

Letter of Transmittal

To Whom It May Concern,

Our team has submitted a final version of the preliminary design for the naphtha catalytic reforming plant, in Kirkuk Iraq.

Product specification of 99% Benzene was met as requested. The design uses advanced methods of safety and process optimization. The goal was to minimize the environmental footprint and decrease the present worth cost of the plant.

Economic analysis was conducted over a 30 year project life under both Iraqi and Kurdish control. Based on the economic results and our initial safety evaluation our team recommends to continue development of the project.

The catalytic reforming plant will require an initial investment of \$31 million. Annual operating costs are projected to be \$1.2 million. The payback period is 1.5 years under Iraqi control and 1.4 years under Kurdsih control. The DCFRORs are 142% and 165% for each tax regime. The price of raw materials has the largest effect on the attractiveness of the project. It is recommended to lock in long term contracts for the feedstocks to ensure economic attractiveness of the plant.

Please let our team know if you have any questions.

Best Regards

AICHE National Student Design Competition Spring 2021

Naphtha Catalytic Reforming Process

April 16, 2021

Group: 1738

Executive Summary

Iraqi Kurdistan is a large exporter of oil yet imports most of its refined fuels ^[1]. Although numerous toppings refineries have emerged in the area due to the lack of local refined fuels, the government of Kurdistan is cracking down on haphazard plants in favor of safer, high output alternatives. In an attempt to capitalize on the recent attitude changes, our team designed a process which takes advantage of an established toppings refinery owned by our client, Mr. Abbasi. Currently, his refinery produces sweet Kiruk crudes which are high in benzene and low in sulfur. We recommend a process which purifies the refinery output to create salable benzene, toluene, xylenes, and various alkanes all while abiding by western standards of operational and environmental loss prevention.

Preliminary design analysis was conducted involving numerous potentially hazardous physical and chemical interactions throughout the process. Many configurations of reactors, heat exchangers, and separation equipment were tested to produce the desired output of aromatics. We decided upon the design which most complied with safety requirements and environmental constraints specified by OSHA and the EPA, respectively. The toppings refinery amendment consists of an upstream reaction train in which reaction streams loop through a fired heater. The downstream section is composed of liquid-liquid extractors, and brief separation before the 99% benzene stream is separated into salable products by distillation.

Upfront capital costs for development of the toppings refinery amendment in Kurdistan is estimated to total \$30.5 million. The overall process consists of 10 heat exchangers, 10 process vessels, 17 pumps, 1 fired heater, 3 tubular reactors, and associated piping and instrumentation. In total the amendment will require 15 operators whose combined salaries will total \$190 thousand annually. There were two cost analyses conducted considering either the Iraqi or the Kurdish tax regimes at 35% or 15%, respectively. Utilities required include saturated steam, cooling water, fuel gas, and electricity which have an estimated combined annual usage cost of \$1.2 million. The project has an annual revenue of \$148 million under Iraqi taxes or \$193 million under Kurdish taxes. An economic analysis life of 30 years was conducted using MACRS depreciation for all equipment. The resulting net present value (NPV) of the project is \$757 million under Iraqi taxes or \$998 million under Kurdish taxes with a discounted cash flow rate of return (DCFROR) equal to 142% or 165% and payback period of 1.5 or 1.4 years under Iraqi or Kurdish tax regimes, respectively. The hurdle rate for this project was 15% and under either tax regime the project is economically favorable. The economics were then subjected to a single variable sensitivity analysis which showed that the largest factor affecting the profit was the raw material cost. Our team recommends that this project be moved into the detailed design phase as it will bring benefit to the engineering firm, Mr. Abbasi, and the local Iraqi Kurdistan region.

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Introduction

The EPC firm is interested in enhancing a small toppings refinery operated by Mr. Abbasi, a local oil COO in Kurdistan, Iraq. As a region that typically imports refined petroleum products, fabricating these products locally could potentially reduce the region's \$3BB annual costs. Politically speaking, there is an urgency to make products that are low in benzene as this byproduct may cause damage to the population and/or the environment. An economic analysis shows promise of installing an amended benzene-enriching process to the toppings refinery already present in Kurdistan. This process would allow the current products of the refinery to remain at their normal capacity while producing a side draw of high-purity benzene, toluene, and xylenes from a naphtha feed. Understanding the local government's restrictions regarding unsafe conditions and products, we pursue a process that is up to code in western safety standards and that minimizes excess machinery or hazardous processing conditions. The additional equipment will be fed crude oil naphtha from the existing refinery. The feed of interest for this investment is denoted "Feed K" which contains 20% naphtha by volume. The feed will be subjected to heat, reaction(s), and various separation units in order to produce fuel gas, gasoline (C5-C8), diesel (>C8), benzene, toluene, and xylenes. The fuel gas may be used to fuel the furnace within the process stream, or it can be piped elsewhere in the plant as a source of readily available energy. The benzene, toluene, and xylenes can all be sold as high purity products. Our goal is to maximize the output of high purity products without compromising safety. This strategy will ensure the economic viability and the longevity of the oil company due to new safety restrictions imposed by the local government. The feed in question will first be pumped to a furnace for preheating before subjection to cracking and dehydrogenation. This will then be sent to a separator of which the tops will be recycled to remove cyclohexane impurities while the bottoms are sent downstream for further separation. The bottoms product is then subjected to various liquid-liquid extractors with water and sulfolane to remove heavy alkanes from the target product. The product stream is then subjected to stripping followed by distillation to remove light alkanes. This is then flashed to remove any remaining heavy alkanes to produce a high purity benzene product hereby known as the "reformate" stream. This stream is sent for further processing to produce toluene and xylenes for a variety of salable products.

Process Description

The process begins with a feed that is mixed with a recycle stream and fed into a furnace to be heated before entering the first reactor. Next the product is fed back into the furnace and through another reactor. Finally that product goes through the furnace one last time before entering the third and final reactor. A heat exchanger cools down the product before going through a separator to produce a recycle stream in the tops and the bottoms are fed into the first extractor. Waste comes out the top and the bottoms are mixed with another recycle stream to be fed into the second extractor. The tops of the second extractor go into a third extractor to produce alkanes while the bottoms are fed into the fourth extractor. The bottoms of the fourth extractor are mixed with a sulfolane stream to be used as the top feed of the second extractor. The tops of the fourth extractor are fed into the first distillation column where the top product is used as a recycle and the bottoms are fed into a component splitter. The bottoms of the component splitter produce alkanes while the tops go into the second distillation column where benzene comes out the tops as a product. The bottoms are fed into a third distillation column to produce end products of toluene in the tops and xylene in the bottoms.

Design Basis

Because the Kurdistan oil company will be extracting the naphtha from their current stock, we can expect to generate quantities of revenue from the brand new product streams. Considering that naphtha constitutes a minimum of 20% of the refinery's streams by volume, we should also expect a reduction in sales from the current production lines. Therefore, in order to

make the entire process feasible we must maximize the outflow of reformat. The following design satisfies these constraints while maintaining safe operating limits. Market analysis confirms that the proposed design is profitable with high purity benzene, toluene, and xylenes generating a brand new profit stream. The feed of interest for this design is designated “Feed K” which is composed of 31.3% (molar basis) cyclohexane. Further feed specifications are listed below in Table 1. The feed’s naphtha will enter at a flow rate of 7,000 barrels per day, at a pressure of 1.2 bara, and a temperature of 70 °C (158.0 °F).

		crude oil	
		TQ1	K
naphtha proxy components	naphtha % volume of crude	28	20
	specific gravity	0.7308	0.749
	n-decane mol%	77.8	59.7
	cyclohexane mol%	20.6	31.3
	benzene mol%	1.6	9

Table 1: Feed Composition(s)

Cyclohexane’s normal boiling point is only 1.2 °F higher than benzene’s normal boiling point. Therefore, these two chemicals will not separate easily based on vapor pressure. In order to meet the specification of 99%+ product purity, we must maximize reactions of which cyclohexane is the key reactant. Through rigorous modeling, it was determined that a furnace temperature of 1750 °F provided sufficient temperatures to crack decane and to dehydrogenation cyclohexane. The feed is sent first through the furnace before entering the first packed-bed reactor (PBR) which is pre-installed with a platinum catalyst. The stream then cycles back to the furnace before entering a second PBR, as well as a third time to maximize reaction extents. The final reacted stream is then cooled before entering separator V-100 (Figure 1) which recycles the vapors and continues the liquids downstream. The recycled vapors are mixed with the feed pre-furnace in order to reduce waste. Now saturated with both light and heavy alkanes, the bottoms of the first separator are sent to a reboiled column (T-100) where light alkanes are separated and sent elsewhere in the facility for use as fuel gas.

Heavy alkanes are removed from the target stream through use of liquid-liquid extraction in extractors T-101 and T-102. Sulfolane is used to first extract the heavy alkanes from the product stream in T-101, and water is used to extract any remaining heavy liquids in T-102 and separate alkanes to be sold as product gas or diesel. The bottoms of T-101 are sent to a reboiled stripper (T-104) for separation of sulfolane and product benzene. The sulfolane is recycled back to extractor T-101 to be reused, and the tops of T-104 are cooled and sent to distillation column T-105 (Figure 1) for further purification. The tops of T-105 contain light alkanes and benzene which are recycled to the beginning of extraction to reclaim the salable product. The bottoms of T-105 are directly flashed in tank X-100 which produces product Benzene and alkanes. This process allows us to theoretically harvest 99.5% benzene from the naphtha feed as well as salable benzene. These specifications allow us to obtain profit from these streams at the following values in Table 2. Benzene was used to model toluene and xylenes. The correct split of product flows was calculated using the following mass balance equations listed in Table 3.

Financial Information	
linear alkane (diesel) [\$/L]	0.98
cyclic hydrocarbons (gasoline) [\$/L]	0.63
local naphtha [\$/L]	0.325
Sulfolane [\$/kg]	5
Benzene [\$/gal]	3.49
gas sale applicable	C5-C8
diesel sale applicable	>C8
consumed in facility (no credit)	<C5

Table 2: Product Economic Variables

$$x_{reformat,benzene} = \frac{\dot{n}_{naphtha}x_{naphtha\ benzene} + (\dot{n}_{reformat} - \dot{n}_{naphtha}x_{naphtha\ benzene}) * 0.11}{\dot{n}_{reformat}}$$

$$x_{reformat,toluene} = \frac{(\dot{n}_{reformat} - \dot{n}_{naphtha}x_{naphtha\ benzene}) * 0.55}{\dot{n}_{reformat}}$$

$$x_{reformat,para-xylene} = \frac{(\dot{n}_{reformat} - \dot{n}_{naphtha}x_{naphtha\ benzene}) * 0.34}{\dot{n}_{reformat}}$$

Table 3: Product Composition Splits According to Reformate Stream Data

This process involves using utilities for various cooling, heating, and electrical needs. The operating costs associated with the above setup were calculated using the following values specified by the economic team in Table 4.

	utilities
electricity	\$0.25 kWhr
steam, 450 psig	\$19.36/1000 kg
steam, 150 psig	\$14.08/1000 kg
steam, 50 psig	\$8.8/1000 kg
natural gas	\$9.43/MMBTU
cooling water, 25° C	\$0.5/GJ

Table 4: Utility Economic Variables

These specifications were considered in our modeling, and our cost calculations. These limits are established by either environmental, governmental, or material constraints. In order to minimize hot zones, we used the smallest viable pressure steam possible for sufficient heat transfer in the column reboilers.

All designs within the report were modeled using Aspen HYSYS simulation software. The NRTL-Ideal fluid package was used for the extraction section of the process, and Peng-Robinson was used everywhere else in the simulation where vapors were modeled. This

was done at the request of management, but more thorough investigation will need to be done for a more accurate representation of the process.

Design Philosophy

Complex design requires frequent use of assumptions to enable process modeling. Physical constraints, chemical interactions, and consequence reduction strategies were pertinent to all design decisions. All major assumptions and constraints used throughout the process are listed below organized by associated process vessels.

Pumps

- Pump Efficiency set at 70%
- 10 foot difference in height between reflux drums and reflux pumps
- Centrifugal pumps were used throughout the process design
- Electric drives were used for each pump

Furnace

- Complete combustion assumed for design purposes
- Fuel gas and air fed at 1:10 methane:oxygen molar ratio
- Radiant heat transfer to all process streams

Reaction Train

- Autoclave packed bed reactors
- Maximum temperature of 600 °C to optimize cyclohexane dehydrogenation
- Platinum catalyst used with void fraction of 70% and catalyst fraction of 30%
- Cooling water supplied at 25 °C
- Steam available at 50, 150, and 450 psig

Heat Exchangers

- Floating head shell and tube format used for all heat exchangers for efficiency
- Counter flow orientation used to maximize heat transfer
- Thin walled assumption used ($d_o=d_i$)

Columns and 2-Phase Separators

- 4 feet spacing between vessel tops and top trays to allow proper vapor disengagement
- 6 feet spacing between vessel bottoms and bottom trays to account sufficient space for reboiler return
- Vertical orientation used to reduce floor area usage and support long term maintenance
- Maximum height set to 83 ft to minimize foundational issues
- Trayed vessels have 2 foot separations between trays to promote separation
- Demister added to all separators
- L/D ratio sought below 5

Extractors

- Horizontal orientations used for liquid-liquid extraction (LLE)
- NRTL fluid package used to model separations for LLE

Miscellaneous

- Carbon steel material of construction due to low sulfur content
- Service factor set to 0.92 for economic analysis
- 99% purity required for salable benzene pre-distillation section
- Wash-out assumption used for all cash flow calculations
- Economic analysis was done over a 30-year period with an effective tax rate of 15% (Kurdish control) or 35% (Iraqi control)

P-101 A/B
Furnace
Feed Pump

F-100
Furnace

R-101
First
Reactor

R-102
Second
Reactor

R-103
Third
Reactor

E-101
Heat Exchanger
To Separator

V-101
Separator

T-101
First
Extractor Column

E-102
First Extractor
Reboiler

P-102 A/B
Pump To
First Extractor

P-103 A/B
Pump To
Second Extractor

Process ID Key

- Temperature: °C
- Pressure: kPa
- Flow Rate: kgmol/hr

Stream Data:

