Letter of Transmittal

16 April, 2021
Mr. Abbasi (COO)
Kirkuk Iraq

Dear Mr. Abbasi,

May this letter find you well. Within the attached report, you will find a preliminary design for a fixed bed continuous catalytic reformer as prepared by the student engineering team. The technical and economic feasibility of this project is evaluated within the report.

The analysis includes a simulation of a system capable of processing naphtha from the local crudes to separate salable benzene without losses in fuels revenue. This simulation is capable of replacing the current small toppings refinery and upgrading heavier naphtha into gasoline while providing benzene, toluene, and xylenes (BTX) as byproducts. This process is technically feasible and economically attractive. The implementation of this design will help to advance the goals of the company by increasing overall profitability. In addition to the request that this facility process feed K, this facility will also be able to process feed TQ1 while meeting all specifications. This facility has been designed to operate seamlessly with either feed.

Safety was a top priority while the preliminary design process for this facility was being completed. The inherently safer design concepts of minimization, substitution, moderation, and simplification were implemented to minimize the inherent risk of day to day operations. Process Hazard Analysis was completed, and engineering and management strategies were included to decrease risk. Several recommendations for the detailed design phase have also been included.

Thank you for the opportunity to complete this preliminary design. Your partnership is appreciated, and we look forward to working with you in the future. Please review the official report and respond with your thoughts. If you have any questions after reviewing the report, the engineering team is happy to be of assistance, and you can reach out to them at your convenience.

Thank you again for your consideration and partnership.

Respectfully,
The EPC Firm

Spring 2021

AIChE Student Design Competition

Fixed Bed Continuous Catalytic Reformer

April 16, 2021

Executive Summary

Mr. Abbasi is seeking a study estimate for a fixed bed continuous catalytic reformer for his small toppings refinery in Kurdistan to avoid government shutdown. Products from the facility can no longer harbor hazardous benzene from the feedstocks. The goal of the reformer design is to be able to upgrade heavier naphtha into gasoline while providing benzene, toluene, and xylenes as lucrative byproducts. A request for investigation of project feasibility and economic profitability was submitted to this engineering construction and procurement firm (EPC).

This design enables either feed K or feed TQ1 to be processed through the same facility while still meeting product and waste requirements. The diesel and gasoline products contain 0.3% benzene, and BTX is isolated to 99% purity. In addition, fuel gas and excess hydrogen are also produced as products. For feed K, approximately 48,970 kg/hr of BTX is produced. This is reduced to 31,200 kg/hr for feed TQ1. There were multiple innovations made to the given design prompt in order to maximize production, increase safety standards, and minimize capital costs. Several concepts of inherently safer design were incorporated into the final project design, including minimization, substitution, moderation, and simplification. Utilization of these concepts enhances the overall safety of the facility for both the equipment and the personnel. It also improves the cost effectiveness of the project. For instance, an example of the inherently safer design concept of minimization follows. The dangers associated with the storage and transportation of flammable and hazardous materials are minimized through the integration of recycle systems. Moreover, all waste products are disposed of in safe concentrations with proper methods.

Four scenarios needed to be analyzed with regards to economic feasibility: feed K under Iraqi tax control, feed K under Kurdish tax control, feed TQ1 under Iraqi tax control, and feed TQ1 under Kurdish tax control. The Net Present Values (NPV) for the four scenarios are: $5,472,285,166; $7,362,227,493; $3,668,727,833; and $5,003,729,441 respectively. The project solution has a projected capital cost of $24,461,577. The projected operating costs for feeds K and TQ1 are $370,201,389 and $400,202,199 respectively. The DCFRORs for each scenario are 67.33%, 74.11%, 51.14%, and 56.44% with discounted project payback periods of 1.73 years, 1.56 years, 2.23 years, and 2.03 years, respectively. The economic factors that had the greatest impact on the economic analysis were production rates and operating costs. Tornado charts detailing the impacts of these variables on the economics of this project have been included in the report for convenience.

After extensive economic evaluation in conjunction with safety and environmental evaluations, this design for a fixed bed continuous catalytic reformer for Mr. Abbasi's small toppings refinery is feasible and economically attractive. Implementing the project is recommended.

Table of Contents

Table of Figures

Table of Tables

Introduction

Iraqi Kurdistan has historically imported expensive refined fuels and exported indigenous oil at low prices, costing the economy $3B annually. To combat this, the Iraqi government is now closing numerous wildcat “teapot” facilities and toppings refineries that have proliferated to supply this hydrocarbon shortfall [1].

The objective of this project is to redesign a small toppings refinery in Iraqi Kurdistan to remove hazardous benzene from the feedstocks and to meet western refining standards. The client has requested a full study estimate for two possible feeds under two possible tax regimes for a fixed bed continuous catalytic reformer that will upgrade naphtha into gasoline while providing benzene, toluene, and xylenes as byproducts to be sold [1]. Based on these designs, the technical and economic feasibility of the reformed refinery is to be determined. The completion of these objectives will help to advance the goals of our client’s company by ensuring their survival under new government regulations, avoiding losses in fuel revenue, and increasing overall profits by isolating lucrative byproducts. The new refinery must be able to process 35,000 BPD of light sweet kirkuk crudes that are rich in benzene and low in sulfur. A block flow diagram of the existing refinery, Figure 1, was provided as the basis to start the design [1].

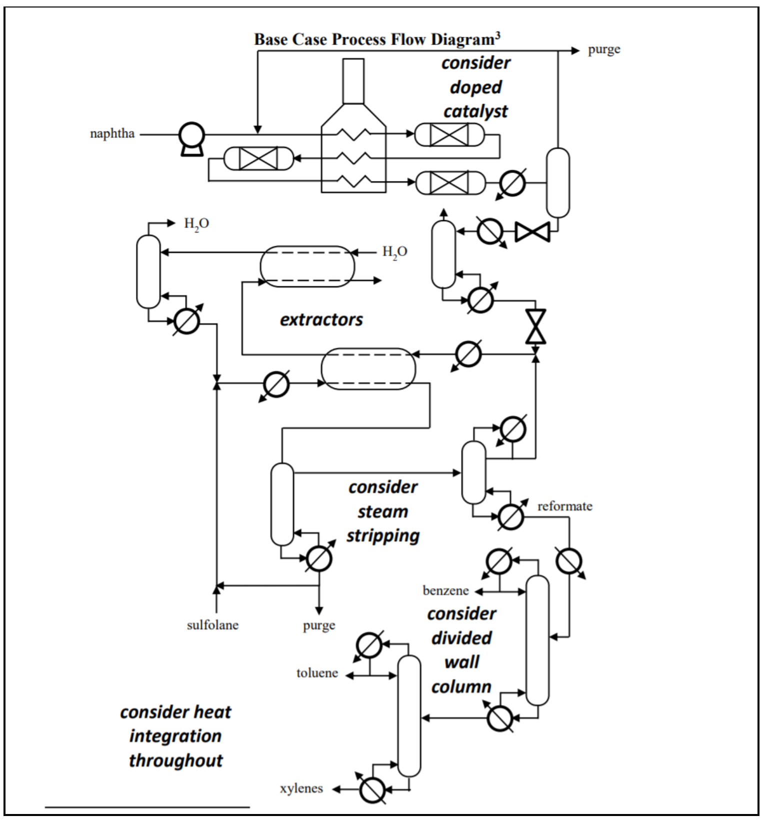


Figure 1: Base-Case Process Flow Sheet [1]

Process Description and Design Basis

This process design was modeled using Aspen HYSYS as a simulation software. Vapors were assumed to be ideal while the NRTL VLE and LLE thermodynamic packages were used for liquid modeling in distillation columns and extractors, as recommended in the AIChE technical objectives [1]. The plant is designed to generate gasoline with benzene, toluene, and para-xylene (BTX) as byproducts from a feed stream composed of hazardous cyclic and polyaromatic hydrocarbons. The naphtha compositions from expected refinery crude feeds K and TQ1 are shown in Table 1. Both feeds are analyzed in this report, but feed K is of greatest interest.

Table 1: Broad Naphtha Composition in Feed Streams [1]			
		Crude Oil	
		TQ1	*K*
	naphtha % volume of crude	28	20
	specific gravity	0.7308	0.749
naphtha proxy components	*n-decane mol %*	77.8	59.7
	cyclohexane mol %	20.6	31.3
	Benzene mol %	1.6	9

Three operations isolate valuable aromatics to produce cleaner fuels. Packed bed reactors convert naphtha cyclohexane into benzene. High reactor temperatures crack feed alkanes like decane into lighter species. A preliminary flow diagram of how these processes work together, Figure 2, was provided as a foundation to begin the process design. In addition to the BTX, linear alkanes, and cyclic hydrocarbons, this process also produces excess hydrogen and fuel gas that can be used as utilities and feed for the plant.

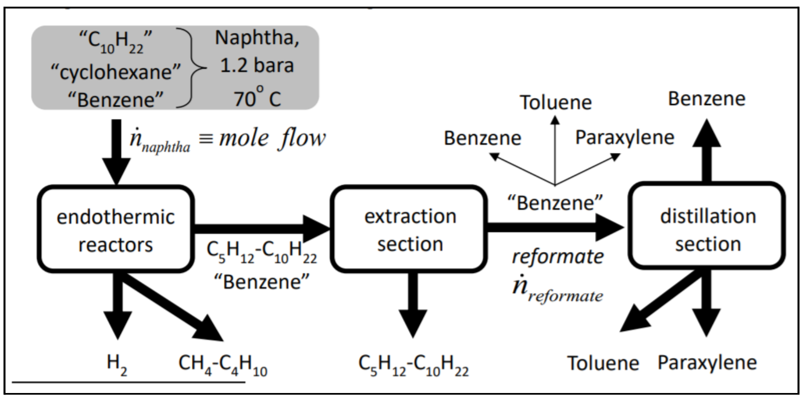


Figure 2: Operations to Isolate Aromatics [1]

All three processes were incorporated into one simulation file. The only "break" in the simulated process occurs before the last two distillation columns, where the BTX feed is instantiated. Before the "reformate" entered into the distillation section, the mole fractions of the chemical components represented by benzene were calculated using *Equations 1-3*.

$$x_{reformate,\,benzene} = \frac{\dot{n}_{naphtha}x_{naphtha,benzene} + (\dot{n}_{reformate} - \dot{n}_{naphtha}x_{naphtha,benzene}}{\dot{n}_{reformate}} \quad \text{(Eq. 1)}$$

$$x_{reformate,\,toluene} = \frac{(\dot{n}_{reformate} - \dot{n}_{naphtha}x_{naphtha,benzene})*0.55}{\dot{n}_{reformate}} \quad \text{(Eq. 2)}$$

$$x_{reformate,\,para-xylene} = \frac{(\dot{n}_{reformate} - \dot{n}_{naphtha}x_{naphtha,benzene})*0.34}{\dot{n}_{reformate}} \quad \text{(Eq. 3)}$$

where: x=mole fraction , ṅ=molar flow rate.

This facility was designed to ensure that operations would continue if either feed K or feed TQ1 was being run through the facility. To ensure that the facility could handle both feeds and still meet the benzene and BTX specifications, separate simulations were constructed with both feeds. Equipment sizing and costs were calculated separately for both simulations. These values were combined into the final capital cost values by going through each line item for both feeds and selecting the higher price for each item. For example if a reflux drum was larger for feed TQ1 than it was for feed K, the design for TQ1 would go into the final design. This is to ensure that both feeds can be safely run through the same facility, as specifically requested in the project description. The operating costs, however, are going to vary based on which feed is being used.

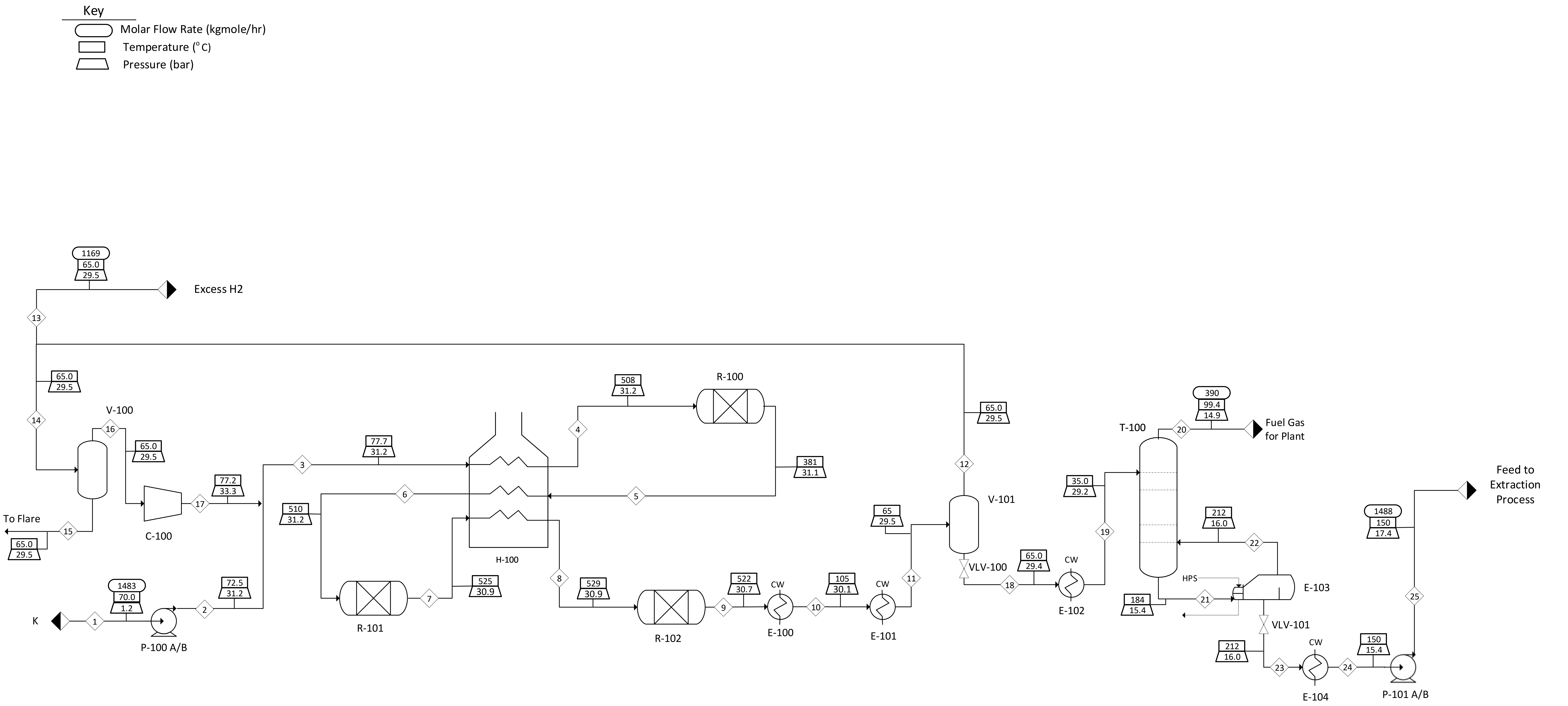
The service factor affected the design philosophy of this project. No service factor was specified by the client, so a service factor of 95% was assumed. This was chosen because any service factor between 91-99% is reasonable for this design [2]. All utility costs were calculated assuming that the plant will be operating 24 hours a day for 95% of the calendar year. Overall production revenue was also affected by the service factor. It was assumed that the manufacturing costs and maintenance costs could be calculated independent of the service factor because maintenance of equipment and installation of any new equipment can occur during the 5% downtime.

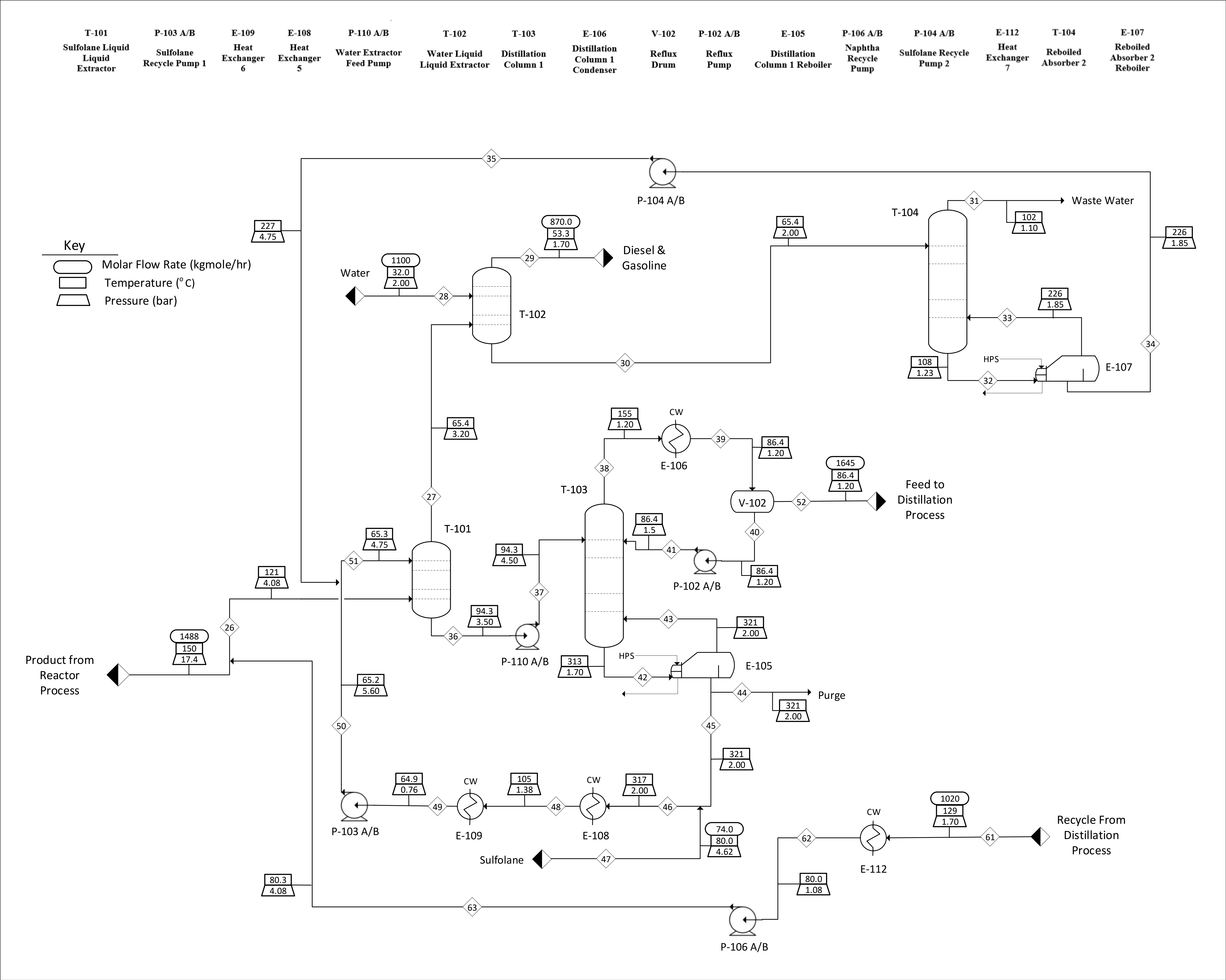
The overall process flow diagram (PFD) was split into three separate diagrams based on the reactor, extractor, and distillation sections for feed K of the process. All of these PFDs are included in their respectives appendices. The first piece of the process is the reactor section which highlights the fired heater and the packed bed reactors. The PFD for this section is included in Appendix A. Feed K of naphtha cyclohexane enters the process and goes through fired heater H-100 and a series of packed bed reactors R-100, R-101, R-102, and R-103. R-103 is a swing reactor not seen in the PFD but looked at more in detail in the Hysys simulation section later in the report. Once past the reactors, the stream travels through two heat exchangers in series, E-100 and E-101, to enter V-101. Hydrogen is separated from the mixture and recycled through compressor knockout drum, V-100, and compressor, C-100. Recycled hydrogen can also be sold as excess hydrogen while the liquid hydrogen from the knockout drum V-100 is sent to flare. The mixture continues from V-101 to heat exchanger E-102 and then enters column T-100. column T-100 produces fuel gas for the plant while sending the bottoms product through heat exchanger E-104 and pump P-101 A/B as the feed for the extraction process.

