  117.5  |                                                            |
87.0
         |                                                            |
         -------------------------------------------------------------
                                    COLD

   ZONE HEAT TRANSFER AND AREA:

   ZONE       HEAT DUTY       AREA       LMTD       AVERAGE U       UA
              BTU/HR          SQFT        F         BTU/HR-SQFT-R
BTU/HR-R
     1    1047856.601      182.7829      38.2968   149.6937
27361.4443

 BLOCK:  E-5      MODEL: HEATX
 -----------------------------
   HOT SIDE:
   ---------
   INLET STREAM:             26
   OUTLET STREAM:            27
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE
   COLD SIDE:
   ----------
   INLET STREAM:             23
   OUTLET STREAM:            24
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                       ***  MASS AND ENERGY BALANCE  ***
                                     IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           3299.04         3299.04
0.00000
       MASS(LB/HR   )           144403.         144403.
0.00000
       ENTHALPY(BTU/HR  )      -0.244845E+09   -0.244845E+09   -
0.243438E-15

                       ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR

                           ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                  50
   CONVERGENCE TOLERANCE
0.000100000
```

```
   FLASH SPECS FOR COLD SIDE:
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                   50
   CONVERGENCE TOLERANCE
0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED COLD OUTLET TEMP
     SPECIFIED VALUE              F                      300.0000
     LMTD CORRECTION FACTOR                                1.00000

   PRESSURE SPECIFICATION:
     HOT  SIDE PRESSURE DROP      PSI                      0.0000
     COLD SIDE PRESSURE DROP      PSI                      0.0000

   HEAT TRANSFER COEFFICIENT SPECIFICATION:
     HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R         149.6937

                        ***  OVERALL RESULTS  ***

   STREAMS:
                  --------------------------------------
                  |                                    |
   26      ----->|                HOT                 |-----> 27
   T=  5.0000D+02 |                                    |       T=
3.1631D+02
   P=  6.8635D+02 |                                    |       P=
6.8635D+02
   V=  1.0000D+00 |                                    |       V=
0.0000D+00
                  |                                    |
   24      <-----|                COLD                |<----- 23
   T=  3.0000D+02 |                                    |       T=
1.7512D+02
   P=  4.6470D+02 |                                    |       P=
4.6470D+02
   V=  1.0000D+00 |                                    |       V=
0.0000D+00
                  --------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY         BTU/HR             20194190.8827
```

```
    CALCULATED (REQUIRED) AREA     SQFT                        609.1590
    ACTUAL EXCHANGER AREA          SQFT                        609.1590
    PER CENT OVER-DESIGN                                         0.0000

  HEAT TRANSFER COEFFICIENT:
    AVERAGE COEFFICIENT (DIRTY)    BTU/HR-SQFT-R               149.6937
    UA (DIRTY)                     BTU/HR-R                  91187.2371

  LOG-MEAN TEMPERATURE DIFFERENCE:
    LMTD CORRECTION FACTOR                                       1.0000
    LMTD (CORRECTED)               F                           221.4585
    NUMBER OF SHELLS IN SERIES                                  1

  PRESSURE DROP:
    HOTSIDE, TOTAL                 PSI                           0.0000
    COLDSIDE, TOTAL                PSI                           0.0000

                     ***  ZONE RESULTS  ***

  TEMPERATURE LEAVING EACH ZONE:

                                        HOT
         -------------------------------------------------------------
         |              |              |              |              |
 HOT IN  |     COND     |     COND     |     COND     |     LIQ      |
HOT OUT
 ------> |              |              |              |
|------>
  500.0  |         500.0|         500.0|         500.0|              |
316.3
         |              |              |              |              |
 COLDOUT |     VAP      |     BOIL     |     LIQ      |     LIQ      |
COLDIN
 <------ |              |              |              |
|<------
  300.0  |         278.4|         277.9|         227.7|              |
175.1
         |              |              |              |              |
         -------------------------------------------------------------
                                        COLD

  ZONE HEAT TRANSFER AND AREA:

  ZONE       HEAT DUTY       AREA        LMTD        AVERAGE U       UA
             BTU/HR          SQFT         F          BTU/HR-SQFT-R
BTU/HR-R
     1    2427749.742       77.0004     210.6239   149.6937
11526.4685
     2    6677343.755      201.0427     221.8769   149.6937
30094.8151
     3    6295970.339      170.7136     246.3720   149.6937
25554.7362
```

```
     4    4793127.048       160.4024     199.6203   149.6937
24011.2173

 BLOCK:  E-10     MODEL: HEATX
 -----------------------------
   HOT SIDE:
   ---------
   INLET STREAM:             34
   OUTLET STREAM:            37
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE
   COLD SIDE:
   ----------
   INLET STREAM:             35
   OUTLET STREAM:            36
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN               OUT
RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            482.968         482.968
0.00000
       MASS(LB/HR   )            10909.8         10909.8
0.00000
       ENTHALPY(BTU/HR  )      -0.574336E+08   -0.574336E+08
0.00000

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E              0.00000      LB/HR
    PRODUCT STREAMS CO2E           0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION    0.00000      LB/HR
    UTILITIES CO2E PRODUCTION      0.00000      LB/HR
    TOTAL CO2E PRODUCTION          0.00000      LB/HR

                          ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                   50
   CONVERGENCE TOLERANCE
0.000100000

   FLASH SPECS FOR COLD SIDE:
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                   50
   CONVERGENCE TOLERANCE
0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED HOT OUTLET TEMP
     SPECIFIED VALUE               F                        98.0000
```

```
  LMTD CORRECTION FACTOR                                       1.00000

PRESSURE SPECIFICATION:
  HOT  SIDE PRESSURE DROP        PSI                          0.0000
  COLD SIDE PRESSURE DROP        PSI                          0.0000

HEAT TRANSFER COEFFICIENT SPECIFICATION:
  HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R              149.6937
  HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R              149.6937
  HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R              149.6937
  HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R              149.6937
  HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R              149.6937
  HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R              149.6937
  HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R              149.6937
  HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R              149.6937
  HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R              149.6937

                    ***  OVERALL RESULTS  ***

STREAMS:
                ---------------------------------------
                |                                     |
34        ----->|                HOT                  |-----> 37
T=  2.4067D+02  |                                     |       T=
9.8000D+01
P=  1.5100D+02  |                                     |       P=
1.5100D+02
V=  0.0000D+00  |                                     |       V=
0.0000D+00
                |                                     |
36        <-----|                COLD                 |<----- 35
T=  1.1752D+02  |                                     |       T=
8.7000D+01
P=  5.0000D+01  |                                     |       P=
5.0000D+01
V=  0.0000D+00  |                                     |       V=
0.0000D+00
                ---------------------------------------

DUTY AND AREA:
  CALCULATED HEAT DUTY           BTU/HR                   262331.9541
  CALCULATED (REQUIRED) AREA     SQFT                         37.7441
  ACTUAL EXCHANGER AREA          SQFT                         37.7441
  PER CENT OVER-DESIGN                                         0.0000

HEAT TRANSFER COEFFICIENT:
  AVERAGE COEFFICIENT (DIRTY)    BTU/HR-SQFT-R              149.6937
  UA (DIRTY)                     BTU/HR-R                  5650.0573

LOG-MEAN TEMPERATURE DIFFERENCE:
  LMTD CORRECTION FACTOR                                       1.0000
  LMTD (CORRECTED)               F                            46.4300
```

```
     NUMBER OF SHELLS IN SERIES                              1

   PRESSURE DROP:
     HOTSIDE, TOTAL               PSI                     0.0000
     COLDSIDE, TOTAL              PSI                     0.0000

                       ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:

                                HOT
         -------------------------------------------------------------
         |                                                           |
  HOT IN |                      LIQ                                  |
HOT OUT
  ------> |
|------>
   240.7  |                                                           |
98.0
         |                                                           |
  COLDOUT |                      LIQ                                  |
COLDIN
  <------ |
|<------
   117.5  |                                                           |
87.0
         |                                                           |
         -------------------------------------------------------------
                                COLD

   ZONE HEAT TRANSFER AND AREA:

   ZONE       HEAT DUTY       AREA       LMTD       AVERAGE U       UA
              BTU/HR          SQFT        F         BTU/HR-SQFT-R
BTU/HR-R
     1     262331.954       37.7441      46.4300   149.6937
5650.0573

  BLOCK:  P-1      MODEL: PUMP
  ----------------------------
   INLET STREAM:          1
   OUTLET STREAM:         2
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           1126.91         1126.91
0.00000
       MASS(LB/HR   )           65646.3         65646.3
0.00000
```

```
      ENTHALPY(BTU/HR  )      -0.718405E+08   -0.718156E+08   -
0.346553E-03

                  ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR

                      ***  INPUT DATA  ***
    OUTLET PRESSURE  PSIA                               160.000
    DRIVER EFFICIENCY                                     1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                          50
    TOLERANCE
0.000100000

                       ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                     1,894.99
    PRESSURE CHANGE  PSI                                 45.3041
    NPSH AVAILABLE   FT-LBF/LB                          247.492
    FLUID POWER  HP                                       6.24370
    BRAKE POWER  HP                                       9.78469
    ELECTRICITY  KW                                       7.29644
    PUMP EFFICIENCY USED                                  0.63811
    NET WORK REQUIRED  HP                                 9.78469
    HEAD DEVELOPED FT-LBF/LB                            188.320

 BLOCK:  P-3      MODEL: PUMP
 ----------------------------
   INLET STREAM:          16
   OUTLET STREAM:         17
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                  ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
      MOLE(LBMOL/HR)           2184.85         2184.85
0.00000
      MASS(LB/HR   )           126540.         126540.
0.00000
      ENTHALPY(BTU/HR  )      -0.130919E+09   -0.130892E+09   -
0.200850E-03

                  ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
```

```
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR

                          ***  INPUT DATA  ***
    OUTLET PRESSURE  PSIA                                        175.000
    DRIVER EFFICIENCY                                              1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                                   50
    TOLERANCE
0.000100000

                           ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                              4,059.78
    PRESSURE CHANGE  PSI                                          25.0000
    NPSH AVAILABLE   FT-LBF/LB                                     0.0
    FLUID POWER  HP                                                7.38142
    BRAKE POWER  HP                                               10.3343
    ELECTRICITY  KW                                                7.70629
    PUMP EFFICIENCY USED                                           0.71426
    NET WORK REQUIRED  HP                                         10.3343
    HEAD DEVELOPED FT-LBF/LB                                     115.499

 BLOCK:  P-5      MODEL: PUMP
 ----------------------------
   INLET STREAM:          22
   OUTLET STREAM:         23
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           2143.52         2143.52
0.00000
       MASS(LB/HR   )           123586.         123586.
0.117747E-15
       ENTHALPY(BTU/HR  )      -0.127930E+09   -0.127607E+09   -
0.253119E-02

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR

                          ***  INPUT DATA  ***
    OUTLET PRESSURE  PSIA                                        464.696
```

```
    DRIVER EFFICIENCY                                                1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                                     50
    TOLERANCE
0.000100000

                           ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                                3,973.44
    PRESSURE CHANGE  PSI                                           313.696
    NPSH AVAILABLE   FT-LBF/LB                                       0.0
    FLUID POWER  HP                                                 90.6510
    BRAKE POWER  HP                                                127.264
    ELECTRICITY  KW                                                 94.9009
    PUMP EFFICIENCY USED                                             0.71231
    NET WORK REQUIRED  HP                                          127.264
    HEAD DEVELOPED FT-LBF/LB                                     1,452.34

 BLOCK:  R-1      MODEL: RSTOIC
 ------------------------------
   INLET STREAM:          24
   OUTLET STREAM:         25
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                       ***  MASS AND ENERGY BALANCE  ***
                              IN          OUT       GENERATION
RELATIVE DIFF.
   TOTAL BALANCE
   MOLE(LBMOL/HR)         2143.52       2143.52       0.00000
0.140218E-05
   MASS(LB/HR   )         123586.       123586.
0.136437E-05
   ENTHALPY(BTU/HR  )   -0.107412E+09 -0.109443E+09
0.185506E-01

                       ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR

                          ***  INPUT DATA  ***
   STOICHIOMETRY MATRIX:

    REACTION #   1:
     SUBSTREAM MIXED   :
     ISOBU-02   1.00    N-BUT-01  -1.00
```

REACTION CONVERSION SPECS: NUMBER=   1
REACTION #   1:
SUBSTREAM:MIXED   KEY COMP:N-BUT-01 CONV FRAC: 0.4860

TWO   PHASE TP FLASH

SPECIFIED TEMPERATURE F		300.000
SPECIFIED PRESSURE	PSIA	444.696
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.000100000
SIMULTANEOUS REACTIONS		
GENERATE COMBUSTION REACTIONS FOR FEED SPECIES		NO

*** RESULTS ***

OUTLET TEMPERATURE	F	300.00
OUTLET PRESSURE	PSIA	444.70
HEAT DUTY	BTU/HR	-0.20285E+07
VAPOR FRACTION		1.0000

REACTION EXTENTS:

REACTION NUMBER	REACTION EXTENT LBMOL/HR
1	639.77

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
WATER	0.10785E-01	0.10220E-01	0.10785E-01	1.2158
PROPA-01	0.12053E-02	0.10698E-02	0.12053E-02	1.2979
1-BUT-01	0.10798E-02	0.10728E-02	0.10798E-02	1.1596
ISOBU-01	0.18772E-02	0.18611E-02	0.18772E-02	1.1620
TRANS-01	0.92729E-02	0.93632E-02	0.92729E-02	1.1409
CIS-2-01	0.27254E-01	0.27766E-01	0.27254E-01	1.1307
ISOBU-02	0.62850	0.62194	0.62850	1.1642
N-BUT-01	0.31566	0.32185	0.31566	1.1299

```
      2:2-D-01         0.19026E-02    0.20588E-02    0.19026E-02
1.0646
      2-MET-01         0.24158E-02    0.27446E-02    0.24158E-02
1.0140
      N-PEN-01         0.42342E-04    0.49239E-04    0.42342E-04
0.99066
      N-HEX-01         0.96118E-06    0.12913E-05    0.96118E-06
0.85750

 BLOCK:  T-1      MODEL: RADFRAC
 -------------------------------
    INLETS   - 25       STAGE  24
               2        STAGE  25
    OUTLETS  - 7        STAGE   1
               16       STAGE  26
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT
RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           3270.43         3270.43       -
0.278097E-15
       MASS(LB/HR   )           189232.         189232.
0.937871E-12
       ENTHALPY(BTU/HR  )      -0.181258E+09   -0.199338E+09
0.907015E-01

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR


                         **********************
                         ****  INPUT DATA  ****
                         **********************

   ****   INPUT PARAMETERS   ****

    NUMBER OF STAGES                                          26
    ALGORITHM OPTION                                        STANDARD
    ABSORBER OPTION                                         NO
    INITIALIZATION OPTION                                   STANDARD
    HYDRAULIC PARAMETER CALCULATIONS                        NO
    INSIDE LOOP CONVERGENCE METHOD                          BROYDEN
    DESIGN SPECIFICATION METHOD                             NESTED
    MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS                   200
    MAXIMUM NO. OF INSIDE LOOP ITERATIONS                     10
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    MAXIMUM NUMBER OF FLASH ITERATIONS                        50
    FLASH TOLERANCE
0.000100000
    OUTSIDE LOOP CONVERGENCE TOLERANCE
0.000100000

   ****   COL-SPECS   ****

    MOLAR VAPOR DIST / TOTAL DIST                              0.0
    MOLAR REFLUX RATIO                                        10.0000
    MOLAR BOILUP RATIO                                         4.60000

   ****    PROFILES   ****

    P-SPEC          STAGE   1  PRES, PSIA                    150.000

                               ******************
                               ****  RESULTS  ****
                               ******************


   ***   COMPONENT SPLIT FRACTIONS   ***

                                  OUTLET STREAMS
                                  --------------
                  7              16
    COMPONENT:
    WATER        .17434E-01   .98257
    PROPA-01     .91207       .87934E-01
    1-BUT-01     .34465       .65535
    ISOBU-01     .37657       .62343
    TRANS-01     .58088E-01   .94191
    CIS-2-01     .18832E-01   .98117
    ISOBU-02     .58369       .41631
    N-BUT-01     .43909E-01   .95609
    2:2-D-01     .21182E-03   .99979
    2-MET-01     .19989E-07   1.0000
    N-PEN-01     .41716E-09   1.0000
    N-HEX-01     0.0000       1.0000