Stream ID	Flow Rate (kgmol/hr)	Temperature (°C)
1	70	120
2	70	1440
3	35.99	116.2
4	241.2	116.2
5	449.6	116.2
6	449.6	116.2
7	599.1	116.2
8	605.9	116.2
9	22.56	116.2
10	22.56	116.2
11	22.56	116.2
12	22.56	116.2
13	20.22	47.26
14	30.22	47.26
15	59.5	620.5
16	118.3	758.4
17	150.8	758.4
18	150.8	758.4
19	150.8	758.4

Figure 1: Reactor PFD

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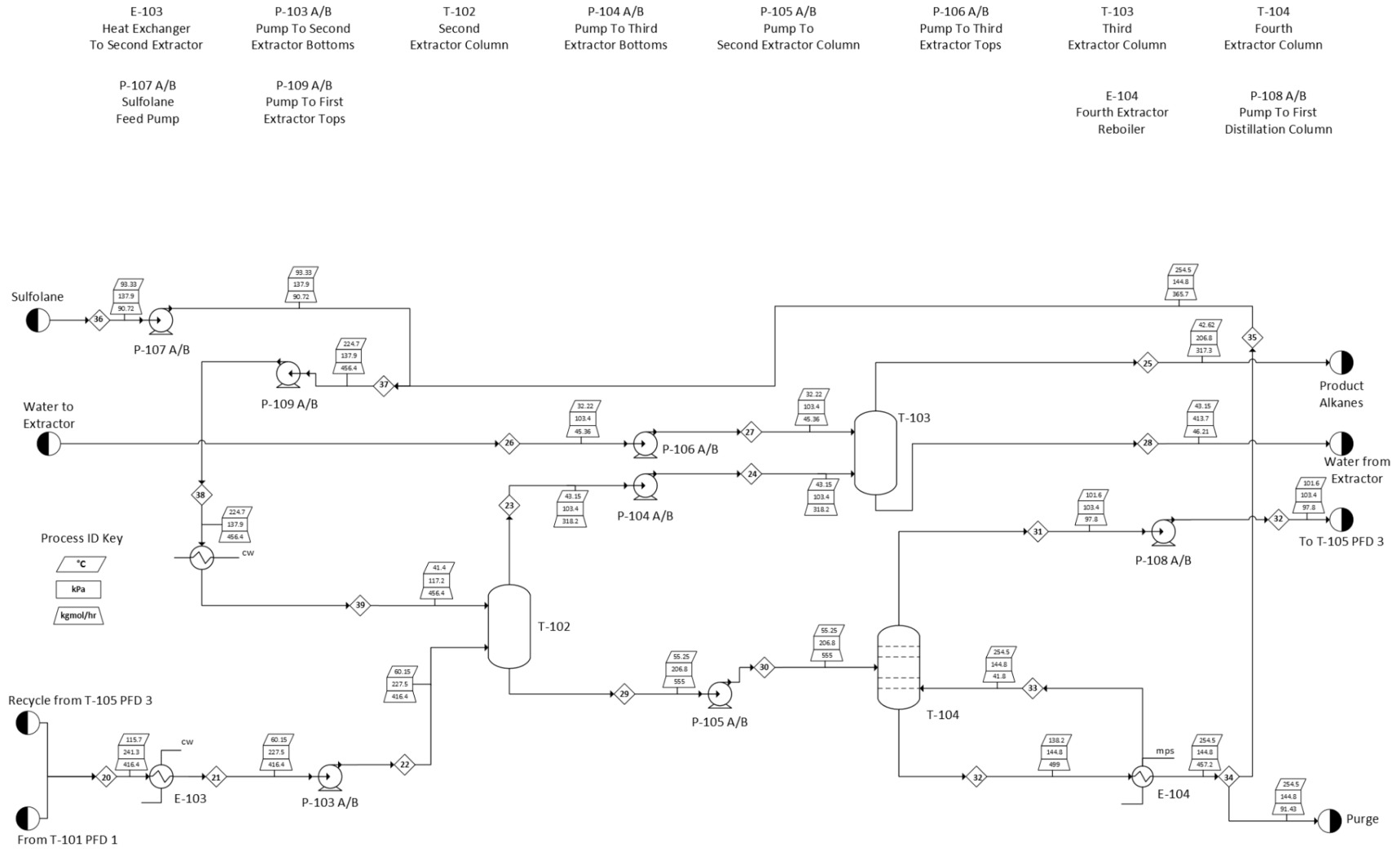


Figure 2: Extraction PFD

- | | | | | | | | | | |
|---|---|--|---|---|--|---|--|---|---|
| E-105
Heat Exchanger To
First Distillation Column | P-110 A/B
Pump To First
Distillation Column | T-105
First Distillation
Column | E-106
First Distillation
Column Condenser | P-111 A/B
Pump To Second
Extractor Column | T-106
Second Distillation
Column | E-108
Second Distillation
Column Condenser | P-116 A/B
Pump for Second
Distillation Tops | P-117 A/B
Pump To Third
Distillation Column | E-110
Third Distillation
Condenser |
| E-107
First Distillation
Reboiler | P-113 A/B
Pump To
Component Splitter | V-102
First Distillation
Condensate Receiver | P-112 A/B
First Distillation
Reflux Pump | V-103
Component
Splitter | P-114 A/B
Pump To Second
Distillation Column | V-104
Second Distillation
Condensate Receiver | P-115 A/B
Second Distillation
Reflux Pump | E-109
Second Distillation
Reboiler | T-107
Third Distillation
Column |
| | | | | | E-111
Third Distillation
Reboiler | P-118 A/B
Third Distillation
Reflux Pump | V-105
Third Distillation
Condensate Receiver | P-119 A/B
Pump For Third
Distillation Tops | P-120 A/B
Pump For Third
Distillation Bottoms |

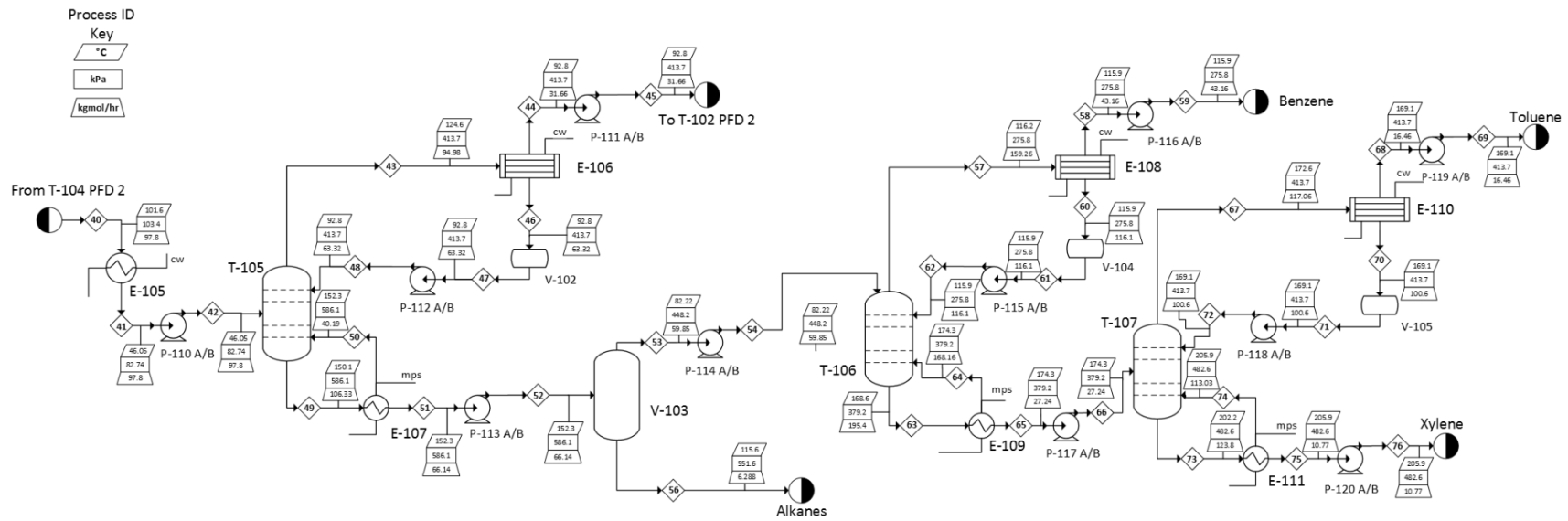


Figure 3: Distillation PFD

Table 5: Stream and Equipment Summary

Stream Number		1	2	3	4	5	6	7	8	9	10	11	12	13
Phase Fraction		0.0	0.9	1.0	1.0	1.0	1.0	1.0	1.0	0.9	1.0	1.0	0.0	0.0
Pressure	kPa	120.0	116.2	116.2	116.2	116.2	116.2	116.2	116.2	116.2	116.2	116.2	116.2	47.3
Temperature	C	70.0	36.0	241.2	241.2	449.6	449.4	599.1	605.9	22.6	22.6	22.6	22.6	20.2
Molar Enthalpy	kJ/kgmole	-211241.2	-84359.4	-61605.5	-61605.6	-33659.7	-33647.1	-11925.8	-11545.7	-81890.7	-70645.4	-70645.4	-162711.2	-162711.2
Mass flow rate	kg/h	34814.6	121862.0	121862.0	121862.0	129432.5	129432.5	129432.5	129432.6	129432.6	91632.9	87047.4	37799.8	37799.8
Molar flow rate	kgmole/h	294.3	3013.4	3013.4	3013.4	3155.5	3156.7	3156.7	3260.6	3260.6	2862.3	2719.2	398.3	398.3
Volumetric Flow rate	m ³ /h	49.0	59658.4	110610.1	110610.1	163091.7	163108.2	197006.3	205085.1	60020.4	59967.6	56967.3	52.8	515.2
Actual Density	kg/m ³	710.6	2.0	1.1	1.1	0.8	0.8	0.7	0.6	2.2	1.5	1.5	716.4	73.4
Std Ideal Lig Vol Flow	m ³ /h	46.4	238.4	238.4	238.4	249.4	249.4	249.4	255.2	255.2	202.1	192.0	53.1	53.1
Component Mole Flow rate														
Cyclohexane	kgmole/h	92.1	92.5	92.5	92.5	92.1	91.7	91.7	0.9	0.9	0.4	0.4	0.5	0.5
Benzene	kgmole/h	26.5	113.2	113.2	113.2	130.0	130.4	130.4	179.1	179.1	91.3	86.8	87.7	87.7
Hydrogen	kgmole/h	0.0	699.4	699.4	699.4	846.7	847.9	847.9	736.3	736.3	736.2	699.4	0.1	0.1
n-Pentane	kgmole/h	0.0	203.8	203.8	203.8	210.7	210.7	210.7	266.8	266.8	214.5	203.8	52.3	52.3
o-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
m-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
p-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Toluene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Hexane	kgmole/h	0.0	17.5	17.5	17.5	0.0	0.0	0.0	33.8	33.8	18.4	17.5	15.4	15.4
n-Heptane	kgmole/h	0.0	20.1	20.1	20.1	42.4	42.4	42.4	80.0	80.0	21.1	20.1	58.9	58.9
n-Octane	kgmole/h	0.0	7.4	7.4	7.4	38.6	38.6	38.6	76.3	76.3	7.8	7.4	68.5	68.5
n-Nonane	kgmole/h	0.0	2.6	2.6	2.6	37.6	37.6	37.6	75.2	75.2	2.7	2.6	72.5	72.5
n-Decane	kgmole/h	175.7	175.7	175.7	175.7	175.7	175.6	175.6	6.2	6.2	0.1	0.1	6.1	6.1
Methane	kgmole/h	0.0	465.2	465.2	465.2	434.0	434.0	434.0	490.0	490.0	489.7	465.2	0.4	0.4
Ethane	kgmole/h	0.0	454.9	454.9	454.9	425.1	425.1	425.1	481.1	481.1	478.8	454.9	2.3	2.3
Propane	kgmole/h	0.0	425.0	425.0	425.0	399.5	399.5	399.5	455.6	455.6	447.3	425.0	8.2	8.2
n-Butane	kgmole/h	0.0	336.2	336.2	336.2	323.2	323.2	323.2	379.3	379.3	353.9	336.2	25.3	25.3
Oxygen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFOLANE	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Stream Number		14	15	16	17	18	19	20	21	22	23	24	25	26
Phase Fraction		0.0	1.0	0.0	0.0	0.0	0.3	0.3	0.0	0.0	0.0	0.0	0.0	0.0
Pressure	kPa	47.3	620.5	758.4	758.4	758.4	241.3	241.3	227.5	227.5	103.4	103.4	206.8	103.4
Temperature	C	20.2	59.5	150.8	150.8	150.8	118.8	115.7	60.1	60.1	43.1	43.1	42.6	32.2
Molar Enthalpy	kJ/kgmole	-162711.2	-99461.4	-134848.7	-134848.7	-134848.7	-134848.7	-124999.2	-145604.7	-145604.7	-200955.8	-200955.8	-200187.7	-284352.4
Mass flow rate	kg/h	37799.8	601.4	37198.3	37198.3	37198.3	37198.3	39567.2	39567.2	39567.2	31813.5	31813.5	31667.4	817.2
Molar flow rate	kgmole/h	398.3	13.4	384.9	384.9	384.9	384.9	416.4	416.4	416.4	318.2	318.2	317.3	45.4
Volumetric Flow rate	m ³ /h	515.2	59.6	63.2	63.2	63.2	1680.0	1769.6	57.4	57.4	46.7	46.7	46.6	0.8
Actual Density	kg/m ³	73.4	10.1	588.4	588.4	588.4	22.1	22.4	689.2	689.2	681.2	681.2	680.2	1001.9
Std Ideal Lig Vol Flow	m ³ /h	53.1	1.2	51.9	51.9	51.9	51.9	54.9	54.9	54.9	45.5	45.5	45.4	0.8
Component Mole Flow rate														
Cyclohexane	kgmole/h	0.5	0.0	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.4	0.4	0.4	0.0
Benzene	kgmole/h	87.7	0.2	87.5	87.5	87.5	87.5	109.6	109.6	109.6	26.1	26.1	26.1	0.0
Hydrogen	kgmole/h	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Pentane	kgmole/h	52.3	0.5	51.7	51.7	51.7	51.7	56.5	56.5	56.5	51.6	51.6	51.6	0.0
o-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
m-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
p-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Toluene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Hexane	kgmole/h	15.4	0.0	15.4	15.4	15.4	15.4	16.1	16.1	16.1	15.1	15.1	15.1	0.0
n-Heptane	kgmole/h	58.9	0.1	58.9	58.9	58.9	58.9	59.1	59.1	59.1	56.8	56.8	56.8	0.0
n-Octane	kgmole/h	68.5	0.0	68.5	68.5	68.5	68.5	68.5	68.5	68.5	66.6	66.6	66.6	0.0
n-Nonane	kgmole/h	72.5	0.0	72.5	72.5	72.5	72.5	72.5	72.5	72.5	70.4	70.4	70.4	0.0
n-Decane	kgmole/h	6.1	0.0	6.1	6.1	6.1	6.1	6.1	6.1	6.1	5.9	5.9	5.9	0.0
Methane	kgmole/h	0.4	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ethane	kgmole/h	2.3	2.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	kgmole/h	8.2	7.7	0.5	0.5	0.5	0.5	0.6	0.6	0.6	0.5	0.5	0.5	0.0
n-Butane	kgmole/h	25.3	2.0	23.3	23.3	23.3	23.3	26.9	26.9	26.9	23.3	23.3	23.3	0.0
Oxygen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.4	45.4
CO	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFOLANE	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.3	1.3	0.0	0.0