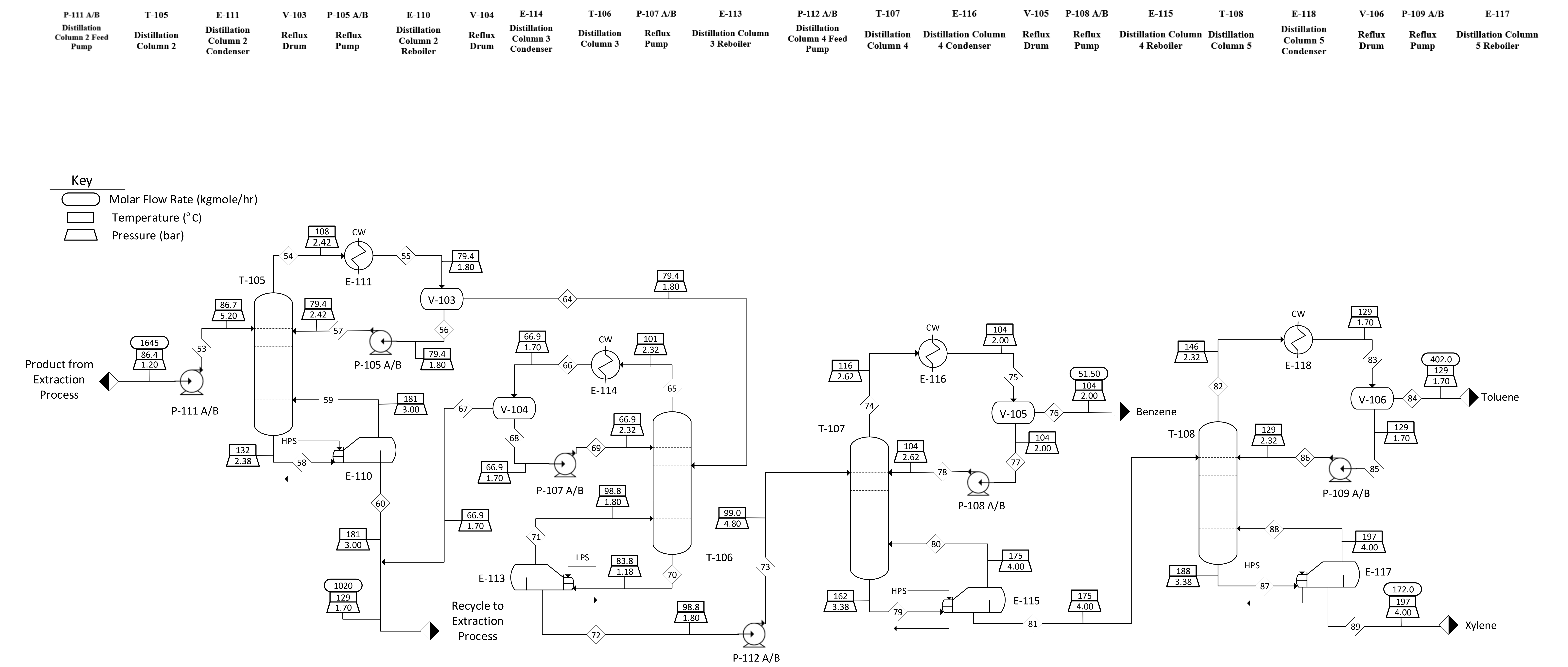
This leads to the second PFD for the extraction section, found in Appendix B. This section highlights the two liquid-liquid separators. Feed from the reactor process enters the first liquid-liquid extractor, T-101. The top products from T-101 enter the second liquid-liquid extractor, T-102, while the bottom product enters pump P-110 A/B. T-102 uses a water feed and the feed from T-101 to produce a top product of diesel and gasoline. The bottoms product from T-102 is sent to column T-104 to produce waste water and a recycle back to T-101. Pump P-110 A/B sends the mixture to the major fractionator T-103, to create a bottoms product sulfolane recycle stream and a distillate product to be used as the feed for the distillation section.

The PFD for the distillation section is found in Appendix C. The distillation section features the production of benzene, toluene, and para-xylene. The product from the extraction process enters P-111 A/B to feed into T-105. This will produce a recycle back to the extraction process and a feed to column T-106. Column T-106 produces a bottoms product of BTX to be fed to column T-107 and a tops product to add to the recycle back to the extraction process. The last two columns T-107 and T-108 separate BTX into benzene, toluene, and para-xylene products.

P-100 A/B	V-100	C-100	R-101	H-100	R-100	R-102	E-100	E-101	V-101	E-102	T-100	E-103	E-104	P-101 A/B
Fired Heater Feed Pump	Compressor Knockout drum	Compressor	Packed Bed Reactor 2	Fired Heater	Packed Bed Reactor 1	Packed Bed Reactor 3	Heat Exchanger 1	Heat Exchanger 2	Separator 1	Heat Exchanger 3	Reboiled Absorber 1	Reboiled Absorber 1 Reboiler	Heat Exchanger	Sulfolane Extractor Feed Pump




T-101
Sulfolane Liquid Liquid Extractor
P-103 A/B
Sulfolane Recycle Pump 1
E-109
Heat Exchanger 6
E-108
Heat Exchanger 5
P-110 A/B
Water Extractor Feed Pump
T-102
Water Liquid Liquid Extractor
T-103
Distillation Column 1
E-106
Distillation Column 1 Condenser
V-102
Reflux Drum
P-102 A/B
Reflux Pump
E-105
Distillation Column 1 Reboiler
P-106 A/B
Naphtha Recycle Pump
P-104 A/B
Sulfolane Recycle Pump 2
E-112
Heat Exchanger 7
T-104
Reboiled Absorber 2
E-107
Reboiled Absorber 2 Reboiler
Key
Molar Flow Rate (kgmole/hr)
Temperature (°C)
Pressure (bar)
35
P-104 A/B
227
4.75
T-104
31
Waste Water
102
1.10
65.4
2.00
226
1.85
870.0
53.3
1.70
29
Diesel &
Gasoline
1100
32.0
2.00
Water
28
T-102
226
1.85
33
34
30
108
1.23
HPS
32
E-107
65.4
3.20
155
1.20
CW
39
86.4
1.20
38
E-106
1645
86.4
1.20
Feed to
Distillation
Process
T-103
V-102
52
27
40
T-101
86.4
1.5
65.3
4.75
94.3
4.50
41
51
P-102 A/B
86.4
1.20
121
4.08
37
94.3
3.50
43
321
2.00
26
36
1488
150
17.4
P-110 A/B
313
1.70
HPS
E-105
Product from
Reactor
Process
42
65.2
5.60
44
Purge
321
2.00
50
45
321
2.00
64.9
0.76
CW
105
1.38
CW
317
2.00
49
48
46
P-103 A/B
E-109
E-108
74.0
80.0
4.62
CW
1020
129
1.70
Sulfolane
47
62
61
Recycle From
Distillation
Process
E-112
80.3
4.08
80.0
1.08
63
P-106 A/B



P-111 A/B Distillation Column 2 Feed Pump
T-105 Distillation Column 2
E-111 Distillation Column 2 Condenser
V-103 Reflux Drum
P-105 A/B Reflux Pump
E-110 Distillation Column 2 Reboiler
V-104 Reflux Drum
E-114 Distillation Column 3 Condenser
T-106 Distillation Column 3
P-107 A/B Reflux Pump
E-113 Distillation Column 3 Reboiler
P-112 A/B Distillation Column 4 Feed Pump
T-107 Distillation Column 4
E-116 Distillation Column 4 Condenser
V-105 Reflux Drum
P-108 A/B Reflux Pump
E-115 Distillation Column 4 Reboiler
T-108 Distillation Column 5
E-118 Distillation Column 5 Condenser
V-106 Reflux Drum
P-109 A/B Reflux Pump
E-117 Distillation Column 5 Reboiler
Key
Molar Flow Rate (kgmole/hr)
Temperature (°C)
Pressure (bar)
Product from Extraction Process
Recycle to Extraction Process
Benzene
Toluene
Xylene
CW
HPS
LPS
T-105
T-106
T-107
T-108
E-110
E-111
E-113
E-114
E-115
E-116
E-117
E-118
V-103
V-104
V-105
V-106
P-111 A/B
P-105 A/B
P-107 A/B
P-108 A/B
P-109 A/B
P-112 A/B


P-100 A/B Fired Heater Feed Pump
V-100 Compressor Knockout drum
C-100 Compressor
R-101 Packed Bed Reactor 2
H-100 Fired Heater
R-100 Packed Bed Reactor 1
R-102 Packed Bed Reactor 3
E-100 Heat Exchanger 1
E-101 Heat Exchanger 2
V-101 Separator 1
E-102 Heat Exchanger 3
T-100 Reboiled Absorber 1
E-103 Reboiled Absorber 1 Reboiler
E-104 Heat Exchanger 4
P-101 A/B Sulfolane Extractor Feed Pump
T-101 Sulfolane Liquid Liquid Extractor
P-103 A/B Sulfolane Recycle Pump 1
E-109 Heat Exchanger 6
E-108 Heat Exchanger 5
P-110 A/B Water Extractor Feed Pump
T-102 Water Liquid Liquid Extractor
T-103 Distillation Column 1
E-106 Distillation Column 1 Condenser
V-102 Reflux Drum
P-102 A/B Reflux Pump
E-105 Distillation Column 1 Reboiler
P-106 A/B Naphtha Recycle Pump
P-104 A/B Sulfolane Recycle Pump 2
P-111 A/B Distillation Column 2 Feed Pump
E-112 Heat Exchanger 7
T-104 Reboiled Absorber 2
E-107 Reboiled Absorber 2 Reboiler
T-105 Distillation Column 2
E-111 Distillation Column 2 Condenser
V-103 Reflux Drum
P-105 A/B Reflux Pump
E-110 Distillation Column 2 Reboiler
V-104 Reflux Drum
E-114 Distillation Column 3 Condenser
T-106 Distillation Column 3
P-107 A/B Reflux Pump
E-113 Distillation Column 3 Reboiler
P-112 A/B Distillation Column 4 Feed Pump
T-107 Distillation Column 4
E-116 Distillation Column 4 Condenser
V-105 Reflux Drum
P-108 A/B Reflux Pump
E-115 Distillation Column 4 Reboiler
T-108 Distillation Column 5
E-118 Distillation Column 5 Condenser
V-106 Reflux Drum
P-109 A/B Reflux Pump
E-117 Distillation Column 5 Reboiler
Key
Molar Flow Rate (kgmole/hr)
Temperature (°C)
Pressure (bar)
Excess H2
To Flare
TQ1 or K
Fuel Gas for Plant
Water
Diesel & Gasoline
Waste Water
Purge
Sulfolane
Benzene
Toluene
Xylene
CW
HPS
LPS

Stream Tables for Feed K – Compositions are given in Mole Fractions

Component	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
n-Decane	0.60	0.60	0.27	0.27	0.19	0.19	0.06	0.06	0.05	0.05	0.05	0.00	0.00	0.00	0.13
Cyclohexane	0.31	0.31	0.14	0.14	0.02	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Benzene	0.09	0.09	0.05	0.05	0.12	0.12	0.13	0.13	0.14	0.14	0.14	0.01	0.01	0.01	0.34
Methane	0.00	0.00	0.07	0.07	0.05	0.05	0.08	0.08	0.08	0.08	0.08	0.12	0.12	0.12	0.01
Oxygen	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen	0.00	0.00	0.39	0.39	0.53	0.53	0.44	0.44	0.44	0.44	0.44	0.72	0.72	0.72	0.01
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulfolane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-Pentane	0.00	0.00	0.00	0.00	0.01	0.01	0.04	0.04	0.04	0.04	0.04	0.01	0.01	0.01	0.08
n-Hexane	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.02	0.01	0.01	0.01	0.00	0.00	0.00	0.02
n-Heptane	0.00	0.00	0.00	0.00	0.00	0.00	0.03	0.03	0.03	0.03	0.03	0.00	0.00	0.00	0.08
n-Octane	0.00	0.00	0.00	0.00	0.00	0.00	0.03	0.03	0.03	0.03	0.03	0.00	0.00	0.00	0.08
n-Nonane	0.00	0.00	0.00	0.00	0.00	0.00	0.03	0.03	0.03	0.03	0.03	0.00	0.00	0.00	0.08
n-Butane	0.00	0.00	0.01	0.01	0.01	0.01	0.04	0.04	0.04	0.04	0.04	0.02	0.02	0.02	0.08
Propane	0.00	0.00	0.03	0.03	0.02	0.02	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.06
Ethane	0.00	0.00	0.04	0.04	0.04	0.04	0.06	0.06	0.06	0.06	0.06	0.08	0.08	0.08	0.04
CO2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Toluene	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
p-Xylene	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00

Component	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
n-Decane	0.00	0.00	0.13	0.13	0.000	0.159	0.159	0.159	0.159	0.159	0.271	0.274	0.000	0.275	0.000
Cyclohexane	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Benzene	0.01	0.01	0.34	0.34	0.032	0.416	0.416	0.416	0.416	0.416	0.408	0.004	0.000	0.004	0.000
Methane	0.12	0.12	0.01	0.01	0.066	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Nitrogen	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Hydrogen	0.72	0.72	0.01	0.01	0.058	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Sulfolane	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.007	0.004	0.000	0.000	0.003
H2O	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.003	0.997
n-Pentane	0.01	0.01	0.08	0.08	0.031	0.098	0.098	0.098	0.098	0.098	0.084	0.168	0.000	0.168	0.000
n-Hexane	0.00	0.00	0.02	0.02	0.003	0.029	0.029	0.029	0.029	0.029	0.022	0.049	0.000	0.049	0.000
n-Heptane	0.00	0.00	0.08	0.08	0.003	0.095	0.095	0.095	0.095	0.095	0.063	0.153	0.000	0.153	0.000
n-Octane	0.00	0.00	0.08	0.08	0.001	0.096	0.096	0.096	0.096	0.096	0.068	0.165	0.000	0.165	0.000
n-Nonane	0.00	0.00	0.08	0.08	0.001	0.097	0.097	0.097	0.097	0.097	0.068	0.165	0.000	0.166	0.000
n-Butane	0.02	0.02	0.08	0.08	0.334	0.010	0.010	0.010	0.010	0.010	0.009	0.017	0.000	0.017	0.000
Propane	0.05	0.05	0.06	0.06	0.285	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Ethane	0.08	0.08	0.04	0.04	0.186	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO2	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Toluene	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
p-Xylene	0.00	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000

Stream Tables for Feed K – Compositions are given in Mole Fractions

Component	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45
n-Decane	0.000	0.000	0.000	0.000	0.000	0.083	0.083	0.269	0.269	0.269	0.269	0.000	0.000	0.000	0.000
Cyclohexane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Benzene	0.000	0.000	0.000	0.000	0.000	0.191	0.191	0.622	0.622	0.622	0.622	0.000	0.000	0.000	0.000
Methane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Hydrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Sulfolane	0.000	0.980	0.980	0.980	0.980	0.695	0.695	0.010	0.010	0.010	0.010	1.000	1.000	1.000	1.000
H2O	1.000	0.020	0.020	0.020	0.020	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-Pentane	0.000	0.000	0.000	0.000	0.000	0.012	0.012	0.039	0.039	0.039	0.039	0.000	0.000	0.000	0.000
n-Hexane	0.000	0.000	0.000	0.000	0.000	0.002	0.002	0.008	0.008	0.008	0.008	0.000	0.000	0.000	0.000
n-Heptane	0.000	0.000	0.000	0.000	0.000	0.005	0.005	0.015	0.015	0.015	0.015	0.000	0.000	0.000	0.000
n-Octane	0.000	0.000	0.000	0.000	0.000	0.005	0.005	0.017	0.017	0.017	0.017	0.000	0.000	0.000	0.000
n-Nonane	0.000	0.000	0.000	0.000	0.000	0.005	0.005	0.016	0.016	0.016	0.016	0.000	0.000	0.000	0.000
n-Butane	0.000	0.000	0.000	0.000	0.000	0.002	0.002	0.005	0.005	0.005	0.005	0.000	0.000	0.000	0.000
Propane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Ethane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Toluene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
p-Xylene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000