   ***    SUMMARY OF KEY RESULTS    ***

    TOP STAGE TEMPERATURE          F                        153.182
    BOTTOM STAGE TEMPERATURE       F                        170.569
    TOP STAGE LIQUID FLOW          LBMOL/HR               12,045.0
    BOTTOM STAGE LIQUID FLOW       LBMOL/HR                2,184.85
    TOP STAGE VAPOR FLOW           LBMOL/HR                    0.0
    BOILUP VAPOR FLOW              LBMOL/HR               10,050.3
    MOLAR REFLUX RATIO                                        11.0956
    MOLAR BOILUP RATIO                                         4.60000
```

CONDENSER DUTY (W/O SUBCOOL) BTU/HR -0.913215+08
REBOILER DUTY BTU/HR 0.732412+08

**** MANIPULATED VARIABLES ****

	BOUNDS		CALCULATED
	LOWER	UPPER	VALUE
MOLAR REFLUX RATIO	5.0000	18.000	11.096

**** DESIGN SPECIFICATIONS ****

NO	SPEC-TYPE	QUALIFIERS	UNIT	SPECIFIED VALUE	CALCULATED VALUE
1	STDVOL-FRAC	STREAMS: 7		0.92000	0.92000
		COMPS: ISOBU-02			

**** MAXIMUM FINAL RELATIVE ERRORS ****

DEW POINT	0.88518E-08	STAGE= 23	
BUBBLE POINT	0.15160E-07	STAGE= 18	
COMPONENT MASS BALANCE	0.40109E-08	STAGE= 23	COMP=PROPA-01
ENERGY BALANCE	0.22790E-06	STAGE= 1	

**** PROFILES ****

NOTE REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS FROM THE STAGE INCLUDING ANY SIDE PRODUCT.

STAGE	TEMPERATURE F	PRESSURE PSIA	ENTHALPY BTU/LBMOL LIQUID	ENTHALPY BTU/LBMOL VAPOR	HEAT DUTY BTU/HR
1	153.18	150.00	-63027.	-55929.	-.91321+08
2	154.49	150.00	-63053.	-56072.	
3	155.34	150.00	-63014.	-56096.	
4	155.97	150.00	-62937.	-56056.	
22	166.01	150.00	-60365.	-53505.	
23	166.50	150.00	-60272.	-53392.	
24	167.59	150.00	-60147.	-53209.	
25	168.95	150.00	-60068.	-53042.	

26	170.57	150.00	-59921.	-52812.	.73241+08

STAGE	FLOW RATE LBMOL/HR		FEED RATE LBMOL/HR			PRODUCT RATE LBMOL/HR	
	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	LIQUID	VAPOR
1	0.1313E+05	0.000				1085.5739	
2	0.1205E+05	0.1313E+05					
3	0.1204E+05	0.1313E+05					
4	0.1202E+05	0.1312E+05					
22	0.1160E+05	0.1270E+05					
23	0.1077E+05	0.1268E+05		2143.5194			
24	0.1073E+05	9709.					
25	0.1224E+05	9671.	1126.9086				
26	2185.	0.1005E+05				2184.8541	

**** MASS FLOW PROFILES ****

STAGE	FLOW RATE LB/HR		FEED RATE LB/HR			PRODUCT RATE LB/HR	
	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	LIQUID	VAPOR
1	0.7583E+06	0.000				.62693+05	
2	0.6975E+06	0.7583E+06					
3	0.6979E+06	0.7602E+06					
4	0.6975E+06	0.7606E+06					
22	0.6694E+06	0.7336E+06					
23	0.6214E+06	0.7321E+06		.12359+06			
24	0.6195E+06	0.5605E+06					
25	0.7076E+06	0.5587E+06	.65646+05				
26	0.1265E+06	0.5811E+06				.12654+06	

**** MOLE-X-PROFILE ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.38556E-03	0.24685E-01	0.11213E-02	0.22390E-02	0.11296E-02
2	0.49367E-03	0.13476E-01	0.12132E-02	0.23889E-02	0.13936E-02
3	0.61599E-03	0.78080E-02	0.12942E-02	0.25160E-02	0.16785E-02
4	0.75489E-03	0.49611E-02	0.13660E-02	0.26244E-02	0.19866E-02
22	0.84376E-02	0.18490E-02	0.12492E-02	0.21717E-02	0.88765E-02
23	0.92703E-02	0.18347E-02	0.11999E-02	0.20861E-02	0.90870E-02
24	0.94344E-02	0.21603E-02	0.11683E-02	0.20392E-02	0.89348E-02

25	0.95510E-02	0.23067E-02	0.11208E-02	0.19662E-02	0.86376E-02
26	0.10797E-01	0.11825E-02	0.10594E-02	0.18417E-02	0.91013E-02

**** MOLE-X-PROFILE ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.10354E-02	0.91369	0.55711E-01	0.97907E-06	0.79259E-09
2	0.13497E-02	0.91046	0.69218E-01	0.15653E-05	0.18357E-08
3	0.17115E-02	0.90044	0.83936E-01	0.24042E-05	0.40138E-08
4	0.21282E-02	0.88620	0.99978E-01	0.36007E-05	0.85380E-08
22	0.22616E-01	0.47048	0.48153	0.10694E-02	0.16899E-02
23	0.24327E-01	0.45362	0.49400	0.13824E-02	0.31345E-02
24	0.24447E-01	0.41211	0.53212	0.15562E-02	0.58973E-02
25	0.24286E-01	0.36700	0.57113	0.17998E-02	0.11859E-01
26	0.26805E-01	0.32379	0.60273	0.22960E-02	0.19701E-01

**** MOLE-X-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.41661E-12	0.12886E-21
2	0.11644E-11	0.81405E-21
3	0.30476E-11	0.47324E-20
4	0.77522E-11	0.26988E-19
22	0.26199E-04	0.39722E-06
23	0.55578E-04	0.20303E-05
24	0.11926E-03	0.11269E-04
25	0.27484E-03	0.66883E-04
26	0.49622E-03	0.20126E-03

**** MOLE-Y-PROFILE ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.29723E-03	0.44766E-01	0.10238E-02	0.20734E-02	0.90411E-03
2	0.38556E-03	0.24685E-01	0.11213E-02	0.22390E-02	0.11296E-02
3	0.48473E-03	0.14403E-01	0.12056E-02	0.23765E-02	0.13717E-02
4	0.59693E-03	0.92041E-02	0.12799E-02	0.24931E-02	0.16331E-02
22	0.70355E-02	0.38140E-02	0.12833E-02	0.22569E-02	0.79880E-02

23	0.77484E-02	0.38038E-02	0.12382E-02	0.21775E-02	0.82133E-02
24	0.79424E-02	0.45286E-02	0.12177E-02	0.21493E-02	0.81562E-02
25	0.81192E-02	0.49005E-02	0.11827E-02	0.20976E-02	0.79837E-02
26	0.92801E-02	0.25510E-02	0.11342E-02	0.19932E-02	0.85368E-02

**** MOLE-Y-PROFILE ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.78422E-03	0.90587	0.44283E-01	0.60422E-06	0.33730E-09
2	0.10354E-02	0.91369	0.55711E-01	0.97907E-06	0.79259E-09
3	0.13237E-02	0.91073	0.68102E-01	0.15169E-05	0.17494E-08
4	0.16556E-02	0.90153	0.81601E-01	0.22864E-05	0.37474E-08
22	0.19225E-01	0.52495	0.43186	0.75271E-03	0.82878E-03
23	0.20768E-01	0.50842	0.44508	0.97792E-03	0.15453E-02
24	0.21076E-01	0.46645	0.48437	0.11131E-02	0.29428E-02
25	0.21197E-01	0.42046	0.52662	0.13049E-02	0.60070E-02
26	0.23738E-01	0.37639	0.56427	0.16919E-02	0.10154E-01

**** MOLE-Y-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.14670E-12	0.20030E-22
2	0.41661E-12	0.12886E-21
3	0.11026E-11	0.75741E-21
4	0.28300E-11	0.43516E-20
22	0.11217E-04	0.70706E-07
23	0.23957E-04	0.36322E-06
24	0.52286E-04	0.20393E-05
25	0.12292E-03	0.12288E-04
26	0.22671E-03	0.37670E-04

**** K-VALUES ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.77092	1.8135	0.91305	0.92601	0.80035
2	0.78100	1.8317	0.92425	0.93726	0.81062
3	0.78692	1.8446	0.93154	0.94456	0.81723

4	0.79075	1.8553	0.93698	0.94996	0.82206
22	0.83382	2.0628	1.0274	1.0393	0.89991
23	0.83583	2.0732	1.0319	1.0438	0.90386
24	0.84186	2.0963	1.0422	1.0540	0.91286
25	0.85009	2.1245	1.0552	1.0668	0.92429
26	0.85951	2.1574	1.0706	1.0823	0.93798

**** K-VALUES ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.75739	0.99144	0.79488	0.61714	0.42556
2	0.76716	1.0035	0.80486	0.62547	0.43178
3	0.77341	1.0114	0.81135	0.63091	0.43585
4	0.77795	1.0173	0.81619	0.63498	0.43891
22	0.85008	1.1158	0.89684	0.70388	0.49042
23	0.85372	1.1208	0.90097	0.70742	0.49298
24	0.86213	1.1318	0.91027	0.71523	0.49900
25	0.87281	1.1457	0.92205	0.72503	0.50654
26	0.88559	1.1624	0.93619	0.73687	0.51539

**** K-VALUES ****

STAGE	N-PEN-01	N-HEX-01
1	0.35212	0.15544
2	0.35778	0.15830
3	0.36179	0.16005
4	0.36506	0.16124
22	0.42813	0.17800
23	0.43105	0.17890
24	0.43842	0.18097
25	0.44725	0.18373
26	0.45688	0.18717

**** MASS-X-PROFILE ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.12028E-03	0.18849E-01	0.10894E-02	0.21753E-02	0.10975E-02

2	0.15360E-03	0.10263E-01	0.11756E-02	0.23149E-02	0.13504E-02
3	0.19141E-03	0.59389E-02	0.12525E-02	0.24350E-02	0.16245E-02
4	0.23444E-03	0.37713E-02	0.13213E-02	0.25385E-02	0.19216E-02
22	0.26332E-02	0.14124E-02	0.12141E-02	0.21107E-02	0.86273E-02
23	0.28936E-02	0.14018E-02	0.11665E-02	0.20280E-02	0.88337E-02
24	0.29432E-02	0.16496E-02	0.11352E-02	0.19813E-02	0.86810E-02
25	0.29752E-02	0.17588E-02	0.10874E-02	0.19075E-02	0.83800E-02
26	0.33585E-02	0.90032E-03	0.10263E-02	0.17842E-02	0.88170E-02

**** MASS-X-PROFILE ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.10060E-02	0.91959	0.56070E-01	0.12232E-05	0.99022E-09
2	0.13079E-02	0.91395	0.69483E-01	0.19505E-05	0.22874E-08
3	0.16564E-02	0.90275	0.84151E-01	0.29921E-05	0.49953E-08
4	0.20585E-02	0.88797	0.10018	0.44786E-05	0.10620E-07
22	0.21981E-01	0.47370	0.48483	0.13365E-02	0.21121E-02
23	0.23649E-01	0.45682	0.49749	0.17281E-02	0.39185E-02
24	0.23753E-01	0.41480	0.53558	0.19444E-02	0.73682E-02
25	0.23561E-01	0.36884	0.57401	0.22453E-02	0.14795E-01
26	0.25967E-01	0.32495	0.60488	0.28603E-02	0.24543E-01

**** MASS-X-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.52049E-12	0.19229E-21
2	0.14510E-11	0.12116E-20
3	0.37928E-11	0.70346E-20
4	0.96423E-11	0.40094E-19
22	0.32745E-04	0.59298E-06
23	0.69477E-04	0.30315E-05
24	0.14900E-03	0.16816E-04
25	0.34288E-03	0.99663E-04
26	0.61817E-03	0.29946E-03

**** MASS-Y-PROFILE ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.93169E-04	0.34346E-01	0.99946E-03	0.20241E-02	0.88262E-03
2	0.12028E-03	0.18849E-01	0.10894E-02	0.21753E-02	0.10975E-02
3	0.15085E-03	0.10971E-01	0.11685E-02	0.23034E-02	0.13295E-02
4	0.18555E-03	0.70030E-02	0.12391E-02	0.24136E-02	0.15810E-02
22	0.21948E-02	0.29124E-02	0.12469E-02	0.21928E-02	0.77611E-02
23	0.24180E-02	0.29055E-02	0.12034E-02	0.21163E-02	0.79825E-02
24	0.24784E-02	0.34589E-02	0.11834E-02	0.20888E-02	0.79266E-02
25	0.25320E-02	0.37407E-02	0.11487E-02	0.20373E-02	0.77542E-02
26	0.28917E-02	0.19457E-02	0.11007E-02	0.19344E-02	0.82848E-02

**** MASS-Y-PROFILE ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.76557E-03	0.91610	0.44784E-01	0.75851E-06	0.42343E-09
2	0.10060E-02	0.91959	0.56070E-01	0.12232E-05	0.99022E-09
3	0.12830E-02	0.91441	0.68377E-01	0.18906E-05	0.21804E-08
4	0.16028E-02	0.90414	0.81837E-01	0.28463E-05	0.46651E-08
22	0.18679E-01	0.52836	0.43467	0.94044E-03	0.10355E-02
23	0.20185E-01	0.51189	0.44812	0.12222E-02	0.19313E-02
24	0.20483E-01	0.46960	0.48764	0.13910E-02	0.36776E-02
25	0.20587E-01	0.42304	0.52985	0.16298E-02	0.75025E-02
26	0.23037E-01	0.37840	0.56728	0.21114E-02	0.12672E-01

**** MASS-Y-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.18416E-12	0.30033E-22
2	0.52049E-12	0.19229E-21
3	0.13742E-11	0.11275E-20
4	0.35231E-11	0.64706E-20
22	0.14014E-04	0.10551E-06
23	0.29941E-04	0.54220E-06
24	0.65342E-04	0.30440E-05

25	0.15352E-03	0.18331E-04
26	0.28293E-03	0.56151E-04

****
***** HYDRAULIC PARAMETERS *****
****

*** DEFINITIONS ***

```
MARANGONI INDEX = SIGMA - SIGMATO
FLOW PARAM = (ML/MV)*SQRT(RHOV/RHOL)
QR = QV*SQRT(RHOV/(RHOL-RHOV))
F FACTOR = QV*SQRT(RHOV)
  WHERE:
  SIGMA IS THE SURFACE TENSION OF LIQUID FROM THE STAGE
  SIGMATO IS THE SURFACE TENSION OF LIQUID TO THE STAGE
  ML IS THE MASS FLOW OF LIQUID FROM THE STAGE
  MV IS THE MASS FLOW OF VAPOR TO THE STAGE
  RHOL IS THE MASS DENSITY OF LIQUID FROM THE STAGE
  RHOV IS THE MASS DENSITY OF VAPOR TO THE STAGE
  QV IS THE VOLUMETRIC FLOW RATE OF VAPOR TO THE STAGE
```

	TEMPERATURE F	
STAGE	LIQUID FROM	VAPOR TO
1	153.18	154.49
2	154.49	155.34
3	155.34	155.97
4	155.97	156.50
22	166.01	166.50
23	166.50	183.98
24	167.59	168.95
25	168.95	170.57
26	170.57	170.57

	MASS FLOW LB/HR		VOLUME FLOW CUFT/HR		MOLECULAR WEIGHT	
STAGE	LIQUID FROM	VAPOR TO	LIQUID FROM	VAPOR TO	LIQUID FROM	VAPOR TO
1	0.75830E+06	0.75830E+06	24710.	0.45575E+06	57.751	57.751
2	0.69750E+06	0.76019E+06	22733.	0.45627E+06	57.902	57.889
3	0.69789E+06	0.76058E+06	22743.	0.45641E+06	57.975	57.956

4	0.69745E+06	0.76014E+06	22723.	0.45635E+06	58.007	57.986
22	0.66941E+06	0.73210E+06	21571.	0.45106E+06	57.728	57.730
23	0.62142E+06	0.68411E+06	20010.	0.44374E+06	57.716	57.719
24	0.61955E+06	0.55865E+06	19931.	0.34562E+06	57.747	57.768
25	0.70759E+06	0.58105E+06	22739.	0.36031E+06	57.833	57.814
26	0.12654E+06	0.0000	4059.8	0.0000	57.917	

	DENSITY LB/CUFT		VISCOSITY CP		SURFACE TENSION DYNE/CM
STAGE	LIQUID FROM	VAPOR TO	LIQUID FROM	VAPOR TO	LIQUID FROM
1	30.688	1.6638	0.12818	0.92640E-02	6.0883
2	30.682	1.6661	0.12851	0.92693E-02	6.0716
3	30.685	1.6664	0.12842	0.92745E-02	6.0597
4	30.693	1.6657	0.12808	0.92798E-02	6.0545
22	31.033	1.6231	0.11541	0.94134E-02	6.0941
23	31.055	1.5417	0.11501	0.96128E-02	6.0995
24	31.084	1.6164	0.11354	0.94401E-02	6.1086
25	31.119	1.6127	0.11197	0.94591E-02	6.1085
26	31.169		0.11078		6.1018

STAGE	MARANGONI INDEX DYNE/CM	FLOW PARAM	QR CUFT/HR	REDUCED F-FACTOR (LB-CUFT)**.5/HR
1		0.23285	0.10912E+06	0.58788E+06
2	-.16654E-01	0.21381	0.10933E+06	0.58894E+06
3	-.11922E-01	0.21383	0.10937E+06	0.58918E+06
4	-.52051E-02	0.21374	0.10932E+06	0.58897E+06
22	0.29621E-01	0.20911	0.10596E+06	0.57465E+06
23	0.54607E-02	0.20239	0.10142E+06	0.55097E+06
24	0.91212E-02	0.25289	80947.	0.43941E+06
25	-.35023	0.27722	84235.	0.45756E+06
26	-.67013E-02		0.0000	0.0000

```
************************************
***** TRAY SIZING CALCULATIONS *****
************************************
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*** SECTION   1 ***
*******************

STARTING STAGE NUMBER                                                2
ENDING STAGE NUMBER                                                 25
FLOODING CALCULATION METHOD                                       FAIR

DESIGN PARAMETERS
-----------------
PEAK CAPACITY FACTOR                                           1.00000
SYSTEM FOAMING FACTOR                                          1.00000
FLOODING FACTOR                                                0.80000
MINIMUM COLUMN DIAMETER         FT                             1.00000
MINIMUM DC AREA/COLUMN AREA                                    0.100000
HOLE AREA/ACTIVE AREA                                          0.12000

TRAY SPECIFICATIONS
-------------------
TRAY TYPE                                                        SIEVE
NUMBER OF PASSES                                                     1
TRAY SPACING                    FT                             2.00000


        ***** SIZING RESULTS @ STAGE WITH MAXIMUM DIAMETER *****

STAGE WITH MAXIMUM DIAMETER                                          3
COLUMN DIAMETER                 FT                            17.4725
DC AREA/COLUMN AREA                                            0.100000
DOWNCOMER VELOCITY              FT/SEC                         0.26348
FLOW PATH LENGTH                FT                            12.0044
SIDE DOWNCOMER WIDTH            FT                             2.73402
SIDE WEIR LENGTH                FT                            12.6957
CENTER DOWNCOMER WIDTH          FT                             0.0
CENTER WEIR LENGTH              FT                             0.0
OFF-CENTER DOWNCOMER WIDTH      FT                             0.0
OFF-CENTER SHORT WEIR LENGTH    FT                             0.0
OFF-CENTER LONG WEIR LENGTH     FT                             0.0
TRAY CENTER TO OCDC CENTER      FT                             0.0


                      **** SIZING PROFILES ****

   STAGE    DIAMETER      TOTAL AREA   ACTIVE AREA    SIDE DC AREA
              FT            SQFT          SQFT            SQFT
      2       17.466        239.58        191.67          23.958
      3       17.472        239.77        191.82          23.977
      4       17.469        239.66        191.73          23.966
      5       17.452        239.21        191.37          23.921
      6       17.438        238.82        191.05          23.882
      7       17.426        238.51        190.81          23.851
      8       17.409        238.03        190.43          23.803
      9       17.391        237.55        190.04          23.755
```

10	17.374	237.07	189.66	23.707
11	17.355	236.57	189.26	23.657
12	17.337	236.06	188.85	23.606
13	17.316	235.51	188.41	23.551
14	17.296	234.96	187.96	23.496
15	17.276	234.40	187.52	23.440
16	17.256	233.86	187.09	23.386
17	17.236	233.32	186.66	23.332
18	17.217	232.80	186.24	23.280
19	17.198	232.29	185.83	23.229
20	17.179	231.78	185.42	23.178
21	17.160	231.28	185.02	23.128
22	17.134	230.58	184.46	23.058
23	16.685	218.65	174.92	21.865
24	15.392	186.07	148.86	18.607
25	15.925	199.18	159.35	19.918