Stream Number		27	28	29	30	31	32	33	34	35	36	37	38	39
Phase Fraction		0.0	0.0	0.0	0.0	1.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Pressure	kPa	103.4	413.7	206.8	206.8	103.4	103.4	144.8	144.8	144.8	137.9	137.9	137.9	117.2
Temperature	C	32.2	43.1	55.3	55.3	101.6	101.6	254.5	254.5	254.5	93.3	224.7	224.7	41.4
Molar Enthalpy	kJ/kgmole	-284352.4	-288086.6	-359089.1	-359089.1	49626.5	49626.5	-390569.7	-390569.7	-390569.7	-440394.3	-400732.0	-400732.0	-443554.6
Mass flow rate	kg/h	817.2	963.2	62331.6	62331.6	7777.4	7777.4	54554.2	54554.2	43643.4	10901.4	54544.8	54544.8	54544.8
Molar flow rate	kgmole/h	45.4	46.2	555.0	555.0	97.8	97.8	457.2	457.2	365.7	90.7	456.4	456.4	456.4
Volumetric Flow rate	m ³ /h	0.8	0.9	54.0	54.0	2946.6	2946.6	181.2	181.2	145.0	9.1	51.1	51.1	44.1
Actual Density	kg/m ³	1001.9	1030.6	1154.8	1154.8	2.6	2.6	301.0	301.0	301.0	1196.1	1068.4	1068.4	1236.1
Std Ideal Lig Vol Flow	m ³ /h	0.8	0.9	52.6	52.6	9.3	9.3	43.3	43.3	34.6	8.6	43.2	43.2	43.2
Component Mole Flow rate														
Cyclohexane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Benzene	kgmole/h	0.0	0.0	90.8	90.8	81.7	81.7	9.1	9.1	7.3	0.0	7.3	7.3	7.3
Hydrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Pentane	kgmole/h	0.0	0.0	4.8	4.8	4.8	4.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0
o-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
m-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
p-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Toluene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Hexane	kgmole/h	0.0	0.0	0.9	0.9	0.9	0.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Heptane	kgmole/h	0.0	0.0	2.3	2.3	2.3	2.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Octane	kgmole/h	0.0	0.0	1.9	1.9	1.9	1.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Nonane	kgmole/h	0.0	0.0	2.1	2.1	2.1	2.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Decane	kgmole/h	0.0	0.0	0.2	0.2	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Methane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ethane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	kgmole/h	0.0	0.0	0.2	0.2	0.2	0.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Butane	kgmole/h	0.0	0.0	3.6	3.6	3.6	3.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Oxygen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	kgmole/h	45.4	44.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFOLANE	kgmole/h	0.0	1.3	448.2	448.2	0.1	0.1	448.1	448.1	358.5	90.7	449.2	449.2	449.2

Stream Number		40	41	42	43	44	45	46	47	48	49	50	51	52
Phase Fraction		1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Pressure	kPa	103.4	82.7	82.7	413.7	413.7	413.7	413.7	413.7	413.7	586.1	586.1	586.1	586.1
Temperature	C	101.6	46.1	46.1	92.8	92.8	92.8	92.8	92.8	92.8	152.3	152.3	152.3	152.3
Molar Enthalpy	kJ/kgmole	49626.5	11658.1	11658.1	-4300.5	-4300.5	-4300.5	-4300.5	-4300.5	-4300.5	40357.9	40357.9	40357.9	40357.9
Mass flow rate	kg/h	7777.4	7777.4	7777.4	2380.2	2380.2	2380.2	2380.2	2380.2	2380.2	5397.2	5397.2	5397.2	5397.2
Molar flow rate	kgmole/h	97.8	97.8	97.8	31.7	31.7	31.7	31.7	31.7	31.7	66.1	66.1	66.1	66.1
Volumetric Flow rate	m ³ /h	2946.6	9.6	9.6	7.3	7.3	7.3	7.3	7.3	7.3	9.6	9.6	9.6	9.6
Actual Density	kg/m ³	2.6	807.2	807.2	325.2	325.2	325.2	325.2	325.2	325.2	561.1	561.1	561.1	561.1
Std Ideal Lig Vol Flow	m ³ /h	9.3	9.3	9.3	3.0	3.0	3.0	3.0	3.0	3.0	6.3	6.3	6.3	6.3
Component Mole Flow rate														
Cyclohexane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Benzene	kgmole/h	81.7	81.7	81.7	22.2	22.2	22.2	22.2	22.2	22.2	59.6	59.6	59.6	59.6
Hydrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Pentane	kgmole/h	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	0.1	0.1	0.1	0.1
o-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
m-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
p-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Toluene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Hexane	kgmole/h	0.9	0.9	0.9	0.7	0.7	0.7	0.7	0.7	0.7	0.3	0.3	0.3	0.3
n-Heptane	kgmole/h	2.3	2.3	2.3	0.3	0.3	0.3	0.3	0.3	0.3	2.0	2.0	2.0	2.0
n-Octane	kgmole/h	1.9	1.9	1.9	0.0	0.0	0.0	0.0	0.0	0.0	1.9	1.9	1.9	1.9
n-Nonane	kgmole/h	2.1	2.1	2.1	0.0	0.0	0.0	0.0	0.0	0.0	2.1	2.1	2.1	2.1
n-Decane	kgmole/h	0.1	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.1	0.1	0.1
Methane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ethane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	kgmole/h	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.0	0.0	0.0	0.0
n-Butane	kgmole/h	3.6	3.6	3.6	3.6	3.6	3.6	3.6	3.6	3.6	0.0	0.0	0.0	0.0
Oxygen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFOLANE	kgmole/h	0.1	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.1	0.1	0.1

Stream Number		53	54	55	56	57	58	59	60	61	62	63	64	65
Phase Fraction		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Pressure	kPa	448.2	448.2	551.6	551.6	275.8	275.8	275.8	275.8	275.8	275.8	379.2	379.2	379.2
Temperature	C	82.2	82.2	115.6	115.6	115.9	115.9	115.9	115.9	115.9	115.9	174.3	174.3	174.3
Molar Enthalpy	kJ/kgmole	56307.6	56307.6	-228172.1	-228172.1	62160.4	62160.4	62160.4	62160.4	62160.4	62160.4	27202.8	27202.8	27202.8
Mass flow rate	kg/h	4677.4	4677.4	719.8	719.8	3377.3	3377.3	3377.3	3377.3	3377.3	3377.3	2654.9	2654.9	2654.9
Molar flow rate	kgmole/h	59.9	59.9	6.3	6.3	43.2	43.2	43.2	43.2	43.2	43.2	27.2	27.2	27.2
Volumetric Flow rate	m ³ /h	5.8	5.8	1.1	1.1	4.4	4.4	4.4	4.4	4.4	4.4	3.7	3.7	3.7
Actual Density	kg/m ³	807.5	807.5	627.5	627.5	768.3	768.3	768.3	768.3	768.3	768.3	711.6	711.6	711.6
Std Ideal Lig Vol Flow	m ³ /h	5.3	5.3	1.0	1.0	3.8	3.8	3.8	3.8	3.8	3.8	3.1	3.1	3.1
Component Mole Flow rate														
Cyclohexane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Benzene	kgmole/h	59.6	59.6	0.0	0.0	42.7	42.7	42.7	42.7	42.7	42.7	0.2	0.2	0.2
Hydrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Pentane	kgmole/h	0.0	0.0	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
o-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
m-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
p-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	10.6	10.6	10.6
Toluene	kgmole/h	0.0	0.0	0.0	0.0	0.4	0.4	0.4	0.4	0.4	0.4	16.5	16.5	16.5
n-Hexane	kgmole/h	0.3	0.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Heptane	kgmole/h	0.0	0.0	2.0	2.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Octane	kgmole/h	0.0	0.0	1.9	1.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Nonane	kgmole/h	0.0	0.0	2.1	2.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Decane	kgmole/h	0.0	0.0	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Methane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ethane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Butane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Oxygen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFOLANE	kgmole/h	0.0	0.0	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Stream Number		66	67	68	69	70	71	72	73	74	75	76
Phase Fraction		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Pressure	kPa	379.2	413.7	413.7	413.7	413.7	413.7	413.7	482.6	482.6	482.6	482.6
Temperature	C	174.3	169.1	169.1	169.1	169.1	169.1	169.1	205.9	205.9	205.9	205.9
Molar Enthalpy	kJ/kgmole	27202.8	37601.3	37601.3	37601.3	37601.3	37601.3	37601.3	17451.7	17451.7	17451.7	17451.7
Mass flow rate	kg/h	2654.9	1522.5	1522.5	1522.5	1522.5	1522.5	1522.5	1132.3	1132.3	1132.3	1132.3
Molar flow rate	kgmole/h	27.2	16.5	16.5	16.5	16.5	16.5	16.5	10.8	10.8	10.8	10.8
Volumetric Flow rate	m ³ /h	3.7	2.1	2.1	2.1	2.1	2.1	2.1	1.7	1.7	1.7	1.7
Actual Density	kg/m ³	711.6	715.6	715.6	715.6	715.6	715.6	715.6	677.6	677.6	677.6	677.6
Std Ideal Lig Vol Flow	m ³ /h	3.1	1.8	1.8	1.8	1.8	1.8	1.8	1.3	1.3	1.3	1.3
Component Mole Flow rate												
Cyclohexane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Benzene	kgmole/h	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.0	0.0	0.0	0.0
Hydrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Pentane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
o-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
m-Xylene	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
p-Xylene	kgmole/h	10.6	0.6	0.6	0.6	0.6	0.6	0.6	10.0	10.0	10.0	10.0
Toluene	kgmole/h	16.5	15.6	15.6	15.6	15.6	15.6	15.6	0.8	0.8	0.8	0.8
n-Hexane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Heptane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Octane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Nonane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Decane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Methane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ethane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
n-Butane	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Oxygen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFOLANE	kgmole/h	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Heat Exchangers	E-101	E-102	E-103	E-104	E-105	E-106
Type	Floating Head	Floating Head	Floating Head	Floating Head	Floating Head	Floating Head
Area(m^2)	498.4	63.1	50.5	85.0	31.2	11.9
Duty(Btu/hr)	217396542.7	11119346.9	8132189.8	24239790.1	3519602.4	2786598.0
Shell						
Temp (C)	120.0	366.0	120.0	480.0	120.0	120.0
Pres. (barg)	11.4	12.1	14.5	19.7	9.3	9.3
Phase	1.0	1.0	1.0	1.0	1.0	0.0
MOC	C.S.	C.S.	C.S.	C.S.	C.S.	C.S.
Tube						
Temp(C)	1123.0	304.0	240.0	400.0	214.9	255.8
Pres. (barg)	11.4	12.1	14.5	19.7	9.3	9.3
Phase	0.0	1.0	0.0	1.0	0.0	0.0
MOC	C.S.	C.S.	C.S.	C.S.	C.S.	C.S.
Vessels/Towers	T-101	T-102	T-103	T-104	T-105	V-101
Temperature (C)	703.7	725.7	760.7	634.6	854.7	22.6
Pres. (bar)	110.0	21.0	85.0	55.0	70.0	3.6
Orientation	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical
MOC	C.S.	C.S.	C.S.	C.S.	C.S.	C.S.
Size						
Height/Length (ft.)	20.0	22.0	22.0	44.0	34.0	6.2
Diameter (ft.)	13.0	8.0	11.0	15.0	15.0	1.8
Internals	5 CS Sieve Trays	6 CS Sieve Trays	6 CS Sieve Trays	17 CS Sieve Trays	12 CS Sieve Trays	NA
Pumps/Compressors	P-101A/B	P-102A/B	P-103A/B	P-104A/B	P-105A/B	P-106A/B
Flow (lb/hr)	76752.2	83333.3	82007.4	70500.6	137036.5	1801.5
Fluid Density (lb/ft.^3)	44.5	44.7	36.7	44.4	72.8	62.6
Power (hp)	52.0	28.5	7.3	7.8	4.7	4.7
Pressure in (bar)	15.8	17.2	90.1	27.8	25.8	28.6
Pressure out (bar)	231.1	126.4	113.8	66.6	44.2	50.9
Temperature (C)	70.0	20.2	150.8	43.2	55.3	32.2
Efficiency	0.7	0.7	0.7	0.7	0.7	0.7
Type/Drive	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric
MOC	C.S.	C.S.	C.S.	C.S.	C.S.	C.S.
Reactors	R-101	R-102	R-103			
Type	Autoclave	Autoclave	Autoclave			
Flow (kgmole/hr)	3013.4	3155.5	3156.7			
Pressure (kPa)	116.2	116.2	116.2			
Temperature (C)	241.2	449.6	599.1			
Volume (m^3)	7.5	7.5	7.5			
Benzene Conversion	0.5	0.5	0.5			
Dehydrogenation Conversion	0.5	0.5	0.5			
Alkane Conversion	1.0	1.0	1.0			
MOC	C.S.	C.S.	C.S.			