Component	46	47	48	49	50	51	52	53	54	55	56	57	58	59	60
n-Decane	0.000	0.000	0.000	0.000	0.000	0.000	0.269	0.269	0.000	0.000	0.000	0.000	0.671	0.671	0.671
Cyclohexane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Benzene	0.000	0.000	0.000	0.000	0.000	0.000	0.622	0.622	0.903	0.903	0.903	0.903	0.200	0.200	0.200
Methane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Hydrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Sulfolane	1.000	1.000	1.000	1.000	1.000	1.000	0.010	0.010	0.000	0.000	0.000	0.000	0.025	0.025	0.025
H2O	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-Pentane	0.000	0.000	0.000	0.000	0.000	0.000	0.039	0.039	0.065	0.065	0.065	0.065	0.000	0.000	0.000
n-Hexane	0.000	0.000	0.000	0.000	0.000	0.000	0.008	0.008	0.013	0.013	0.013	0.013	0.000	0.000	0.000
n-Heptane	0.000	0.000	0.000	0.000	0.000	0.000	0.015	0.015	0.010	0.010	0.010	0.010	0.023	0.023	0.023
n-Octane	0.000	0.000	0.000	0.000	0.000	0.000	0.017	0.017	0.000	0.000	0.000	0.000	0.042	0.042	0.042
n-Nonane	0.000	0.000	0.000	0.000	0.000	0.000	0.016	0.016	0.000	0.000	0.000	0.000	0.039	0.039	0.039
n-Butane	0.000	0.000	0.000	0.000	0.000	0.000	0.005	0.005	0.008	0.008	0.008	0.008	0.000	0.000	0.000
Propane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Ethane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Toluene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
p-Xylene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000

Stream Tables for Feed K – Compositions are given in Mole Fractions

Component	61	62	63	64	65	66	67	68	69	70	71	72	73	74	75
n-Decane	0.434	0.434	0.434	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Cyclohexane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Benzene	0.398	0.398	0.398	0.903	0.761	0.761	0.761	0.761	0.761	0.99	0.99	0.99	0.99	0.900	0.900
Methane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Hydrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Sulfolane	0.016	0.016	0.016	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
H2O	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
n-Pentane	0.063	0.063	0.063	0.065	0.177	0.177	0.177	0.177	0.177	0.00	0.00	0.00	0.00	0.000	0.000
n-Hexane	0.012	0.012	0.012	0.013	0.035	0.035	0.035	0.035	0.035	0.00	0.00	0.00	0.00	0.005	0.005
n-Heptane	0.016	0.016	0.016	0.010	0.004	0.004	0.004	0.004	0.004	0.01	0.01	0.01	0.01	0.061	0.061
n-Octane	0.027	0.027	0.027	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
n-Nonane	0.025	0.025	0.025	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
n-Butane	0.008	0.008	0.008	0.008	0.023	0.023	0.023	0.023	0.023	0.00	0.00	0.00	0.00	0.000	0.000
Propane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Ethane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000
Toluene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.034	0.034
p-Xylene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.00	0.00	0.000	0.000

Component	76	77	78	79	80	81	82	83	84	85	86	87	88	89
n-Decane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Cyclohexane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Benzene	0.900	0.900	0.900	0.013	0.013	0.013	0.015	0.015	0.015	0.015	0.015	0.000	0.000	0.000
Methane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Hydrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Sulfolane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H2O	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-Pentane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-Hexane	0.005	0.005	0.005	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-Heptane	0.061	0.061	0.061	0.007	0.007	0.007	0.007	0.007	0.007	0.007	0.007	0.000	0.000	0.000
n-Octane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-Nonane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-Butane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Propane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Ethane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Toluene	0.034	0.034	0.034	0.606	0.606	0.606	0.675	0.675	0.675	0.675	0.675	0.050	0.050	0.050
p-Xylene	0.000	0.000	0.000	0.374	0.374	0.374	0.303	0.303	0.303	0.303	0.303	0.950	0.950	0.950

Stream Tables for Feed TQ1 – Compositions are given in Mole Fractions

Component	1	3	5	7	9	12	15	16	18	20	21	26	27	30	32	33
n-Decane	0.778	0.439	0.249	0.085	0.072	0.000	0.138	0.000	0.138	0.000	0.190	0.288	0.266	0.000	0.000	0.000
Cyclohexane	0.206	0.116	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Benzene	0.016	0.010	0.092	0.108	0.113	0.003	0.211	0.003	0.211	0.018	0.285	0.333	0.003	0.000	0.000	0.000
Methane	0.000	0.086	0.083	0.107	0.107	0.196	0.026	0.196	0.026	0.094	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Hydrogen	0.000	0.277	0.373	0.296	0.306	0.634	0.011	0.634	0.011	0.039	0.000	0.000	0.000	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Sulfolane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.006	0.004	0.003	0.980	0.980
H2O	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.996	0.020	0.020
n-Pentane	0.000	0.002	0.021	0.051	0.052	0.005	0.095	0.005	0.095	0.029	0.120	0.096	0.168	0.000	0.000	0.000
n-Hexane	0.000	0.000	0.018	0.016	0.009	0.000	0.017	0.000	0.017	0.002	0.023	0.017	0.031	0.000	0.000	0.000
n-Heptane	0.000	0.000	0.018	0.047	0.048	0.000	0.090	0.000	0.090	0.003	0.123	0.082	0.166	0.000	0.000	0.000
n-Octane	0.000	0.000	0.018	0.047	0.048	0.000	0.090	0.000	0.090	0.001	0.124	0.085	0.173	0.000	0.000	0.000
n-Nonane	0.000	0.000	0.018	0.047	0.047	0.000	0.090	0.000	0.090	0.001	0.124	0.084	0.174	0.000	0.000	0.000
n-Butane	0.000	0.007	0.024	0.054	0.055	0.015	0.091	0.015	0.091	0.304	0.010	0.009	0.014	0.000	0.000	0.000
Propane	0.000	0.020	0.034	0.063	0.064	0.046	0.080	0.046	0.080	0.289	0.000	0.000	0.000	0.000	0.000	0.000
Ethane	0.000	0.043	0.052	0.079	0.079	0.100	0.061	0.099	0.061	0.219	0.000	0.000	0.000	0.000	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Toluene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
p-Xylene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000

Component	36	38	52	55	58	64	65	72	74	81	86	87
n-Decane	0.075	0.304	0.304	0.000	0.588	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Cyclohexane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Benzene	0.143	0.581	0.581	0.881	0.300	0.881	0.707	0.99	0.90	0.036	0.052	0.000
Methane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Nitrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Hydrogen	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Sulfolane	0.757	0.011	0.011	0.000	0.021	0.000	0.000	0.00	0.00	0.000	0.000	0.000
H2O	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
n-Pentane	0.010	0.042	0.042	0.087	0.000	0.087	0.234	0.00	0.00	0.000	0.000	0.000
n-Hexane	0.001	0.005	0.005	0.011	0.000	0.011	0.029	0.00	0.00	0.000	0.000	0.000
n-Heptane	0.004	0.018	0.018	0.010	0.025	0.010	0.003	0.01	0.06	0.010	0.014	0.000
n-Octane	0.004	0.018	0.018	0.000	0.035	0.000	0.000	0.00	0.00	0.000	0.000	0.000
n-Nonane	0.004	0.016	0.016	0.000	0.031	0.000	0.000	0.00	0.00	0.000	0.000	0.000
n-Butane	0.001	0.005	0.005	0.010	0.000	0.010	0.026	0.00	0.00	0.000	0.000	0.000
Propane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Ethane	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
CO2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.000	0.000	0.000
Toluene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.03	0.590	0.821	0.050
p-Xylene	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00	0.00	0.364	0.113	0.950

Economic Analysis and Sensitivities

Under direction from AIChE, a standard 10 year project evaluation life and a straight line MACRS depreciation was used to analyze the economics of this project [1]. Because this project primarily concerns chemical plant equipment, a 5 year MACRS depreciation life was used [3]. Four economic analyses were completed for processing through the same facility: feed K under Iraqi tax control, feed K under Kurdish tax control, feed TQ1 under Iraqi tax control, and feed TQ1 under Kurdish tax control. A washout assumption for the costs and revenues was used for all economic analyses [3].

Available utilities on site and costing for those utilities can be seen in Table 2. Heat exchangers used throughout the design use cooling water and pressurized steam to accomplish the desired heating or cooling needed. The pumps and compressors use the on-site electricity. The fired heaters use air and utility fuel gas in combination with the fuel gas that is produced in the catalytic reformers.

Table 2: Utility Costs [1]

Utility	Cost	Units
Electricity kWhr	0.25	$/kWhr
Steam, 450 psig	19.36	$/1000kg
Steam 150 psig	14.08	$/1000kg
Steam, 50 psig	8.8	$/1000kg
Natural gas	9.43	$/MMBTU
Cooling water, 25℃	0.5	$/GJ

The process has five product streams: fuel gas, a gasoline and diesel mixture, benzene, toluene, xylenes, and fuel gas. The only product stream specification was for the BTX to be separated to 99% purity, which was achieved in this design for both feeds. Pricing information for the feed streams and product streams is included in Table 3 [1].

Table 3: Feed and Product Pricing Information

Designation	Stream	Component	Price (2021)	Units
Reactant	*1*	Naphtha	0.325	$/L
Solvent	*47*	Sulfolane	5	$/kg
Platinum Catalyst		Platinum	51000	$/kg
Product	*29*	Linear Alkane (Diesel)	0.98	$/L
		Cyclic Hydrocarbon (Gasoline)	0.63	$/L
Product	*76*	Benzene	3.49	$/gal
Product	*84*	Toluene	2.79	$/gal
Product	*89*	Xylenes	2.79	$/gal
Product	*20*	Fuel Gas	*Consumed in facility operations*	

Five directly sellable products are produced. Of these, the product with the highest gross revenue is the linear alkanes product stream, which can be sold as diesel for $0.98/L. The products with the least gross revenue are the xylenes and toluenes which both sell for $2.79/gal [1]. Additionally, fuel gas and hydrogen products can be utilized in plant operations to reduce operating costs. The production volumes were determined using the HYSYS simulation and are summarized in Table 4.

Table 4: Product Production and Gross Revenue				
Product	**Feed K Production Values**	**Feed TQ1 Production Values**	**Feed K Gross Revenue ($)**	**Feed TQ1 Gross Revenue ($)**
Diesel (Linear Alkanes, L/Yr)	1,253,030,400	1,436,289,600	$1,227,969,792	$1,407,563,808
Gasoline (Cyclic Hydrocarbons, L/Yr)	3,781,692	4,914,360	$2,382,465	$3,096,047
Benzene (L/Yr)	61,214,880	38,920,680	$213,639,931	$135,833,173
Toluene (L/Yr)	558,099,600	354,867,600	$1,557,097,884	$990,080,604
Para-Xylene (L/Yr)	81,835,920	52,034,400	$228,322,216	$145,175,976
Total	**1,957,962,492**	**1,887,026,640**	**$3,229,412,288**	**$2,681,749,608**

Cash Flow

Tables 5 and 6 show the cash flow analysis used to determine the economic viability of the project, specifically for feeds K and TQ1 under the Iraqi tax regime. It is important to note that the value for the operating costs includes the total price for labor and for sulfolane per year. The utility prices per year included in the economic evaluation were electrical, fuel gas, steam, and cooling water. The overall goal for the cash flow tables was to determine the Discounted Cash Flow Rate of Return (DCFROR) and the Net Present Value (NPV). These values help us determine the overall economic attractiveness of the project. The DCFROR for feed K under the Iraqi tax regime was 67.33%, which is greater than the minimum rate of return of 15%, making the project economically attractive. The DCFROR for feed TQ1 also under the Iraqi tax regime was 51.14%, making that process economically attractive as well. The DCFRORs for feeds K and TQ1 for the Kurdish tax regime were calculated to be 74.11% and 56.44%. All of the DCFRORs are summarized below the cash flow tables in Table 7. Based on these four economic analyses, it was concluded that operating the plant for each feed is economically feasible if the plant is taxed with either rate. This therefore makes the overall project economically attractive and worth the significant investment. Another important factor the company should consider is that feed K is more economically advantageous than feed TQ1.

Project Title	**Table 5: Cash Flow Table for Feed K, Taxes - Iraqi Control**										
*Minimum Rate of Return, i**	15%	0.15	*Tax Rate*	35%	0.35						
End of Year	**0 (2021)**	**1 (2022)**	**2 (2023)**	**3 (2024)**	**4 (2025)**	**5 (2026)**	**6 (2027)**	**7 (2028)**	**8 (2029)**	**9 (2030)**	**10 (2031)**
Diesel Production		1,227,969,792	1,227,969,792	1,227,969,792	1,227,969,792	1,227,969,792	1,227,969,792	1,227,969,792	1,227,969,792	1,227,969,792	1,227,969,792
Gasoline Production		2,382,466	2,382,466	2,382,466	2,382,466	2,382,466	2,382,466	2,382,466	2,382,466	2,382,466	2,382,466
Benzene Production		213,639,931	213,639,931	213,639,931	213,639,931	213,639,931	213,639,931	213,639,931	213,639,931	213,639,931	213,639,931
Toluene Production		1,557,097,884	1,557,097,884	1,557,097,884	1,557,097,884	1,557,097,884	1,557,097,884	1,557,097,884	1,557,097,884	1,557,097,884	1,557,097,884
P-Xylene Production		228,322,217	228,322,217	228,322,217	228,322,217	228,322,217	228,322,217	228,322,217	228,322,217	228,322,217	228,322,217
Service Factor	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95
Sales Revenue	0	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675
Net Revenue	0	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675	3,067,941,675
(-) Raw Materials Cost	(1,945,333,173)					(1,945,333,173)					(1,945,333,173)
(-) Other Op Costs	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)	(370,201,389)
(-) Electrical + Gas Utilities	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)	(22,030,923)
(-) Steam Utilities	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)	(21,440,703)
(-) Cooling Water Utilities	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)	(7,476,038)
(-) Depreciation	(4,892,315)	(7,827,705)	(4,696,623)	(2,817,974)	(2,817,974)	(1,408,987)					
Taxable Income	(2,371,374,541)	2,638,964,917	2,642,095,999	2,643,974,648	2,643,974,648	700,050,462	2,646,792,622	2,646,792,622	2,646,792,622	2,646,792,622	701,459,449
(-) Tax @ 35%	829,981,089	(923,637,721)	(924,733,600)	(925,391,127)	(925,391,127)	(245,017,662)	(926,377,418)	(926,377,418)	(926,377,418)	(926,377,418)	(245,510,807)
Net Income	(1,541,393,452)	1,715,327,196	1,717,362,399	1,718,583,521	1,718,583,521	455,032,800	1,720,415,204	1,720,415,204	1,720,415,204	1,720,415,204	455,948,642
(+) Depreciation	4,892,315	7,827,705	4,696,623	2,817,974	2,817,974	1,408,987					
(+) Writeoff											883,325,205
(-) Working Capital	(883,325,205)										
(-) Fixed Capital	(24,461,577)										
Cash Flow	(2,444,287,918)	1,723,154,901	1,722,059,022	1,721,401,495	1,721,401,495	456,441,787	1,720,415,204	1,720,415,204	1,720,415,204	1,720,415,204	1,339,273,847
(P/F) i,n Factor	1	0.87	0.756	0.658	0.572	0.497	0.432	0.376	0.327	0.284	0.247
Discounted Cash Flow	(2,444,287,918)	1,498,395,566	1,302,124,024	1,131,849,426	984,216,892	226,932,238	743,782,969	646,767,799	562,406,782	489,049,376	331,048,012
NPV @ i*	**5,472,285,166**										
DCFROR	**67.33%**										

Project Title	**Table 6: Cash Flow Table for Feed TQ1, Taxes - Iraqi Control**										
*Minimum Rate of Return, i**	15%	0.15	*Tax Rate*	35%	0.35						
End of Year	**0 (2021)**	**1 (2022)**	**2 (2023)**	**3 (2024)**	**4 (2025)**	**5 (2026)**	**6 (2027)**	**7 (2028)**	**8 (2029)**	**9 (2030)**	**10 (2031)**
Diesel Production	0	1,407,563,808	1,407,563,808	1,407,563,808	1,407,563,808	1,407,563,808	1,407,563,808	1,407,563,808	1,407,563,808	1,407,563,808	1,407,563,808
Gasoline Production	0	3,096,047	3,096,047	3,096,047	3,096,047	3,096,047	3,096,047	3,096,047	3,096,047	3,096,047	3,096,047
Benzene Production	0	135,833,173	135,883,173	135,833,173	135,833,173	135,833,173	135,833,173	135,833,173	135,833,173	135,833,173	135,833,173
Toluene Production	0	990,080,604	990,080,604	990,080,604	990,080,604	990,080,604	990,080,604	990,080,604	990,080,604	990,080,604	990,080,604
P-Xylene Production	0	145,175,976	145,175,976	145,175,976	145,175,976	145,175,976	145,175,976	145,175,976	145,175,976	145,175,976	145,175,976
Service Factor	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95
Sales Revenue	0	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128
Net Revenue	0	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128	2,547,662,128
(-) Raw Materials Cost	(1,945,333,173)					(1,945,333,173)					(1,945,333,173)
(-) Other Op Costs	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)	(400,202,199)
(-) Electrical + Gas Utilities	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)	(22,143,462)
(-) Steam Utilities	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)	(19,151,196)
(-) Cooling Water Utilities	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)	(6,824,020)
(-) Depreciation	(4,892,315)	(7,827,705)	(4,696,623)	(2,817,974)	(2,817,974)	(1,408,987)					
Taxable Income	(2,398,546,365)	2,091,513,546	2,094,644,628	2,096,523,277	2,096,523,277	152,599,091	2,099,314,251	2,099,314,251	2,099,314,251	2,099,314,251	154,008,078
(-) Tax @ 35%	839,491,228	(732,029,741)	(733,125,620)	(733,783,147)	(733,783,147)	(53,409,682)	(734,769,438)	(734,769,438)	(734,769,738)	(734,769,438)	(53,902,827)
Net Income	(1,559,055,137)	1,359,483,805	1,361,519,008	1,362,740,130	1,362,740,130	99,189,409	1,364,571,813	1,364,571,813	1,364,571,813	1,364,571,813	100,105,251
(+) Depreciation	4,892,315	7,827,705	4,696,623	2,817,974	2,817,974	1,408,987					
(+) Writeoff											
(-) Working Capital	(883,325,205)										
(-) Fixed Capital	(24,461,577)										
Cash Flow	(2,461,949,604)	1,367,311,510	1,366,215,631	1,365,558,104	1,365,558,104	100,598,396	1,364,571,813	1,364,571,813	1,364,571,813	1,364,571,813	983,430,456
(P/F) i,n Factor	1	0.87	0.756	0.658	0.572	0.497	0.432	0.376	0.327	0.284	0.247
Discounted Cash Flow	(2,461,949,604)	1,188,966,530	1,033,055,298	897,876,620	780,762,278	50,015,182	589,942,051	512,993,088	446,080,946	387,896,475	243,088,968
NPV @ i*	**3,668,727,833**										
DCFROR	**51.14%**										

Payback Periods

A payback period is a useful financial analysis tool for determining when the company will start truly earning money. Table 7 details the payback period for each feed as well as the average payback period for all of the feeds. On average, it will take between 1.5 and 2 years for the project to recover the cash spent on project manufacturing and development. Table 7 also allows the payback periods to be compared to the economic analysis tools, the DCFROR and the NPV. This allows us to see the relationships between the different analysis tools.