```
 BLOCK:  T-2      MODEL: RADFRAC
 -------------------------------
    INLETS   - 17       STAGE   4
    OUTLETS  - 22       STAGE   1
               34       STAGE  31
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN               OUT
RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           2184.85         2184.85       -
0.270577E-14
       MASS(LB/HR   )           126540.         126540.       -
0.298997E-14
       ENTHALPY(BTU/HR  )      -0.130892E+09   -0.130817E+09   -
0.571553E-03

                    ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR


                       **********************
                       ****  INPUT DATA  ****
                       **********************

   ****   INPUT PARAMETERS   ****

    NUMBER OF STAGES                                            31
    ALGORITHM OPTION                                          NEWTON
```

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    INITIALIZATION OPTION                                         STANDARD
    HYDRAULIC PARAMETER CALCULATIONS                              NO
    DESIGN SPECIFICATION METHOD                                   SIMULT
    MAXIMUM NO. OF NEWTON ITERATIONS                               200
    MAXIMUM NUMBER OF FLASH ITERATIONS                              50
    FLASH TOLERANCE
0.000100000
    COLUMN EQUATIONS CONVERGENCE TOLERANCE                           0.100000-
06

   ****   COL-SPECS   ****

    MOLAR VAPOR DIST / TOTAL DIST                                    0.0
    MOLAR REFLUX RATIO                                               2.35000
    MOLAR BOILUP RATIO                                               5.00000

   ****    PROFILES   ****

    P-SPEC          STAGE   1  PRES, PSIA                          151.000

                                  *******************
                                  ****  RESULTS  ****
                                  *******************


   ***   COMPONENT SPLIT FRACTIONS   ***

                                     OUTLET STREAMS
                                     --------------
                     22           34
    COMPONENT:
    WATER        .98001       .19991E-01
    PROPA-01    1.0000        .21700E-13
    1-BUT-01     .99999       .10910E-04
    ISOBU-01     .99999       .77410E-05
    TRANS-01     .99957       .42615E-03
    CIS-2-01     .99761       .23940E-02
    ISOBU-02    1.0000        .90474E-06
    N-BUT-01     .99964       .36095E-03
    2:2-D-01     .81298       .18702
    2-MET-01     .12030       .87970
    N-PEN-01     .83715E-01   .91628
    N-HEX-01     .46853E-02   .99531


   ***    SUMMARY OF KEY RESULTS    ***

    TOP STAGE TEMPERATURE          F                             170.122
    BOTTOM STAGE TEMPERATURE       F                             240.673
    TOP STAGE LIQUID FLOW          LBMOL/HR                    5,037.28
    BOTTOM STAGE LIQUID FLOW       LBMOL/HR                       41.3318
    TOP STAGE VAPOR FLOW           LBMOL/HR                        0.0
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    BOILUP VAPOR FLOW               LBMOL/HR              6,572.45
    MOLAR REFLUX RATIO                                       2.35000
    MOLAR BOILUP RATIO                                     159.017
    CONDENSER DUTY (W/O SUBCOOL)    BTU/HR                    -
0.522751+08
    REBOILER DUTY                   BTU/HR
0.523499+08


   ****   MANIPULATED VARIABLES   ****

                                                BOUNDS
CALCULATED
                                          LOWER        UPPER
VALUE
   MOLAR BOILUP RATIO                     1.0000       2100.0
159.02

   ****    DESIGN SPECIFICATIONS   ****

 NO  SPEC-TYPE     QUALIFIERS             UNIT       SPECIFIED
CALCULATED
                                                       VALUE
VALUE
  1  STDVOL-FRAC   STREAMS: 34                       0.10000E-01
0.10000E-01
                   COMPS:   N-BUT-01

   ****   MAXIMUM FINAL RELATIVE ERRORS   ****

    DEW POINT                      0.40025E-10  STAGE= 24
    BUBBLE POINT                   0.40240E-10  STAGE= 24
    COMPONENT MASS BALANCE         0.76336E-08  STAGE= 31 COMP=CIS-2-
01
    ENERGY BALANCE                 0.17121E-09  STAGE= 31


   ****    PROFILES   ****

   **NOTE** REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE
FLOWS
            FROM THE STAGE INCLUDING ANY SIDE PRODUCT.

                                          ENTHALPY
 STAGE TEMPERATURE   PRESSURE             BTU/LBMOL          HEAT DUTY
       F             PSIA            LIQUID        VAPOR        BTU/HR

   1   170.12        151.00        -59682.      -52713.      -.52275+08
   2   171.53        151.00        -59432.      -52402.
   3   172.50        151.00        -59293.      -52207.
   4   173.22        151.00        -59234.      -52094.
   5   174.25        151.00        -59024.      -51827.
```

```
  30   238.68        151.00       -69998.      -62225.
  31   240.67        151.00       -69851.      -62034.       .52350+08

 STAGE     FLOW RATE                    FEED RATE               PRODUCT
RATE
           LBMOL/HR                      LBMOL/HR
LBMOL/HR
       LIQUID     VAPOR       LIQUID    VAPOR    MIXED      LIQUID
VAPOR
   1  7181.      0.000                                      2143.5224
   2  5017.      7181.
   3  5002.      7161.
   4  7168.      7145.      2184.8541
   5  7148.      7127.
  30  6614.      6554.
  31  41.33      6572.                                        41.3317

    ****  MASS FLOW PROFILES  ****

 STAGE     FLOW RATE                    FEED RATE               PRODUCT
RATE
           LB/HR                         LB/HR                    LB/HR
       LIQUID     VAPOR       LIQUID    VAPOR    MIXED      LIQUID
VAPOR
   1 0.4140E+06  0.000                                      .12359+06
   2 0.2891E+06 0.4140E+06
   3 0.2882E+06 0.4127E+06
   4 0.4132E+06 0.4118E+06  .12654+06
   5 0.4115E+06 0.4102E+06
  30 0.4689E+06 0.4595E+06
  31  2954.     0.4659E+06                                  2953.6455

                         ****   MOLE-X-PROFILE     ****
   STAGE     WATER         PROPA-01      1-BUT-01      ISOBU-01
TRANS-01
      1    0.10785E-01   0.12053E-02   0.10798E-02   0.18772E-02
0.92729E-02
      2    0.12554E-01   0.55671E-03   0.10046E-02   0.17283E-02
0.98488E-02
      3    0.13920E-01   0.34352E-03   0.94738E-03   0.16183E-02
0.10189E-01
      4    0.14961E-01   0.27354E-03   0.90464E-03   0.15383E-02
0.10372E-01
      5    0.17190E-01   0.12370E-03   0.82617E-03   0.13907E-02
0.10813E-01
     30    0.17836E-01   0.45945E-14   0.11094E-05   0.13873E-05
0.33773E-03
     31    0.11410E-01   0.13564E-14   0.61096E-06   0.75364E-06
0.20502E-03

                         ****   MOLE-X-PROFILE     ****
```

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.27256E-01	0.33003	0.61413	0.19026E-02	0.24157E-02
2	0.30666E-01	0.28282	0.65350	0.25680E-02	0.46554E-02
3	0.33071E-01	0.25225	0.67671	0.31661E-02	0.76007E-02
4	0.34725E-01	0.23258	0.68916	0.36983E-02	0.11444E-01
5	0.38351E-01	0.19556	0.71908	0.47113E-02	0.11602E-01
30	0.52996E-02	0.30127E-04	0.19078E-01	0.30089E-01	0.90210
31	0.33921E-02	0.15485E-04	0.11500E-01	0.22699E-01	0.91615

**** MOLE-X-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.42342E-04	0.96114E-06
2	0.91453E-04	0.51064E-05
3	0.16360E-03	0.20325E-04
4	0.26871E-03	0.75711E-04
5	0.27241E-03	0.76086E-04
30	0.20440E-01	0.47916E-02
31	0.24035E-01	0.10589E-01

**** MOLE-Y-PROFILE ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.91930E-02	0.25734E-02	0.11460E-02	0.20138E-02	0.86216E-02
2	0.10785E-01	0.12053E-02	0.10798E-02	0.18772E-02	0.92729E-02
3	0.12025E-01	0.75086E-03	0.10271E-02	0.17729E-02	0.96764E-02
4	0.12980E-01	0.60204E-03	0.98711E-03	0.16960E-02	0.99139E-02
5	0.14982E-01	0.27512E-03	0.90988E-03	0.15472E-02	0.10431E-01
30	0.27458E-01	0.15507E-13	0.19956E-05	0.25303E-05	0.55030E-03
31	0.17876E-01	0.46148E-14	0.11125E-05	0.13913E-05	0.33856E-03

**** MOLE-Y-PROFILE ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.23928E-01	0.38024	0.56964	0.13898E-02	0.12348E-02
2	0.27256E-01	0.33003	0.61413	0.19026E-02	0.24157E-02

3	0.29645E-01	0.29696	0.64171	0.23688E-02	0.39850E-02
4	0.31326E-01	0.27558	0.65794	0.27871E-02	0.60453E-02
5	0.34907E-01	0.23393	0.69309	0.35881E-02	0.61967E-02
30	0.81855E-02	0.58133E-04	0.31317E-01	0.39411E-01	0.87384
31	0.53116E-02	0.30219E-04	0.19125E-01	0.30135E-01	0.90201

**** MOLE-Y-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.19216E-04	0.17825E-06
2	0.42342E-04	0.96114E-06
3	0.76752E-04	0.38655E-05
4	0.12722E-03	0.14516E-04
5	0.13088E-03	0.14739E-04
30	0.17070E-01	0.21035E-02
31	0.20417E-01	0.47551E-02

**** K-VALUES ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.85237	2.1351	1.0613	1.0728	0.92976
2	0.85910	2.1650	1.0748	1.0862	0.94153
3	0.86381	2.1858	1.0842	1.0955	0.94972
4	0.86757	2.2010	1.0912	1.1025	0.95587
5	0.87154	2.2241	1.1013	1.1126	0.96462
30	1.5395	3.3752	1.7988	1.8238	1.6294
31	1.5667	3.4022	1.8209	1.8461	1.6513

**** K-VALUES ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.87791	1.1521	0.92755	0.73045	0.51114
2	0.88882	1.1669	0.93975	0.74091	0.51891
3	0.89642	1.1772	0.94828	0.74819	0.52429
4	0.90213	1.1849	0.95470	0.75360	0.52827
5	0.91018	1.1962	0.96385	0.76160	0.53409

30	1.5445	1.9296	1.6416	1.3098	0.96868
31	1.5659	1.9514	1.6630	1.3276	0.98456

**** K-VALUES ****

STAGE	N-PEN-01	N-HEX-01
1	0.45384	0.18546
2	0.46299	0.18822
3	0.46915	0.19019
4	0.47346	0.19173
5	0.48046	0.19371
30	0.83512	0.43900
31	0.84950	0.44906

**** MASS-X-PROFILE ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.33700E-02	0.92184E-03	0.10508E-02	0.18268E-02	0.90239E-02
2	0.39246E-02	0.42600E-03	0.97814E-03	0.16827E-02	0.95890E-02
3	0.43522E-02	0.26289E-03	0.92249E-03	0.15758E-02	0.99209E-02
4	0.46759E-02	0.20926E-03	0.88056E-03	0.14974E-02	0.10096E-01
5	0.53798E-02	0.94759E-04	0.80528E-03	0.13555E-02	0.10540E-01
30	0.45323E-02	0.28578E-14	0.87800E-06	0.10980E-05	0.26729E-03
31	0.28764E-02	0.83700E-15	0.47969E-06	0.59172E-06	0.16097E-03

**** MASS-X-PROFILE ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.26524E-01	0.33271	0.61911	0.23810E-02	0.30231E-02
2	0.29857E-01	0.28526	0.65912	0.32151E-02	0.58286E-02
3	0.32202E-01	0.25445	0.68260	0.39643E-02	0.95171E-02
4	0.33801E-01	0.23452	0.69491	0.46291E-02	0.14324E-01
5	0.37381E-01	0.19746	0.72608	0.59051E-02	0.14542E-01
30	0.41943E-02	0.24700E-04	0.15641E-01	0.30622E-01	0.91809
31	0.26633E-02	0.12595E-04	0.93538E-02	0.22918E-01	0.92498

**** MASS-X-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.52987E-04	0.14366E-05
2	0.11450E-03	0.76361E-05
3	0.20485E-03	0.30397E-04
4	0.33635E-03	0.11319E-03
5	0.34144E-03	0.11391E-03
30	0.20802E-01	0.58246E-02
31	0.24266E-01	0.12770E-01

**** MASS-Y-PROFILE ****

STAGE	WATER	PROPA-01	1-BUT-01	ISOBU-01	TRANS-01
1	0.28711E-02	0.19673E-02	0.11147E-02	0.19588E-02	0.83860E-02
2	0.33700E-02	0.92184E-03	0.10508E-02	0.18268E-02	0.90239E-02
3	0.37585E-02	0.57447E-03	0.99991E-03	0.17259E-02	0.94198E-02
4	0.40574E-02	0.46065E-03	0.96100E-03	0.16511E-02	0.96517E-02
5	0.46889E-02	0.21076E-03	0.88689E-03	0.15081E-02	0.10167E-01
30	0.70551E-02	0.97530E-14	0.15969E-05	0.20248E-05	0.44038E-03
31	0.45428E-02	0.28706E-14	0.88053E-06	0.11012E-05	0.26796E-03

**** MASS-Y-PROFILE ****

STAGE	CIS-2-01	ISOBU-02	N-BUT-01	2:2-D-01	2-MET-01
1	0.23275E-01	0.38314	0.57398	0.17383E-02	0.15444E-02
2	0.26524E-01	0.33271	0.61911	0.23810E-02	0.30231E-02
3	0.28859E-01	0.29947	0.64714	0.29653E-02	0.49885E-02
4	0.30498E-01	0.27793	0.66355	0.34892E-02	0.75682E-02
5	0.34025E-01	0.23621	0.69985	0.44975E-02	0.77672E-02
30	0.65504E-02	0.48192E-04	0.25962E-01	0.40556E-01	0.89923
31	0.42040E-02	0.24777E-04	0.15681E-01	0.30671E-01	0.91805

**** MASS-Y-PROFILE ****

STAGE	N-PEN-01	N-HEX-01
1	0.24036E-04	0.26630E-06
2	0.52987E-04	0.14366E-05
3	0.96081E-04	0.57797E-05
4	0.15927E-03	0.21706E-04
5	0.16405E-03	0.22066E-04

30	0.17566E-01	0.25854E-02
31	0.20780E-01	0.57806E-02

****
***** HYDRAULIC PARAMETERS *****
****

*** DEFINITIONS ***

MARANGONI INDEX = SIGMA - SIGMATO
FLOW PARAM = (ML/MV)*SQRT(RHOV/RHOL)
QR = QV*SQRT(RHOV/(RHOL-RHOV))
F FACTOR = QV*SQRT(RHOV)
WHERE:
SIGMA IS THE SURFACE TENSION OF LIQUID FROM THE STAGE
SIGMATO IS THE SURFACE TENSION OF LIQUID TO THE STAGE
ML IS THE MASS FLOW OF LIQUID FROM THE STAGE
MV IS THE MASS FLOW OF VAPOR TO THE STAGE
RHOL IS THE MASS DENSITY OF LIQUID FROM THE STAGE
RHOV IS THE MASS DENSITY OF VAPOR TO THE STAGE
QV IS THE VOLUMETRIC FLOW RATE OF VAPOR TO THE STAGE

	TEMPERATURE F	
STAGE	LIQUID FROM	VAPOR TO
1	170.12	171.53
2	171.53	172.50
3	172.50	173.22
4	173.22	174.25
5	174.25	175.17
30	238.68	240.67
31	240.67	240.67

	MASS FLOW LB/HR		VOLUME FLOW CUFT/HR		MOLECULAR WEIGHT	
STAGE	LIQUID FROM	VAPOR TO	LIQUID FROM	VAPOR TO	LIQUID FROM	VAPOR TO
1	0.41401E+06	0.41401E+06	13311.	0.25589E+06	57.656	57.656
2	0.28913E+06	0.41272E+06	9281.6	0.25569E+06	57.628	57.636
3	0.28822E+06	0.41180E+06	9241.3	0.25552E+06	57.622	57.632
4	0.41319E+06	0.41023E+06	13236.	0.25542E+06	57.642	57.562

5	0.41146E+06	0.40851E+06	13162.	0.25523E+06	57.563	57.483
30	0.46887E+06	0.46592E+06	14849.	0.25534E+06	70.893	70.890
31	2953.6	0.0000	93.762	0.0000	71.462	

	DENSITY		VISCOSITY		SURFACE TENSION
	LB/CUFT		CP		DYNE/CM
STAGE	LIQUID FROM	VAPOR TO	LIQUID FROM	VAPOR TO	LIQUID FROM
1	31.103	1.6179	0.11045	0.94838E-02	6.0940
2	31.151	1.6142	0.10909	0.94977E-02	6.1009
3	31.188	1.6116	0.10825	0.95079E-02	6.1010
4	31.217	1.6061	0.10777	0.95256E-02	6.0987
5	31.260	1.6005	0.10679	0.95429E-02	6.1030
30	31.576	1.8247	0.11285	0.99236E-02	5.4625
31	31.502		0.11223		5.3977

	MARANGONI INDEX	FLOW PARAM	QR	REDUCED F-FACTOR
STAGE	DYNE/CM		CUFT/HR	(LB-CUFT)**.5/HR
1		0.22808	59942.	0.32549E+06
2	0.68695E-02	0.15947	59772.	0.32485E+06
3	0.77394E-04	0.15910	59646.	0.32438E+06
4	0.70221E-03	0.22846	59487.	0.32370E+06
5	0.43088E-02	0.22791	59290.	0.32290E+06
30	-.87159E-01	0.24192	63235.	0.34491E+06
31	-.64793E-01		0.0000	0.0000

```
    ************************************
    ***** TRAY SIZING CALCULATIONS *****
    ************************************
```

```
******************
*** SECTION   1 ***
******************
```

STARTING STAGE NUMBER	2
ENDING STAGE NUMBER	30
FLOODING CALCULATION METHOD	FAIR

DESIGN PARAMETERS

PEAK CAPACITY FACTOR	1.00000

SYSTEM FOAMING FACTOR		1.00000
FLOODING FACTOR		0.80000
MINIMUM COLUMN DIAMETER	FT	1.00000
MINIMUM DC AREA/COLUMN AREA		0.100000
HOLE AREA/ACTIVE AREA		0.12000