Heat Exchangers	E-107	E-108	E-109	E-110		
Type	Floating Head	Floating Head	Floating Head	Floating Head		
Area(m^2)	16.8	53.4	7.8	11.5		
Duty(Btu/hr)	4467298.2	5152407.1	3431913.3	3494624.6		
Shell						
Temp (C)	120.0	366.0	120.0	480.0		
Pres. (barg)	9.3	19.7	9.3	19.7		
Phase	1.0	0.0	1.0	0.0		
MOC	C.S.	C.S.	C.S.	C.S.		
Tube						
Temp(C)	250.0	345.7	345.0	402.6		
Pres. (barg)	9.3	19.7	9.3	19.7		
Phase	1.0	0.0	1.0	0.0		
MOC	C.S.	C.S.	C.S.	C.S.		
Vessels/Towers	V-102	V-103	V-104	V-105		
Temperature (C)	92.8	152.3	115.9	169.1		
Pres. (bar)	6.9	6.6	5.2	6.6		
Orientation	Vertical	Horizontal	Horizontal	Horizontal		
MOC	C.S.	C.S.	C.S.	C.S.		
Size						
Height/Length (ft.)	82.96	23.24	1.50	12.16		
Diameter (ft.)	0.50	1.50	2.00	2.00		
Internals	NA	NA	NA	NA		
Pumps/Compressors	P-107A/B	P-108A/B	P-109A/B	P-110A/B	P-111A/B	P-112A/B
Flow (lb/hr)	24033.2	120500.5	16452.3	4998.9	11453.5	9997.7
Fluid Density (lb/ft.^3)	74.7	66.7	50.6	20.3	35.6	20.3
Power (hp)	0.1			2.2	7.0	2.7
Pressure in (bar)	8.2	23.4	20.6	60.6	55.3	61.4
Pressure out (bar)	62.6	52.1	84.0	88.2	88.7	88.4
Temperature (C)	93.3	101.6	224.7	46.1	92.8	92.8
Efficiency	0.7	0.7	0.7	0.7	0.7	0.7
Type/Drive	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric
MOC	C.S.	C.S.	C.S.	C.S.	C.S.	C.S.
Pumps/Compressors	P-113A/B	P-114A/B	P-115A/B	P-116A/B	P-117A/B	
Flow (lb/hr)	13298.4	7445.6	20103.0	5852.9	20475.1	
Fluid Density (lb/ft.^3)	50.7	48.0	48.0	44.4	44.7	
Power (hp)	0.9	1.5	0.9	0.1	0.6	
Pressure in (psi)	62.3	32.4	32.4	37.8	52.9	
Pressure out (psi)	64.7	61.2	76.5	83.7	86.2	
Temperature (F)	152.3	82.2	115.9	115.9	174.3	
Efficiency	0.7	0.7	0.7	0.7	0.7	
Type/Drive	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	Centrifugal/Electric	
MOC	C.S.	C.S.	C.S.	C.S.	C.S.	

Economic Analysis

Capital Cost

The itemized bare module costs for the reactor train, extractor section, and distillation section are itemized in the appendix. The catalyst made up the largest portion of the initial investment followed by the reactors and distillation columns. Cost estimation methods are projected to be within plus or minus 20% of the actual cost [14]. Capital costing for all sections was done using a modified Guthrie method that took into account the design pressure, materials of construction, and a capacity factor. The purchased and installed costs were calculated using equations [16] and [17]. The resulting values were escalated from 2001 to 2020 using the Chemical Engineering Plant Cost Index [17]. The 2021 Plant Index value has not been released yet, but is projected to be higher than 2020 [17]. This could cause the capital costs to exceed the proposed amounts. Further explanation of capital costs can be found in the Appendix.

Table 6 shows the direct manufacturing costs for the naphtha reforming plant. Raw materials make up the largest portion of the total cost followed by waste treatment. All direct manufacturing costs were completed referencing [5] and using equation [24].

Direct Manufacturing Costs	USD (\$)
Raw Materials	141,632,068
Waste Treatment	26,970,044
Fixed Capital Investment	10,294,710
Patents and Royalties	5,424,062
Utilities	1,033,415
Maintenance and repairs	617,683
Operating Labor	191,096
Supervisor and clerical labor	34,397
Laboratory charges	28,664

Table 6: Direct Manufacturing Costs

Operating Costs

The operating costs for the utilities are shown in Table 7. The heat exchangers had the largest energy cost followed by the furnace. Utility costs for the heat exchangers were based on the duty required to achieve the desired temperature. Operating costs for the heat exchangers were calculated using equations [28] and [30]. The furnace utility cost was based on the yearly consumption of fuel and was calculated using equation [27]. The pumps used the provided electricity and were cost based on equation [26]. Further explanation of utility costs can be found in the Appendix.

Utility Cost breakdown	
Heat Exchangers	553,148
Furnace	424,343
Pumps	55,924

Table 7: Utility Operating Cost

The labor costs were calculated using equation [23] and the number of required operators was determined by equation [25]. The recommended number of operators for the plant is 15. An average operator pay of \$32 was used as the basis for calculation [22]. Per the memorandum salaries in the U.S. are 438% higher than in Iraq. Based on the U.S. operator cost the estimated yearly cost of the operators in Iraq is \$191,000.

Labor Costs	
Vessels	4
Pumps	17
Heat Exchangers	10
Towers	5
Reactors	3
Total Equipment	39
Equipment that Needs an Operator	18
N_{oL}	3.2
Total Number of Operators Required	14.53
Final Number of Required Operators	15
Average pay per hour U.S. Operator (\$)	32.3
Operator Cost U.S. (\$/yr)	837,000
Operator Cost Iraqi (\$/y)	191,095.89

Table 8: Labor Costs

Revenue Estimates

The product price of benzene and diesel had the greatest impact on the revenue generated by our process. Since, these two prices led to the largest financial gain the process was optimized to increase these two streams flow rates. Cyclic h-carbons had the lowest contribution to the total revenue. The revenue estimates located in Table 7 were calculated over a yearly basis assuming a service factor of 0.93.

Revenue	USD (\$)
Benzene Sales	118,319,519
Diesel	113,113,325
Local gasoline	62,958,457
Toluene Sales	56,743,848
Xylene Sales	49,234,492
Cyclic h-carbons	2,860,946
Total Sales	403,230,587

Table 9: Revenue Sources

Table 10: Cash Flow Table Iraqi Control

Project Title: AICHe Project
 Corporate financial situation: Stand Alone
 Minimum rate of return, i[^]*= 0.15 or 15 %
 Other Relevant Project Info: 30 Year Life MACRS Assume Washout
 1=\$1

End of Year	2021	2022	2023	2024	2025	2026
Production x Sales Price, \$/unit			201,615,293	403,230,587	403,230,587	403,230,587
Sales Revenue			201,615,293	403,230,587	403,230,587	403,230,587
+Salvage Value						
-Royalties			(2,712,031)	(5,424,062)	(5,424,062)	(5,424,062)
Net Revenue			198,903,262	397,806,525	397,806,525	397,806,525
-Raw Material Costs			(70,816,034)	(141,632,068)	(141,632,068)	(141,632,068)
-Other Op Costs			(14,437,650)	(28,875,299)	(28,875,299)	(28,875,299)
-Depreciation			(1,029,471)	(1,853,048)	(1,482,438)	(1,185,951)
-Amortization						
-Depletion						
-Loss Forward						
-Writeoff						
Taxable Income			112,620,108	225,446,109	225,816,719	226,113,207
-tax @ 35% Iraqi Control			(39,417,038)	(78,906,138)	(79,035,852)	(79,139,622)
Net Income			73,203,070	146,539,971	146,780,867	146,973,584
+Depreciation			1,029,471	1,853,048	1,482,438	1,185,951
+Amortization						
+Depletion						
+Loss Forward						
+Writeoff						
-Working Capital	(20,250,000)					
-Fixed Capital	(10,294,710)					
Cash Flow	(30,544,710)	0	74,232,541	148,393,019	148,263,306	148,159,535
Discount Factor (P/F)	1	0.87	0.76	0.66	0.57	0.50
Discounted Cash Flow	(30,544,710)	0	56,130,466	97,570,819	84,770,026	73,661,474
NPV @ i* (15%)	757,237,594					
DCFROR	142%					
Payback Period (Years)	1.41					
DC Payback Period (Years)	1.54					

Table 10: Cash Flow Iraqi Control Continued

2027	2028	2029	2030	2031	2032	2033	2034
403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587
403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587
(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)
397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525
(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)
(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)
(949,172)	(758,720)	(674,303)	(674,303)	(675,333)	(674,303)	(337,666)	
226,349,985	226,540,437	226,624,854	226,624,854	226,623,824	226,624,854	226,961,491	227,299,157
(79,222,495)	(79,289,153)	(79,318,699)	(79,318,699)	(79,318,338)	(79,318,699)	(79,436,522)	(79,554,705)
147,127,490	147,251,284	147,306,155	147,306,155	147,305,486	147,306,155	147,524,969	147,744,452
949,172	758,720	674,303	674,303	675,333	674,303	337,666	
148,076,662	148,010,004	147,980,458	147,980,458	147,980,819	147,980,458	147,862,635	147,744,452
0.43	0.38	0.33	0.28	0.25	0.21	0.19	0.16
64,017,627	55,642,443	48,375,074	42,065,282	36,578,595	31,807,397	27,636,584	24,012,604

Table 10: Cash Flow Iraqi Control Continued

2035	2036	2037	2038	2039	2040	2041	2042	2043	2044
403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587
403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587
(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)
397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525
(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)
(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)
227,299,157	227,299,157	227,299,157	227,299,157	227,299,157	227,299,157	227,299,157	227,299,157	227,299,157	227,299,157
(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)
147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452
147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452
0.14	0.12	0.11	0.09	0.08	0.07	0.06	0.05	0.05	0.04
20,880,525	18,156,978	15,788,677	13,729,284	11,938,508	10,381,311	9,027,227	7,849,763	6,825,881	5,935,548

Table 10: Cash Flow Iraqi Control Continued

2045	2046	2047	2048	2049	2050	2051
403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587
403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587	403,230,587
(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)	(5,424,062)
397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525	397,806,525
(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)	(141,632,068)
(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)	(28,875,299)
227,299,157	227,299,157	227,299,157	227,299,157	227,299,157	227,299,157	227,299,157
(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)	(79,554,705)
147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452
						20,250,000
147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	147,744,452	167,994,452
0.03	0.03	0.03	0.02	0.02	0.02	0.02
5,161,346	4,488,127	3,902,719	3,393,669	2,951,017	2,566,101	2,537,229

DCFROR and NPV Analysis

Per the memorandum the hurdle rate was given to be 15% and the project was evaluated over 30 years. MACRS depreciation life was used over a 10 year period for oil refining equipment [14]. The equipment was fully depreciated and the working capital invested in 2021 was written off in the final project year. Construction of the plant was assumed to occur at the end of 2021 and was assumed to be completed in mid year 2023. Half year convention was used in 2023 for revenue and operating costs. The washout assumption was used to maintain a constant annual revenue. The washout assumption assumes that the escalation of costs and revenues are different, but the margin between the two remains constant.

An NPV greater than 0 and a DCFROR value higher than the minimum rate of return show that a project is economically attractive [14]. The cash flow table for Iraqi control is shown in Figure XX. The cash flow table for Kurdish control was not included due to the only difference being the tax rate. From the Iraqi control analysis the NPV was determined to be \$757 million with a DCFROR of 142%. The payback period under this tax regime was 1.5 years. From the Kurdish control analysis the NPV was determined to be \$998 million with a DCFROR of 165%. The payback period under this tax regime was 1.4 years.

MACRS 10 Year Property	Oil Refining Equipment
Year 1	0.1
Year 2	0.18
Year 3	0.144
Year 4	0.1152
Year 5	0.0922
Year 6	0.0737
Year 7	0.0655
Year 8	0.0655
Year 9	0.0656
Year 10	0.0655
Year 11	0.0328

Table 11: MACRS Depreciation

Sensitivity Analysis

Sensitivity analysis was conducted to determine the effect of parameters of the NPV of the plant. This analysis helps to show the uncertainty of the economics, not the probability that they will occur. Each parameter was varied while the others, raw materials, utilities, etc. were kept constant. The result is shown in Figure 4. Sensitivities for annual profit, raw materials, project life, waste treatment, initial investment, and utilities were all included in the analysis.

The largest factor on the NPV was the annual profit. According to [5] the annual profit can reach zero which led to the lowest NPV. The raw material cost had the largest effect on the annual profit and the second highest factor of the NPV. Waste treatment variance was similar to initial investment. The utilities had the smallest impact of any of the parameters. The utility costs are so small compared to raw material costs that they have an impact range of \$50 million on the NPV.

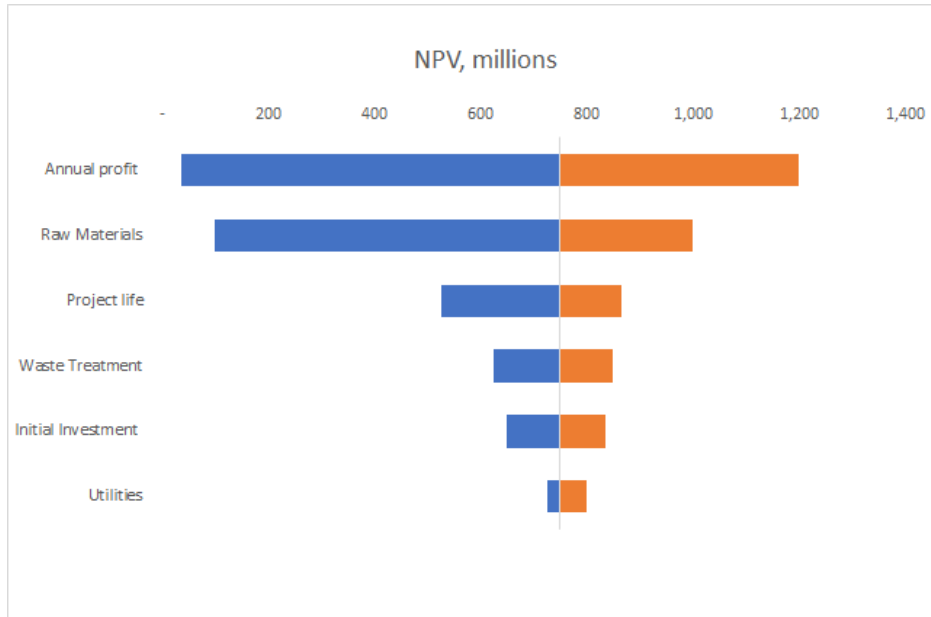


Figure 4: Tornado Chart

The break even revenue was calculated using equation [22] and recorded in Table 10. The initial capital investment of \$30.5 million will need to be countered with ~\$44 million in 2023 when the plant is able to generate revenue.