Table 7: Payback Period, NPV, DCFROR

Feed	Undiscounted Payback Period (Years)	Discounted Payback Period (Years)	Net Present Value	Discounted Cash Flow Rate of Return
K (Kurdish)	1.30	1.56	$7,362,227,493	74.11%
K (Iraqi)	1.42	1.73	$5,472,285,166	67.33%
TQ1 (Kurdish)	1.65	2.03	$5,003,729,441	56.44%
TQ1 (Iraqi)	1.80	2.23	$3,668,727,833	51.14%
Average	1.54	1.89	$5,376,742.483	62.26%

Sensitivity Analysis

In order to better understand the effects of some of the potentially variable economic factors when conducting the economic analysis on the overall DCFROR and the NPV, a sensitivity analysis was conducted. The values for production, tax rate, operating costs, and capital costs were varied by up to 20% and new DCFROR and NPV values were calculated. The results of these calculations for feed K with the Iraqi tax regime is shown in Table 8. This data shows that for feed K (Iraqi) a change in production would have the largest impact on both the DCFROR and the NPV. The sensitivity analysis for feed K (Kurdish) and Feeds TQ1 (Kurdish and Iraqi) concur that production rate has the highest impact on DCFROR and NPV. We then used this data to create a tornado chart to visually show the effects of a +/- 20% change in production, tax rate, operating cost, and capital cost dollars. The value of 20% was chosen to simulate the potential uncertainty in the preliminary design. The base case, or original DCFROR for this particular feed was 67.34%, which is shown as the middle axis in Table 8. The base case for Table 8 was $5,472,285,166. The tornado charts show how the new values that the DCFROR and NPV would take on if the values for the categories listed were increased or decreased by 20%. With a minimum rate of return of 15%, even with a 20% decrease in production (which severely decreases revenue), the project is still economically attractive. The tornado charts for feed K (Kurdish) and Feeds TQ1 (Kurdish and Iraqi) also show that even with a 20% decrease in production, each feed will still produce enough revenue to surpass the project's breakeven point.

Table 8 Sensitivity Analysis for Feed K - Iraqi Tax Regime							
Capital Cost	**Value**	**NPV**	**DCFROR**	**Tax Rate**	**Value**	**NPV**	**DCFROR**
+20%	(29,353,892)	4,568,751,893	67.21%	+20%	0.42	4,810,805,351	64.29%
+10%	(26,907,735)	5,470,518,529	67.27%	+10%	0.39	5,094,296,700	65.65%
0	(24,461,577)	5,472,285,166	67.33%	0	0.35	5,472,285,166	67.33%
-10%	(22,015,419)	5,474,051,802	67.39%	-10%	0.32	5,755,776,515	68.51%
-20%	(19,569,262)	5,475,818,438	67.45%	-20%	0.28	6,133,764,980	69.98%
Production				**Operating Costs**			
+20%	3,681,530,010	7,473,932,791	84.52%	+20%	(444,241,667)	5,182,624,820	63.87%
+10%	3,374,735,843	6,473,108,978	75.97%	+10%	(407,221,528)	5,327,454,993	65.59%
0	3,067,941,675	5,472,285,166	67.33%	0	-370,201,389	5,472,285,166	67.33%
-10%	2,761,147,508	4,471,461,353	58.56%	-10%	(333,181,250)	5,617,115,338	69.11%
-20%	2,454,353,340	4,470,637,540	49.61%	-20%	(296,161,111)	5,761,945,511	70.92%

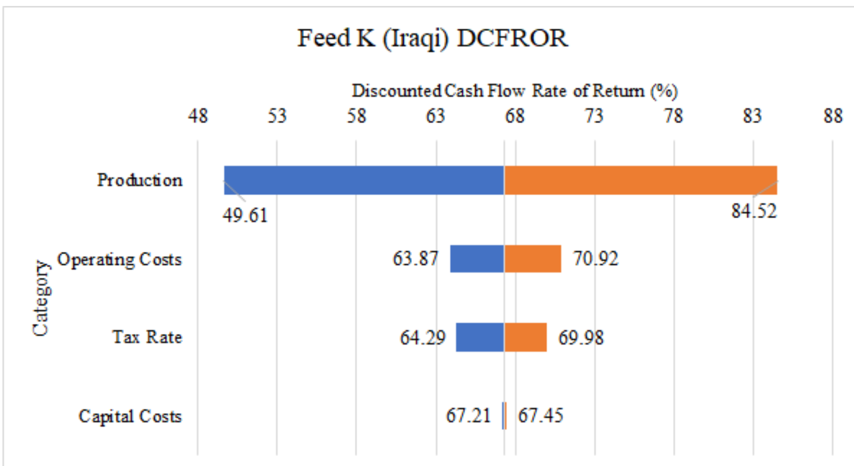


Figure 3 Feed K Iraqi Discounted Cash Flow Rate of Return Tornado Chart

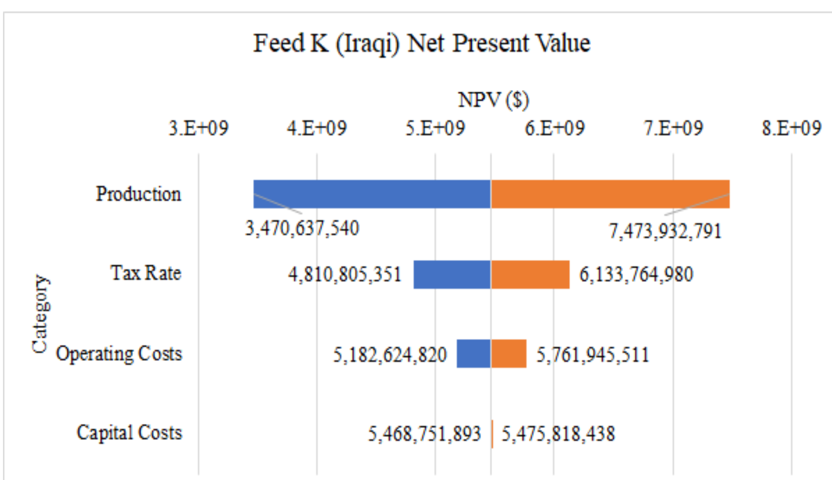


Figure 4 Feed K Iraqi Net Present Value Tornado Chart

Operating Costs & Capital Costs

Direct operating costs and raw material costs were considered for this project. Direct operating costs included the utilities, the operating labor, and the maintenance and repairs. All the waste vapor streams will go to flare, and the wastewater streams can be recycled and reused as feed for the water liquid-liquid extractor. As a result, waste processing did not need to be accounted for in the cost estimation of this facility. Labor was a significant consideration when calculating manufacturing costs. To calculate the number of operators needed per shift *Equation 4* was used. Once the number of operators per shift was calculated it was multiplied by 4.5 because that is how many operators need to be accounted for every single operator [2]. The labor calculations for each of the three process sections are included in the respective appendices. United States wages are 438% higher than Iraqi salaries. The salary per year in 2016 for an operator in the United States was $66,910 [2]. Consequently, the average hourly wage for an Iraqi worker would be $7.99/hr.

$$N_{OL}=(6.29+31.7P^2+0.23N_{np})^{0.5} \qquad \text{(Eq. 4)}$$

N_{OL} = number of operators per shift, P = number of particulate solids processing steps, N_{np} = equipment.

To account for the maintenance and repairs for the refinery, the fixed capital investment was multiplied by 0.1 and that value was used per year [2]. Taking the service factor of 95% into account, the yearly operating costs for each process section are detailed in the respective appendix. *Equations 5-8* were used to calculate the purchased cost, pressure factor, and bare module equipment cost. The material factor and bare module factors were also taken from graphs in Turton [2]. The operating costs, labor costs, and capital costs are listed in the appendix and are divided up by their respective process sections (reactor, distillation, extraction).

$$\log_{10}(C_P^0) = K_1 + K_2\log_{10}(A) + K_3[\log_{10}(A)]^2 \qquad \text{(Eq. 5)}$$

C_P^0 = purchased cost of equipment , A = capacity/ size parameter for the equipment

$$F_{P,vessel} = \frac{\frac{PD}{2[850-0.6(P)]}+0.00315}{0.0063} \qquad \text{(Eq. 6)}$$

$F_{P,vessel}$ = pressure factor for process vessel , P = pressure (barg), D = vessel diameter (m)

$$\log_{10}(F_P) = C_1 + C_2\log_{10}(P) + C_3[\log_{10}(P)]^2 \qquad \text{(Eq. 7)}$$

F_P = pressure factor , P = pressure (barg)

$$C_{BM} = (C_P^0)(F_{BM}) = C_P^0(B_1+B_2F_MF_P) \qquad \text{(Eq. 8)}$$

C_{BM} = bare module factor , F_{BM} = bare module cost factor , F_M = material factor

Process Safety

Environment, Health, and Safety (EHS) aspects are critical to the economic viability, sustainability, and social responsibility of the chemical sector investment and operations. At all of the preliminary stages of design, the main concern is the identification and elimination of hazards. However, some elements that pose hazards are permanent and inseparable parts of this process, so they cannot be eliminated. The hazards associated with them can, however, be eliminated, mitigated, or reduced. This is the foundation of "inherently safer design." The dangerous elements of the process are either eliminated entirely or adjusted to reach an acceptable level of risk. This is completed through a different philosophy of addressing safety issues than usual. Instead of adding protective equipment, hazards are addressed based on the physical and chemical properties of the system [4].

Inherently Safer Design

Inherently safer design is completed through five strategies: minimization, moderation or attenuation, substitution, simplification, or limitation of effects [4]. All of these strategies were considered for this project, and four of them were utilized, as summarized below in Table 9.

Table 9: Inherently Safer Design Evaluation

Substitution	**Heat Integration:** As the original project description recommended, the team had originally considered heat integration as a way to lower utility costs. However, for heat integration to be effective, the transportation distances and storage of hydrocarbons would be significantly increased. There is only one area where it might be beneficial to incorporate heat integration, and that would be crossing one of the hotter streams later in the distillation or extraction process with the feeds entering each fired heater. This would increase the risk of loss of containment for the flammable streams. Instead, utility water was used for each of the heat exchangers. This was also more practical because all of the shell and tube heat exchangers were cooling, rather than heating, the process material. The decision to use cooling water also decreased the required area for the heat exchangers, which decreases both cost and risk.
Minimization	**Fuel Gas Recycle:** Fuel gas is produced in the catalytic reforming process and used as a utility for the fired heater. This decreases the transport of flammable materials. Fuel gas is less than 1% of overall fuel gas being consumed per hour, so the produced fuel gas has no effect on the composition of other fuel gas being burned.
	Sulfolane Recycle: A sulfolane recovery and recycle system was incorporated to reduce the amount of fresh sulfolane that needs to be added to the extraction process per hour. This significantly reduces the amount of sulfolane that needs to be stored and transported onsite both as feed and as a waste product.
	Hydrogen Recycle: Hydrogen is both a reactant and a product of the catalytic reforming process. At the end of the reactor process, the excess hydrogen is separated from the linear and cyclic hydrocarbons and is recycled back to the reactor feed. This reduces the amount of fresh hydrogen that needs to be stored, processed, and transported onsite. Hydrogen is extremely flammable and must be handled with care.
Moderation	**Vacuum Distillation:** The project team determined that T-103 needs to be operated as a vacuum distillation column. This would significantly reduce the high temperatures seen in the reboiler. This would have a safety benefit, as the current required temperatures are near the decomposition temperature for hydrocarbons. This would have an economic benefit, as it would reduce the heat duty of the reboiler and the subsequent sulfolane heat exchangers.
	Low Pressure Columns: For the primary fractionator and the columns in the distillation section, the operating pressures were optimized to be as low as possible. This has multiple benefits to safety and cost. It decreases the intensity of an overpressure event that a header and pressure relief system would be required to manage, and it decreases the reboiler duty of a column. Lower pressures also reduce the boiling point for each substance in the distillation column.
	Reactor Bypass Lines and Temperature Control Systems: The outlet temperatures of each of the PBRs is likely to vary based on the reaction rates. Events such as coking can cause this variation, so it will likely be possible during normal operation. To avoid overheating the process streams, bypass lines and a temperature control system will be installed before and after each reactor section. Sections of these bypass lines will also include a cooling water heat exchanger. These systems will serve to prevent runaway reactions.
Simplification	**Steam Stripping:** Steam stripping was considered, but ultimately not utilized for the sulfolane recovery process due to the added complexity it gave to the process. Eliminating the steam stripper removed two decanters and two heat exchangers from the process. It also allowed carbon steel to be used as a material of construction, since carbon steel cannot be used if water is present in a hydrocarbon stream.

Although these innovations go a long way towards improving the inherent safety of this process, there is still additional work that needs to be done as the project transitions into the detailed design phase. The project team recommends that in the next design phase, dikes are incorporated, as well as fire and explosion resistance barricades. From a controls standpoint, the control rooms should be placed away from operations, and they should be barricaded. All control and operational systems should be well labeled and easy to understand.

Process Safety Management

Professional Engineers hold the safety, health, and welfare of the public and the environment in paramount when performing duties. In order to fulfil these requirements engineers become familiar with the fundamentals of chemicals and process safety, and commit to a culture of loss prevention. Several steps have been taken to make this process safer.

Process Hazards

The primary hazards associated with potential human exposure to process materials are with regards to benzene, carbon monoxide, n-nonane and toluene. This is because these chemicals have the lowest permissible exposure limits in ppm out of the chemicals involved in this process. However, a table of OSHA chemical exposure limits, NFPA diamond classifications, and lethal dose (LD_{50}) limits for all process chemicals in this design - including intermediates - is included below as Table X. This is to provide safety and awareness for all workers that might come into contact with these chemicals. Additionally, Table 11 details a potential consequence summary for the release of various process materials. The consequence summary includes ratings for equipment damage, environmental compliance, loss of life, disruption of other business units, legal and public relations, and community impact.