TRAY SPECIFICATIONS

TRAY TYPE		SIEVE
NUMBER OF PASSES		1
TRAY SPACING	FT	2.00000

***** SIZING RESULTS @ STAGE WITH MAXIMUM DIAMETER *****

STAGE WITH MAXIMUM DIAMETER		30
COLUMN DIAMETER	FT	13.6661
DC AREA/COLUMN AREA		0.100000
DOWNCOMER VELOCITY	FT/SEC	0.28120
FLOW PATH LENGTH	FT	9.38931
SIDE DOWNCOMER WIDTH	FT	2.13841
SIDE WEIR LENGTH	FT	9.92994
CENTER DOWNCOMER WIDTH	FT	0.0
CENTER WEIR LENGTH	FT	0.0
OFF-CENTER DOWNCOMER WIDTH	FT	0.0
OFF-CENTER SHORT WEIR LENGTH	FT	0.0
OFF-CENTER LONG WEIR LENGTH	FT	0.0
TRAY CENTER TO OCDC CENTER	FT	0.0

**** SIZING PROFILES ****

STAGE	DIAMETER	TOTAL AREA	ACTIVE AREA	SIDE DC AREA
	FT	SQFT	SQFT	SQFT
2	12.410	120.96	96.770	12.096
3	12.394	120.64	96.510	12.064
4	13.000	132.73	106.19	13.273
5	12.973	132.18	105.75	13.218
6	12.948	131.68	105.34	13.168
7	12.923	131.17	104.93	13.117
8	12.897	130.64	104.51	13.064
9	12.869	130.07	104.06	13.007
10	12.842	129.53	103.62	12.953
11	12.812	128.91	103.13	12.891
12	12.778	128.23	102.59	12.823
13	12.740	127.48	101.98	12.748
14	12.698	126.64	101.31	12.664
15	12.651	125.71	100.56	12.571
16	12.596	124.62	99.692	12.462
17	12.530	123.32	98.652	12.332
18	12.454	121.82	97.453	12.182
19	12.374	120.25	96.202	12.025

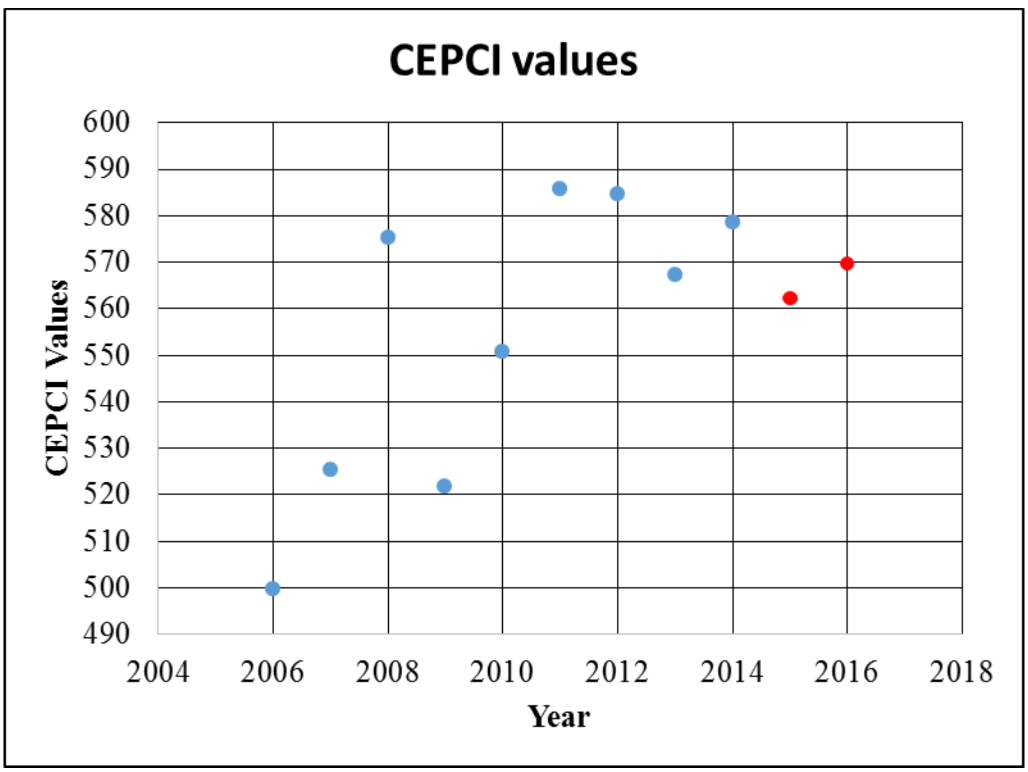
20	12.305	118.91	95.128	11.891
21	12.267	118.19	94.550	11.819
22	12.253	117.92	94.332	11.792
23	12.313	119.08	95.265	11.908
24	12.452	121.78	97.427	12.178
25	12.670	126.08	100.87	12.608
26	12.928	131.27	105.02	13.127
27	13.170	136.22	108.98	13.622
28	13.397	140.96	112.77	14.096
29	13.556	144.34	115.47	14.434
30	13.666	146.68	117.35	14.668

Sample Calculation(s)

CEPCI Values

Year	CEPCI	
2006	499.6	
2007	525.4	
2008	575.4	
2009	521.9	
2010	550.8	
2011	585.7	
2012	584.6	
2013	567.3	
2014	578.6	
2015	562.337	Extrapolated using the growth function =GROWTH(B2:B10,A2:A10,A11:A12,FALSE)
2016	**569.777**	Extrapolated using the growth function =GROWTH(B3:B11,A3:A11,A12:A13,FALSE)

Reference: http://accessintelligence.imirus.com/Mpowered/book/vche14/i11/p186



Graph A-4: *CEPCI Values*

Reactor R-1 Design

Reactor (Horizontal Packed Bed)	R-1	Notes
T (°F)	300	Given range 300-400 F in the memo
F(T)	0.666	F(T) = 0.867 - 0.00067*T Given in the memo
iC4	707.4358	Isobutane molar flow rate entering the reactor
nC4	1316.396	N-butane molar flow rate entering the reactor
iC4/nC4	0.537403486	Molar ratio at the reactor inlet
% Conv	48.65072355	100(F(T)-(iC4/nC4)*(1-F(T))] Given in the memo
Pressure (psig)	430	Specified in the memo = 450 psig - 20 psi (pressure drop)
Volume necessary (ft^3)	900	Specified in the memo
Volume (m^3)	25.48512	Using the conversion factor 1 ft3 = 0.0283168 m3
L/D Ratio	4.5	Turton Heuristic
density (lb/ft3)	5.586717	Aspen Simulation RFEED
m (lb/h)	123586	Aspen Simulation RFEED
Q_V (ft^3/s)	6.144833262	Calculated by dividing m/density and converting to s
Actual Height (ft)	29.5	Found via optimization
Diameter (ft)	6.555555556	Found via optimization
Diameter rounded (ft)	6.25	Rounded to nearest .25 in
Actual Diameter (m)	1.905	1 ft = 0.3048 m
Actual Volume (ft^3)	905.0486649	Calculated using pi*r^2*h; solver
Actual Volume (m^3)	25.62808203	Using the conversion factor 1 ft3 = 0.0283168 m3
Pressure (psia)	444.7	Converted to psia from psig
	489.17	1.1 * Column Pressure
	494.7	50 psi + Column Pressure
Design Pressure (psia)	494.7	Largest of the 1.1*CP or 50+CP
Pressure (barg)	29.64745635	psig converted to barg using 1 psig = 0.0689475729 barg
L/D ratio Check	4.72	Must be between 2.5<L/D<5

Capital Cost		Notes
$F_{P,vessel}$	6.071836577	Turton Equation A.2 for tvessel > 0.0063 m
$F_{P,vessel}$	1	If tvessel < 0.0063 m and Fp,vessel < 1
$C_p°$	21995.93561	Turton Table A.1 (min size = 0.3 and max size = 520 m3)
K_1	3.4974	Turton Table A.1 Vertical Process Vessel
K_2	0.4485	Turton Table A.1 Vertical Process Vessel
K_3	0.1074	Turton Table A.1 Vertical Process Vessel
B_1	2.25	Turton Table A.4 Vertical Process Vessel
B_2	1.82	Turton Table A.4 Vertical Process Vessel

F_M	1	Turton Figure A.18 ID 18 CS
C_{BM} (2001)	1755373.663	6*CBM of Process Vessel Note from Dr. Ramsey
C_{BM} (2016)	**$2,538,608.92**	Using extrapolated CEPCI value of 569.8 for 2016

Cost of the Catalyst (Platinum)		**Notes**
Catalyst (lbm)	44200	Specified in the memo
Price/lbm	13.25	Specified in the memo
Cost/yr	**$585,650.00**	

De-Isobutanizer Tower T-1 Design

De-Isobutanizer Tower	**T-1**	**Notes**
# of Theoretical Stages	25	Does not include the total condenser
# of Computational Stages	26	Entered into Aspen; optimized number
# of Trays Inside the Column	24	Excluding the condenser and the reboiler
Column Diameter (ft)	17.47	Calculated from Aspen Plus sizing function
Column Diameter rounded (ft)	18	Rounded from above
Column Diameter (m)	5.4864	1 ft = 0.3048 m
Column Diameter rounded (m)	5.49	Rounded from above
Tray Spacing/tray (ft)	2	Turton [1] Table 11.14 Heuristics for Tray Towers #1
$h_{Overhead}$ (ft)	3	Dr. Wagner & Dr. Whiteley Distillation Course Notes Column Tray Spacing + 1 ft
h_{trays} (ft)	60	Dr. Wagner & Dr. Whiteley Distillation Course Notes Nactual*tray spacing
h_{Btm} (ft)	7	Dr. Wagner & Dr. Whiteley Distillation Course Notes Rule of Thumb
h_{skirt} (ft)	8	CHE 4224 Exam 1 Problem Solution 1
h_{total} (ft)	78	Sum of hoverhead + htrays + hbtm + hskirt
h for volume (ft)	70	Sum of hoverhead + htrays + hbtm
Volume (ft^3)	17812.83035	Volume of the column using pi*r^2*h
Volume (m^3)	504.4023543	1 ft^3 = 0.0283168 m3
Column Pressure (psig)	135.3	Column pressure specified on Aspen after optimization
	148.83	1.1 * Column Pressure
	185.3	50 psi + Column Pressure
Design Pressure (psig)	185.3	Largest of the 1.1*CP and 50+CP
Design Pressure (barg)	12.77598524	1 psig = 0.0689475728 barg
Column Area (ft^2)	254.4690049	pi*r^2
Downcomer area/column area (ft^2)	0.1	DA/CA ratio from Aspen Simulation
Downcomer Area (ft^2)	25.44690049	Column Area * DA/CA ratio
Tray Area (ft^2)	229.0221044	Column area - downcomer area
Tray Area (m^2)	21.27667338	1 ft2 = 0.09290227 m2
Overall Efficiency	0.8	Using Figure 7.32 (O'Connel Correlation) of CHE 4224 Course Notes by Dr. Whiteley
Nactual	30	(Nstages/Eo)=Nactual

Capital Cost: Tower	Notes

$F_{P,vessel}$	7.630982773	Turton Equation A.2 for tvessel > 0.0063 m
$C_p°$	311998.8331	Turton Equation A.1 (Volume limits 0.3 - 520 m3)
K_1	3.4974	Turton Table A.1
K_2	0.4485	Turton Table A.1
K_3	0.1074	Turton Table A.1
B_1	2.25	Turton Table A.4 Vertical Towers
B_2	1.82	Turton Table A.4 Vertical Towers
F_M	1	Figure A.18 CS ID 18
C_{BM} (2001)	5035158.426	Turton Equation A.4
C_{BM} (2016)	**$7,281,810.33**	CBM*(2014/2001); Using extrapolated CEPCI value of 569.8 for 2016

Capital Cost: Trays (Sieve)		Notes
$C_p°$	19331.4022	Turton Equation A.1 (Area limits 0.07 - 12.30 m2)
K_1	2.9949	Turton Table A.1
K_2	0.4465	Turton Table A.1
K_3	0.3961	Turton Table A.1
N	24	# of trays inside the column
F_Q	1	Turton Table A.5 Fq = 1 since N >= 20
F_{BM}	1	Turton Figure A.19 ID 60 CS
C_{BM} (2001)	463953.6528	Turton Table A.5 Sieve trays
C_{BM} (2016)	**$670,966.48**	CBM*(2014/2001); Using extrapolated CEPCI value of 569.8 for 2016
TOTAL COST	**$7,952,776.81**	Total cost of the column and trays

De-Butanizer Tower T-2 Design

De-Butanizer Tower	T-2	Notes
# of Theoretical Stages	30	Does not include the total condenser
# of Computational Stages	31	Entered into Aspen; optimized number
# of Trays Inside the Column	29	Excluding the condenser and the reboiler
Column Diameter (ft)	13.67	Calculated from Aspen Plus sizing function
Column Diameter rounded (ft)	14	Rounded from above
Column Diameter (m)	4.2672	1 ft = 0.3048 m
Column Diameter rounded (m)	4.27	Rounded from above
Tray Spacing/tray (ft)	2	Turton [1] Table 11.14 Heuristics for Tray Towers #1
$h_{Overhead}$ (ft)	3	Dr. Wagner & Dr. Whiteley Distillation Course Notes Column Tray Spacing + 1 ft
h_{trays} (ft)	68	Dr. Wagner & Dr. Whiteley Distillation Course Notes Nactual*tray spacing
h_{Btm} (ft)	7	Dr. Wagner & Dr. Whiteley Distillation Course Notes Rule of Thumb
h_{skirt} (ft)	8	CHE 4224 Exam 1 Problem Solution 1
h_{total} (ft)	86	Sum of hoverhead + htrays + hbtm + hskirt
h for volume (ft)	78	Sum of hoverhead + htrays + hbtm
Volume (ft^3)	12007.16712	Volume of the column using pi*r^2*h
Volume (m^3)	340.00455	1 ft3 = 0.0283168 m3
Column Pressure (psig)	136.3	Column pressure specified on Aspen after optimization
	149.93	1.1 * Column Pressure
	186.3	50 psi + Column Pressure
Design Pressure (psig)	186.3	Largest of the 1.1*CP and 50+CP
Design Pressure (barg)	12.84493281	1 psig = 0.0689475728 barg
Column Area (ft^2)	153.93804	pi*r^2
Downcomer area/column area (ft^2)	0.1	DA/CA ratio from Aspen Simulation
Downcomer Area (ft^2)	15.393804	Column Area * DA/CA ratio
Tray Area (ft^2)	138.544236	Column area - downcomer area
Tray Area (m^2)	12.87107402	1 ft2 = 0.09290227 m2
Overall Efficiency	0.85	Using Figure 7.32 (O'Connel Correlation) of CHE 4224 Course Notes by Dr. Whiteley
Nactual	34	(Nstages/Eo)=Nactual

Capital Cost: Tower	Notes

$F_{P,vessel}$	6.074352731	Turton Equation A.2 for tvessel > 0.0063 m
$C_p°$	209427.9611	Turton Equation A.1 (Volume limits 0.3 - 520 m3)
K_1	3.4974	Turton Table A.1
K_2	0.4485	Turton Table A.1
K_3	0.1074	Turton Table A.1
B_1	2.25	Turton Table A.4 Vertical Towers
B_2	1.82	Turton Table A.4 Vertical Towers
F_M	1	Figure A.18 CS ID 18
C_{BM} (2001)	2786506.452	Turton Equation A.4
C_{BM} (2016)	**$4,029,825.83**	CBM*(2014/2001); Using extrapolated CEPCI value of 569.8 for 2016

Capital Cost: Trays (Sieve)		Notes
$C_p°$	9506.904657	Turton Equation A.1 (Area limits 0.07 - 12.30 m2)
K_1	2.9949	Turton Table A.1
K_2	0.4465	Turton Table A.1
K_3	0.3961	Turton Table A.1
N	29	# of trays inside the column
F_Q	1	Turton Table A.5 Fq = 1 since N >= 20
F_{BM}	1	Turton Figure A.19 ID 60 CS
C_{BM} (2001)	275700.2351	Turton Table A.5 Sieve trays
C_{BM} (2016)	**$398,715.72**	CBM*(2014/2001); Using extrapolated CEPCI value of 569.8 for 2016
TOTAL COST	**$4,428,541.55**	Total cost of the column and trays

Condensate Receiver for De-Isobutanizer Tower D-1 Design

		Notes
L/D Ratio	3	Turton Table 11.6 heuristic; 2.5<L/D<5
m (lb/h)	7.58E+05	Aspen results vapor to stage 1
density (lb/ft^3)	30.688	Aspen results liquid from stage 1
Q_V (ft^3/s)	6.863884544	Converting mass flow rate to volumetric flow rate using density
Volume (ft^3)	2059.165363	Volume stored in five minutes (Turton Table 11.6, heuristic #5); holdup time is 5 min for half-full reflux drums
Diameter (ft)	12	For horizontal drum, diameter is the height of the drum; calculated from L/D ratio
Diameter (m)	3.6576	1 ft = 0.3048 m
Diameter (ft) - rounded	12.25	Round up to nearest 3" increment
Diameter (m) - rounded	3.7338	Converting to m using 0.3048
Length (ft)	36	Guess and check length until the volume of drum is twice volume of condensed vapor after 5 minutes
Voume drum (ft^3)	4242.917228	Calculated volume of drum
Voume drum (m^3)	120.1458386	1 ft3 = 0.0283168 m3
Voume necessary (ft^3)	4118.330726	(Volume after 5 minutes)*2 = volume necessary for drum
Pressure (psig)	135.3	From Aspen Simulation
Pressure (barg)	9.3286066	1 psig = 0.0689475728 barg

Capital Cost: Horizontal Process Vessel

$F_{P,vessel}$	4.127281507	Turton Equation A.2 for tvessel > 0.0063 m
$F_{P,vessel}$	1	If tvessel < 0.0063 m and Fp,vessel < 1
$C_p°$	54103.77509	Turton Equation A.1 (Volume limits 0.1 - 628 m3)
K1	3.5565	Turton Table A.1
K2	0.3776	Turton Table A.1
K3	0.0905	Turton Table A.1