Initial Cost (\$)	Breakeven Reveune(\$)
(30,544,710)	44,219,961

Table 12: Break Even Revenue

A best and west case scenario was created for the naphtha reforming plant. The best case scenario conditions result from a 10% increase in yearly profit and a 10% decrease in the initial investment [5]. For the best case portion the Kurdish tax regime was assumed to be in control because the 15% tax rate led to a higher NPV and DCFROR. The best case NPV was \$ and the DCFROR was %. For the worst case scenario a profit of \$0 was assumed [5] and the resulting DCFROR was 0%.

	Best Case	Expected Case	Worst Case
Initial Investment (\$)	(27,490,239)	(30,544,710)	(36,653,652)
Annual Profit (\$)	212,830,200	193,482,000	0
Project Life (yrs)	30	30	30
DCFROR (%)	215	142	0
NPV (\$)	1,200,000,000	757,000,000	(36,653,652)

Table 13: Best and Worst Case

Monte Carlo Analysis

Monte Carlo analysis is a tool that quantitatively explains the risk and probability of an economic project[14]. For our preliminary design the Monte Carlo Methods were done for both the NPV and DCFROR. The results are shown in Figure 5 and Figure 6. This method uses central limit theory which states that an infinite amount of samples will have a normal distribution

curve [14]. This means that the actual result will most likely trend towards the center of a histogram. 500 samples were used to construct our graphs and a t-statistic of 1.96 was used [18]. The standard deviation, 95% confidence intervals, and mean were calculated. The results are shown in Table 12 and Table 13. To calculate the NPV and DCFROR a P/A value was used [14] and random values were generated from the profit min and max. The percentage that the profits varied came from [5]. All costs associated with the plants were varied according to [5]. Once the values of cost and revenue were found the NPV and DCFROR could be calculated over a 30 year period. Both histograms of the Monte Carlo analysis are shown in figures [5] and [6].

500 Samples	
Central Limit Theory	
t statistic	1.96
P/A Constant	6.57
A/P Value	0.15
Prop of NPV>0	0.93
Standard Deviation	998
Mean NPV	1,477
95% Confidence Lower Level	(480)
95% Confidence Upper Level	3,434

Table 14: Monte Carlo Data NPV

500 Samples	
Central Limit Theory	
t statistic	1.96
P/A Constant	6.57
A/P Value	0.15
Prop of DCFROR > i*	0.92
Standard Deviation	0.06
Mean DCFROR	1.26
95% Confidence Lower Level	0.23
95% Confidence Upper Level	1.38

Table 15: Monte Carlo Data DCFROR

The average NPV was \$825 million and the average DCFROR was 126%. The NPV and DCFROR had a greater than 90% probability of being economically attractive shown in tables [14] and [15]. For the NPV this meant that the NPV greater than 0 and the DCFROR was greater than the hurdle rate [14]. This tool shows that in most cases the project will be economically favorable and should move to the detailed design phase.

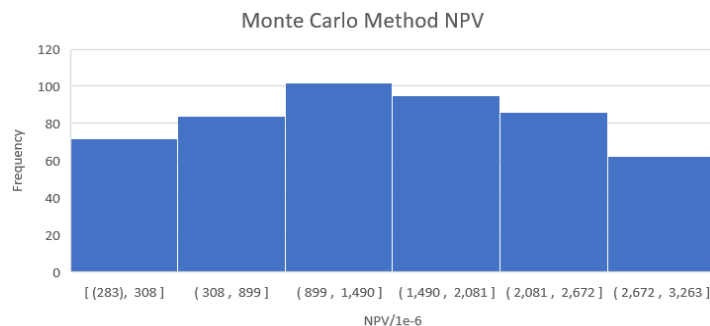


Figure 5: Monte Carlo Method NPV

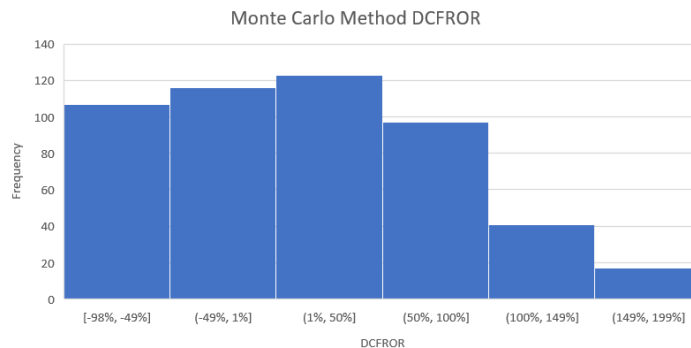


Figure 6: Monte Carlo Method DCFROR

Process Safety

The safety of the plant workers as well as the environment were key considerations for our design philosophy. Considering that Kurdistan is cracking down on outdated processes which pervade the region, we ensured that all parts of the plant design followed the guidelines outlined by the Occupational Safety and Health Administration (OSHA) as well as the Environmental Protection Agency (EPA) as listed in the United States. This both supports the longevity of the plant and, more importantly, the safety of the plant workers. All employees will be trained regarding proper PPE, emergency procedures, and operational discipline on a routine basis prior to any operation(s). Proper PPE includes steel toed boots, eye protection, ear protection, gloves, and flame resistant clothing which is to be worn at all times on site [2]. Operational discipline includes instructing proper lock-out-tag-out and workplace procedures outlined by SOPs and reinforced by leadership.

The overall process involves standard refinery chemistry and multiple effluent material streams which must be made known to all personnel. Alkane streams are intended for sale or for use as fuel in house. Effluent from the fired heater is composed of combustion byproducts which will be released to the atmosphere. Every other exit stream that we did not account for sales was sent to the refinery's flare so as to reduce hydrocarbon emission to the local population. Due to the high volume of organic material exposed to high temperatures, loss of containment is the highest concern of the entire process.

Inherent Safety Evaluation

As part of the commitment to loss prevention associated with this process, inherent safety is crucial to reducing the potential for harm to human and environmental health. The processes we include fall under the categories of minimization, moderation, substitution, and/or simplification.

Minimization was used in process areas where energy reduction contributed to loss prevention. Purge streams are used in both the upstream and downstream process design in order to reduce high concentrations of vapor products. This allows the process equipment to depressurize even after reactions create many more moles of flow (e.g. alkane cracking produces many moles of smaller hydrocarbons from heavier ones). Several separation steps were included in the downstream process in order to reduce buildup of hydrocarbons. These steps also become profitable as we separate diesel from gas which are both salable. Within the furnace, risk of incomplete combustion is combated with a high air:fuel ratio such that plenty of oxygen is available for complete reaction which in turn reduces the amount of carbon monoxide produced by our process.

Moderation was used in process areas where hazardous materials were subjected to less hazardous conditions. Most significantly we designed the upstream reaction train to take advantage of reaction temperatures to both maximize reaction extents while minimizing excess temperature rise. Particularly, alkane cracking and cycloalkane dehydrogenation showed peak reaction rates in the range of 400-600 °C, with the reaction extents reaching a theoretic max near 600 °C. Therefore, we designed the furnace with a flame temperature of 950 °C resulting in a successfully efficient reactor train which nearly reacts all cyclohexane to form either alkanes or benzene (depending on the reaction pathway). The purge streams mentioned in the minimization paragraph double as a moderation strategy as it reduces pressure within the system. By reducing pressure we increase safety and reduce potential equipment wear.

Substitution was used frequently in the process design as this includes reaction chemistry, solvents, and heat transfer media. We designed the series-oriented reactors as PBRs with a platinum catalyst which allowed us to run cracking reactions at much lower temperatures than without catalysis. By substituting this in, we are able to reduce hotspots upstream in the process which ultimately reduces risk of runaway at all points in the plant. For all heat transfer media we used water in order to reduce the amount of flammable or combustible material in the plant. For all reboilers we modeled heat transfer using the lowest possible pressure steam that would still provide sufficient heat flux. In this way we reduce energy consumption which increases the sustainability of the proposed process. Also, lower pressure steam means lower temperatures and a smaller risk of damage to plant workers if steam were to escape containment. All around, this provides benefits both economically and health-wise. Recycle streams were used frequently in our process rather than flaring or uptaking more process material (e.g. sulfolane) to both reduce environmental impacts as well as reduce economic costs.

Simplification was used in process areas where excess materials or vessels were not necessary. An analysis was conducted for each staged vessel such that any stages of zero separation were removed from the model design. In this way, we reduce capital cost while also minimizing the amount of area and volume for the vessels to leak from.

Process Safety Management (PSM)

Preservation of the plant's assets and the workers' health is pertinent to the PSM designed for this project. The proposed plant will operate fully within OSHA and EPA guidelines as they are listed in the United States in order to keep on-site and off-site risks at a minimum. Proper management of process safety requires involvement from management, plant workers, and contractors, as well as smooth communication between extrinsic safety elements and management of potential changes.

Before any operation, a pre-startup safety review will be conducted for any new on-site workers as well as for any process changes. Standard operating procedures (SOP) will be provided to all onsite workers to increase awareness of process conditions and norms. The system of direct reporting should be established such that any deviation from process norms are quickly and directly reported to management. This work culture should also reflect the importance of safe work practices, and should allow the plant worker to shut down operation in the event of drastic changes in process temperature or pressure. Process safety information will be communicated by process monitor equipment as shown below in the P&ID of the major fractionator. Monitor equipment is placed throughout the design in order to mitigate potential hazards that pervade the model plant. Hazard analyses are listed below while taking inherent safety into account to reduce associated risk of the full design.

Process hazards

There are numerous hazards that emerge in the process design which include both physical and chemical variety. Petroleum refineries commonly house high temperature and

pressure processes of which many reactions are likely to occur, and of which process vessels are required to withstand conditions of high stress. In order to address the safety concerns in the process, we conducted a full process hazard analysis to identify, evaluate, and analyze the largest potential hazards. Hazards in our process design stem from either high temperature, high pressure, or chemical hazards.

High Temperatures correspond to high internal energy within heated vessels and piping. Temperatures in our plant model range anywhere from 22 to 950 °C. The highest temperature is found in the furnace effluent stream which is composed of combustion byproducts such as water, carbon dioxide, and excess air. Actual process streams top out at a maximum temperature of 600 °C which is reached near the end of the reactor train. At these temperatures, reactions are highly likely to occur which result in sharp increases of molar flow. Alkane cracking results in the breakdown of larger hydrocarbons into smaller ones, and this will directly affect the pressure of our process downstream of reaction. This poses a risk of runaway reaction in which a positive feedback loop of exothermic reactions could potentially lead to process disaster. Because of this, we modeled the process with platinum catalysis and provided immediate cooling downstream of the reaction train in order to keep temperatures low for the now high molar flow stream.

High pressures correspond to high stress on process vessels, and provide considerable driving force for mass transport. Pressures in our plant model range anywhere from 12 to 209 psia. The highest pressure occurs directly downstream of the feed pump which provides the driving force for flow of the entire process. As the inlet material stream travels through process equipment its pressure slowly drops by way of friction. Powered process vessels such as distillation column pumps provide the extraneous flow that is required for the entire process. Vacuum conditions only occur once in the process in the downstream process for which a column process pump brings it back to above atmospheric pressures. To deal with the high pressures sizing of relief valves were done for both the pumps and vessels in the system. It was found to use pressure relief valves that are 50mm x 75mm for the pumps and ones that are 75mm x 100mm for the vessels. This was done by using the pressure, flow rate, and area of the piping and solving for the correct size area. This value was then used to get the sizes that are listed above.

Chemical hazards complicate the hazards analysis for immiscible, combustible, and oxidizing chemicals are found at various positions in the process. The process itself already has multiple reactions occurring in the upstream section. We compiled data regarding all chemicals potentially involved in the total model plant and compared how each chemical would interact under the worst case scenario condition of total loss of containment (LOC). In this way, we are able to see what conditions must be specifically avoided in order to reduce the probability of process failure. Table 14 below contains hazards for all chemicals in the model as well as important information obtained from their respective SDS sheets. This table also includes occupational discipline and safety measures intended to reduce consequence in the case of accidental release. We found that few interactions occur between the chemicals other than the mixture of combustible materials and oxygen. Otherwise, chemicals that are miscible with water will generate heat upon mixing, so temperature must be monitored in areas where potential heating may occur.