Table 10: NFPA Diamond Classifications and OSHA Exposure Limits

Substance	**NFPA**				**OSHA**
	Health Hazard	**Fire Hazard**	**Reactivity Hazard**	**Special**	**8-hour Time Weighted Average Permissible Exposure Limit in ppm**
Benzene	2	3	0	N/A	50
Carbon Dioxide	2	0	0	SA	5000
Carbon Monoxide	3	4	0	N/A	50
Cyclohexane	1	3	0	N/A	300
Ethane	1	4	0	N/A	N/A
Hydrogen	0	4	0	N/A	N/A
Methane	2	4	0	N/A	N/A
n-Butane	1	4	0	N/A	N/A
n-Decane	1	2	0	N/A	500
n-Heptane	1	3	0	N/A	500
n-Hexane	2	3	0	N/A	500
n-Nonane	1	3	0	N/A	200
n-Octane	1	3	0	N/A	500
n-Pentane	1	4	0	N/A	1000
Nitrogen	0	0	0	SA	N/A
Oxygen	3	0	0	OX	N/A
Sulfolane	1	1	0	N/A	N/A
Toluene	2	3	0	N/A	200
Water	0	0	0	N/A	N/A
Xylenes	2	3	0	N/A	100

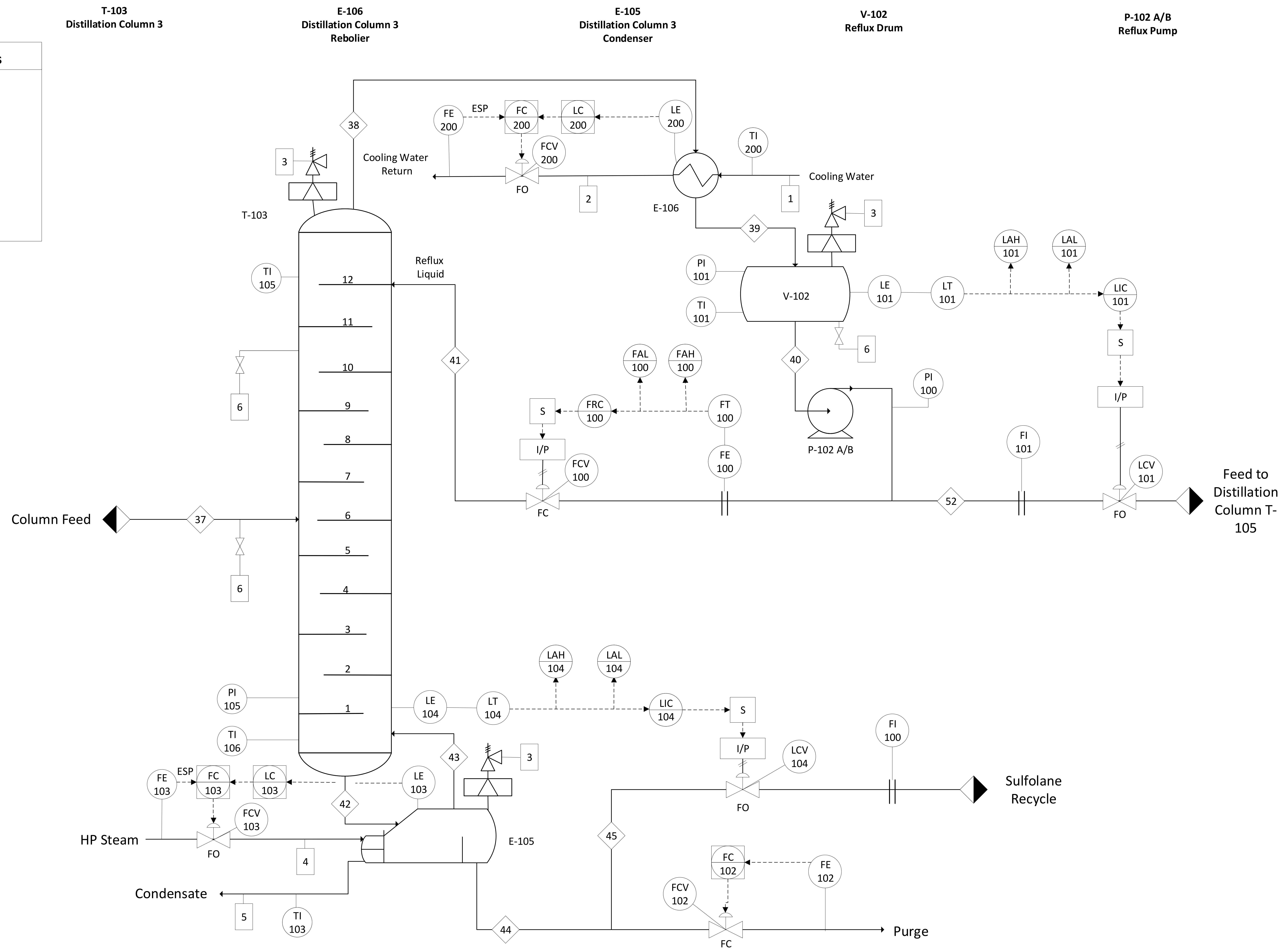
Table 11: Potential Consequence Summary						
Hazard	**Equipment Damage**	**Environmental Compliance**	**Loss of Life**	**Disruption of other Business Units**	**Legal/PR**	**Community Impact**
Hydrocarbon Vapor Cloud Explosion	High	Low	High	High	High	Medium
Hydrocarbon Toxic Release	Low	High	Low	Medium	Medium	Low
BTX Release	Low	High	Medium	Medium	Medium	Low
Sulfolane Release	Low	Medium	Low	Low	Low	Low

Each individual operating in the plant shall be required to wear personal protective equipment (PPE) due to the risks associated with the hazardous process components and dangerous on-site situations. The baseline requirements for most oil and gas plants include eye protection (safety glasses), head protection (helmet), feet protection (steel toe boots), and body protection (fire resistant clothing) [5]. Hearing protection may also be required based on the proximity to certain noisy equipment pieces. Hand protection, including impact or leather gloves may be required for certain physical activities. All individuals should be vigilant of the PPE needs for their daily activities. All persons within the fenceline perimeter of the plant will also be required to wear a gas monitor, the standard 4 gas monitor (O_2, H_2S, CO, and lower explosive limits) is recommended [6]. These gas monitors will be linked to the individual, and will have the capability to notify central plant management if an alarm is set off, or if the person is inactive for a prolonged period (man down alarm). All individuals working onsite will be required to undergo safety training, including but not limited to first aide, CPR, lock out tag out, and site specific emergency procedures. Additional training will be provided for confined space, hot work, and any elevated work that needs to be performed at the plant.

It is important to establish comprehensive management systems for plant maintenance and system changes to mitigate some of these disasters. A lock out tag out system will be implemented to ensure that dangerous items of equipment are properly shut off and not able to be started again with approval. This ensures that an equipment item that has been damaged or that is undergoing routine maintenance is not used without approval from the proper authorities. A management of change documentation system will also be used to ensure that all individuals, including contractors, are up to date on the current system changes within the plant. These systems are usually run by the EHS management team. After any system shutdown occurs, whether unplanned or as a result of maintenance, a Pre-startup Safety Review (PSSR) must be completed before operations can resume.

Utility Connections

1. Cooling Water Supply
2. Cooling Water Return
3. Vent to Header
4. HP Steam Supply
5. Steam Condensate Return
6. Sample Port



Control Systems

The P&ID of column T-103 features the use of multiple control systems, pressure relief valves, rupture disks, sample ports, alarm systems, and interlocks. There are local pressure and temperature indicators, TI-105, PI-105, and TI-106 located along the length of the column. These are useful for maintaining and continuously monitoring the tower. Reflux drum V-102 also uses pressure and temperature indicators PI-101 and TI-101 to monitor the vessel. Pump P-102 A/B uses a pressure indicator PI-100 on the discharge line to monitor the pressure going to reflux and the feed to distillation column T-105. The flow control loop FC-103 is the secondary loop in a cascade configuration. The primary controller is the level controller LC-103, whose purpose is to maintain a constant liquid level in E-105. Output from the primary controller, LC-103, creates the set point for the secondary loop FC-103. If a disturbance in the steam supply flow rate occurs, FC-103 will act very quickly to hold the steam flow rate at its set point. There is also a temperature indicator, TI-103, on the reboiler to help maintenance and production operators keep track of the temperature of the outlet flow of the condensate. Similarly, E-106 uses a control loop featuring FC-200 as the secondary loop controller and LC-200 as the primary controller. LE- 200 reads the level of E-106 and sends a signal to LC-200 which has a set point already programmed. If the reading from LE-200 deviates from the set point for the liquid level in the E-106, then a command is sent to FC-200 to alleviate the disturbance by adjusting control valve FCV-200. FE-200 is set to read the outlet cooling water return flow rate and send a signal to FC-200 which looks for any disturbances that deviate from the set point that LC-200 commands. If a disturbance occurs, then FC-200 sends an order to FCV-200 to either open or close more to mitigate the disruption to the set point. The heat exchanger, E-106 also has a temperature indicator on the inlet of the cooling water stream. Lastly, Stream 44 contains a simple feedback control system that controls the flow rate of the purge. FC-102 has an automated set point and controls FCV-102 based on the reading it receives from FE-102.

Rupture Disks and Pressure Relief Valves

A major tool incorporated to ensure a safer process are the three pressure relief systems that include a rupture disk and a pressure relief valve in series. Overpressure of T-103, V-102, and E-105 are major concerns for the process and therefore must be mitigated by relief systems. If the actual operating pressure of the equipment is above the safe or desired limits, the rupture disk will burst and the relief valve spring will compress, allowing the pressure to vent to header. If the pressure is too low, the rupture disk will serve as a safeguard and not allow flow while the relief valve will close should the rupture disk fail. Every control valve except valves FCV-100 and FCV-102 will fail open in case of failure or loss of power so that no pressure is trapped in the column and no flow is going uncontrolled to purge.

There are three total pressure relief valves on the major factionator. One is on the tower body itself, one is on the reboiler, and one is on the overhead reflux drum. They were all sized as conventional spring type relief valves since they discharge to header instead of directly to the atmosphere [7]. They were sized using the ASME 3% and the ASME 10% rules [4]. The greatest normal operating pressures in the tower, reboiler, and drum were 1.7 bar, 2 bar, and 1.2 bar, respectively. Because these values are so close together, they had similar calculated orifice areas required for the valves. The orifice areas are supposed to be 5.98 in^2, 6.06 in^2, and 6.06in^2, respectively, and they all have valve body sizes with (inlet x outlet) diameters of 4.6 in^2. Sample ports are located on the column feed line, the tower, and V-102 as places to take a sample and ensure they meet the product quality standards of the process.

Alarm and Interlock Systems

Safety Instrumented Systems (SIS) are implemented by companies to adhere to safety standards and to help prevent dangerous situations and potential disasters. Interlocks have switches high or low similar to alarm low and alarm high that will trigger under abnormally high or low process conditions. Solenoid switches are another type of interlock that are also tripped by extreme process conditions. The major fractionator, T-103, has multiple interlocks dispersed in different loops to help the process reach safe conditions. Process alarms are another line of defense used in multiple streams to help alert operators of abnormal or dangerous situations. An alarm is generated automatically when a measured variable exceeds a specified high or low limit [8]. The alarm then takes the appropriate corrective action to activate an annunciator either a visual display or an audible sound [8]. Stream 45 contains a feedback loop with alarms and an interlock system. The liquid level in T-103 is measured using LE-104 and has to pass through alarms LAH-104 and LAL-104 which will trigger if the levels in the column are too high or too low. If the alarms are triggered, a message will be sent to the controls room in the plant and will also trip a solenoid switch. The solenoid switch, located between the controller LIC-104 and control valve LCV-104, causes the air pressure in the pneumatic control valve to be vented. The venting of the control valve causes the control valve to revert to its fail-open position. Alarms and an interlock system are implemented on this loop to help prevent the overpressure and equipment damage of the reboiler. Downstream from Loop 104 is a flow indicator, FI-104, that will monitor the flow rate of the sulfolane recycle. Loop 100 uses an alarm and interlock system to control the amount of reflux going back in column T-103. For the feedback loop 100, FE-100 reads the reflux flow rate and sends a signal that first has to pass through the two alarms, FAL-100 and FAH-100, before reaching controller FRC-100. Similar to loop 104, if the alarms are triggered a message will be sent to the control room and trip the solenoid switch will cause FRC-100 to fail-close. Controller FRC-100 is key to preventing the overpressurization of T-103 which is why it is armed with alarms and interlocks systems. The last loop that uses alarms and an interlock system is feedback loop 101. Level element controller, LE-101, reads the level of the reflux drum, V-102, and sends a signal through alarms LAH-101 and LAL-100 to liquid indicator controller LIC-101. If the alarms are tripped, a message will be sent to the control room about the abnormal and dangerous liquid level in V-102. The solenoid switch will trigger and cause LCV-101 to fail open to resolve the levels in the reflux drum. Alarms and interlock systems are important for this loop so that equipment damage and overpressure of V-102, equipment damage of cooling water due to overfilling or fouling, and deviation from the setpoint for the reflux flow rate to T-103 downstream are all avoided. A flow indicator, FI-101, is also used to help monitor the flow rate of the feed to Distillation Column T-105.

Uncongested Vapor Cloud Deflagration

To highlight the extreme risks in hydrocarbon processing, a TNT equivalency calculation was performed for the atmospheric detonation of all chemicals from the largest, by inventory, process distillation column, T-103 [1]. The mass of flammable materials in T-103 was calculated using the HYSYS mass flow rate estimates through each stage of the column. For the TNT estimation, the mass flow on each stage of the column was tabulated for exactly one second. These values were then summed to get the total mass that would be in the columnar the instant of an explosion. It was found to be approximately 2,783 kg of material. After the total quantity of flammable material was found, the explosion efficiency was estimated and the scaling law and a figure from Reference [1]was used to estimate the peak side-on overpressure at various blast radii [4]. In a worst-case scenario when all fractionator contents instantly, and gaseous, vent to the atmosphere, one would need to stand 99 m away to be at a “safe distance” with a probability of 0.95 of no serious damage below this value. There will still be some damage to house ceilings and 10% of window glass will still be broken [4]. A map of the blast radius and its impacts is shown below in Figure 5 where the center of the map is the point of the detonation. A table of relevant distances from the explosion is also included as Table 12. Additionally, a table of lower explosive limits (LEL) and upper explosive limits (UEL) has been included to inform the installation of facility sensors and alarms. This is shown below as Table 13.

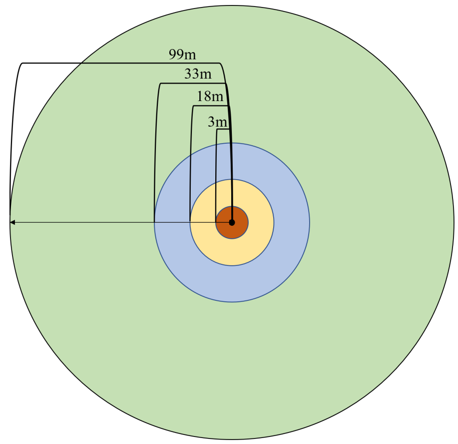


Figure 5: Uncongested Vapor Cloud Deflagration Blast Radius Map

Table 12: Vapor Cloud Distances

Approximate Distance (m)	**Damage Description**
3.0	Limit of Crater Lip
18	Probable Total Destruction of Buildings, Heavy Machine Tools (7000 lb) Moved and Badly Damaged, Very Heavy Machine Tools (12,000 lb) Survive
33 m	Lower Limit of Serious Structural Damage
99 m	“Safe Distance”

Table 13: Explosive Limits for Components in the Major Fractionator (T-103) [9] [4]

Component	**Mol Fraction**	**LEL %**	**UEL %**
n-Decane	8.16E-02	0.8	2.6
Cyclohexane	1.06E-05	1.3	8.4
Benzene	1.69E-01	1.4	8.0
Methane	1.45E-15	5.0	15.0
Hydrogen	6.13E-23	4.0	75.0
n-Pentane	1.12E-02	7.8	1.5
n-Hexane	2.21E-03	1.2	7.5
n-Heptane	4.44E-03	1.0	7.0
n-Octane	4.73E-03	1.0	6.5
n-Nonane	4.29E-03	0.8	2.9
n-Butane	1.48E-03	1.9	8.5
Propane	7.52E-09	2.1	9.5
Ethane	2.45E-10	2.9	13.0
Sulfolane	7.21E-01	2.22	14.1
H2O	2.56E-05	N/A	N/A

A risk assessment was completed on the major fractionator, T-103, including a "What If" hazard analysis in Table 14 [10]. The "What If" approach looks at what could go wrong with the system and shows the consequence, line of defense, hazard rating, and recommendation for the specific incident. To compensate for the risk and consequences, several lines of defense were implemented. T-103 has multiple alarms systems, rupture disks, and relief valves implemented in high risk areas in order to prevent overpressure events. Each failure is also ranked by class based on the risk formula of the severity of the event's consequence multiplied by the frequency of the event. This three tier risk classification system ranging from Class 1, acceptable risk, to Class 3, unacceptable risk, helps to prioritize hazards into categories. Lastly, the table describes the responsible party under the "recommendations or action plan/ by whom" column which details who is in charge should a failure occur. The worst case scenario of an hydrocarbon explosion occurring was also looked at in order to make the process safer as well as be conciencess of the community and environment surrounding the plant.

Table 14: A "What-If" Analysis on the Major Fractionator: T-103

Hazard/ Hazardous Event and Mechanism	**Consequence**	**Lines of Defense**	**Hazard Rating/ Risk Analysis/ Acceptance of Risk**	**Recommendations or Action Plan/ by Whom**
Fire Exposure	Overpressure; Explosion	Multiple Pressure Relief Valves, Rupture Disks, and Alarms	Class 1	Management, EMS, and Operators
Electric Power Failure Affecting P-102	Immediate Loss of Reflux to Fractionator	Pressure Indicator on P-102 and FRC 100	Class 1	Operators
Tube Rupture in E-105	Flashing; Transient Overpressure	LE 103, FC 103, Level and Pressure Indicator on Column, FC 102, and LY 104	Class 1	Operators
Cooling Water Failure in E-106	Column Overpressure, Reversal of Flow	Temperature Indicator, FC 200, Pressure Relief Valve, and Rupture Disk	Class 1	Operators
LIC 101 Failure	Loss of Reflux to Column, and Chance of Overpressure	Level Alarm for V-102, Pressure Relief Valve, Rupture Disk, and Safety Interlock System	Class 1	Operators
LC 103 Failure	Overfill in the reboiler	LE 103, FE 103, and FC 103, FC 102, FI 100	Class 1	Operators
LIC 104 Failure	Abnormal Liquid level in the Distillation Column	Level Alarm System for the Column, Pressure Relief, Rupture Disk, and Safety Interlock System	Class 1	Operators
LC 200 Failure	Column Overpressure and Reversal of Flow	FE 200, FC 20, Pressure Relief Valve, and Rupture Disks	Class 1	Operators
FRC 100 Failure	Loss of Reflux to Column	Flow Alarm System And Safety Interlock System for Stream 41	Class 1	Operators
FC 102 Failure	Flow Reversal, Equipment Damage to E-105	LE 103, LC 103, LY 104	Class 1	Operators
Control Valve Failure	Might Fail in an Undesired Position	Choosing either fail open or fail close	Class 1	Operators
Control Loop May Fail	Blocked Discharge (from Vessel/ Pump/ other Equipment System)	Having a fail open or fail close for the control valve when a control loop fails	Class 1	Operators

Safety Summary

In summary, this analysis has identified and analysed the most significant hazards that the plant design presents. In response, substantial safety measures have been integrated into the design, and additional safeguards have been recommended for the upcoming detailed design phase. For this plant, it is paramount that all efforts be made to prevent the loss of containment, as the materials involved are highly flammable. This included the prevention of over-pressure events. This analysis shows that there are no significant safety hazards that would prevent this project from proceeding to the next design phase.

Conclusions

This preliminary design allows for both feeds to be processed while meeting the specifications for benzene in the gasoline and diesel products as well as the benzene purity. This plant has been sized to allow for the fluctuating operating conditions and process needs of both feed K and feed TQ1. The economic analysis proved that this process design would be profitable for either an Iraqi or Kurdish tax regime. Additional design steps were taken to identify and mitigate risks by using the inherently safer design process process hazard analysis, and process safety management.