B_1	1.49	Turton Table A.4
B_2	1.52	Turton Table A.4
F_M	1	Figure A.18 ID 18 CS
C_{BM}	420032.9206	For 2001
C_{BM} (2016)	**$607,448.62**	Using extrapolated CEPCI at 2016 of 569.8 and 2001 CEPCI of 394

Capital Cost: Vertical Process Vessel

$F_{P,vessel}$	4.127281507	Turton Equation A.2 for tvessel > 0.0063 m
$F_{P,vessel}$	1	If tvessel < 0.0063 m and Fp,vessel < 1
$C_p°$	78464.75127	Turton Equation A.1 (Volume limits 0.3 - 520 m3)
K1	3.4974	Turton Table A.1
K2	0.4485	Turton Table A.1
K3	0.1074	Turton Table A.1
B_1	2.25	Turton Table A.4
B_2	1.82	Turton Table A.4
F_M	1	Figure A.18 ID 18 CS
C_{BM}	765945.623	For 2001
C_{BM} (2016)	**$1,107,705.12**	CBM*(2014/2001); Using extrapolated CEPCI value of 569.8 for 2016

Condensate Receiver for De-Butanizer Tower D-2 Design

		Notes
L/D Ratio	3	Turton Table 11.6 heuristic; 2.5<L/D<5
m (lb/h)	4.14E+05	Aspen results vapor to stage 1
density (lb/ft3)	31.103	Aspen results liquid from stage 1
Q_V (ft^3/s)	3.69748184	Converting mass flow rate to volumetric flow rate using density
Volume (ft3)	1109.24455	Volume stored in five minutes (Turton Table 11.6, heuristic #5); holdup time is 5 min for half-full reflux drums
Diameter (ft)	10	For horizontal drum, diameter is the height of the drum; calculated from L/D ratio
Diameter (m)	3.048	1 ft = 0.3048 m
Diameter (ft) - rounded	10	Round up to nearest 3" increment
Diameter (m) - rounded	3.05	
Length (ft)	30	Guess and check length until the volume of drum is twice volume of condensed vapor after 5 minutes
Voume drum (ft^3)	2356.19449	Calculated volume of drum
Voume drum (m^3)	66.7198881	1 ft3 = 0.0283168 m3
Voume necessary (ft^3)	2218.48911	(Volume after 5 minutes)*2 = volume necessary for drum
Pressure (psig)	136.3	From Aspen Simulation
Pressure (barg)	9.39755417	1 psig = 0.0689475728 barg

Capital Cost: Horizontal Process Vessel

$F_{P,vessel}$	3.48291445	Turton Equation A.2 for tvessel > 0.0063 m
$F_{P,vessel}$	1	If tvessel < 0.0063 m and Fp,vessel < 1
$C_p°$	35197.7115	Turton Equation A.1 (Volume limits 0.1 - 628 m3)
K1	3.5565	Turton Table A.1
K2	0.3776	Turton Table A.1
K3	0.0905	Turton Table A.1

B_1	1.49	Turton Table A.4
B_2	1.52	Turton Table A.4
F_M	1	Figure A.18 ID 18 CS
C_{BM}	238782.33	For 2001
C_{BM} (2016)	**$345,325.31**	Using extrapolated CEPCI at 2016 of 569.8 and 2001 CEPCI of 394

Capital Cost: Vertical Process Vessel

$F_{P,vessel}$	3.48291445	Turton Equation A.2 for tvessel > 0.0063 m
$F_{P,vessel}$	1	If tvessel < 0.0063 m and Fp,vessel < 1
$C_p°$	47097.2733	Turton Equation A.1 (Volume limits 0.3 - 520 m3)
K1	3.4974	Turton Table A.1
K2	0.4485	Turton Table A.1
K3	0.1074	Turton Table A.1
B_1	2.25	Turton Table A.4
B_2	1.82	Turton Table A.4
F_M	1	Figure A.18 ID 18 CS
C_{BM}	404513.973	For 2001
C_{BM}	**$592,294.70**	CBM*(2014/2001); Using extrapolated CEPCI value of 569.8 for 2016

Given:
Tube OD: 1"
Single Pass
5/8" fins, 10 fins/in
2 1/4" triangular pitch

De-Isobutanizer Tower

Condenser (Air-cooled HEX) E-1 Design

		Notes for GPSA Method (extended area)
Distillate Flow Rate (lb/hr)	62692.53	From Aspen Simulation for T1Top
Reflux Ratio	11.0956	From Aspen Simulation solved using Spec & Vary Feature
ṁ (lb/hr) (Vapor Flow Rate)	758303.766	Calculated from RR = L/D, V = L+D
P_{inlet} (psia)	150	Pressure of the vapor at the condenser inlet
U_x (Btu/hr ft^2 ^oF)	4.66	From Fig. 10-10, Hydrocarbon condensers, interpolated for Ux at 1.31 F range
$T_{H,in}$ (^oF) (T_1 - GPSA notation)	154.49	From Aspen, temperature at stage 2 (temperature going into the condenser); see attached stream summary
$T_{H,out}$ (^oF) (T_2 - GPSA notation)	153.18	From Aspen, temperature at stage 1 (temperature of stream at condenser outlet); see attached stream summary table
$T_{C,in}$ (^oF) (t_1 - GPSA notation)	105	Air inlet temperature; given in Utility Specifications
$T_{C,out}$ (^oF) (t_2 - GPSA notation)	114.49	Calculated by using the approach temperature; ΔT (approach) = Th,in - Tc,out --> 173.29 -38.29
ΔT_1 (^oF)	48.18	Approach temperature = Th,out - Tc,in
ΔT_2 (^oF)	40	Approach temperature heuristic from Turton, Table 11.11 for air coolers
LMTD	43.9632391	LMTD = (ΔT1-ΔT2)/(LN(ΔT1/ΔT2))
Δt_a (oF)	27.64061	Using GPSA equation from step 3 of section 10 (p. 10-10)

P	0.19175591	GPSA Fig. 10-8 1-Pass Cross Flow, Both Fluids Unmixed
R	0.13804004	GPSA Fig. 10-8 1-Pass Cross Flow, Both Fluids Unmixed
F	0.97	Read off the Fig. 10-8 using P and R
CMTD	42.644342	CMTD = LMTD * F
Condenser Duty (Btu/hr)	91,417,500	From Aspen Report; see attached stream summary table
Required Extended Surface Area, A_x (ft^2)	**460026**	GPSA Step 4 pg. 10-10 using Ux
Required Extended Surface Area, A_x (m^2)	42737.7955	Conversion to m2 (1 ft2 = 0.092903 m2)
APSF	118.8	Fig. 10-11; 5/8 in by 10, APSF 2 1/4 in. triangular pitch with 4 rows
Face Area, F_a (ft^2)	3872.27273	GPSA Step 5, section 10 (p. 10-10)
Assumed Tube Length (ft)	249	GPSA Example provided Fa=467 with Tube Length = 30 ft, so scaled to Fa = 187 for reasonable assumption
Width (ft)	15.6	GPSA Step 6, Section 10 (p 10-10), rounded to 1 decimal
APF	5.58	GPSA from figure 10-11 for 5/8" by 10 fins
Number of Tubes, N_t	332	GPSA step 7, p. 10-10, rounded to nearest tube
Tube Area, A_t (in^2)	0.5945	GPSA Figure 9-25; Tube OD 1" (given in PW#6 problem statement), and assume BWG 16
Tube Area, A_t (ft^2)	0.00412847	Converted from in2 to ft2 using At/144
Tube Side Mass Velocity, G_t (lb/ft^2*s)	153.678553	GPSA Step 8, p. 10-10, where Gt = (vapor mass flow rate)/(3600*Nt*At)
Tube ID, D_i (in)	0.87	Using characteristics of tubing BWG 16 from GPSA
Tube ID, D_i (ft)	0.0725	Using characteristics of tubing BWG 16 from GPSA
$\dot{m}_{air}$ (lb/hr), W_a - GPSA notation	13780674.5	Skipping to step 12 to estimate air quantity; Wa = Q/0.24*Δta1; Cpair = 0.24 Btu/lb*oF
Air Face Mass Velocity, G_a (lb/ft^2*h)	3558.80784	GPSA Step 13, p 10-10
Air-side film coefficient, h_a (Btu/h*ft2*°F)	10	GPSA Figure 10-17 read using Ga
AR	21.4	GPSA Figure 10-11 for 5/8" tubes by 10

GPSA Method (bare tube area) | **Notes**

U_b (Btu/hr ft^2 °F)	99.724	Converted to bare tube area using Ux (extended surface)*AR ratio
Required Bare Surface Area, A_b (ft^2)	**21497**	GPSA Step 4 pg. 10-10 using Ub
Required Bare Surface Area, A_b (m^2)	1997.13579	Conversion to m2 (1 ft2 = 0.092903 m2)

GPSA Method (fan sizing) | **Notes**

Number of Fans, N_f	1	Assumed that the process uses one fan
Minimum fan area, FAPF (ft^2)	1548.90909	GPSA Step 16, p 10-14
Fan diameter (ft)	45	GPSA Step 17, p 10-14, round up the fan diameter to nearest foot
Pressure drop factor, Fp	0.16	GPSA Figure 10-18, using air face mass velocity
T_a, avg (°F)	109.745	Average air temperature
Number of rows	4	Assumed; also assumed for APSF
Air density ratio, Dr	0.91	GPSA Figure 10-16; Using average air temperature at 992.33 ft from Utility Specifications
Air static pressure drop, ΔP_a	0.7032967	GPSA Step 18, p 10-14
Actual air volume (ACFM)	3369737.07	GPSA Step 19, p 10-14; Using equation ACFM = Wa/(Dr*60*0749)
Fan total pressure, PF (in H_2O)	0.95797889	GPSA step 20, p 10-14
BHP, per fan	725.554473	GPSA step 21, p 10-14; Assuming 70% fan efficiency; conversion factor used (6356 = (33000 ft*lb/min*hp)*(12/ft)*(ft3/62.3 lb)
Efficiency of fan motor speed reducer	0.92	Given in step 21 for GPSA
Actual fan motor needed (bhp)	788.646166	Actual fan motor/efficiency of speed reducer

Motor size, rounded to discrete size, bhp	800	Using discrete motor sizes from Pump Notes
Motor size, kW	596.56	Converting to kw (1 hp = 0.7457 kw)

Capital Cost

$C_p°$	222634.326	Turton Equation A.1; Turton analysis uses bare surface area
K_1	4.0336	Turton Table A.1
K_2	0.2341	Turton Table A.1
K_3	0.0497	Turton Table A.1
B_1	0.96	Turton Table A.4
B_2	1.21	Turton Table A.4
F_M	1	Turton from from Figure A.18 (ID# 10); Assumed carbon steel
Pressure (psia)	150	Assumed from condenser pressure given
Pressure (barg)	9.30792234	conversion to barg ((200-15)*0.0689475729)
$F_{P,vessel}$	0.99304736	Turton using equation A.3
C_1	-0.125	Turton Table A.2
C_2	0.15361	Turton Table A.2
C_3	-0.02861	Turton Table A.2
C_{BM} (2001)	$481,243.53	Turton Equation A.4
C_{BM} (2016)	**$695,970.98**	Using extrapolated CEPCI value of 569.8 for 2016

Motor Capital Cost

$C_p°$	100151.876	Turton Equation A.1
K_1	2.4604	Turton Table A.1
K_2	1.4191	Turton Table A.1
K_3	-0.1798	Turton Table A.1

F_{BM}	1.5	Turton, Figure A.19 (ID# 13)
Shaft Power, kW	554.8008	Using Pump Notes: brake hp (15)*ηmotor (0.93)
C_{BM} (2001)	150227.815	Turton Equation A.4
C_{BM} (2016)	**$217,258.40**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost

Purchased (hp)	860.215054	Using Pump Notes: brake hp (15)*ηmotor (0.93)
Purchased (kW)	641.462366	Converting purchased (hp) to kW using 1 hp = 0.7457 kW conversion factor
Service Factor	0.96	Given in problem statement PW #5 last semester
$ per kWh	0.07	Specified in the Memorandum
Operating Cost Per Year	**$377,610.93**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.07/kWh * purchase kW
Total Capital Cost	**$913,229.38**	

Reboiler (Kettle) for De-Isobutanizer E-2 Design		Notes
Reboiler Duty (Btu/hr)	7.32E+07	From Aspen Simulation
$T_{H,in}$ (°F)	550	Using Process Conditions medium pressure recommended value (Ch. 6) Class Notes
$T_{H,out}$ (°F)	550	This temperature is the same because the steam is being condensed; all phase changes occur at constant temperatures
$T_{C,in}$ (°F)	168.955	From Aspen Simulation; going into the reboiler Stage 25
$T_{C,out}$ (°F)	170.577	From Aspen Simulation; coming out of the reboiler Stage 26
Approach Temp (°F)	20	Heuristic for steam reboiler approach temperature Gas Purification by Cole & Nielsen
LMTD	380.2334234	Using the LMTD equation given in class from Dr. Aichele
U (Btu/hr ft² °F)	140	Heuristic given in Dr. Aichele's class notes on estimating U; U chosen for steam reboilers; lowest for conservative purposes
F (correction factor)	0.9	Assumed using the heuristics in Table 11.11
Area (ft²)	**1529**	Rearraging the HEX design equation; Q = F*LMTD*U*A
Area (m²)	142.048687	Converting from ft2 to m2 using the conversion factor 1 ft2 = 0.092903 m^2
Column Pressure (psia)	150	From Aspen Simulation Last Stage Pressure
Computational Stages	26	From Aspen Simulation Input
Trays Inside the Column	24	
Overall Efficiency	0.8	Using Figure 7.32 (O'Connel Correlation) of CHE 4224 Course Notes by Dr. Whiteley/Wagner
Nactual	30	(Nstages/Eo)=Nactual
Final Reboiler Pressure (psia)	153	Column P + Nactual*0.1 (0.1 psi pressure drop per stage) from PW#4 last semester
Reboiler Pressure (barg)	9.51476506	Converting to barg ((Pressure (psia)-15)*0.0689475729)
ΔT_{steam}	0	Because steam is being condensed; phase changes occur at constant temperatures
$T_{avg, steam}$	550	Same average temperature due to phase change
λ_{steam} (Btu/lbm)	970.4	Latent heat of vaporization for steam (engineeringtoolbox.com)
$\dot{m}_{steam}$ (lb/hr)	75475.57708	Solving for the mass flow rate of steam using Q = mdot*λ

Capital Cost

$C_p°$	144912.1637	Turton Equation A.1
K_1	4.4646	Turton Table A.1
K_2	-0.5277	Turton Table A.1
K_3	0.3955	Turton Table A.1
B_1	1.63	Turton Table A.4
B_2	1.66	Turton Table A.4
F_M	1	Turton from from Figure A.18 (ID# 10); Assumed carbon steel for shell-and-tubes
Pressure (psia)	194.7	Boiler feed water pressure given in the Memorandum
Design Pressure (psia)	244.7	Add 50 psi for design pressure (greater than 1.1*operating pressure)
Pressure (barg)	15.8372575	conversion to barg ((200-15)*0.0689475729)
$F_{P,vessel}$	1.019860005	Turton using equation A.3 using the design pressure
C_1	-0.00164	Turton Table A.2
C_2	-0.00627	Turton Table A.2
C_3	0.0123	Turton Table A.2
C_{BM} (2001)	$481,538.43	Turton Equation A.4
C_{BM} (2016)	**$696,397.45**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost

Hours/yr	8409.6	365 days/yr * 24 hrs/day * 0.96 service factor
Service Factor	0.96	Given in problem statement PW #5
Mass of steam/yr	634719413	Multiplying hours/yr by mdot
$ per lb steam	$0.00330	Given in the memorandum
Operating Cost Per Year	**$2,094,574.06**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.06/kWh * purchase kW

Shell-and-Tube Cooler (Floating Head) - Isobutane Product HEX E-3 Design

		Notes
Cooler Duty (Btu/hr)	1.05E+06	From Aspen Simulation - Cooler Duty
Molar flow rate (lbmol/hr)	1085.574	From Aspen Simulation - molar flow rate
Cp (Btu/lbmol F)	39.7919	From Aspen Simulation - under material properties
$\Delta T_{Bottoms}$ (^oF)	25.182	Hot in - Hot out of the bottoms stream through a cooler
$T_{H,in}$ (^oF)	153.182	From Aspen Simulation T1TOP
$T_{H,out}$ (^oF)	128	From Aspen Simulation COOLPROD
$T_{C,in}$ (^oF)	87	Specified in the Memorandum
$T_{C,out}$ (^oF)	120	Specified in the Memorandum
LMTD	36.95326844	Using the LMTD equation given in class from Dr. Aichele
U (Btu/hr ft^2 ^oF)	80	Heuristic given in Dr. Aichele's class notes on estimating U; lowest for conservative purposes
F (correction factor)	0.9	Assumed using the heuristics in Table 11.11
Area (ft^2)	**394**	Rearraging the HEX design equation; Q = F*LMTD*U*A
Area (m^2)	36.603782	Converting from ft2 to m2 using the conversion factor 1 ft2 = 0.092903 m^2
Pressure (psia)	150	Aspen Simulation
Pressure (barg)	9.307922342	Converting to barg ((Pressure (psia)-15)*0.0689475729)
$\dot{m}_{water}$ (lbmol/hr)	1767.098498	Solving for the mass flow rate of water using Q/CpdT of water
$\dot{m}_{water}$ (lb/hr)	31807.77297	Converting lbmol/hr flow rate to lb/hr using 1 lbmol = 18 lbm of water
ρ_{water} (lb/ft3)	62.42	Density of water from Dr. Whiteley's Unit Conversions Packet
$\dot{m}_{water}$ (gpm)	63.53163388	Converting from lb/hr to gpm dividing by density and 1 gpm = 8.020833 ft3/hr
Cp (Btu/lbmol F)	17.9691	From ChemCad simulation for water at 90 F
ΔT_{water} (^oF)	33	Cold out - cold in for water

Capital Cost (Shell-and-Tube - Floating Head)

$C_p°$	19023.65933	Turton Equation A.1
K_1	4.8306	Turton Table A.1
K_2	-0.8509	Turton Table A.1
K_3	0.3187	Turton Table A.1
B_1	1.63	Turton Table A.4
B_2	1.66	Turton Table A.4
F_M	1	Turton from from Figure A.18 (ID# 1); Assumed carbon steel for shell-and-tubes
Pressure (psia)	150	Medium pressure steam given in Chapter 6 notes under recommended value
Design Pressure (psia)	200	Add 50 psi for design pressure (greater than 1.1*operating pressure)
Pressure (barg)	12.75530099	conversion to barg ((200-15)*0.0689475729)
$F_{P,vessel}$	1.014996882	Turton using equation A.3 using the design pressure
C_1	-0.00164	Turton Table A.2
C_2	-0.00627	Turton Table A.2
C_3	0.0123	Turton Table A.2
C_{BM} (2001)	$63,061.43	Turton Equation A.4
C_{BM} (2016)	**$91,198.99**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost

Converting to per min basis	504576	365 days/yr * 24 hrs/day *60 min* 0.96 service factor
Service Factor	0.96	Given in problem statement PW #5 last semester
Cooling water (gpm)	63.53163388	Imported from above
$/ gpm year cooling water	120	From Dr. Aichele & Dr. Ramsey Memorandum
$/ gpm	2.28E-04	$/gpm year / 365 * 60 * 24
Operating Cost Per Year	**$7,318.84**	gpm*$/gpm*365 days/yr * 24 hrs/day *60 min* 0.96 service factor

Given:
Tube OD: 1"
Single Pass
5/8" fins, 10 fins/in
2 1/4" triangular pitch

De-Butanizer Tower

Condenser (Air-cooled HEX)
E-4 Design

		Notes for GPSA Method (extended area)
Distillate Flow Rate (lb/hr)	123586	From Aspen Simulation for T2Top
Reflux Ratio	2.35	From Aspen Simulation solved using Spec & Vary Feature
$\dot{m}$ (lb/hr) (Vapor Flow Rate)	414013.1	Calculated from RR = L/D, V = L+D
P_{inlet} (psia)	151	Pressure of the vapor at the condenser inlet
U_x (Btu/hr ft^2 ^{o}F)	4.658	From Fig. 10-10, Hydrocarbon condensers, interpolated for Ux at 1.40 F range
$T_{H,in}$ (^{o}F) (T_1 - GPSA notation)	171.53	From Aspen, temperature at stage 2 (temperature going into the condenser); see attached stream summary
$T_{H,out}$ (^{o}F) (T_2 - GPSA notation)	170.13	From Aspen, temperature at stage 1 (temperature of stream at condenser outlet); see attached stream summary table
$T_{C,in}$ (^{o}F) (t_1 - GPSA notation)	105	Air inlet temperature; given in Utility Specifications
$T_{C,out}$ (^{o}F) (t_2 - GPSA notation)	131.53	Calculated by using the approach temperature; ΔT (approach) = Th,in - Tc,out --> 173.29 -38.29
ΔT_1 (^{o}F)	65.13	Approach temperature = Th,out - Tc,in
ΔT_2 (^{o}F)	40	Approach temperature heuristic from Turton, Table 11.11 for air coolers
LMTD	51.5481027	LMTD = (ΔT1-ΔT2)/(LN(ΔT1/ΔT2))
Δt_a (oF)	37.246614	Using GPSA equation from step 3 of section 10 (p. 10-10)
P	0.39876747	GPSA Fig. 10-8 1-Pass Cross Flow, Both Fluids Unmixed
R	0.05277045	GPSA Fig. 10-8 1-Pass Cross Flow, Both Fluids Unmixed

F	0.99	Read off the Fig. 10-8 using P and R
CMTD	51.0326217	CMTD = LMTD * F
Condenser Duty (Btu/hr)	52,255,200	From Aspen Report; see attached stream summary table
Required Extended Surface Area, A_x (ft^2)	**219828**	GPSA Step 4 pg. 10-10 using Ux
Required Extended Surface Area, A_x (m^2)	20422.6807	Conversion to m2 (1 ft2 = 0.092903 m2)
APSF	118.8	Fig. 10-11; 5/8 in by 10, APSF 2 1/4 in. triangular pitch with 4 rows
Face Area, F_a (ft^2)	1850.40404	GPSA Step 5, section 10 (p. 10-10)
Assumed Tube Length (ft)	119	GPSA Example provided Fa=467 with Tube Length = 30 ft, so scaled to Fa = 187 for reasonable assumption
Width (ft)	15.5	GPSA Step 6, Section 10 (p 10-10), rounded to 1 decimal
APF	5.58	GPSA from figure 10-11 for 5/8" by 10 fins
Number of Tubes, N_t	332	GPSA step 7, p. 10-10, rounded to nearest tube
Tube Area, A_t (in^2)	0.5945	GPSA Figure 9-25; Tube OD 1" (given in PW#6 problem statement), and assume BWG 16
Tube Area, A_t (ft^2)	0.00412847	Converted from in2 to ft2 using At/144
Tube Side Mass Velocity, G_t (lb/ft^2*s)	83.9042832	GPSA Step 8, p. 10-10, where Gt = (vapor mass flow rate)/(3600*Nt*At)
Tube ID, D_i (in)	0.87	Using characteristics of tubing BWG 16 from GPSA
Tube ID, D_i (ft)	0.0725	Using characteristics of tubing BWG 16 from GPSA
$\dot{m}_{air}$ (lb/hr), W_a - GPSA notation	5845632.04	Skipping to step 12 to estimate air quantity; Wa = Q/0.24*Δta1; Cpair = 0.24 Btu/lb*oF
Air Face Mass Velocity, G_a (lb/ft^2*h)	3159.11115	GPSA Step 13, p 10-10
Air-side film coefficient, h_a (Btu/h*ft2*°F)	9.5	GPSA Figure 10-17 read using Ga
AR	21.4	GPSA Figure 10-11 for 5/8" tubes by 10

GPSA Method (bare tube area)		**Notes**
U_b (Btu/hr ft^2 °F)	99.6812	Converted to bare tube area using Ux (extended surface)*AR ratio
Required Bare Surface Area, A_b (ft^2)	**10273**	GPSA Step 4 pg. 10-10 using Ub
Required Bare Surface Area, A_b (m^2)	954.392519	Conversion to m2 (1 ft2 = 0.092903 m2)

GPSA Method (fan sizing)		**Notes**
Number of Fans, N_f	1	Assumed that the process uses one fan
Minimum fan area, FAPF (ft^2)	740.161616	GPSA Step 16, p 10-14
Fan diameter (ft)	31	GPSA Step 17, p 10-14, round up the fan diameter to nearest foot
Pressure drop factor, Fp	0.15	GPSA Figure 10-18, using air face mass velocity
T_a, avg (°F)	118.265	Average air temperature
Number of rows	4	Assumed; also assumed for APSF
Air density ratio, Dr	0.89	GPSA Figure 10-16; Using average air temperature at 992.33 ft from Utility Specifications
Air static pressure drop, ΔP_a	0.6741573	GPSA Step 18, p 10-14
Actual air volume (ACFM)	1461532.24	GPSA Step 19, p 10-14; Using equation ACFM = Wa/(Dr*60*0749)
Fan total pressure, PF (in H_2O)	0.8822109	GPSA step 20, p 10-14
BHP, per fan	289.80034	GPSA step 21, p 10-14; Assuming 70% fan efficiency; conversion factor used (6356 = (33000 ft*lb/min*hp)*(12/ft)*(ft3/62.3 lb)
Efficiency of fan motor speed reducer	0.92	Given in step 21 for GPSA
Actual fan motor needed (bhp)	315.000369	Actual fan motor/efficiency of speed reducer
Motor size, rounded to discrete size, bhp	350	Using discrete motor sizes from Pump Notes
Motor size, kW	260.995	Converting to kw (1 hp = 0.7457 kw)

Capital Cost

$C_p°$	148740.461	Turton Equation A.1; Turton analysis uses bare surface area
K_1	4.0336	Turton Table A.1
K_2	0.2341	Turton Table A.1
K_3	0.0497	Turton Table A.1
B_1	0.96	Turton Table A.4
B_2	1.21	Turton Table A.4
F_M	1	Turton from from Figure A.18 (ID# 10); Assumed carbon steel
Pressure (psia)	151	Assumed from condenser pressure given
Pressure (barg)	9.37686991	conversion to barg ((200-15)*0.0689475729)
$F_{P,vessel}$	0.99376643	Turton using equation A.3
C_1	-0.125	Turton Table A.2
C_2	0.15361	Turton Table A.2
C_3	-0.02861	Turton Table A.2
C_{BM} (2001)	$321,644.91	Turton Equation A.4
C_{BM} (2016)	**$465,160.58**	Using extrapolated CEPCI value of 569.8 for 2016

Motor Capital Cost

$C_p°$	66416.393	Turton Equation A.1
K_1	2.4604	Turton Table A.1
K_2	1.4191	Turton Table A.1
K_3	-0.1798	Turton Table A.1
F_{BM}	1.5	Turton, Figure A.19 (ID# 13)
Shaft Power, kW	242.72535	Using Pump Notes: brake hp (15)*ηmotor (0.93)
C_{BM} (2001)	99624.5896	Turton Equation A.4

C_{BM} (2016)	**$144,076.37**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost

Purchased (hp)	376.344086	Using Pump Notes: brake hp (15)*ηmotor (0.93)
Purchased (kW)	280.639785	Converting purchased (hp) to kW using 1 hp = 0.7457 kW
Service Factor	0.96	Given in problem statement PW #5 last semester
$ per kWh	0.07	Specified in the Memorandum
Operating Cost Per Year	**$165,204.78**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.07/kWh * purchase kW
Total Capital Cost	**$609,236.95**	

Reactor Heat Exchanger (Shell & Tube HEX) E-5 Design

		Notes
Heater Duty (Btu/hr)	2.02E+07	From Aspen Simulation - Heater Duty before the reactor
$T_{H,in}$ (^oF)	550	Medium Pressure Steam from memorandum
$T_{H,out}$ (^oF)	550	Medium Pressure Steam from memorandum
$T_{C,in}$ (^oF)	175.129	From Aspen Simulation of the process fluid inlet
$T_{C,out}$ (^oF)	300	From Aspen Simulation of the process fluid outlet
Approach Temp (^oF)	20	Heuristic for steam reboiler approach temperature, from Gas Purification by Cole & Nielsen
LMTD	308.2313307	Using the LMTD equation given in class from Dr. Aichele
U (Btu/hr ft^2 ^oF)	140	Heuristic given in Dr. Aichele's class notes on estimating U; U chosen for steam reboilers; lowest for conservative purposes
F (correction factor)	0.9	Assumed using the heuristics in Table 11.11
Area (ft^2)	**520**	Rearraging the HEX design equation; Q = F*LMTD*U*A
Area (m^2)	48.30956	Converting from ft2 to m2 using the conversion factor 1 ft2 = 0.092903 m^2
Final Pressure (psia)	444.696	From Aspen Simulation for Heat Exchanger before the reactor
Pressure (barg)	29.62649628	Converting to barg ((Pressure (psia)-15)*0.0689475729)
ΔT_{steam}	0	Because steam is being condensed; phase changes occur at constant temperatures
$T_{avg, steam}$	550	Same average temperature due to phase change
λ_{steam} (Btu/lbm)	970.4	Latent heat of vaporization for steam (engineeringtoolbox.com)
$\dot{m}_{steam}$ (lb/hr)	20810.2629	Solving for the mass flow rate of steam using Q = mdot*λ

Capital Cost

$C_p°$	49838.02892	Turton Equation A.1
K_1	4.4646	Turton Table A.1
K_2	-0.5277	Turton Table A.1
K_3	0.3955	Turton Table A.1

B_1	1.63	Turton Table A.4
B_2	1.66	Turton Table A.4
F_M	1	Turton from from Figure A.18 (ID# 10); Assumed carbon steel for shell-and-tubes
Pressure (psia)	194.7	Boiler feed water pressure given in the Memorandum
Design Pressure (psia)	244.7	Add 50 psi for design pressure (greater than 1.1*operating pressure)
Pressure (barg)	15.8372575	conversion to barg ((200-15)*0.0689475729)
$F_{P,vessel}$	1.019860005	Turton using equation A.3 using the design pressure
C_1	-0.00164	Turton Table A.2
C_2	-0.00627	Turton Table A.2
C_3	0.0123	Turton Table A.2
C_{BM} (2001)	$165,610.16	Turton Equation A.4
C_{BM} (2016)	**$239,504.23**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost

Hours/yr	8409.6	365 days/yr * 24 hrs/day * 0.96 service factor
Service Factor	0.96	Given in problem statement PW #5
Mass of steam/yr	175005986.9	Multiplying hours/yr by mdot
$ per lb steam	$0.00330	Given in the memorandum
Operating Cost Per Year	**$577,519.76**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.06/kWh * purchase kW

Reboiler (Kettle) for De-Butanizer E-6 Design

		Notes
Reboiler Duty (Btu/hr)	5.24E+07	From Aspen Simulation PW #4
$T_{H,in}$ (°F)	550	Using Process Conditions medium pressure recommended value (Ch. 6) Class Notes
$T_{H,out}$ (°F)	550	This temperature is the same because the steam is being condensed; all phase changes occur at constant temperatures
$T_{C,in}$ (°F)	238.68	From Aspen Simulation; going into the reboiler Stage 30
$T_{C,out}$ (°F)	240.671	From Aspen Simulation; coming out of the reboiler Stage 31
Approach Temp (°F)	20	Heuristic for steam reboiler approach temperature, Gas Purification by Cole & Nielsen
LMTD	310.3234355	Using the LMTD equation given in class from Dr. Aichele
U (Btu/hr ft^2 °F)	140	Heuristic given in Dr. Aichele's class notes on estimating U; U chosen for steam reboilers; lowest for conservative purposes
F (correction factor)	0.9	Assumed using the heuristics in Table 11.11
Area (ft^2)	**1339**	Rearraging the HEX design equation; Q = F*LMTD*U*A
Area (m^2)	124.397117	Converting from ft2 to m2 using the conversion factor 1 ft2 = 0.092903 m^2
Column Pressure (psia)	151	From Aspen Simulation Last Stage Pressure
Computational Stages	31	From Aspen Simulation Input
Trays Inside the Column	29	
Overall Efficiency	0.85	Using Figure 7.32 (O'Connel Correlation) of CHE 4224 Course Notes by Dr. Whiteley/Wagner
Nactual	34	(Nstages/Eo)=Nactual
Final Reboiler Pressure (psia)	154.4	Column P + Nactual*0.1 (0.1 psi pressure drop per stage) from PW#4 last semester
Reboiler Pressure (barg)	9.611291662	Converting to barg ((Pressure (psia)-15)*0.0689475729)
ΔT_{steam}	0	Because steam is being condensed; phase changes occur at constant temperatures
$T_{avg, steam}$	550	Same average temperature due to phase change
λ_{steam} (Btu/lbm)	970.4	Latent heat of vaporization for steam (engineeringtoolbox.com)
$\dot{m}_{steam}$ (lb/hr)	53946.9291	Solving for the mass flow rate of steam using Q = mdot*λ

Capital Cost

$C_p°$	124369.4356	Turton Equation A.1
K_1	4.4646	Turton Table A.1
K_2	-0.5277	Turton Table A.1
K_3	0.3955	Turton Table A.1
B_1	1.63	Turton Table A.4
B_2	1.66	Turton Table A.4
F_M	1	Turton from from Figure A.18 (ID# 10); Assumed carbon steel for shell-and-tubes
Pressure (psia)	194.7	Boiler feed water pressure given in the Memorandum
Design Pressure (psia)	244.7	Add 50 psi for design pressure (greater than 1.1*operating pressure)
Pressure (barg)	15.8372575	conversion to barg ((200-15)*0.0689475729)
$F_{P,vessel}$	1.019860005	Turton using equation A.3 using the design pressure
C_1	-0.00164	Turton Table A.2
C_2	-0.00627	Turton Table A.2
C_3	0.0123	Turton Table A.2
C_{BM} (2001)	$413,275.61	Turton Equation A.4
C_{BM} (2016)	**$597,676.24**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost

Hours/yr	8409.6	365 days/yr * 24 hrs/day * 0.96 service factor
Service Factor	0.96	Given in problem statement PW #5
Mass of steam/yr	453672095	Multiplying hours/yr by mdot
$ per lb steam	$0.00330	Given in the Memorandum
Operating Cost Per Year	**$1,497,117.91**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.06/kWh * purchase kW

Shell-and-Tube Cooler (Floating Head) - Natural Gasoline Product E-7 Design

		Notes
Cooler Duty (Btu/hr)	2.62E+05	From Aspen Simulation - Cooler Duty
Molar flow rate (lbmol/hr)	41.33176	From Aspen Simulation - molar flow rate
Cp (Btu/lbmol F)	39.1191	From Aspen Simulation - under material properties
$\Delta T_{Bottoms}$ (°F)	142.673	Hot in - Hot out of the bottoms stream through a cooler
$T_{H,in}$ (°F)	240.673	From Aspen Simulation T2BOTTOM
$T_{H,out}$ (°F)	98	From Aspen Simulation NGASCOOL
$T_{C,in}$ (°F)	87	Specified in the Memorandum
$T_{C,out}$ (°F)	120	Specified in the Memorandum
LMTD	45.78886836	Using the LMTD equation given in class from Dr. Aichele
U (Btu/hr ft^2 °F)	80	Heuristic given in Dr. Aichele's class notes on estimating U; lowest for conservative purposes
F (correction factor)	0.9	Assumed using the heuristics in Table 11.11
Area (ft^2)	**80**	Rearraging the HEX design equation; Q = F*LMTD*U*A
Area (m^2)	7.43224	Converting from ft2 to m2 using the conversion factor 1 ft2 = 0.092903 m^2
Pressure (psia)	151	Aspen Simulation
Pressure (barg)	9.376869914	Converting to barg ((Pressure (psia)-15)*0.0689475729)
$\dot{m}_{water}$ (lbmol/hr)	442.3955281	Solving for the mass flow rate of water using Q/CpdT of water
$\dot{m}_{water}$ (lb/hr)	7963.119507	Converting lbmol/hr flow rate to lb/hr using 1 lbmol = 18 lbm of water
ρ_{water} (lb/ft3)	62.42	Density of water from Dr. Whiteley's Unit Conversions Packet
$\dot{m}_{water}$ (gpm)	15.90523152	Converting from lb/hr to gpm dividing by density and 1 gpm = 8.020833 ft3/hr
Cp (Btu/lbmol F)	17.9691	From ChemCad simulation for water at 90 F
ΔT_{water} (°F)	33	Cold out - cold in for water

Capital Cost (Shell-and-Tube - Floating

Head)