Component	Flammable	Irritant	Health Hazard	Environmental Damage	Oxidizer
Cyclohexane	X	X	X	X	
Benzene	X	X	X		
toluene	X	X	X		
o-xylene	X	X	X		
m-xylene	X	X	X		
p-xylene	X	X	X		
n-decane	X	X	X		
n-nonane	X	X	X		
n-octane	X	X	X		
n-heptane	X	X	X	X	
n-hexane	X	X	X		
n-pentane	X	X	X		
n-butane	X				
propane	X				
ethane	X				
methane	X				
sulfolane		X		X	
water					
carbon monoxide	X	X	X		
carbon dioxide					
Hydrogen	X				
oxygen	Required for Combustion				X
nitrogen					

Table 16(a): Chemical Hazards According to SDS Sheets

Component	Vapor Density (Air = 1.0)	Toxicity LD50	Mitigation For Loss of Containment
Cyclohexane		2.9 inhalation 13.9 mg/L/4h	Combustibles: Ensure process equipment is set up with proper ventilation so as to reduce chemical accumulation if leaking were to occur. Wear proper PPE including eye protection and respirators to prevent chronic respiratory exposure to irritating and toxic chemicals. Ensure process streams are not exposed to any ignition source to prevent mass combustion.
Benzene		2.7 inhalation 10000 ppm/7h	
toluene		3.1 inhalation 12,500-28,800 mg/m3/4h	
o-xylene		3.7 inhalation 4330 ppm/6h	
m-xylene		3.66 inhalation 4330 ppm/6h	
p-xylene		3.66 inhalation 4330 ppm/6h	
n-decane		4.9 inhalation 72.3 mg/L/4h	
n-nonane		4.41 inhalation 3200 ppm/4h	
n-octane		3.9 inhalation 25260 ppm/4h	
n-heptane		3.5 inhalation 103 g/m3/4h	
n-hexane		2.97 inhalation 48000 ppm/4h	
n-pentane		2.5 inhalation 364 g/m3/4h	
n-butane		2.1 inhalation 658 g/m3/4h	
propane		1.6 N/A	
ethane		1.1 N/A	
methane		0.6 N/A	
sulfolane	Liquid SG = 1.260	Oral 1941 mg/kg dermal >3800 mg/kg	Heavy Liquids: Ensure employees are aware of heavy liquid use in process and provide sufficient PPE with regard to bodily and facial protection. Recycle Sulfolane to minimize release outside of plant.
water	Liquid SG=1.00	N/A	
carbon monoxide		0.97 inhalation 3760 ppm/1h	Furnace Gases: Ensure all furnace combustion streams have routine maintenance so as to not mix with hydrocarbon process streams. Ensure that all combustion byproducts are immediately sent to the atmosphere to reduce residence time near plant workers, and reduce potential mixture with combustibles.
carbon dioxide		1.53 N/A	
Hydrogen		0.07 N/A	
oxygen		1.1 N/A	
nitrogen		0.967 N/A	

Table 16(b): Chemical Density, Toxicity Limits, and Mitigation Strategies

	Normal Flash Point [C]	LFL [%]	UFL [%]
Cyclohexane	-17.99	1.3	8
Benzene	-11	1.2	7.8
toluene	4	0.7	1.2
o-xylene	31	0.9	6.7
m-xylene	31	1.1	7
p-xylene	31	1.1	7
n-decane	46	0.7	5.4
n-nonane	31	0.8	2.9
n-octane	13	0.8	6.5
n-heptane	-4	1	6.7
n-hexane	-22	1.1	7.5
n-pentane	-49	1.5	7.8
n-butane	-60	1.8	8.4
propane	-104	1.8	8.4
ethane	-104	1.8	8.4
methane	-104	5	14
sulfolane	165	N/A	N/A
water	N/A	N/A	N/A
carbon monoxide	-191.5	10.9	74.2
carbon dioxide	N/A	N/A	N/A
Hydrogen	N/A	4	76
oxygen	N/A	N/A	N/A
nitrogen	N/A	N/A	N/A

Table 16(c): Chemical Flashpoints and Flammability Limits

According to Table 14, we find that carbon monoxide poses the greatest toxicity risk with an LD50 rating ranked at only 1 hour of exposure [3]. The combustible materials also pose a toxicity risk after prolonged exposure, but their potential hazards are most significant with regard to flammability when exposed to oxygen as shown in chemical interaction Table 14. The toxicity risk is not to be overlooked, however, as most of the hydrocarbons have vapor densities much higher than air which means that upon release the vast quantities of combustibles will stay low to the ground and spread to local communities. The heavy liquids pose the least amount of chemical risk in the plant, but do so more when considering the nearby water supply due to sulfolane's high solubility and specific gravity with respect to water. After discussion among the engineering team, we concluded on six high-priority hazards which are listed below in Table 15.

Hazard	Equipment Damage	Environmental Compliance	Loss of Life	Community Damage
Hydrocarbon explosion	High	Medium	Medium	Medium
Incomplete Combustion (Furnace)	Low	Low	Low	Low
LOC (Hydrocarbons, Pre-reactors)	Low	High	Medium	High
LOC (Hydrocarbons, Post-reactors)	Low	High	Medium	High
LOC (Heavy Liquids)	Low	Medium	Low	Low
LOC (Furnace Gases)	Low	Low	Low	Low

Table 17: Potential Hazard Identification

The above hazards all stem from concerns over harsh processing conditions and potential release and/or unintended reaction of process chemicals. Keeping these in mind, we

implemented numerous inherent safety concepts to mitigate the risk and promote a safer, simpler design for the Kurdistan oil company. Our consequence reduction strategy is listed in Table 16 below.

Hazard	Inherent Safety Concept	Consequence Reduction Strategy
Hydrocarbon explosion	Minimization, Moderation, Substitution, Simplification	Purge streams located at plant upstream, downstream, and their connection to reduce concentration and pressure buildup of combustible materials. Water is used as the only cooling and heating media to minimize excess combustible material. All trayed process equipment were optimized such that there are no "dead zones" are present in the process which reduces areas of potentially hazardous concentration or process error can occur.
Incomplete Combustion (Furnace)	Minimization	Ensure furnace fuel stream is overcompensated with air flow to minimize carbon monoxide production and maximize complete combustion.
LOC (Hydrocarbons, in reactor train)	Moderation, Substitution	Temperature and pressure reduction via furnace temperature. Utilization of Platinum Catalyst in each reactor makes processing condition requirements less severe. Furnace temperature set to 950 C in order to reduce excessive heating and minimize potential hot spots.
LOC (Hydrocarbons, post-reactors)	Minimization	Numerous separation steps in the downstream process are used to dilute concentrations of combustible materials. Also creates product alkane streams which will be sent to sales.
LOC (Heavy Liquids)	Substitution	Utilize recycle streams in the extraction section to reduce heavy liquid mass requirements and therefore output.
LOC (Furnace Gases)	Substitution, Minimization	Carbon monoxide is lighter than air, therefore it will rise due to molecular mass and will dilute according to the furnace output concentration gradient and heat gradient. Excess air used to reduce outflow of toxic gas.

Table 18: Consequence Reduction Strategy for Process Hazards

P&ID of the Major Factionator

The basic control system used in the process was the use of feedback control and is shown in the P&ID below in Figure 7. This control system is the best way to control the different flows of the streams and can be used in many different areas. Alarms were also installed to increase the safety of the process while being able to run at the correct process conditions. The use of controls also has indicators for temperature, pressure, and flow in order to ensure that these values are maintained from the control room.

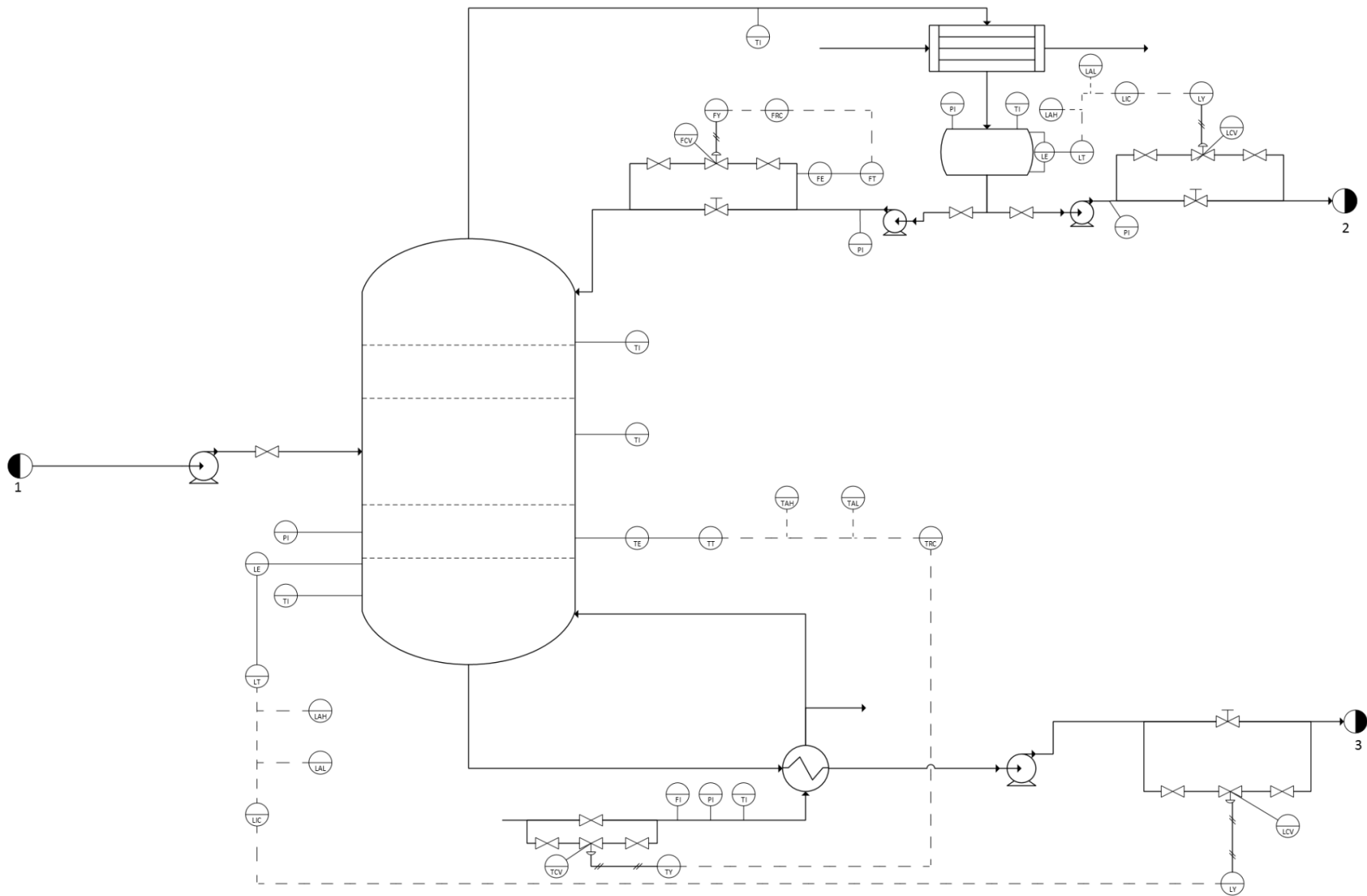


Figure 7: P&ID of the Major Fractionator

Uncongested Vapor Cloud Deflagration

The worst case scenario for any process should be taken into consideration in order to properly prepare for that situation. For the case of a petroleum refinery, the worst case scenario would be total loss of containment followed by the formation of a combustible vapor cloud. In our process model, there are no stored chemicals as the process is entirely continuous. According to a review article which covered a vapor cloud explosion in Buncefield, UK there was a short window of 20 minutes of chemical release before the cloud ignited^[4]. Therefore, we conducted calculations to find the TNT equivalency of a total release for twenty minutes. The results of which are listed below in Table 17. We then estimated overpressure values assuming yearly average atmospheric pressure conditions for Kurdistan (14.81 psia). The resulting overpressures and associated effects at various distances from the blast center are listed further below in Table 18 with values obtained from chemical process safety correlation tables^[5]. Calculations were based on a TNT equivalent of 1120 kcal/kg.

Material	Maximum Quantity in Plant [lb/hr]	Heat of Combustion [kJ/lb]	Energy Release (Total Combustion) [kJ/hr]
Cyclohexane	1.71E+04	3.77E+04	-1.75E+09
Benzene	1.00E+04	2.21E+04	-8.97E+08
toluene	3.36E+03	7.40E+03	-3.15E+08
o-xylene			-2.37E+08
m-xylene	2.50E+03	5.50E+03	
p-xylene			
n-decane	5.51E+04	1.21E+05	-5.79E+09
n-nonane	2.05E+04	4.53E+04	-2.16E+09
n-octane	1.74E+04	3.83E+04	-1.83E+09
n-heptane	1.32E+04	2.92E+04	-1.40E+09
n-hexane	3.11E+03	6.86E+03	-3.31E+08
n-pentane	1.00E+04	2.21E+04	-1.07E+09
n-butane	5.51E+03	1.21E+04	-5.51E+08
propane	2.97E+03	6.55E+03	-3.02E+08
ethane	1.74E+03	3.84E+03	-1.81E+08
methane	1.60E+04	3.54E+04	-1.77E+09
sulfolane	2.41E+04	5.31E+04	-4.88E+08
water	37831+92000 (utility)	N/A	N/A
carbon monoxide	0.00E+00	0.00E+00	0.00E+00
carbon dioxide	4.40E+04	N/A	N/A
Hydrogen	1.63E+02	3.59E+02	-4.19E+07
oxygen	6.72E+04	N/A	N/A
nitrogen	2.21E+05	N/A	N/A
Total Value	6.65E+05	Total Energy Release [kJ/hr]	-1.91E+10
Total Value (Combustibles)	2.03E+05	TNT Equivalent [kg/hr]	1.71E+07
		TNT Equivalent [kg/20 min]	5.69E+06

Table 19: TNT Equivalent Calculations

Distance from Blast [m]	Overpressure [psig]	Damage: 20 minutes of accumulation pre-explosion
10	66	Probable total destruction of buildings, heavy machine tools moved and badly damaged, very heavy machine tools survive
25	33	Probable total destruction of buildings, heavy machine tools moved and badly damaged, very heavy machine tools survive
50	9.9	Probable total destruction of buildings, heavy machine tools moved and badly damaged, very heavy machine tools survive
100	2.75	50% destruction of brickwork of houses, concrete or cinder block walls shatter, serious structural damage
200	0.44	Limited minor structural damage, large/small windows usually shatter
300	0.198	Within safe distance, near projectile limit and glass breakage limit
500	0.066	Loud noise (143 dB), sonic boom, glass failure
750	0.033	Occasional breaking of large glass windows already under strain
1000	0.022	Annoying noise (1.37 dB)
1250	0.0165	Annoying noise (<1.37 dB)
1500	0.0143	Not impacted

Table 20: Impact of Uncongested Vapor Cloud Deflagration at Atmospheric Pressure

Safety Summary

We found that the highest risks associated with the process are the loss of containment of hydrocarbons as well as their mixture with air resulting in a high volumetric explosive vapor cloud. In order to address the potential hazards, we conducted a full process hazard analysis and incorporated inherent safety concepts to reduce potential consequence frequency and severity. We used minimization in the form of purge streams to prevent high pressure buildup throughout the process, as well as optimizing our furnace fuel:air ratio so as to reduce the amount of carbon monoxide released to the atmosphere. Numerous separation steps have been used in the process in order to reduce concentrations of potentially hazardous and combustible materials. These will both lead to less severe processing conditions and effective environmental compliance. Substitution was used in the form of a platinum catalyst such that reaction temperatures and pressures could be drastically reduced, and water was used as the only heat transfer media to reduce the amount of flammable material on site. Recycle streams were utilized rather than flaring, and they were also used to reduce the intake and output of sulfolane within our process. Simplification was established as we optimized all trayed equipment to reduce areas of potential error or chemical buildup.