Appendix A: Reactor Train Detail

When modeling the reactors, it was determined that the factor with the greatest impact on reactor performance (including conversion) was the rate laws that had been predetermined in the project description. Therefore, the remaining factors that the team were able to manipulate to optimize the reactors were the operating temperature, the catalyst properties, the reactor size, and the void fraction. Of these factors, the one that had the greatest impact on the conversion rate was void fraction, which was determined through optimization, to be 0.55. Operating pressures must be between 25-35 bar and the reaction must be carried out at 400-500℃ to ensure that there is less opportunity for undesirable hydrogenolysis or hydrocracking in the rate-determining isomerization step of the reactions; these conditions were determined from industry literature [14]. Catalyst particle diameter must be between 2-5 mm in fixed bed catalytic reactors [2]. These particles were modeled as perfect spheres. The heat capacity used was the standard heat capacity for solids which is 250 kJ/kg-k. The bulk density of the catalyst was 770 kg/m^3, which was pulled from industry data for a Platinum - Iridium Catalyst [11].

Reaction kinetics for cycloalkane ("cyclohexane" proxy component) dehydrogenation were provided and are shown in *Equations 4* and *5*. Reaction kinetics for cycloalkane ("cyclohexane" proxy component") cyclization were also provided and are shown in *Equations 11* and *12*.

Reaction 1 $$C_6H_{12} \rightarrow C_6H_6 + 3H_2 \quad \text{(Eq. 9)}$$

$$rate = 9.4928 * 10^{13} e^{\frac{-160506.4}{8.314T}} P_{C_6H_{12}} - 8.2728 * 10^{-4} e^{\frac{52170.4}{8.314T}} P_{C_6H_6} P_{H_2}^3 \quad \text{(Eq. 10)}$$

Reaction 2 $$C_6H_{12} + 2H_2 \rightarrow 0.4C_5H_{12} + 0.4C_4H_{10} + 0.4C_3H_8 + 0.4C_2H_6 + 0.4CH_4 \quad \text{(Eq.11)}$$

$$rate = 3.6704 * 10^{21} e^{\frac{-287756.8}{8.314T}} P_{C_6H_{12}} \quad \text{(Eq. 12)}$$

Reaction kinetics for alkane ("decane" proxy component) cracking were provided and are shown in *Equations 8* and *9*. Reaction kinetics for cycloalkane ("cyclohexane" proxy component") cyclization were also provided and are shown in *Equations 15* and *16*.

Reaction 3 $$4.5C_{10}H_{22} + 4.5H_2 \rightarrow C_9H_{20} + C_8H_{18} + C_6H_{14} + C_5H_{12} + C_4H_{10} + C_3H_8 + C_2H_6 + CH_4 \quad \text{(Eq. 13)}$$

$$rate = 3.6704 * 10^{21} e^{\frac{-287756.8}{8.314T}} P_{C_{10}H_{22}} \quad \text{(Eq. 14)}$$

Reaction 4 $$C_6H_{12} + H_2 \rightarrow C_6H_{14} \quad \text{(Eq.115)}$$

$$rate = 3.33674 * 10^{19} e^{\frac{-275285.8}{8.314T}} P_{C_6H_{12}} P_{H_2} - 4.198816 * 10^{21} e^{\frac{-312237.9}{8.314T}} P_{C_6H_{12}} \quad \text{(Eq. 116)}$$

Where: Rate ≡ kmol/(m^3/hr) , P ≡ MPa ,T ≡ Kelvin

The kinetic parameters for each of the three reactors for feed K are plotted in the figures below. Additionally, a table of the reaction extents and the actual percent conversions for the feed K packed bed reactors is included below as Table 15.

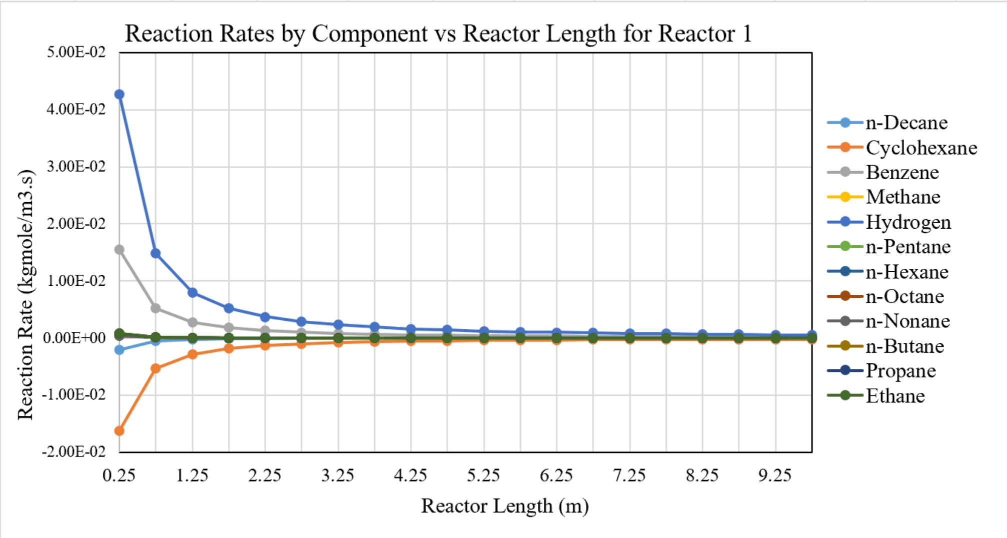


Figure 6: Reaction Rates by Component vs Reactor Length - Reactor 1

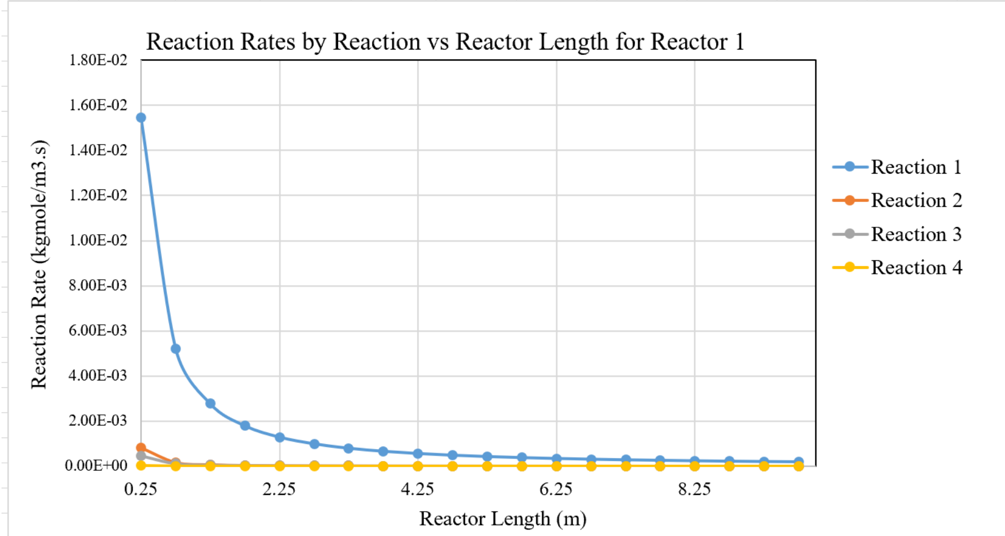


Figure 7: Reaction Rates by Reaction vs Reactor Length -Reactor 1

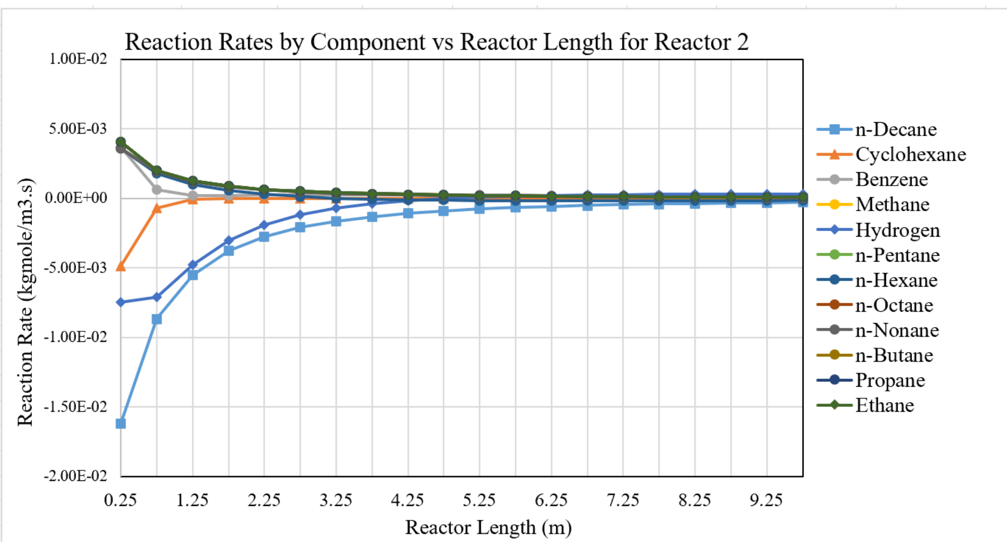


Figure 8: Reaction Rates by Component vs Reactor Length - Reactor 2

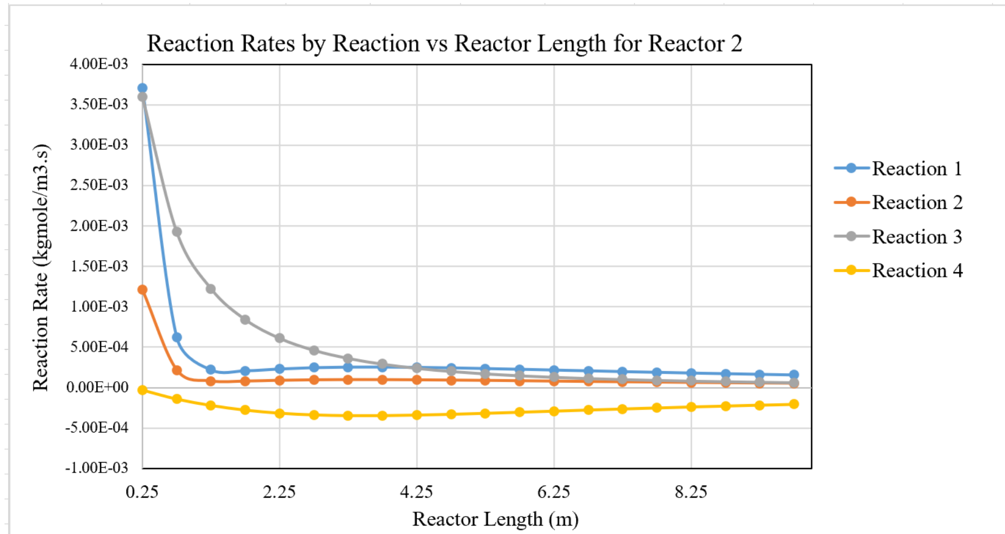


Figure 9: Reaction Rates by Reaction vs Reactor Length - Reactor 2

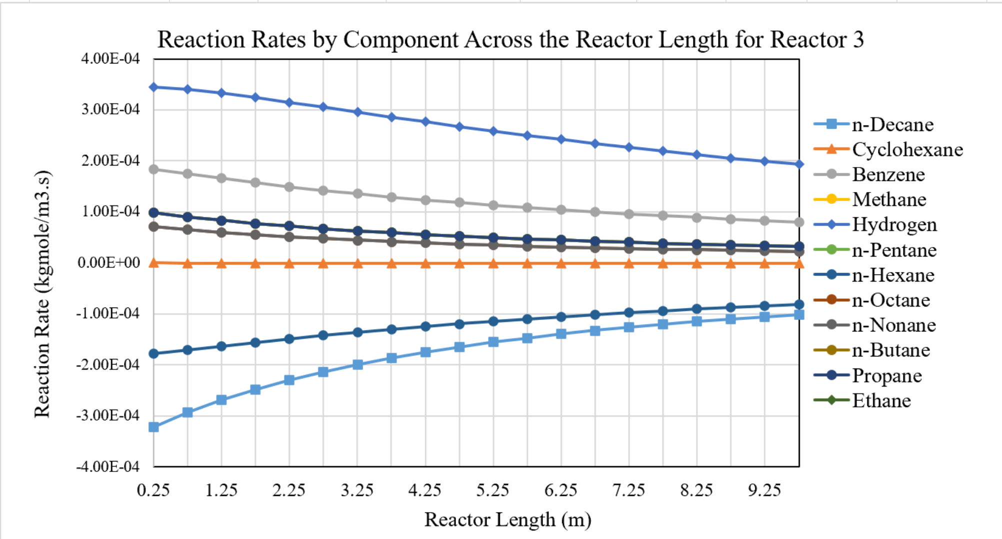


Figure 10: Reaction Rates by Component vs Reactor Length - Reactor 3

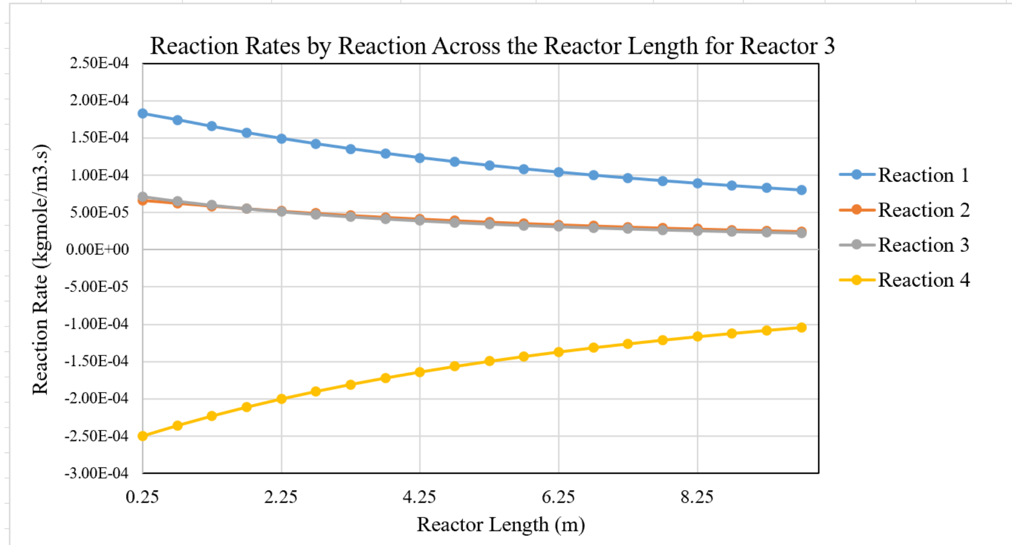


Figure 11: Reaction Rates by Reaction vs Reactor Length - Reactor 3

Table 15: Reaction Extents & Conversion Percent for Feed K PBRs

		Reactor 1		**Reactor 2**		**Reactor 3**	
Reaction	Base Component	Act. % Conv	Rxn Extent	Act. % Conv	Rxn Extent	Act. % Conv	Rxn Extent
Reaction 1	*Cyclohexane*	82.53	383.20	143.40	95.65	98.05	28.41
Reaction 2	*Cyclohexane*	2.95	13.71	49.51	33.03	33.10	9.54
Reaction 3	*n-Decane*	4.34	8.55	67.10	126.3	14.93	9.24
Reaction 4	*Cyclohexane*	0.14	0.68	-93.34	-62.27	-13.12	-3.78

Over time, the catalyst in each reactor will undergo catalyst deactivation and will need to be replaced with fresh catalyst. This brings about the need to have a swing reactor integrated into the design. This swing reactor will be able to bypass any of the three main reactors (R-100, R-101, and R-102) so that the catalyst in that reactor can be replaced without complete shutdown of the catalytic reforming process. An example of this is shown in Figure 12. Over time each of the reactors will undergo catalyst deactivation due to coking [12]. It is possible to delay the deactivation by running the reactor at a higher temperature (282.2℃), but eventually one of the reactors will need to be shut down. Then, the catalyst will need to undergo regeneration. Although regeneration has been experimentally proven to be a successful method to reverse catalyst deactivation, replacement catalyst was also factored into the economic evaluation every five years.

The price of the catalyst was calculated using the bulk density, catalyst volume, and reactor size. Additional catalyst was added for the swing reactor when costing the catalyst. Industry pricing information gave the price for platinum reforming catalyst as $51,000 per kg [13]. For all four reactors this added up to a total cost of $884 million.

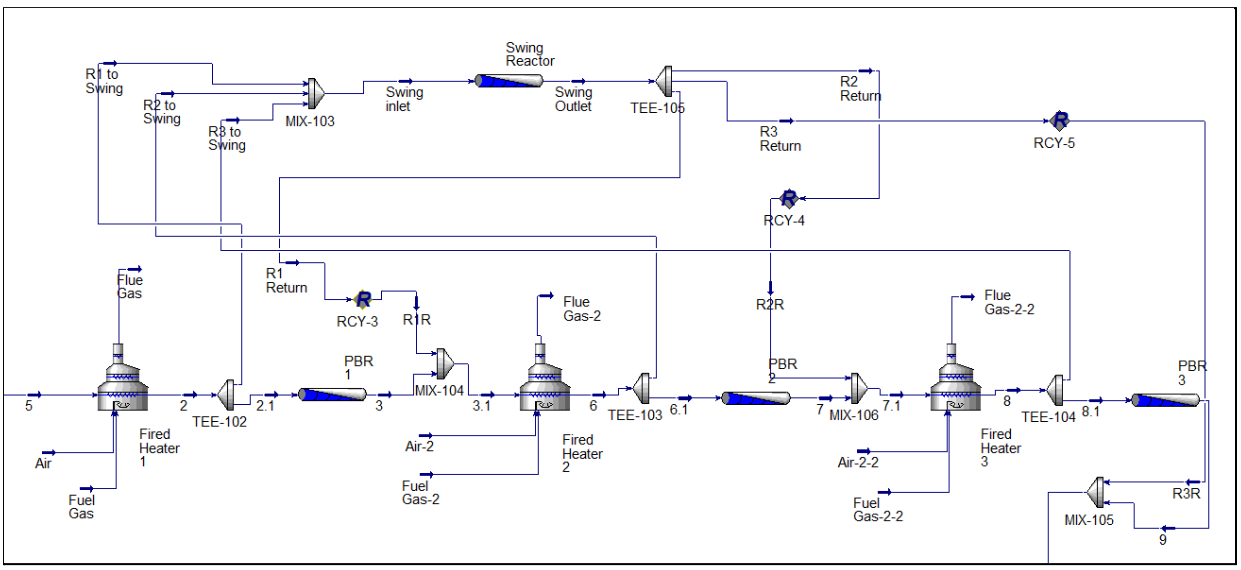


Figure 12: Aspen Simulation with Swing Reactor Integration

Hydrogen Recycle Ratio

The hydrogen recycle ratio was determined by modifying the simulation and measuring the results. Initially a fresh stream of hydrogen was used as feed in excess. This flow rate was adjusted to maximize the conversion in the reactors while also minimizing the amount of hydrogen in excess. Once the total moles of hydrogen needed as feed were determined, a recycle system was incorporated by taking the vapor stream from V-100, which has a mole fraction of hydrogen of 0.72, and returning it to the operating temperature and pressure of the original feed stream using compressor C-100. The set of four reactions in the reactors produce hydrogen as a product, which means that there is more hydrogen at the end of the reactor set than there is at the start. The amount of hydrogen flow sent back to the reactor system is equal to the original molar hydrogen feed divided by 0.72. For feed K this is equal to 1,262 kgmole/hr of 72% hydrogen feed. The excess in stream 13, is sent to another area of the plant for use or disposal.