$C_p°$	21439.24985	Turton Equation A.1
K_1	4.8306	Turton Table A.1
K_2	-0.8509	Turton Table A.1
K_3	0.3187	Turton Table A.1
B_1	1.63	Turton Table A.4
B_2	1.66	Turton Table A.4
F_M	1	Turton from from Figure A.18 (ID# 1); Assumed carbon steel for shell-and-tubes
Pressure (psia)	151	Medium pressure steam given in Chapter 6 notes under recommended value
Design Pressure (psia)	201	Add 50 psi for design pressure (greater than 1.1*operating pressure)
Pressure (barg)	12.82424856	conversion to barg ((200-15)*0.0689475729)
$F_{P,vessel}$	1.015111569	Turton using equation A.3 using the design pressure
C_1	-0.00164	Turton Table A.2
C_2	-0.00627	Turton Table A.2
C_3	0.0123	Turton Table A.2
C_{BM} (2001)	$71,072.94	Turton Equation A.4
C_{BM} (2016)	**$102,785.18**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost

Converting to per min basis	504576	365 days/yr * 24 hrs/day *60 min* 0.96 service factor
Service Factor	0.96	Given in problem statement PW #5 last semester
Cooling water (gpm)	15.90523152	Imported from above
$/ gpm year cooling water	120	From Dr. Aichele & Dr. Ramsey Memorandum
$/ gpm	2.28E-04	$/gpm year / 365 * 60 * 24
Operating Cost Per Year	**$1,832.28**	gpm*$/gpm*365 days/yr * 24 hrs/day *60 min* 0.96 service factor

T-1 Feed Pump, P-1 Design

		Notes
$P_{suction}$ (psia)	114.7	From Memorandum
$P_{feed,DC}$ (psia)	160	Aspen Simulation
$P_{frictional\ losses}$ (psi)	15	5 psi for frictional losses from Dr. Ramsey/Dr. Aichele last semester, 10 psi for CV
Number of stages	32.5	Number of actual stages is theoretical stages (26/0.8) divided by 80% overall efficiency
Feed stage from ground	2	See Optimum feed location from optimization
Height of feed stage (ft)	19	26(2 ft) + 7 ft + 8 ft. 2 ft for tray spacing and 7 ft for reboiler and 8 ft for skirt .
SPG feed	0.57784	Aspen output of feed density
$P_{hydrostatic}$ (psi)	4.7527965	Using Pump Notes dP equation for converting height to psi
$P_{discharge}$ (psia)	179.7528	Sum of the feed DC Pressure, frictional losses, and hydrostatic head
	197.72808	1.1 * Pdischarge
	229.7528	50 psi + Pdischarge
Design P (psia)	229.7528	Largest of the 1.1*DiscP and 50+DiscP
Design P (barg)	14.82789	(Design P - 14.7)= (psig)*(1psi=0.06895bar)

Finding Efficiences

Q (ft^3/hr)	1894.99	Determined from Aspen simulation
Q (gpm)	196.7263	Converting from ft3/hr to gpm using 1 ft3/hr = 0.1038138999 gpm
ΔP (psi)	65.052797	Change in pressure is the discharge pressure - suction pressure
Hydraulic (hp)	7.4621552	Using Pump Notes dP equation for finding hydraulic hp
η_{pump}	0.6	Using Figure 14-37 from Pump Notes using Q (gpm)
Brake hp (bhp)	12.436925	Using Pump Notes equation BHP
η_{motor}	0.87	Using Figure 14-38 from Pump Notes using BHP
Purchase hp	17.241379	Using Pump Notes equation for P HP
Motor (bhp)	15	Using Pump Notes: Pump Drivers; chosen by using BHP

Capital Cost (Pump)	Centrifugal	
Shaft Power (kW)	9.7313833	Converting motor (bhp) to kW using motor (bhp)*0.745699872 conversion factor * ηmotor
$C_p°$	3911.5615	Using Equation A.1 from Turton (Shaft power limits 1 - 300 kW)
K_1	3.3892	Table A.1 for Centrifugal pump
K_2	0.0536	Table A.1 for Centrifugal pump
K_3	0.1538	Table A.1 for Centrifugal pump

B_1	1.89	Table A.5 for Centrifugal pump
B_2	1.35	Table A.5 for Centrifugal pump
F_M	1	Figure A.18 using ID # 38 for CS
P_{design} (barg)	14.82789	From above
$F_{P,vessel}$	1.1662635	Using Equation A.3 from Turton 10 < P < 100 barg
C_1	-0.3935	Table A.1 for Centrifugal Pump
C_2	0.3957	Table A.1 for Centrifugal Pump
C_3	-0.00226	Table A.1 for Centrifugal Pump
C_{BM} (2001)	$13,551.43	Cost of the pump is the Cp0*(B1+(B2*FP*FM))
C_{BM} (2016)	**$19,597.98**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost (Pump)	Centrifugal	
Purchased (hp)	17.241379	Using Pump Notes: brake hp (40)/ηmotor (0.89)
Purchased (kW)	12.856897	Converting purchased (hp) to kW using 1 hp = 0.7457 kW conversion factor
Service Factor	0.96	Given in problem statement PW #5 last semester
$ per kWh	0.07	Specified in Memorandum
Operating Cost Per Year	**$7,568.50**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.07/kWh * purchase kW

With Spare Pumps		
Capital Cost (Pump)	$39,195.97	Regular pump capital cost multiplied by 2
Total Capital Cost	**$39,195.97**	

T-1 Reflux Pump, P-2 Design

		Notes
$P_{suction}$ (psia)	150	From Aspen Simulation
$P_{feed,DC}$ (psia)	155	From Aspen Simulation
Exit flow rate (ft^3/s)	6.86	From Aspen Simulation
P drop (psi)	15	10 psi CV and 5 psi line losses
$P_{frictional\ losses}$ (psi)	16.5	10% upsized for reflux pumps Heuristic #10 Turton Table 11.13
Number of stages	32.5	Number of actual stages is theoretical stages (28) divided by 80% overall efficiency
Optimal Feed Stage	25	See Optimum feed location from optimization
Height of feed stage (ft)	65	2 ft for tray spacing and 7 ft for reboiler and 8 ft for skirt
SPG feed	0.563	Aspen output of density = 36.9646 lb/ft3/62.4 lb/ft3
$P_{hydrostatic}$ (psi)	15.841991	Using Pump Notes dP equation for converting height to psi
$P_{discharge}$ (psia)	187.34199	Sum of the feed DC Pressure, frictional losses, and hydrostatic head
	206.07619	1.1 * Pdischarge
	237.34199	50 psi + Pdischarge
Design P (psia)	237.34199	Largest of the 1.1*DiscP and 50+DiscP
Design P (barg)	15.351165	(Design P - 14.7)= (psig)*(1psi=0.06895bar)

Finding Efficiences

Q (ft^3/hr)	22734.2	Determined from Aspen simulation
Q (gpm)	2360.126	Converting from ft3/hr to gpm using 1 ft3/hr = 0.1038138999 gpm
ΔP (psi)	37.341991	Change in pressure is the discharge pressure - suction pressure
Hydraulic (hp)	51.388807	Using Pump Notes dP equation for finding hydraulic hp
η_{pump}	0.8	Using Figure 14-37 from Pump Notes using Q (gpm)
Brake hp (bhp)	64.236008	Using Pump Notes equation BHP
η_{motor}	0.9	Using Figure 14-38 from Pump Notes using BHP
Purchase hp	83.333333	Using Pump Notes equation for P HP
Motor (bhp)	75	Using Pump Notes: Pump Drivers; chosen by using BHP

Capital Cost (Pump)	Centrifugal	
Shaft Power (kW)	50.334741	Converting motor (bhp) to kW using motor (bhp)*0.745699872 conversion factor * ηmotor
$C_p°$	8431.0057	Using Equation A.1 from Turton (Shaft power limits 1 - 300 kW)
K_1	3.3892	Table A.1 for Centrifugal pump
K_2	0.0536	Table A.1 for Centrifugal pump

K_3	0.1538	Table A.1 for Centrifugal pump
B_1	1.89	Table A.5 for Centrifugal pump
B_2	1.35	Table A.5 for Centrifugal pump
F_M	1	Figure A.18 using ID # 38 for CS
P_{design} (barg)	15.351165	From above
$F_{P,vessel}$	1.1821606	Using Equation A.3 from Turton 10 < P < 100 barg
C_1	-0.3935	Table A.1 for Centrifugal Pump
C_2	0.3957	Table A.1 for Centrifugal Pump
C_3	-0.00226	Table A.1 for Centrifugal Pump
C_{BM} (2001)	$29,389.78	Cost of the pump is the Cp0*(B1+(B2*FP*FM))
C_{BM} (2016)	**$42,503.30**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost (Pump)	Centrifugal	
Purchased (hp)	83.333333	Using Pump Notes: brake hp (40)/ηmotor (0.89)
Purchased (kW)	62.141667	Converting purchased (hp) to kW using 1 hp = 0.7457 kW conversion factor
Service Factor	0.96	Given in problem statement PW #5 last semester
$ per kWh	0.07	Specified in Memorandum
Operating Cost Per Year	**$36,581.06**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.07/kWh * purchase kW

With Spare Pumps		
Capital Cost (Pump)	$85,006.59	Regular pump capital cost multiplied by 2
Total Capital Cost	**$85,006.59**	

T-2 Feed Pump, P-3

Design Notes

$P_{suction}$ (psia)	150	Aspen Simulation
$P_{feed,DC}$ (psia)	175	Aspen Simulation
$P_{frictional\ losses}$ (psi)	15	5 psi for frictional losses from Dr. Ramsey/Dr. Aichele last semester
Number of stages	36.470588	Number of actual stages is theoretical stages (28) divded by 80% overall efficiency PW # 4
Feed stage from ground	27	See Optimum feed location from optimization
Height of feed stage (ft)	69	2 ft for tray spacing and 7 ft for reboiler and 8 ft for skirt .
SPG feed	0.580286	Aspen output of feed density
$P_{hydrostatic}$ (psi)	17.333218	Using Pump Notes dP equation for converting height to psi
$P_{discharge}$ (psia)	207.33322	Sum of the feed DC Pressure, frictional losses, and hydrostatic head
	228.06654	1.1 * Pdischarge
	257.33322	50 psi + Pdischarge
Design P (psia)	257.33322	Largest of the 1.1*DiscP and 50+DiscP
Design P (barg)	16.72956	(Design P - 14.7)= (psig)*(1psi=0.06895bar)

Finding Efficiences

Q (ft^3/hr)	4060	Determined from Aspen simulation
Q (gpm)	421.48443	Converting from ft3/hr to gpm using 1 ft3/hr = 0.1038138999 gpm
ΔP (psi)	57.333218	Change in pressure is the discharge pressure - suction pressure
Hydraulic (hp)	14.090413	Using Pump Notes dP equation for finding hydraulic hp
η_{pump}	0.7	Using Figure 14-37 from Pump Notes using Q (gpm)
Brake hp (bhp)	20.129162	Using Pump Notes equation BHP
η_{motor}	0.87	Using Figure 14-38 from Pump Notes using BHP
Purchase hp	28.735632	Using Pump Notes equation for P HP
Motor (bhp)	25	Using Pump Notes: Pump Drivers; chosen by using BHP

Capital Cost (Pump) Centrifugal

Shaft Power (kW)	16.218972	Converting motor (bhp) to kW using motor (bhp)*0.745699872 conversion factor * ηmotor
$C_p°$	4777.982	Using Equation A.1 from Turton (Shaft power limits 1 - 300 kW)
K_1	3.3892	Table A.1 for Centrifugal pump
K_2	0.0536	Table A.1 for Centrifugal pump
K_3	0.1538	Table A.1 for Centrifugal pump

B_1	1.89	Table A.5 for Centrifugal pump
B_2	1.35	Table A.5 for Centrifugal pump
F_M	1	Figure A.18 using ID # 38 for CS
P_{design} (barg)	16.72956	From above
$F_{P,vessel}$	1.2225026	Using Equation A.3 from Turton 10 < P < 100 barg
C_1	-0.3935	Table A.1 for Centrifugal Pump
C_2	0.3957	Table A.1 for Centrifugal Pump
C_3	-0.00226	Table A.1 for Centrifugal Pump
C_{BM} (2001)	$16,915.87	Cost of the pump is the Cp0*(B1+(B2*FP*FM))
C_{BM} (2016)	**$24,463.60**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost (Pump)	Centrifugal	
Purchased (hp)	28.735632	Using Pump Notes: brake hp (40)/ηmotor (0.89)
Purchased (kW)	21.428161	Converting purchased (hp) to kW using 1 hp = 0.7457 kW conversion factor
Service Factor	0.96	Given in problem statement PW #5
$ per kWh	0.07	Specified in Memorandum
Operating Cost Per Year	**$12,614.16**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.07/kWh * purchase kW

With Spare Pumps		
Capital Cost (Pump)	$48,927.21	Regular pump capital cost multiplied by 2
Total Capital Cost	**$48,927.21**	

T-2 Reflux Pump, P-4

Design

		Notes
$P_{suction}$ (psia)	151	From Aspen Simulation
$P_{feed,DC}$ (psia)	155	From Aspen Simulation
P drop (psi)	25	10 psi x 2 per CV, 5 psi line losses
$P_{frictional\ losses}$ (psi)	27.5	10% upsized for reflux pumps Heuristic #10 Turton Table 11.13
Number of stages	36.470588	Number of actual stages is theoretical stages (28) divded by 80% overall efficiency PW # 4
Optimal Feed Stage	30	From Aspen Simulation
Height of feed stage (ft)	75	2 ft for tray spacing and 7 ft for reboiler and 8 ft for skirt .
SPG feed	0.579297	Aspen output of density = 36.9646 lb/ft3/62.4 lb/ft3
$P_{hydrostatic}$ (psi)	18.808344	Using Pump Notes dP equation for converting height to psi
$P_{discharge}$ (psia)	201.30834	Sum of the feed DC Pressure, frictional losses, and hydrostatic head
	221.43918	1.1 * Pdischarge
	251.30834	50 psi + Pdischarge
Design P (psia)	251.30834	Largest of the 1.1*DiscP and 50+DiscP
Design P (barg)	16.314145	(Design P - 14.7)= (psig)*(1psi=0.06895bar)

Finding Efficiences

Q (ft^3/hr)	5017.59	Determined from Aspen simulation
Q (gpm)	520.89559	Converting from ft3/hr to gpm using 1 ft3/hr = 0.1038138999 gpm
ΔP (psi)	50.308344	Change in pressure is the discharge pressure - suction pressure
Hydraulic (hp)	15.280113	Using Pump Notes dP equation for finding hydraulic hp
η_{pump}	0.7	Using Figure 14-37 from Pump Notes using Q (gpm)
Brake hp (bhp)	21.828733	Using Pump Notes equation BHP
η_{motor}	0.875	Using Figure 14-38 from Pump Notes using BHP
Purchase hp	28.571429	Using Pump Notes equation for P HP
Motor (bhp)	25	Using Pump Notes: Pump Drivers; chosen by using BHP

Capital Cost (Pump)	Centrifugal	
Shaft Power (kW)	16.312185	Converting motor (bhp) to kW using motor (bhp)*0.745699872 conversion factor * ηmotor
C_p°	4789.6657	Using Equation A.1 from Turton (Shaft power limits 1 - 300 kW)
K_1	3.3892	Table A.1 for Centrifugal pump
K_2	0.0536	Table A.1 for Centrifugal pump
K_3	0.1538	Table A.1 for Centrifugal pump

B_1	1.89	Table A.5 for Centrifugal pump
B_2	1.35	Table A.5 for Centrifugal pump
F_M	1	Figure A.18 using ID # 38 for CS
P_{design} (barg)	16.314145	From above
$F_{P,vessel}$	1.2105669	Using Equation A.3 from Turton 10 < P < 100 barg
C_1	-0.3935	Table A.1 for Centrifugal Pump
C_2	0.3957	Table A.1 for Centrifugal Pump
C_3	-0.00226	Table A.1 for Centrifugal Pump
C_{BM} (2001)	$16,880.05	Cost of the pump is the Cp0*(B1+(B2*FP*FM))
C_{BM} (2016)	**$24,411.81**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost (Pump)	Centrifugal	
Purchased (hp)	28.571429	Using Pump Notes: brake hp (40)/ηmotor (0.89)
Purchased (kW)	21.305714	Converting purchased (hp) to kW using 1 hp = 0.7457 kW conversion factor
Service Factor	0.96	Given in problem statement PW #5 last semester
$ per kWh	0.07	Specified in Memorandum
Operating Cost Per Year	**$12,542.08**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.07/kWh * purchase kW

With Spare Pumps		
Capital Cost (Pump)	$48,823.63	Regular pump capital cost multiplied by 2
Total Capital Cost	**$48,823.63**	

Reactor Feed Pump, P-5 Design

		Notes
$P_{suction}$ (psia)	196.331682	From Aspen Simulation
$P_{feed,DC}$ (psia)	464.7	From Aspen Simulation
P drop (psi)	10	To upsize the pump
$P_{frictional\ losses}$ (psi)	15	5 psi for frictional losses + the Pdrop
Number of stages	30.5882353	Number of actual stages is theoretical stages (28) divded by 80% overall efficiency PW # 4
Optimal Feed Stage	1	From Aspen Simulation
Height of feed stage (ft)	17	2 ft for tray spacing and 7 ft for reboiler and 8 ft for skirt .
SPG feed	0.579297	Aspen output of density = 36.9646 lb/ft3/62.4 lb/ft3
$P_{hydrostatic}$ (psi)	4.26322468	Using Pump Notes dP equation for converting height to psi
$P_{discharge}$ (psia)	483.963225	Sum of the feed DC Pressure, frictional losses and hydrostatic head
	532.359547	1.1 * Pdischarge
	533.963225	50 psi + Pdischarge
Design P (psia)	533.963225	Largest of the 1.1*DiscP and 50+DiscP
Design P (barg)	35.8031993	(Design P - 14.7)= (psig)*(1psi=0.06895bar)

Finding Efficiences

Q (ft^3/hr)	3973.67	Determined from Aspen simulation
Q (gpm)	412.52218	Converting from ft3/hr to gpm using 1 ft3/hr = 0.1038138999 gpm
ΔP (psi)	287.631543	Change in pressure is the discharge pressure - suction pressure
Hydraulic (hp)	69.1862338	Using Pump Notes dP equation for finding hydraulic hp
η_{pump}	0.7	Using Figure 14-37 from Pump Notes using Q (gpm)
Brake hp (bhp)	98.8374769	Using Pump Notes equation BHP
η_{motor}	0.92	Using Figure 14-38 from Pump Notes using BHP
Purchase hp	135.869565	Using Pump Notes equation for P HP
Motor (bhp)	125	Using Pump Notes: Pump Drivers; chosen by using BHP