Our process design has also been thoroughly equipped with process instrumentation in order to properly maintain process conditions and monitor potential hazards. Regardless of intrinsic or extrinsic safety measures, the worst case scenario showed the effects of deflagration can break glass out to 750 meters, launch projectiles near 300 meters, minor structural damage out to 200 meters, and probable total destruction at 50 meters. The estimated safe distance is

outside of 300 meters, so a risk management plan must be put in place to provide workers with instructions in case of total loss of containment. Proper PPE includes respirators, flame retardant clothing, eye protection, helmet gear, and gloves. Every worker must be equipped with proper training, procedural discipline, and pre-startup safety reviews must be held before plant startup.

Conclusions

To meet new Western refining standards our team completed the preliminary design for a naphtha catalytic reforming plant. All process units were sized and cost appropriately and economic analysis was done to analyze the feasibility of the project.

The main focus of this project was putting forth a plant that has identified all of the potential safety hazards to keep all of the involved parties safe. The equipment was designed to keep hydrocarbons in the pipes and mitigate potential risks. All of the equipment was built with oversize factors to prevent overpressure or abnormal conditions. PHA information and SDS were created and tabulated for detailed design.

A variety of optimization techniques were utilized to provide a safer and cheaper plant. Heat integration was used to eliminate the need for two heat exchangers. Process conditions were chosen to keep energy costs low and mitigate the risk of an accident. An entire stripping column was removed to avoid over complication. A divided wall column was used in the distillation section to simplify the system and lead to a lower present worth cost.

An initial investment of \$30.5 million is required to complete the design of the plant. The annual revenue was projected to be \$403 million and the annual operating was projected to be \$181 million. Both Kurdish and Iraqi control tax brackets were analyzed. A 30 year project evaluation life was used and the resulting NPV was \$757 million under Iraqi control. The DCFROR was calculated to be 142% with a payback period of 1.5 years. A 30 year project evaluation life was used and the resulting NPV was \$998 million under Kurdish control. The DCFROR was calculated to be 165% with a payback period of 1.4 years. MACRS 10 year depreciation life was used for the refining equipment. Under either tax regime the project is economically attractive under the current conditions.

Sensitivity analysis was performed to analyze the largest contributing factors to the economic viability of the project. The largest factor that contributed to the project being unfavorable was yearly profit. The profit varied the most based on the feedstock and product cost. The worst case scenario could lead to an NPV of \$(36) million and a DCFROR of 0% . The best case scenario would lead to an NPV of \$1.2 billion and a DCFROR of 215% under Kurdish control. Overall, our team believes that this project will bring further value to the company.

Recommendations

- Detailed design should begin on the naphtha catalytic reforming plant because the project has been determined to be economically attractive.
- The toluene-xylene distillation column consumes a large amount of energy and was under optimized. This could be a future starting point for optimization.
- Research and test additional catalysts that achieve the same conversion and selectivity, but require less hazardous process conditions.
- If sulfur analysis is included in the scope of the detailed design phase, the materials of construction should be reevaluated for long term effectiveness.
- Adjust the project life on the economic evaluation to a more realistic timeline, 5 years.
- Add a scrubber to the process design to prevent catalyst poisoning from sulfur.
- Lower the sulfolane feed flow rate to reduce feedstock costs.
- Research new solvents for T-102 to potentially eliminate any hazards associated with sulfolane.
- Optimize T-104 to reduce reboiler duty

- Lock in long term contracts with feedstock companies to prevent drastic price fluctuation.
- Further improve heat integration within the plant to reduce energy costs. E-101 is the most expensive heat exchanger and should be analyzed first.

Acknowledgements

Our team would like to acknowledge the representative at UOP who assisted us with our catalyst pricing and sizing. Without this information our project would have been less accurate and less effective.

Appendix

Reactor Train Detail

Overview

The reactor train section is made up of R-101, R-102, R-103, R-104, V-101, F-100, and T-101. This section represents the front end of the catalytic reforming plant and takes in Feed K. R-101, R-102, and R-103 represent the three main reactors. R-104 represents the swing reactor. F-100 is the furnace and V-101 and T-101 are the separator and stripper for this portion of the plant. Aspen HYSYS was the software used to model the reforming process. Peng Robinson was used as the fluid package to model the binary interactions in the reactors. The largest innovations for this section were the use of a swing reactor and the less harsh process conditions. Swing reactors are added to provide a spare reactor for catalyst maintenance[18]. The kinetic parameters for each reaction were calculated using equation [32] and the values are listed in Tables 21-24.

Cycloalkane Dehydrogenation	
Activation Energy (kJ/mol)	180
Rate Constant (kmol/hr*kmol) ⁻¹	0.35
Rate at Reactor Temperature (kmol/m ³ hr)	0.05

Table 21: Kinetic Parameters Cycloalkane Dehydrogenation

Kinetic Parameters Cycloalkane Cracking	
Activation Energy (kJ/mol)	145
Rate Constant (kmol/hr*kmol) ⁻¹	0.46
Rate at Reactor Temperature (kmol/m ³ hr)	0.06

Table 22: Kinetic Parameters Cycloalkane Cracking

Kinetic Parameters Alkane Cracking	
Activation Energy (kJ/mol)	170
Rate Constant (kmol/hr*kmol) ⁻¹	0.15
Rate at Reactor Temperature (kmol/m ³ hr)	0.09

Table 23: Kinetic Parameters Alkane Cracking

Kinetic Parameters Cycloalkane Cyclization	
Activation Energy (kJ/mol)	90
Rate Constant (kmol/hr*kmol) ⁻¹	0.27
Rate at Reactor Temperature (kmol/m ³ hr)	0.09

Table 24: Kinetic Parameters Cycloalkane Cyclization

Bare Module Cost

To enhance the safety of the process max allowable operating pressures were determined and applied to the design. Reactors were designed with a 50 psi addition to the design pressure[5]. This was done to prevent possible pressure buildups and safety issues. Process pumps were oversized to prevent abnormal conditions from upsetting the units. All pumps were provided a spare to decrease the down time of an upset.

Bare Module	USD (\$)
Catalyst	20,500,000
Reactors	3,062,597
Furnace	1,448,624
Heat Exchanger	466,636
Separators	19,579
Pumps	10,243

Table 25: Reactor Train Bare Module Cost 2021

Operating Cost	USD (\$/yr)
Waste Treatment	2,697,004
Maintenance	339,725
Utilities	212,143
Labor	79,623

Table 26: Reactor Train Yearly Operating Cost

Reactor Design and Catalyst Deactivation

A swing reactor was used to assist with the deactivation of the catalyst [18]. As previously mentioned, the catalyst takes up a majority of the reactor train investment. Maintaining the catalyst is vital to the plant's economics[5]. A swing reactor is an additional reactor outside of the three main reactors[19]. This reactor is put online when one of the main reactors, R-101-R-103 is down for catalyst maintenance. This allows the plant to operate continuously without down time. The swing reactor can also be used as a safety measure if one of the main reactors is having issues[22]. Our swing reactor R-104 will be used in this manner.

Catalyst deactivation is a common problem for many chemical reactors. Deactivation can occur from a variety of sources, but mainly results from carbon formation or sulfur poisoning[21]. Scrubbers are typically used to remove the sulfur from the sulfur from the feed. In our case a

scrubber is out of the project scope. To prevent catalyst deactivation our reactor temperatures were kept above 800°F in two of our reactors. The temperature in R-101 is lower, ~500°F, than the other two reactors. This could potentially lead to future catalyst deactivation. Our team analyzed several catalysts with Nickel additions to the crystal lattice. This process has lengthened the life of catalysts[23]. However, kinetic data could not be found for these catalysts and will be a future point of interest if the project moves into the detailed design phase.

Operating Conditions

The pressure drop in the reactors was calculated using equation[31]. Operating pressure was assumed to be ~17 psia based on[22]. Operating temperatures for the reactors were chosen to be over 1000°F by the final reactor[21]. Below this temperature the reactions occur at a very slow rate. At these temperatures conversion is the highest for the reactions and catalyst deactivation due to carbon formation is less likely[18]. The first and second reactor operating temperatures are 470°F and 840°F. Once the pressures and temperatures were set for the reactors the dimensions of the catalyst were input. These dimensions were taken from[20] and caused the simulation to converge. Reactors lengths were chosen from[20] and were optimized to increase the conversion of the desired products. The hydrogen recycle ratio was assumed to be 2/3 based on [21]. A larger amount of hydrogen present with the catalyst can lead to a higher rate of reaction[21].

Extractor Section Detail

Overview

The extractor section of the catalytic reforming plant is made up of T-102, T-103, and T-104. These represent the heavy liquid stripper and both extractors. Aspen HYSYS was the software used to model the reforming process. Per the memorandum NRTL VLE and LLE fluid packages were used to model the binary interactions. Sulfolane was used in the first extractor , T-102, as the solvent and water was used as the solvent in the second extractor, T-103. The largest innovation in the extraction section was the removal of the sulfolane water stripper. The bare module and operating costs are shown in Tables 27 and 28.

Bare Module Cost	USD (\$)
Heat Exchnagers	224059
Columns	210761
Pumps	149084
Extractors	134271

Table 27: Extractor Bare Module Cost 2021

Operating Cost	USD (\$/yr)
Waste Treatment	24,273,039
Utilities	137,964
Maintenance	123,537
Labor	63,699

Table 28: Extractor Yearly Operating Cost

Sulfolane Recovery

The original PFD for the catalytic reforming process included a stripper that would recover sulfolane from water. This stripper would recycle the sulfolane back into the process to be used as a solvent. Our original HYSYS model contained this stripper. The simulation could not converge because some of the water would get recycled along with sulfolane. Aqueous water in the reformate feed was not feasible with the VLE fluid package that was assigned to the tower. Before our team looked at removing the tower the economic attractiveness of sulfolane recovery was analyzed. The composition of sulfolane in the extractor stream was 2%.

The cost of the tower and the operating costs are shown in Table 27. The financial gain, on a yearly basis, of sulfolane is also shown. The column was estimated to cost \$135,000 with a yearly operating cost of \$98,000. Sulfolane benefit was calculated to be \$86,000 per year. This illustrates that the cost of the tower is not outweighed by the gain from recovering sulfolane. The decision was made to remove the tower to simplify the process, save money, and make the plant safer. Removing the column caused the simulation to converge due to the aqueous water in the reformate feed to be removed.

Sulfolane Recovery (\$/yr)	46208
Cost of Column (\$)	135,000
Column Operating Costs (\$/yr)	98000

Table 29: Sulfolane Yearly Benefit

First Extractor

T-102 was designed to extract product alkanes in sulfolane and send benzene and other valuable aromatic hydrocarbons downstream. To do this, the extractor feed, Stream 22 was cooled through E-103 to a desired temperature of 144 psia. Several model simulations were analyzed to determine the design temperature and pressure of the extractor. These references used feed temperatures in the range of 100°F-130°F [21] and [22]. The sulfolane recycle stream was cooled to a temperature of 107°F prior to entering the extractor. The pressure of the extractor was set to be 15 psia based on similar simulations. The pressure drop for each tray was taken to be 0.1 psi per tray[5] and an efficiency of 0.8 was used[5]. The temperature distribution based on tray location is shown below in Figure 8. Since, operating costs are not affected by a liquid liquid extractor only the number of stages was optimized. Through an iterative method the optimum point was determined to be 5 theoretical stages. Internal packing, in liquid liquid extractors, can lead to greater efficiency of the tower. For this reason pall rings were used for their cost effectiveness and higher efficiency compared to sieve trays.

Several issues occurred during our simulation of T-102. Each time the simulation would run Aspen HYSYS would show a fatal error. To combat this issue the binary interaction parameter of sulfolane and benzene was adjusted. Using[23] a value of 114 the simulation was able to converge. This parameter is calculated wrong in Aspen HYSYS and has the potential to cause future issues.

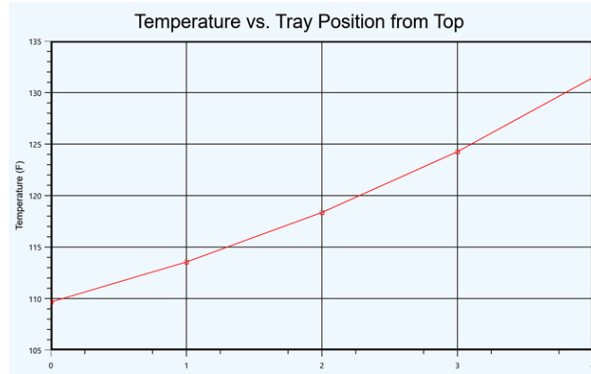


Figure 8: Temperature vs. Tray Position T-102

Second Extractor

T-103 was designed to extract product alkanes in water. A water flow rate of 1802 (lb/hr) was used as the solvent. The water feed entered the column at 90°F and 15 psia. Several model simulations were analyzed to determine the design temperature and pressure of the extractor. These references used feed temperatures in the range of 100°F-130°F [21] and [22]. The pressure of the extractor was set to be 30 psia based on similar simulations. The pressure drop for each tray was taken to be 0.1 psi per tray[5] and an efficiency of 0.8 was used[5]. The temperature distribution based on tray location is shown below in Figure 9. Since, operating costs are not affected by a liquid liquid extractor only the number of stages was optimized. Through an iterative method the optimum point was determined to be 5 theoretical stages. Internal packing, in liquid liquid extractors, can lead to greater efficiency of the tower. For this reason pall rings were used for their cost effectiveness and higher efficiency compared to sieve trays.

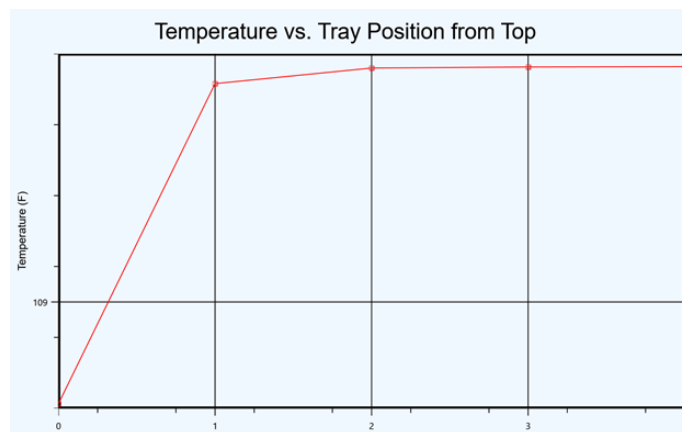


Figure 9: Temperature vs. Tray Position T-103

Heavy Liquid Stripper

The heavy liquid stripper, T-104, was idealized to maximize the recovery of benzene being sent to the reformate column. A 90% recovery of benzene was achieved. Reboiler pressure was specified to be 21 psia which set the temperature at 450 °F. Pressure drops across the reboiler was taken to be 2 psi [7]. The temperature distribution of the column based on tray position is included below in Figure 10. The column was optimized to have the lowest present worth cost. The optimum point was found to be 5 stages.