Table 16: Equipment Constraints & Bare Module Costs - Reactor Process											
Equipment Type	**Description**	**k_1**	**k_2**	**k_3**	**Sizing Units**	**F_M**	**Material Name**	**F_{BM}**	**F_P**	**Feed K Bare Module Cost 2021**	**Feed TQ1 Bare Module Cost 2021**
C-100	Compressor	2.2897	1.3604	-0.1027	kW	1	C.S.	1	1	$340,441	$133,456
C-100 Drive	Electric, Explosion Proof	2.4604	1.4191	-0.1798	kW	1	C.S.	1.5	1	$168,303	$74,672
E-100	Floating Head; Cooling HEX; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.5	1.13	$461,755	$491,395
E-101	Floating Head; Cooling HEX; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.5	1.12	$223,363	$268,231
E-102	Floating Head; Cooling HEX; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	5.0	1.12	$273,870	$228,622
E-103	Floating Head; Reboiler; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.4	1.04	$369,467	$429,671
E-104	Floating Head; Cooling HEX; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.4	1.05	$120,977	$123,402
H-100	Non-Reactive Fired Heater	7.83488	-1.1666	0.2028	kW	1	C.S	2.1	1	$10,200,077	$9,778,242
P-100	Centrifugal Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	5.2	1.54	$867,425	$410,599
P-101	Centrifugal Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	4.5	1.21	$134,331	$134,291
R-100	Packed Bed Reactor; Jacketed, Non Agitated	3.3496	0.7235	0.0025	m^3	1	C.S	3.8	1	$241,926	$364,148
R-101	Packed Bed Reactor; Jacketed, Non Agitated	3.3496	0.7235	0.0025	m^3	1	C.S	3.8	1	$241,926	$364,148
R-102	Packed Bed Reactor; Jacketed, Non Agitated	3.3496	0.7235	0.0025	m^3	1	C.S	3.8	1	$241,926	$364,148
T-100	Reboiled Absorber	3.4974	0.4485	0.1074	m^3	1	C.S	16.8	7.97	$1,305,902	$1,344,947
V-100	Vertical Process Vessel: Knockout Drum	3.4974	0.4485	0.1074	m^3	1	C.S	5.7	1.87	$14,784	$15,103
V-101	Vertical Process Vessel: Separator	3.4974	0.4485	0.1074	m^3	1	C.S	4.1	1	$60,401	$71,981
Total	-	-	-	-	-	-	-	-	-	**15,266,900**	**14,004,300**

Table 17: Equipment Constraints 2- Reactor Process						
Equipment Type	**C_1**	**C_2**	**C_3**	**Pressure Range (barg)**	**B1**	**B2**
E-100 to E-104	0.03881	-0.11272	0.08183	5<P<140	1.63	1.66
H-100	0	0	0	P<10	N/A	N/A
P-100 to P-101	-0.3935	0.3957	-0.00226	10<P<100	1.89	1.35
T-100	N/A	N/A	N/A	N/A	2.25	1.82
V-100 to V-101	N/A	N/A	N/A	N/A	2.25	1.85

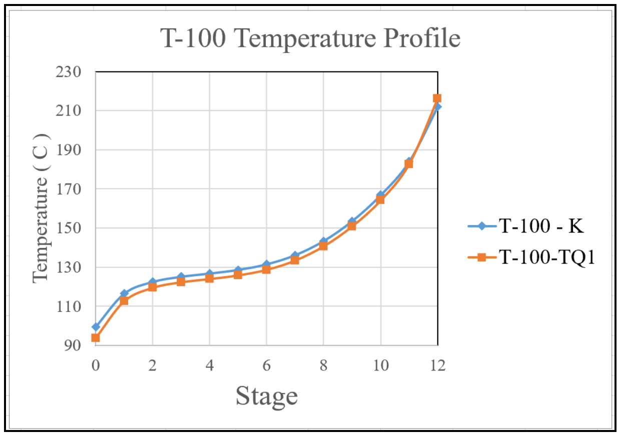


Figure 13: T-100 Temperature Profile

Table 18: Reactor Labor Cost	
Total Equipment Number	7
NOL	2.81
Total Number of Operators	13
Cost Per Hour	$7.99
Annual Labor Cost	$205,269

Table 19: Equipment Operating Costs - Reactor Process

Equipment Type	**Electricity (kW)**	**Steam (kJ/s)**	**Cooling Water (kJ/s)**	**Natural Gas (MMBTU/hr)**	**Total Operating Costs ($)**
C-100	181.25	-	-	-	$396,938
E-100	-	-	77255.96	-	$2,436,344
E-101	-	-	6853.38	-	$216,128
E-102	-	-	2964.49	-	$93,488
E-103	-	7.14	-	-	$4,356,311
E-104	-	-	6889.38	-	$217,264
H-100	-	-	-	248	$20,486,486
P-100	342.50	-	-	-	$750,075
P-101	23.60	-	-	-	$51,690
Total	**547.35**	**7.14**	**93963.20**	**248**	**29,004,724**

Appendix B: Extractor Section Detail

The liquid product exiting the naphtha catalytic reforming process is treated with a solvent to preferentially dissolve aromatic compounds. Sulfolane was chosen because it produces an aromatic stream with less than 1% non-aromatics in the extract. The non-aromatic linear alkanes and cyclic hydrocarbons are mostly retained in the raffinate [14]. This simulation uses static tray liquid-liquid extractors. These liquid-liquid extractors have higher capacities and lower cost than other styles of extractors. They also have a familiar sizing method, but they do have a lower stage efficiency [2]. There was some difficulty in the simulation of these extractors in Aspen Hysys due to some errors in the way that Hysys estimated the binary coefficients. This error was eventually resolved.

Sulfolane Recycle

There are two ways that sulfolane is recovered and recycled in this process section. A small amount of sulfolane is retained in the linear alkane product stream. This is removed from the stream in a second liquid-liquid extractor that uses water as the solvent. The water removes the sulfolane from the final diesel and gasoline product stream. This water and sulfolane stream goes to a reboiled absorber where the water and sulfolane are separated to 99% purity. This sulfolane is pumped back to the feed stream for the first liquid-liquid extractor.

The second sulfolane recycle system is used to recover the sulfolane that contains the BTX product. This sulfolane exits the first liquid-liquid extractor and is sent to distillation column T-103. The bottoms product of this distillation column is 100% pure sulfolane. 2% of the flowrate of this stream is purged in stream and sent to waste treatment. An equivalent amount of fresh sulfolane is added into the stream to bring it back to the original flow rate. For feed K, this is equivalent to 8892 kg/hr of fresh sulfolane. A flow rate of 44,600 kg/hr of sulfolane is required as the feed for the first liquid-liquid extractor in order to maintain less than 0.4% mole percent of benzene is the gasoline and diesel product stream. A summary of the equipment used for these recycle systems is included in Table 20.

Table 20: Equipment Pricing for Sulfolane Recycle System.

Equipment Type	Description	Bare Module Cost 2021 (K)	Annual Utility Cost (K)
E-108	Heat Exchanger 5	$482,883	$1,774,334
E-109	Heat Exchanger 6	$300,852	$286,745
P-103 A/B	Sulfolane Recycle Pump 1	$266,620 (Includes Spare)	$93,075
P-104 A/B	Sulfolane Recycle Pump 2	$38,613 (Includes Spare)	$2,190
Total	**-**	**$1,394,201**	**$2,156,344**

Evaluated over the course of five years, the sulfolane recovery system will cost 12.18 million dollars. Comparatively, without a recycle system, the cost of additional sulfolane will be $178,540 per hour which adds up to $7.82 billion over the course of 5 years. This results in a net savings of $7.808 billion.

As requested in the project description, the team considered steam stripping as an option for T-103 (sulfolane recovery column). A separate Hysys model was made using steam stripping and two decanters, and two additional heat exchangers to recover the sulfolane. Ultimately this method was not used in the final design due to the safety and material concerns of adding water into the system. Since water would be present in the same streams as our BTX and naphtha components, carbon steel could not be used as a material of construction without the threat of corrosion. Additionally, water draw systems had to be added to all downstream distillation columns as the water was causing phase errors to occur in the condensers. The team was able to generate similar energy consumption rates and compositions by using a regular distillation column rather than a steam stripper.

In the sample process flow diagram provided by the design prompt, a reboiled absorber is used for sulfolane recovery. However, this team decided to use a distillation column with a very small reflux ratio for the same purpose. When a reboiled absorber was originally used, a condensing heat exchanger had to be used on the vapor product, and there were not enough degrees of freedom to properly simulate the absorber to meet the design constraints. Using a distillation column to model this column resolved these issues.

Table 21: Equipment Constraints- Extractor Section											
Equipment Type	**Description**	**k_1**	**k_2**	**k_3**	**Sizing Units**	**F_M**	**Material Name**	**F_{BM}**	**F_P**	**Feed K Bare Module Cost 2021**	**Feed TQ1 Bare Module Cost 2021**
E-105	Floating Head; Reboiler; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S.	3.4	1	$707,398	$707,398
E-106	Floating Head; Condenser; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S.	3.3	1	$334,716	$286,140
E-107	Floating Head; Reboiler; 2 Shells	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$323,864	$284,081
E-108	Floating Head; Cooling HEX; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$482,883	$520,890
E-109	Floating Head; Cooling HEX; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$300,852	$314,232
E-112	Floating Head; Cooling HEX; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$291,471	$138,110
P-102	Centrifugal Reflux Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	2.3	0.21	$30,375	$30,382
P-103	Centrifugal Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	3.5	0.76	$219,244	$318,113
P-104	Centrifugal Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	3.3	0.68	$47,798	$46,424
P-106	Centrifugal Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	3.2	0.62	$112,108	$109,099
P-110	Centrifugal Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	3.3	0.66	$134,180	$133,707
T-101	Sulfolane Liquid-Liquid Extractor Horizontal Process Vessel	3.5565	0.3776	0.0905	m^3	1	C.S	1.8	0.22	$11,689	$26,022
T-102	Water Liquid-Liquid Extractor Horizontal Process Vessel	3.5565	0.3776	0.0905	m^3	1	C.S	3.0	1	$13,645	$21,544
T-103	Distillation Column	3.4974	0.4485	0.1074	m^3	1	C.S	6.0	2.07	$2,003,565	$1,484,276
T-104	Reboiled Absorber	3.4974	0.4485	0.1074	m^3	1	C.S	4.1	1	$172,674	$169,719
V-102	Horizontal Process Vessel: Reflux Drum	3.5565	0.3776	0.0905	m^3	1	C.S	3.01	1	$17,312	$24,022
Total	-	-	-	-	-	-	-	-	-	**$5,203,774**	**$4,614,159**

Table 22: Equipment Constraints 2- Extractor Section						
Equipment Type	**C_1**	**C_2**	**C_3**	**Pressure Range (barg)**	**B1**	**B2**
E-105	0.03881	-0.11272	0.08183	5<P<140	1.63	1.66
E-106 to E-109 and E-112	0	0	0	P<5	1.63	1.66
P-102 to P-104 and P-106,10	0	0	0	P<10	1.89	1.35
T-101 and T-102	N/A	N/A	N/A	N/A	1.49	1.52
T-103 and T-104	N/A	N/A	N/A	N/A	2.25	1.82
V-102	N/A	N/A	N/A	N/A	1.49	1.52

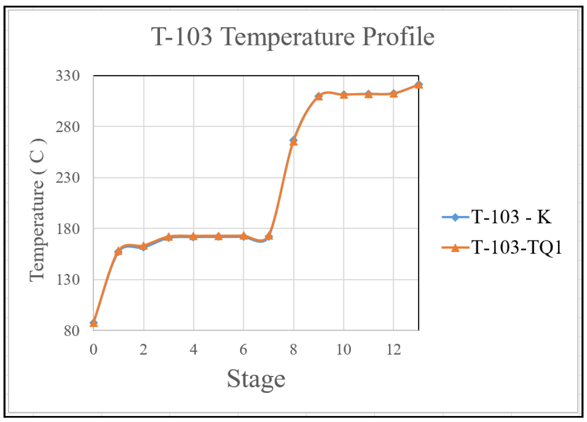


Figure 14: T-103 Temperature Profile

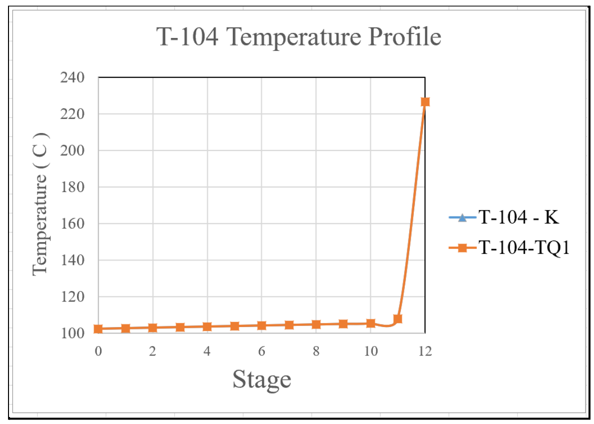


Figure 15: T-104 Temperature Profile

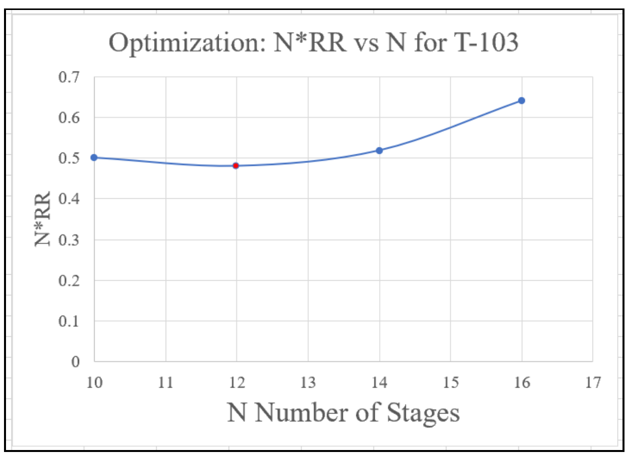


Figure 16: Column Optimization Chart for T-103

Table 23: Extractor Labor Costs	
Total Equipment Number	5
NOL	2.73
Total Number of Operators	13
Cost Per Hour	$7.99
Annual Labor Cost	$205,269

Table 24: Equipment Operating Costs - Extractor Process				
Equipment Type	**Electricity (kW)**	**Steam (kJ/s)**	**Cooling Water (kJ/s)**	**Total Operating Costs ($)**
E-105	-	7.14	-	$4,356,311
E-106	-	-	23032.57	$726,355
E-107	-	4.72	-	$2,883,012
E-108	-	-	56263.77	$1,774,334
E-109	-	-	9092.63	$286,745
E-112	-	-	5719.53	$180,371
P-102	0.63	-	-	$1,372
P-103	42.50	-	-	$93,075
P-104	1.00	-	-	$2,190
P-106	17.40	-	-	$38,106
P-110	29.57	-	-	$64,748
Total	91.10	11.86	94108.50	$10,406,619

Appendix C: Distillation Section Detail

In order to isolate the BTX to 99% purity, two distillation columns had to be used in series. The first distillation column separated the BTX from the C7+ components, and was optimized to retain the greatest amount of benzene in the vapor product. The second distillation column isolated benzene from the C5 and lighter components. This column was carefully optimized to send the smallest amount of naphtha to the recycle stream, as high recycle rates easily escalated the operating and capital costs of both the extraction and distillation sections. The most difficult separation was the isolation of benzene from the C5 and C6 components due to similar relative volatilities.