Capital Cost (Pump) Centrifugal

Shaft Power (kW)	85.7554853	Converting motor (bhp) to kW using motor (bhp)*0.745699872 conversion factor * ηmotor
$C_p°$	11685.5884	Using Equation A.1 from Turton (Shaft power limits 1 - 300 kW)
K_1	3.3892	Table A.1 for Centrifugal pump
K_2	0.0536	Table A.1 for Centrifugal pump
K_3	0.1538	Table A.1 for Centrifugal pump

B_1	1.89	Table A.5 for Centrifugal pump
B_2	1.35	Table A.5 for Centrifugal pump
F_M	1	Figure A.18 using ID # 38 for CS
P_{design} (barg)	35.8031993	From above
$F_{P,vessel}$	1.6441056	Using Equation A.3 from Turton 10 < P < 100 barg
C_1	-0.3935	Table A.1 for Centrifugal Pump
C_2	0.3957	Table A.1 for Centrifugal Pump
C_3	-0.00226	Table A.1 for Centrifugal Pump
C_{BM} (2001)	$48,022.42	Cost of the pump is the Cp0*(B1+(B2*FP*FM))
C_{BM} (2016)	**$69,449.69**	Using extrapolated CEPCI value of 569.8 for 2016

Operating Cost (Pump)	Centrifugal	
Purchased (hp)	135.869565	Using Pump Notes: brake hp (40)/ηmotor (0.89)
Purchased (kW)	101.317935	Converting purchased (hp) to kW using 1 hp = 0.7457 kW conversion factor
Service Factor	0.96	Given in problem statement PW #5 last semester
$ per kWh	0.07	Specified in Memorandum
Operating Cost Per Year	**$59,643.03**	Which is the 365 days/yr * 24 hrs/day * 0.96 * 0.07/kWh * purchase kW

With Spare Pumps		
Capital Cost (Pump)	$138,899.37	Regular pump capital cost multiplied by 2
Total Capital Cost	**$138,899.37**	

Service Factor Calculations

Service Factor (%) = (Operating hrs/yr)/(365 x 24 hr/yr) x 100%

Operating hours/year	8409.6
Service Factor	0.96

Additional Tables

Operating Costs/year		
Butamer Reactor	R-1	$0
De-Isobutanizer Tower	T-1	$0
De-Butanizer Tower	T-2	$0
Condensate Receiver 1	D-1	$0
Condensate Receiver 2	D-2	$0
De-Isobutanizer Reboiler	E-2	($2,094,574)
De-Butanizer Reboiler	E-6	($1,497,118)
Butamer Reactor Heater	E-5	($577,520)
Isobutane Cooler	E-3	($7,319)
De-Isobutanizer Feed Pump	P-1	($7,568)
De-Isobutanizer Reflux Pump	P-2	($36,581)
De-Butanizer Feed Pump	P-3	($12,614)
De-Butanizer Reflux Pump	P-4	($12,542)
Butamer Reactor Feed Pump	P-5	($59,643)
De-Isobutanizer Air-Cooler	E-1	($377,611)
De-Butanizer Air-Cooler	E-4	($165,205)
Natural Gas Cooler	E-7	($1,832)
Total		**($4,848,295)**

Table 40: *Operating Costs Summary*

Component	Purchase/Sale Location	Price ($/BBL)	Purchased/ Produced (BBL/day)	Produced (BBL/yr)	Total ($)
Mixed Butanes	Conway	$27.50	7800	2847000	($78,292,500)
Normal Butane	Mt. Belleview	$31.78	0	0	$0
Isobutane	Mt. Belleview	$37.67	7642.428282	2789486.32	$105,079,950
Natural Gas	Oklahoma	$55.85	324.5349607	118455.261	$6,615,726
Transportation (mixed butanes)	Conway to OK	$1.10	7800	2847000	($3,131,700)
Transportation (isobutane)	OK to Mt.Belleview	$1.35	7642.428282	2789486.32	($3,765,807)

Isobutane product produced (BPD)	7642.428282	Raw Material Costs	($85,190,007)
Natural Gas product produced (BPD)	324.5349607	Sales	$111,695,676

Table 41: *Raw Material and Sales Costs Summary*

Type of Cost	Factors	$/Year
Fixed Capital Cost, C_{TM}	1.18*Fixed Capital	($22,869,341)
Raw Materials		($85,190,007)
Utilities		($4,848,295)
Operating Labor	C_{OL}	($606,760)
Direct Supervisory and Clerical Labor	$0.18*C_{OL}$	($109,217)
Maintenance and Repairs	$0.06*C_{TM}$	($1,372,160)
Operating Supplies	$0.009*C_{TM}$	($205,824)
Laboratory charges	$0.15*C_{OL}$	($91,014)
Total Fixed Manufacturing Costs	$0.708*C_{OL}$ + $0.068*C_{TM}$	($1,984,701.25)
Total General Manufacturing Costs (TGMC)	$0.177*C_{OL}$ + $0.009*C_{TM}$	($313,220.59)
Total Operating Expenses (Sum Utilities to TGMC)		($9,531,192)

Table 42: *Manufacturing Costs Sample Calculations*

Year	2016	2017	2018	2019	2020	2021
End of Year	0	1	2	3	4	5
Production Natural Gas, "BBL/yr"	0	59227.63033	118455.2607	118455.2607	118455.2607	118455.2607
Production Iso-butane, "BBL/yr"	0	1,394,743	2,789,486	2,789,486	2,789,486	2,789,486
Natural Gas Sales Price, $/"BBL"	0	$55.85	$55.85	$55.85	$55.85	$55.85
Iso-butane Sales Price, $/"BBL"	0	$37.67	$37.67	$37.67	$37.67	$37.67
Sales Revenue	0	$55,847,838	$111,695,676	$111,695,676	$111,695,676	$111,695,676
Mixed Butanes, "BBL/yr"	0	1,423,500	2,847,000	2,847,000	2,847,000	2,847,000
Mixed Butanes Transportation, "$/BBL"	0	($1.10)	($1.10)	($1.10)	($1.10)	($1.10)
Isobutane Transportation, "$/BBL"	0	($1.35)	($1.35)	($1.35)	($1.35)	($1.35)
Mixed Butanes Purchased, "$/BBL"	0	($27.50)	($27.50)	($27.50)	($27.50)	($27.50)
— Mixed Butanes Purchased, "$/BBL"	0	($40,712,100)	($81,424,200)	($81,424,200)	($81,424,200)	($81,424,200)
— Isobutane Transportation, "$/BBL"	0	($1,882,903)	($3,765,807)	($3,765,807)	($3,765,807)	($3,765,807)
Net Revenue	0	$13,252,835	$26,505,670	$26,505,670	$26,505,670	$26,505,670
— Total Operating Costs	0	(4,765,596)	(9,531,192)	(9,531,192)	(9,531,192)	(9,531,192)
— Depreciation		($2,286,934)	($4,116,481)	($3,293,185)	($2,634,548)	($2,108,553)
— Write off						($8,429,639)
Taxable Income	0	$6,200,305	$12,857,996	$13,681,292	$14,339,929	$6,436,285
— Tax @ 40%	0	($2,480,122)	($5,143,198)	($5,472,517)	($5,735,972)	($2,574,514)
Net Income	0	$3,720,183	$7,714,798	$8,208,775	$8,603,958	$3,861,771
+ Depreciation		$2,286,934	$4,116,481	$3,293,185	$2,634,548	$2,108,553
+ Write off						$8,429,639
— Catalyst		($585,650)				
— Fixed Capital	($22,869,341)					
Cash Flow	($22,869,341)	$5,421,467	$11,831,279	$11,501,960	$11,238,506	$14,399,963
Discount Factor ($P/F_{i^*,n}$)	1.0000	0.8696	0.7561	0.6575	0.5718	0.4972
Discounted Cash Flow	($22,869,341)	$4,714,319	$8,946,147	$7,562,726	$6,425,652	$7,159,327
Discount Factor ($F/P_{i^*,n}$)	3.059022863	2.66001988	2.313060766	2.011357188	1.74900625	1.520875
NPV @ i* =	**$11,938,829**	The project is economically attractive since NPV > 0				
DCFROR =	**33%**	The project is economically attractive since DCFROR > i*				
PWNR @ i*	**$34,808,170**					
PWNC @ i*	**($22,869,341)**					
BCR =	**1.52204519**	The project is economically attractive since BCR > 1				
FWR	**$106,478,988**					
GROR =	**36.02%**	-0.000217958	This project is economically attractive since GROR > i*			

Table 43: *Sample Economics with Manufacturing Costs*

Year	2016	2017	2018	2019	2020	2021	2022
End of Year	0	1	2	3	4	5	6
Production Natural Gas, "BBL/yr"	0	59227.63033	118455.2607	118455.2607	118455.2607	118455.2607	
Production Iso-butane, "BBL/yr"	0	1,394,743	2,789,486	2,789,486	2,789,486	2,789,486	
Natural Gas Sales Price, $/"BBL"	0	$55.85	$55.85	$55.85	$55.85	$55.85	
Iso-butane Sales Price, $/"BBL"	0	$37.67	$37.67	$37.67	$37.67	$37.67	
Sales Revenue	0	$55,847,838.05	$111,695,676	$111,695,676	$111,695,676	$111,695,676	
— Raw Material, "$/BBL"	0	($42,595,003.27)	($85,190,007)	($85,190,007)	($85,190,007)	($85,190,007)	
Net Revenue	0	$13,252,835	$26,505,670	$26,505,670	$26,505,670	$26,505,670	
— Total Operating Costs	0	($2,727,527.56)	($5,455,055)	($5,455,055)	($5,455,055)	($5,455,055)	
— Depreciation		($4,573,868)	($7,318,189.03)	($4,390,913)	($2,634,548)	($2,634,548)	($1,317,274)
— Write off						($1,317,274)	
Taxable Income	0	$5,951,439	$13,732,425	$16,659,701	$18,416,066	$17,098,792	
— Tax @ 40%	0	($2,380,576)	($5,492,970)	($6,663,880)	($7,366,427)	($6,839,517)	
Net Income	0	$3,570,863	$8,239,455	$9,995,821	$11,049,640	$10,259,275	
+ Depreciation		$4,573,868	$7,318,189	$4,390,913	$2,634,548	$2,634,548	
+ Write off						$1,317,274	
— Catalyst		($585,650)					
— Fixed Capital	($22,869,341)						
Cash Flow	($22,869,341)	$7,559,082	$15,557,644	$14,386,734	$13,684,188	$14,211,097	
Discount Factor ($P/F_{i*,n}$)	1.0000	0.8696	0.7561	0.6575	0.5718	0.4972	0.4323
Discounted Cash Flow	($22,869,341)	$6,573,114	$11,763,814	$9,459,511	$7,823,979	$7,065,427	0
Discount Factor ($F/P_{i*,n}$)	3.059022863	2.66001988	2.313060766	2.011357188	1.74900625	1.520875	1.3225
NPV @ i* =	**$19,816,505**	The project is economically attractive since NPV > 0					
DCFROR =	**44%**	The project is economically attractive since DCFROR > i*					
PWNR @ i*	**$42,685,846**						
PWNC @ i*	**($22,869,341)**						
BCR =	**1.866509674**	The project is economically attractive since BCR > 1					
FWR	**$130,576,978**						
GROR =	**74.48%**	-10.46088792	This project is economically attractive since GROR > i*				

Table 44: *Sample Cash Flow Table with 5-year MACRS Depreciation*

Year	2016	2017	2018	2019	2020	2021	2022	2023	2024
End of Year	0	1	2	3	4	5	6	7	8
Production Natural Gas, "BBL/yr"	0	59227.63033	118455.2607	118455.2607	118455.2607	118455.2607			
Production Iso-butane, "BBL/yr"	0	1,394,743	2,789,486	2,789,486	2,789,486	2,789,486			
Natural Gas Sales Price, $/"BBL"	0	$55.85	$55.85	$55.85	$55.85	$55.85			
Iso-butane Sales Price, $/"BBL"	0	$37.67	$37.67	$37.67	$37.67	$37.67			
Sales Revenue	0	$55,847,838.05	$111,695,676	$111,695,676	$111,695,676	$111,695,676			
— Raw Material, "$/BBL"	0	($42,595,003.27)	($85,190,007)	($85,190,007)	($85,190,007)	($85,190,007)			
Net Revenue	0	$13,252,835	$26,505,670	$26,505,670	$26,505,670	$26,505,670			
— Total Operating Costs	0	($2,727,527.56)	($5,455,055)	($5,455,055)	($5,455,055)	($5,455,055)			
— Depreciation		($3,268,029)	($5,600,701.54)	($3,999,848)	($2,856,381)	($2,042,232)	($2,039,945)	($2,042,232)	($1,019,973)
— Write off						($5,102,150)			
Taxable Income	0	$7,257,278	$15,449,913	$17,050,767	$18,194,234	$13,906,232			
— Tax @ 40%	0	($2,902,911)	($6,179,965)	($6,820,307)	($7,277,694)	($5,562,493)			
Net Income	0	$4,354,367	$9,269,948	$10,230,460	$10,916,540	$8,343,739			
+ Depreciation		$3,268,029	$5,600,702	$3,999,848	$2,856,381	$2,042,232			
+ Write off						$5,102,150			
— Catalyst		($585,650)							
— Fixed Capital	($22,869,341)								
Cash Flow	($22,869,341)	$7,036,746	$14,870,649	$14,230,308	$13,772,921	$15,488,121			
Discount Factor ($P/F_{i^*,n}$)	1.0000	0.8696	0.7561	0.6575	0.5718	0.4972	0.4323		
Discounted Cash Flow	($22,869,341)	$6,118,909	$11,244,347	$9,356,658	$7,874,712	$7,700,334	0		
Discount Factor ($F/P_{i^*,n}$)	3.059022863	2.66001988	2.313060766	2.011357188	1.74900625	1.520875	1.3225		
NPV @ i* =	**$19,425,620**	The project is economically attractive since NPV > 0							
DCFROR =	**43%**	The project is economically attractive since DCFROR > i*							
PWNR @ i*	**$42,294,961**								
PWNC @ i*	**($22,869,341)**								
BCR =	**1.849417589**	The project is economically attractive since BCR > 1							
FWR	**$129,381,253**								
GROR =	**74.48%**	-10.513173	This project is economically attractive since GROR > i*						

Table 45: *Sample Cash Flow Table with 7-year MACRS Depreciation*

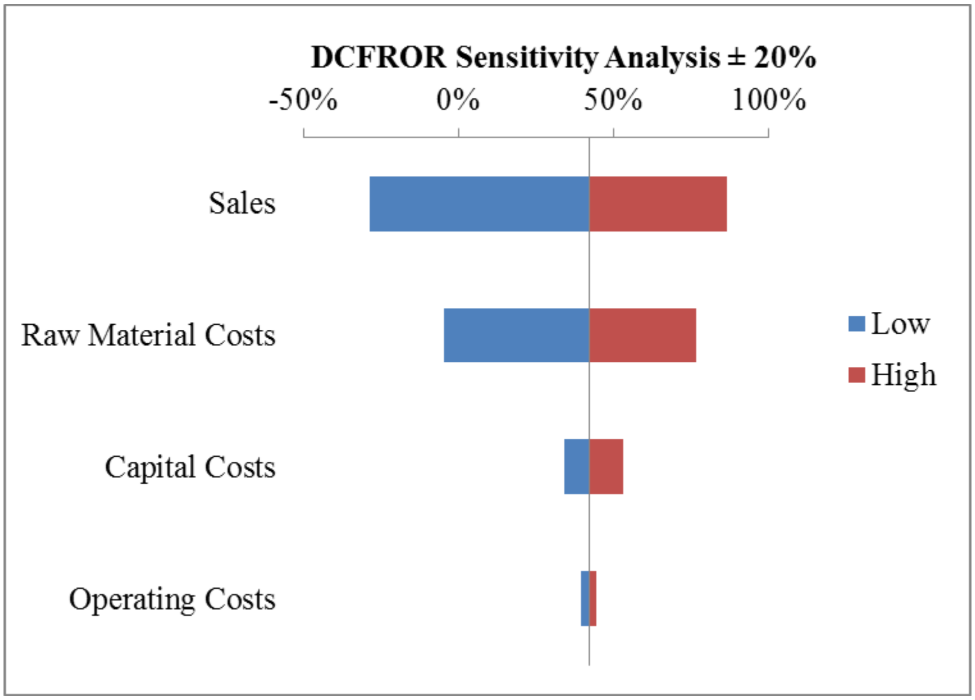


Figure 7: *DCFROR Sensitivity Analysis ± 20%*

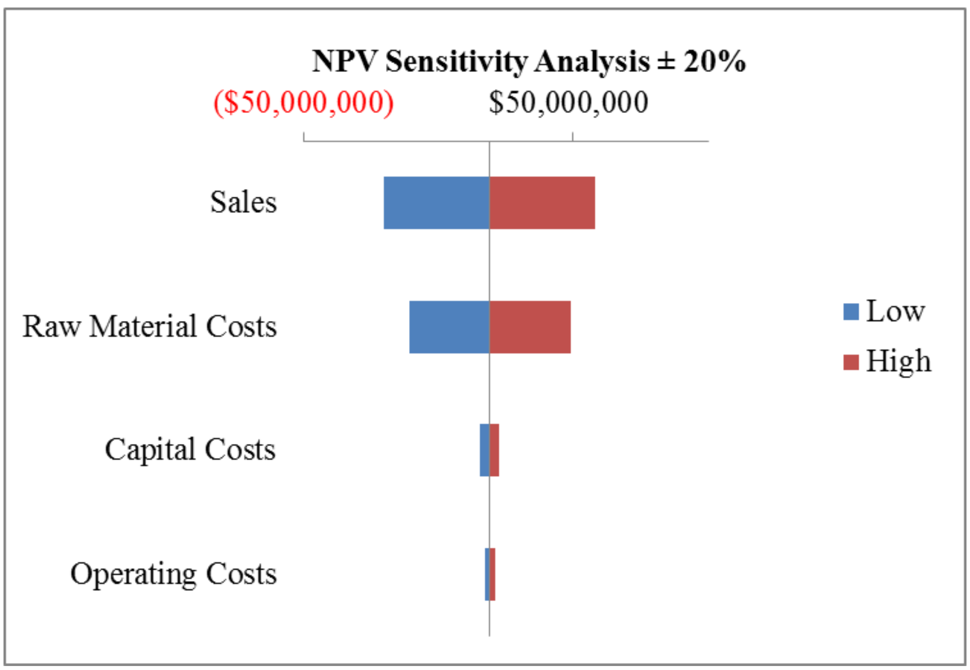


Figure 8: *NPV Sensitivity Analysis ± 20%*