Tray efficiency for the column was set to 0.9 based on [5]. Our tower was given a 0.1 psi per tray[5] pressure drop. The reboiler heated the process stream to 450°F. High pressure steam was used to achieve this temperature. NRTL VLE fluid package was used to converge the simulation.

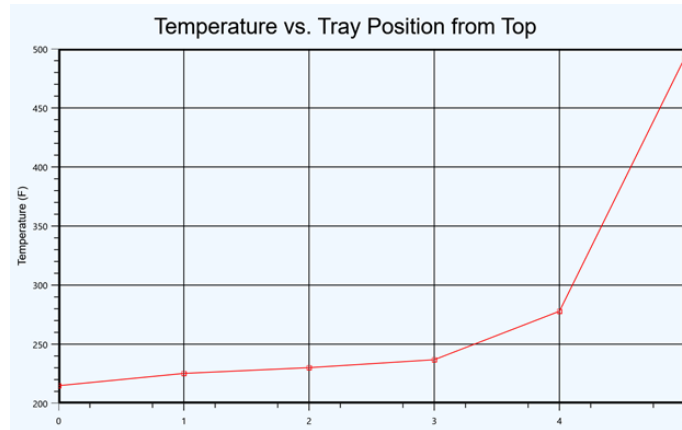


Figure 10: Temperature vs. Tray Position T-104

Distillation Section Detail

Overview

The distillation section of the catalytic reforming plant is composed of T-105, T-106, and T-107. These represent the reformate, benzene, and toluene-xylene distillation columns. Aspen HYSYS was the software used to model the reforming process. Per the memorandum NRTL VLE and LLE fluid packages were used to model the binary interactions. The 99% benzene specification was met before the process was fed to the benzene and toluene-xylene ,BTX, separators. The separation of BTX was extremely difficult due to the small difference in relative volatilities[23]. Due to this energy consumption was high and the towers were larger to achieve the desired separation. The largest innovation of the distillation section was the use of a divided wall column. Bare module and operating costs for the distillation section are shown in Tables XX and XX.

Equipment	USD (\$)
Columns	1,675,153
Heat Exchangers	591,034
Pumps	85,528
Reflux Drums	38,391
Seperators	20,238

Table 30: Distillation Bare Module Cost 2021

Operating Cost	USD (\$/yr)
Utilities	226,041
Maintenance	154,421
Labor	47,774
Waste Treatment	0

Table 31: Distillation Yearly Operating Cost

Reformate Column

The reformate column, T-105, was idealized to achieve a 99% composition of benzene in the bottoms of the tower. The distillate stream was recycled in the process to extract any remaining valuable components. Cooling water was used in the condenser. Condenser pressure was specified to be 60 psia which set the temperature at 199 °F. For this process cooling water was supplied at 77 °F and returned at 120°F which led to an accurate approach temperature of 79°F. Pressure drops across the condenser and reboiler were taken to be 3 psi and 2 psi[7]. The temperature distribution of the column based on tray position is included below in Figure 11.

Tray efficiency for the column was set to 0.9 based on [5]. Our tower was given a 0.1 psi per tray[5] pressure drop. The reboiler heated the process stream to 306°F. Heat integration was used to eliminate the need for a reboiler. Stream 55 was used to provide the needed heat duty to achieve the desired temperature.

The largest problem we had with T-105 was aqueous phases present in the feed to the column. This would not allow the column to converge with a viable solution. To combat this problem E-106, reformate feed heat exchanger, was optimized to provide the desired process conditions[7]. This method paired with the removal of the sulfolane recycle tower led to the convergence of the tower. Another issue of the column was achieving the desired purity of benzene. Originally a reactor temperature of 800°F was used, but this led to unreacted cyclohexane. This caused the benzene composition to be 92%. The reactor temperature was raised and a separator was added after the column. This separator flashed the stream leading to a 99% purity of benzene in the distillate. Adding the separator was justified because the heat exchangers, on the tower, were consuming 45% more energy per hour. Adding the separator led to a savings of \$80,000 per year.

T-105 was designed with sieve trays due to their flexibility and low cost[23]. To optimize the column the number of stages was varied from 3 to 20. The reflux ratio was calculated in HYSYS. The optimization point was determined to be 5 stages with a reflux ratio of 2. This minimized the present worth cost of the tower.

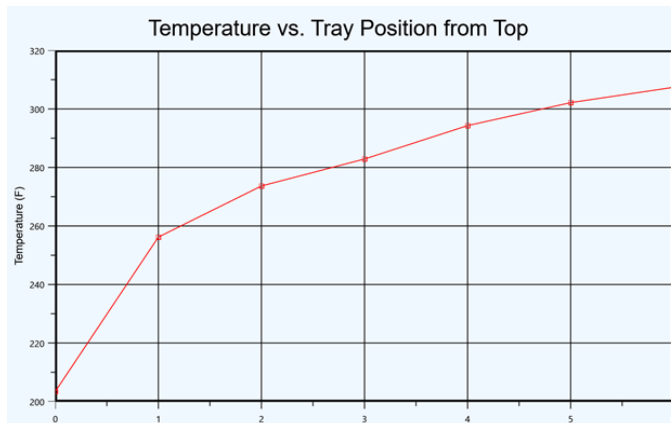


Figure 11: Temperature vs. Tray Position T-105

Benzene Column

The benzene column, T-106, was idealized to maximize the recovery of benzene in the distillate. Benzene has the highest product pricing, \$3.49/gal so our team's goal was to maximize this stream. A composition of 99% benzene was achieved along with a 99% recovery of the benzene from the reformat tower. T-106 was calculated to be the largest fractionator by volume. Our goal for the distillation section was to optimize the major fractionator.

Cooling water was used in the condenser and medium pressure steam, 150 psig, was used in the reboiler. Condenser pressure was specified to be 40 psia which set the temperature at 240 °F. For this process cooling water was supplied at 77 °F and returned at 120°F which led to an approach temperature of 120°F. Pressure drops across the condenser and reboiler were taken to be 3 psi and 2 psi[7]. The temperature distribution of the column based on tray position is included below in Figure 12. Tray efficiency for the column was set to 0.9 based on [5]. Our tower was given a 0.1 psi per tray[5] pressure drop. The reboiler heated the process stream to 345°F.

The largest problem we had with T-106 was an extremely high energy cost. Both the condenser and reboiler contributed to 45% of the distillation utility cost. To address this problem a divided wall column was used. Divided wall columns have become common in newer plant designs and can reduce yearly energy costs by 30%[20]. Divided wall columns are more expensive than traditional distillation towers from an initial capital investment standpoint, but can make up that cost difference in the first year[20]. Switching to a divided wall column led to a 25% more expensive column, but led to a lower present worth cost.

T-106 was designed with sieve trays due to their flexibility and low cost[23]. To optimize the column the number of stages was varied from 10 to 30. The reflux ratio was calculated in HYSYS. The optimization point was determined to be 15 stages with a reflux ratio of 2.7. This minimized the present worth cost of the tower.

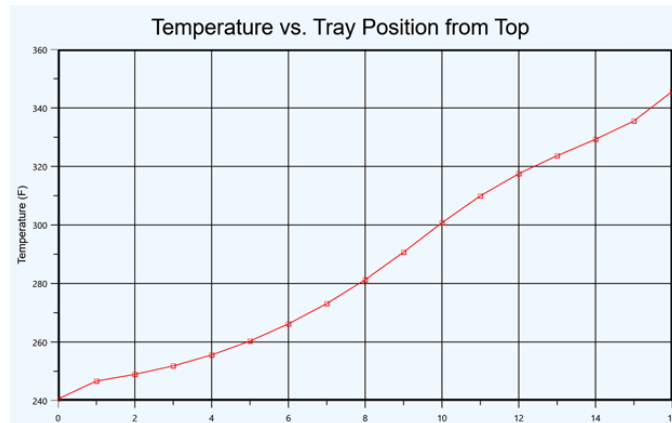


Figure 12: Temperature vs. Tray Position T-106

Tol-Xylene Column

The toluene-xylene column, T-107, was optimized to recover the largest amount of toluene in the distillate and the largest amount of xylene in the bottoms. Cooling water was used in the condenser and high pressure steam, 450 psig, was used in the reboiler. Condenser pressure was specified to be 60 psia which set the temperature at 337 °F. For this process cooling water was supplied at 77 °F and returned at 120°F which led to an approach temperature of 217°F. Pressure drops across the condenser and reboiler were taken to be 3 psi and 2 psi[7]. The temperature distribution of the column based on tray position is included below in Figure 13.

Tray efficiency for the column was set to 0.9 based on [5]. Our tower was given a 0.1 psi per tray[5] pressure drop. The reboiler heated the process stream to 402°F.

T-107 was designed with sieve trays due to their flexibility and low cost[7]. To optimize the column the number of stages was varied from 7 to 20. The reflux ratio was calculated in HYSYS and was created. The optimization point was determined to be 10 stages with a reflux ratio of 6.1. This minimized the present worth cost of the tower.

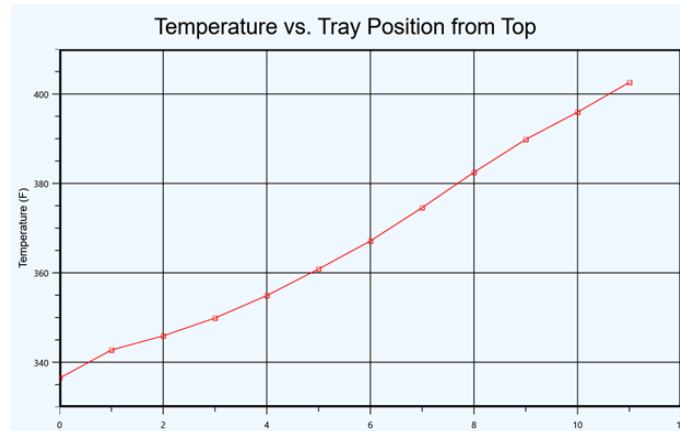


Figure 13: Temperature vs. Tray Position T-107

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Equations
Heat Exchanger

$$1) \quad Q = U_o A_o \Delta T_{lm} F$$

$$2) \Delta T_{lm, counterflow} = \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\left(\frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)^{-1}}$$

$$3) U_o = \left(\frac{d_o}{h_i d_i} + R_{f,i} \frac{d_o}{d_i} + \frac{d_o}{2k_w} \left(\frac{d_o}{d_i} \right) + R_{f,o} + \frac{1}{h_o} \right)^{-1}$$

$$4) P = \frac{T_{ho} - T_{hi}}{T_{ci} - T_{hi}}$$

$$5) R = \frac{T_{ci} - T_{co}}{T_{ho} - T_{hi}}$$

Pump

$$6) P_d = P_{Tower} + \frac{\rho g h_1}{g_c} + \Delta P_{Line Loss} + \Delta P_{Valves}$$

$$7) P_s = P_{Condenser} + \frac{\rho g h_2}{g_c} - \Delta P_{Line Loss} - \Delta P_{Va}$$

$$8) \Delta P = P_d - P_s$$

$$9) H = \frac{2.31 * \Delta P}{S.G.}$$

$$10) hhp_{Actual} = \frac{Q_{Actual} * \Delta P}{1715}$$

$$11) hhp_{Run Out} = \frac{Q_{Run Out} * \Delta P}{1715}$$

$$12) bhp_{Actual} = \frac{hhp_{Actual}}{\eta_{Pump}}$$

$$13) bhp_{Run Out} = \frac{hhp_{Run Out}}{\eta_{Pump}}$$

$$14) Actual hp = \frac{bhp_{Run Out}}{\eta_{Motor}}$$

Reflux Drum

$$15) V = \frac{\pi}{4} D^2 L$$

Capital Costing

$$16) C_p = C_p^o F_m F_p$$

$$17) \log C_p^o = K_1 + K_2 \log A + K_3 [\log A]^2$$

$$18) F_p = \frac{\frac{(P+1)^D}{2[850-0.6(P+1)]} + 0.00315}{0.0063}$$

$$19) \log F_p = C_1 + C_2 \log P + C_3 [\log P]^2$$

$$20) C_{BM} = C_p^o (B_1 + B_2 F_m F_p)$$

$$21) C_{BM} = C_p^o N F_{BM} F_q$$

$$22) C_{TM} = 1.18 \sum C_{BM}$$

$$23) Cost in Year A = Cost in Year B \left[\frac{Cost Index in Year A}{Cost Index in Year B} \right]$$

Manufacturing Costs

$$24) COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM})$$

Operating Labor

$$25) N_{OL} = (6.29 + 31.7P^2 + 0.23N_{np})^{0.5}$$

Utilities

$$26) Cost of Electricity = Purchased HP (Cost of Electricity)$$

$$27) Q = \dot{m}_{steam} \Delta H_{vap}$$

$$28) Annual Cost of Steam = Q(Cost of Steam)(Service Factor)$$

$$29) Q = \dot{m}_{water} \Delta T_{p,water}$$

$$30) Annual Cost of Cooling Water = Q(Cost of Cooling Water)$$

Reactor

$$31) \frac{\Delta p}{L} = \frac{150\mu(1-\epsilon)^2 u_0}{\epsilon^3 d_p^2} + \frac{1.75(1-\epsilon)\rho u_0^2}{\epsilon^3 d_p}$$

$$32) k = Ae^{-\frac{E_a}{RT}}$$

$$33) V = \int_{F_{A0}}^{F_A} \frac{dF_A}{r_A}$$

Extractor

$$34) U_i = 4Vi / D_T$$

$$35) U_D / U_C = m_D / m_C * (p_C / p_D)$$

$$36) D_T = (4m_D / p_D U_D)^{0.5}$$