After the BTX was isolated to 99% purity, the benzene, toluene, and para-xylene were separated using two distillation columns. The first column separated the benzene from the toluene & para-xylene. The relative volatility of benzene to toluene at 1 atm is 2.5, making this separation the first that should be performed. The second column separated the toluene and para-xylene, which have an approximate relative volatility of 0.43, making this a more difficult separation. A divided wall column was simulated as another potential solution for this separation, however it was only able to separate these three components to a maximum of 65 % purity. The two column solution was able to achieve 90% purity or higher for all three components. For the detailed design considerations, the engineers may want to consider a lower BTX purity, as even small fluctuations in product purity had drastic effects on the recycle flow rate (0.5% purity change increased/decreased recycle flow rate by as much as 50%)

Table 25: Equipment Constraints & Bare Module Cost - Distillation Section [2]											
Equipment Type	**Description**	**k_1**	**k_2**	**k_3**	**Sizing Units**	**F_M**	**Material Name**	**F_{BM}**	**F_P**	**Feed K Bare Module Cost 2021**	**Feed TQ1 Bare Module Cost 2021**
E-110	Floating Head; Reboiler; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$291,471	$188,682
E-111	Floating Head; Condenser; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$432,497	$370,423
E-113	Floating Head; Reboiler; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$152,716	$159,174
E-114	Floating Head; Condenser; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$274,965	$209,373
E-115	Floating Head; Reboiler; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$323,864	$125,985
E-116	Floating Head; Condenser; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$156,441	$142,129
E-117	Floating Head; Reboiler; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$167,634	$167,634
E-118	Floating Head; Condenser; 1 Shell	4.8306	-0.8509	0.3187	m^2	1	C.S	3.3	1	$219,290	$173,319
P-105	Centrifugal Reflux Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	2.7	0.37	$45,676	$39,615
P-107	Centrifugal Reflux Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	2.6	0.35	$37,222	$35,113
P-108	Centrifugal Reflux Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	2.8	0.40	$36,489	$34,511
P-109	Centrifugal Reflux Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	2.6	0.35	$41,211	$38,760
P-111	T-105 Centrifugal Feed Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	3.4	0.71	$176,425	$143,155,149
P-112	T-107 Centrifugal Feed Pump & Spare; Electric Explosion Proof Drives	3.3892	0.0536	0.1538	kW	1.58	C.S	3.3	0.68	$69,049	$54,568
T-105	Distillation Column	3.4974	0.4485	0.1074	m^3	1	C.S	4.1	1	$466,069	$264,459
T-106	Distillation Column	3.4974	0.4485	0.1074	m^3	1	C.S	4.1	1	$348,380	$213,775
T-107	Distillation Column	3.4974	0.4485	0.1074	m^3	1	C.S	4.1	1	$225,544	$145,593
T-108	Distillation Column	3.4974	0.4485	0.1074	m^3	1	C.S	4.1	1	$260,414	$212,911
V-103	T-105 Reflux Drum	3.5565	0.3776	0.0905	m^3	1	C.S	3.0	1	$17,758	$18,196
V-104	T-106 Reflux Drum	3.5565	0.3776	0.0905	m^3	1	C.S	3.0	1	$20,299	$11,471
V-105	T-107 Reflux Drum	3.5565	0.3776	0.0905	m^3	1	C.S	3.0	1	$15,481	$16,603
V-106	T-108 Reflux Drum	3.5565	0.3776	0.0905	m^3	1	C.S	3.0	1	$18,679	$11,453
Total	-	-	-	-	-	-	-	-	-	**$3,797,574**	**$145,788,896**

Table 26: Equipment Constraints 2- Distillation Section [2]						
Equipment Type	**C_1**	**C_2**	**C_3**	**Pressure Range (barg)**	**B1**	**B2**
E-110, 11,13, 14, 15, 16, 17, 18	0	0	0	P<5	1.63	1.66
P-105, 7, 8, 9, 11, 12	0	0	0	P<10	1.89	1.35
T-105 to T-108	N/A	N/A	N/A	N/A	2.25	1.82
V-103 to V-106	N/A	N/A	N/A	N/A	1.49	1.52

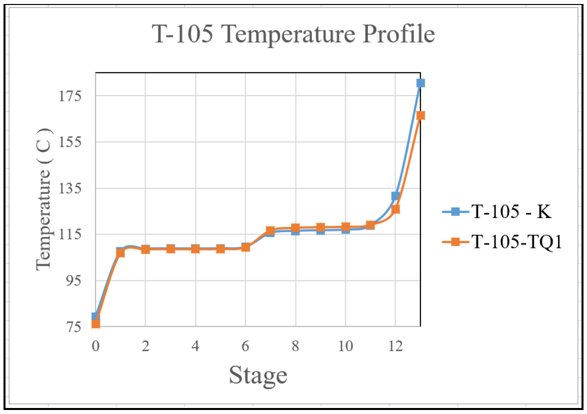


Figure 17: T-105 Temperature Profile

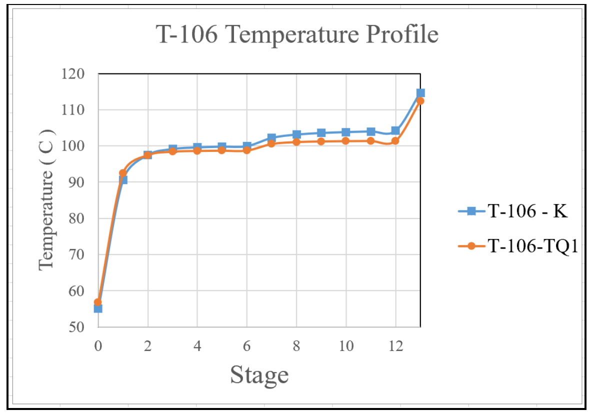


Figure 18: T-106 Temperature Profile

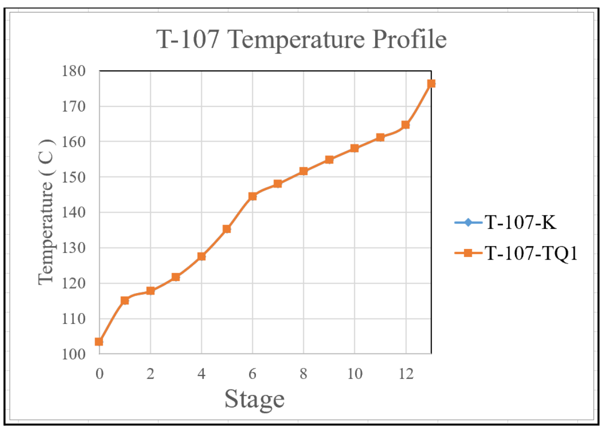


Figure 19: T-107 Temperature Profile

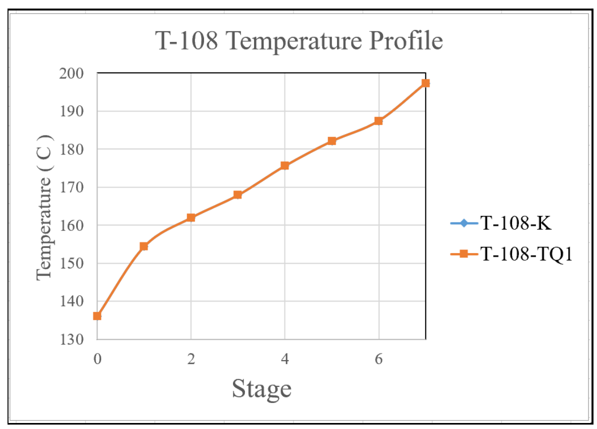


Figure 20: T-108 Temperature Profile

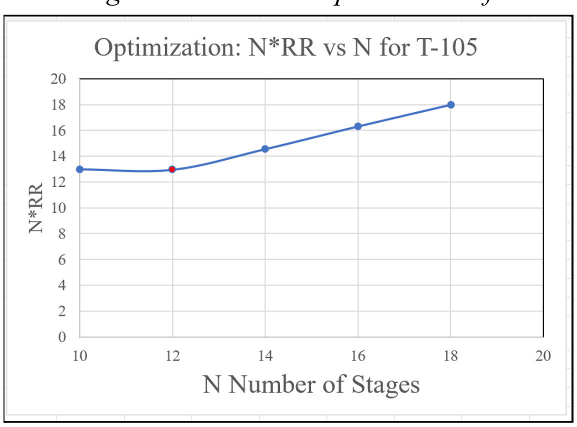


Figure 21: T-105 Optimization Profile

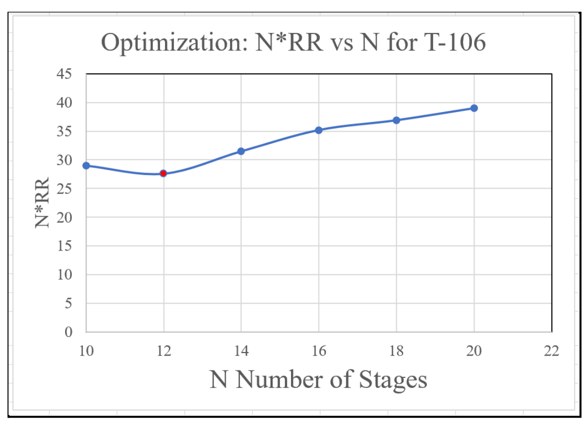


Figure 22: T-106 Optimization Profile

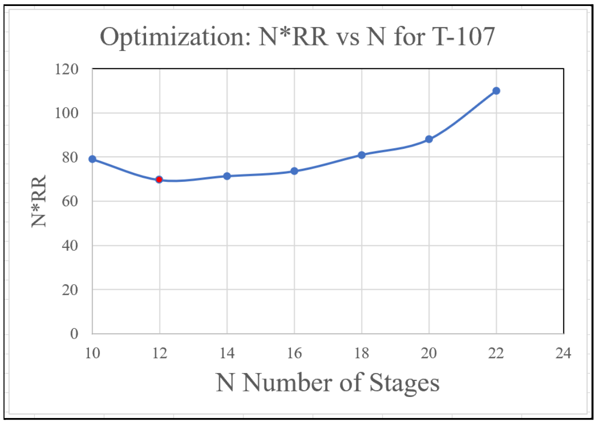


Figure 23: T-107 Optimization Profile

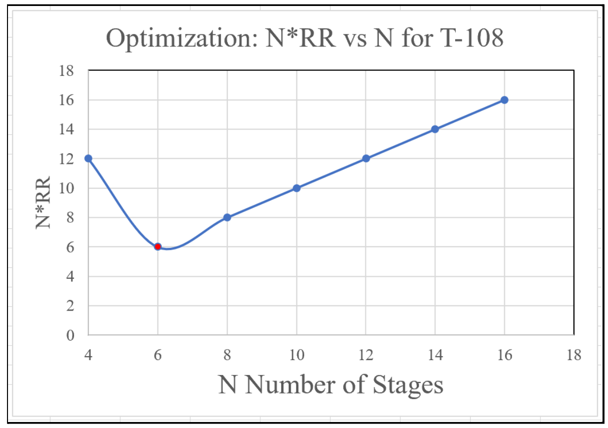


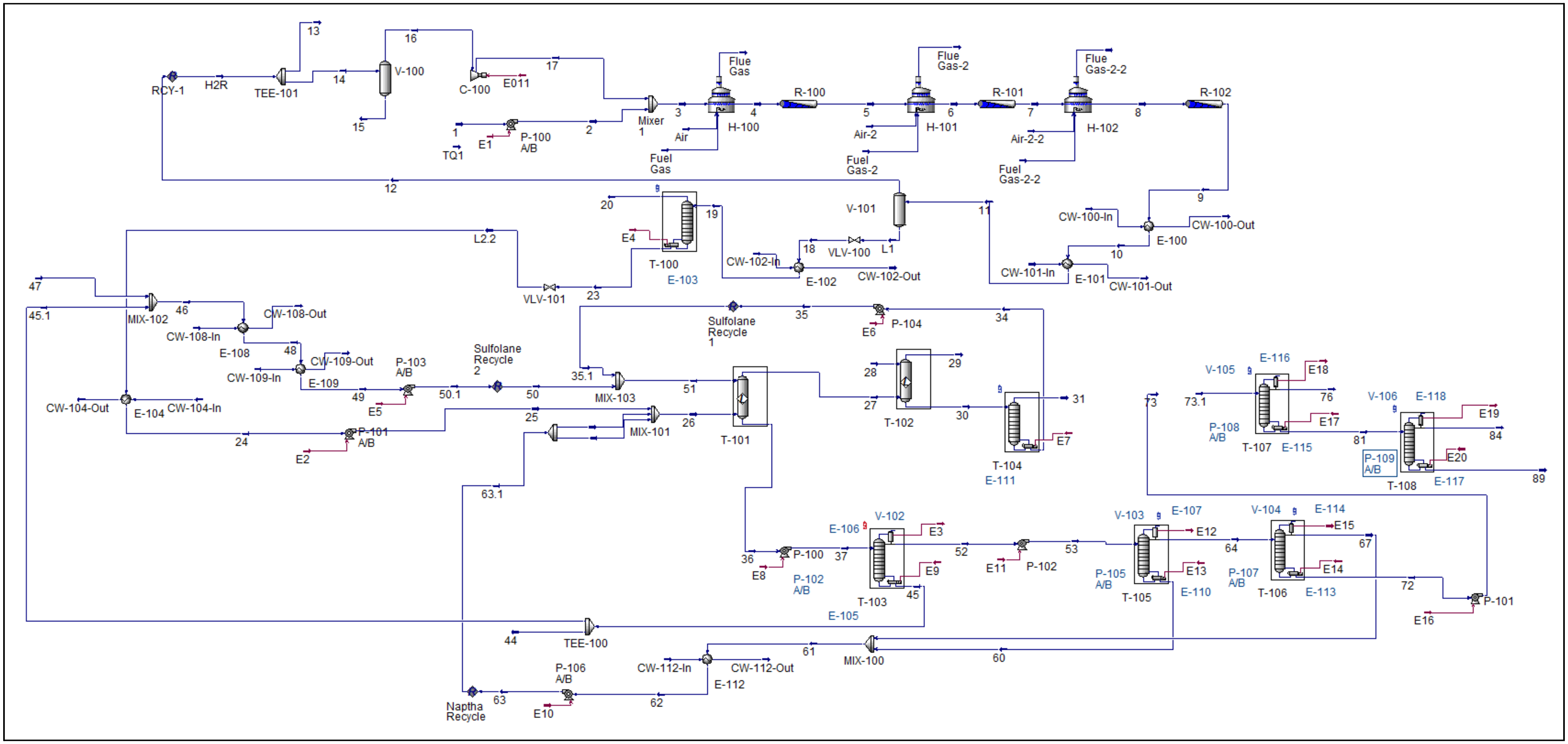
Figure 24: T-108 Optimization Profile

Table 27: Distillation Labor Cost Estimate	
Total Equipment Number	4
NOL	2.685
Total Number of Operators	13
Cost Per Hour	$7.99
Annual Labor Cost	$205,269

Table 28: Equipment Operating Costs - Distillation Process

Equipment Type	Electricity (kW)	Steam (kJ/s)	Cooling Water (kJ/s)	Total Operating Costs ($)
E-110	-	9.04	-	$5,517,073
E-111	-	-	20056.90	$632,514
E-113	-	2.75	-	$762,683
E-114	-		11750.34	$370,559
E-115	-	3.42	-	$2,087,867
E-116	-		5450.30	$171,881
E-117	-	2.42	-	$1,477,445
E-118	-	-	11734.36	$370,055
P-105	3.56	-	-	$7,804
P-107	3.33	-	-	$7,283
P-108	2.93	-	-	$6,418
P-109	3.60	-	-	$7,883
P-111	43.81	-	-	$95,952
P-112	9.54	-	-	$20,902
Total	66.78	17.62	48,991.90	**11,536,319**

Appendix D: HYSYS Flowsheet



HYSYS Mass & Energy Balances

Table 29: Aspen Hysys Mass Balances			
Inlet Materials Streams	**Mass Flow (kg/hr)**	**Outlet Materials Streams**	**Mass Flow (kg/hr)**
1	175,000	Flue Gas	254,061
Air	243,731	Flue Gas-2	58,419
Fuel Gas	10,330	Flue Gas-2-2	3,691
Air-2	52,002	20	17,272
Fuel Gas-2	6,417	29	97,125
Air-2-2	2,889	31	17,766
Fuel Gas-2-2	802	31.1	18
28	19,817	Waste Water2	1
73.1	59,701	Waste Water3	0.00
47	8,891	44	8,892
CW-100-In	162,136	76	5,342
CW-112-In	90,076	84	47,931
CW-104-In	10,809	89	6,428
CW-102-In	4,342	13	12,107
CW-101-In	162,136	15	0.00
CW-108-In	162,136	CW-100-Out	162,136
CW-109-In	180,151	CW-112-Out	90,076
		CW-104-Out	10,809
		CW-102-Out	4,342
		CW-101-Out	162,136
		CW-108-Out	162,136
		CW-109-Out	180,151
		73	48,969
Total:	1,351,866	**Total:**	1,351,809
Imbalance	-57.15 kg/hr		
Relative Imbalance	0.00%		

Table 30: Aspen Hysys Energy Balances			
Inlet Streams	**Energy Flow (kcal/hr)**	**Outlet Streams**	**Energy Flow (kcal/hr)**
1	-75,100,322	Flue Gas	-81,249,958
E1	236,125	Flue Gas-2	-26,704,386
Air	0	Flue Gas-2-2	-1,560,182
Fuel Gas	-11,526,794	20	-8,358,383
Air-2	0	29	-49,280,267
Fuel Gas-2	-7,160,620	31	-62,507,274
Air-2-2	0	31.1	-66,794
Fuel Gas-2-2	-895,077	E15	5,760,017
E4	17,207,372	Waste Water2	-4,636
28	-74,762,163	E12	14,510,362
E7	11,387,997	Waste Water3	0
E2	14,770	44	-6,752,794
E6	34	E18	7,978,017
E14	6,364,532	76	620,745
E13	18,714,617	E19	12,848,116
E10	12,610	84	2,427,528
E17	10,055,680	89	210,581
73.1	2,116,831	13	-6,259,925
E20	11,911,748	15	0
E011	194,178	CW-100-Out	-546,3665,409
47	-7,841,108	CW-112-Out	-335,859,293
CW-100-In	-612,838,028	CW-104-Out	-34,928,103
CW-112-In	-340,465,571	CW-102-Out	-13,859,925
CW-104-In	-40,855,896	CW-101-Out	-606,941,235
CW-102-In	-16,410,630	CW-108-Out	-564,430,572
CW-101-In	-612,838,028	CW-109-Out	-673,108,135
CW-108-In	-612,838,028	E3	17,709,845
CW-109-In	-680,931,143	73	8,473,106
E5	56,167		
E9	68,708,362		
E8	17,813		
E16	6,112		
E11	24,630		
Total:	(2,947,433,804)	**Total:**	(2,947,428,953)
Imbalance	4851 kcal/hr		
Relative Imbalance	0.00%		

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