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Toppings Refinery Retrofit

by

Conner Meyer Gullledge

An Honors Capstone

submitted in partial fulfillment of the requirements

for the Honors Diploma

to

The Honors College

of

The University of Alabama in Huntsville

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2020-2021 AIChE Student Design Competition

Toppings Refinery Retrofit

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Executive Summary

Our engineering team has been commissioned to investigate the development of a crude fractionation facility in order to meet the demand for gasoline in Iraqi Kurdistan. As western refining standards are quickly becoming the gold standard in Iraq, benzene, toluene, and xylene are to be removed and sold as additional byproducts. In this study, we investigate the profitability of a catalytic cracking and fractionation facility, accounting for utilities costs, market prices of products, and equipment costs. By achieving western standards for aromatic concentrations in fuels, the available market expands to include exportation. As such, product specifications are to be a minimum of US standards.

The following documentation presents the project design and means to accomplish the product specification goals in an economically viable manner. Each section of the refinery is broken down in the process flow diagrams included in the process description section. Additionally, an emphasis is placed on the identification of hazards in the process, attributed to the use of hazardous chemicals, through a What-If analysis safety review as a part of the project safety management program. Each consequence was characterized for each hazardous scenario and a proposed control method was established in the event that the scenarios were to occur. Finally, the retrofit design is assessed for economic feasibility and compliance with new local regulations to separate salable benzene without loss in revenue.

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Process Description

The process begins with the reactor system, as shown in **Figure 1**. The naphtha fraction from 35,000 barrels of oil per day (stream 1) is catalytically cracked at high temperatures to produce lower molecular weight alkanes, marketable as gasoline and diesel. In addition to aromatics already in the system, benzene is produced from cyclohexane present in the naphtha. The resulting mixture is flashed, producing a hydrogen-rich gas (stream 3) that is utilized in facility operations. The bottoms of the flash vessel go on to the extraction section.

The main objective of the extraction section, shown in **Figure 2**, is to separate aromatics from the other hydrocarbons. The mixture first enters the major fractionator, which removes C₁-C₄ alkanes (stream 5) that are also utilized in facility operations. Liquid-liquid extraction (LLE) is achieved with sulfolane solvent, which selectively removes aromatics from the mixture. The remaining hydrocarbons, which soak in some of the solvent, are separated from that solvent to form a mixture of water and C₅-C₁₀ alkanes (stream 9), which is to be decanted and fractionated into diesel and gas with an existing toppings refinery. The aromatics are distilled from the solvent and residual alkanes are removed, with these components being returned to the LLE, and the aromatics proceed to the distillation section.

As seen in **Figure 3**, the distillation section purifies the aromatics mixture into its constituent fractions. Benzene is first distilled from the BTX to 99% purity (stream 17). The remaining aromatics, toluene and xylenes, are both separated to 99% purity afterward (streams 19 and 20, respectively). The conditions of important input and output streams are shown in **Table 1**. A full list of stream compositions is available in the appendix.

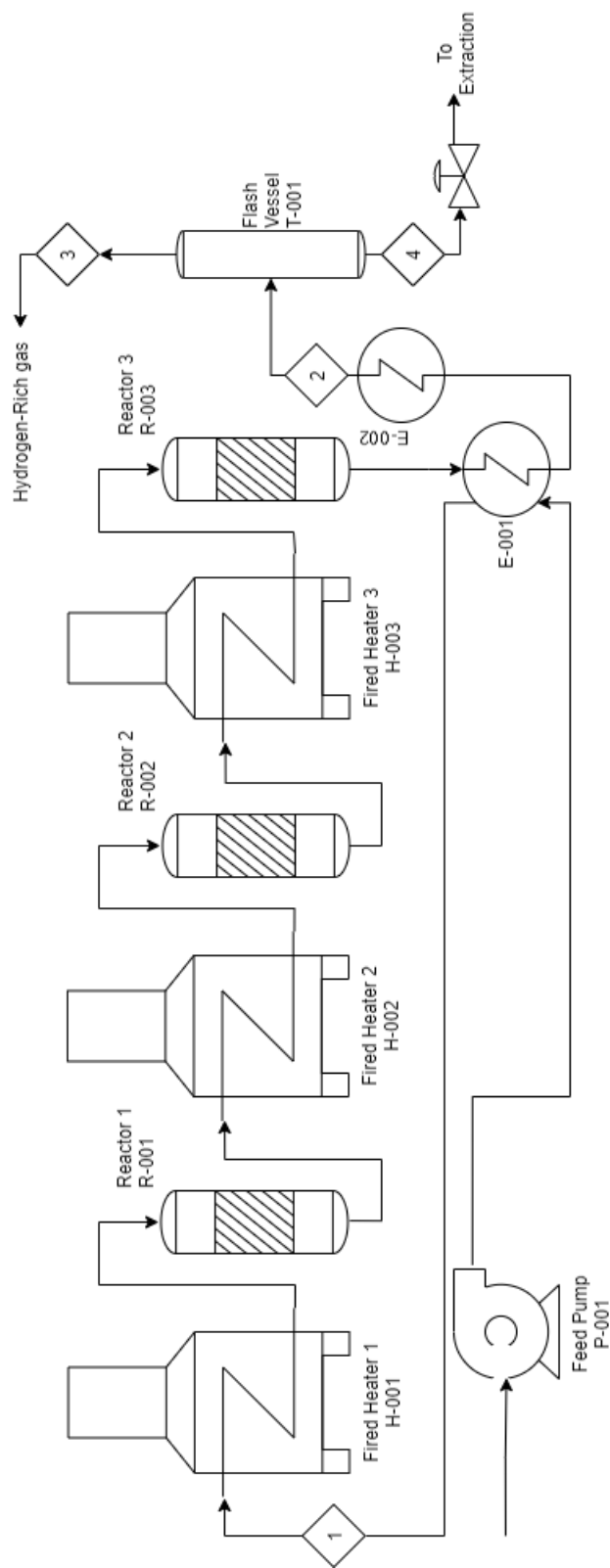


Figure 1: Reactor Section Process Flow Diagram

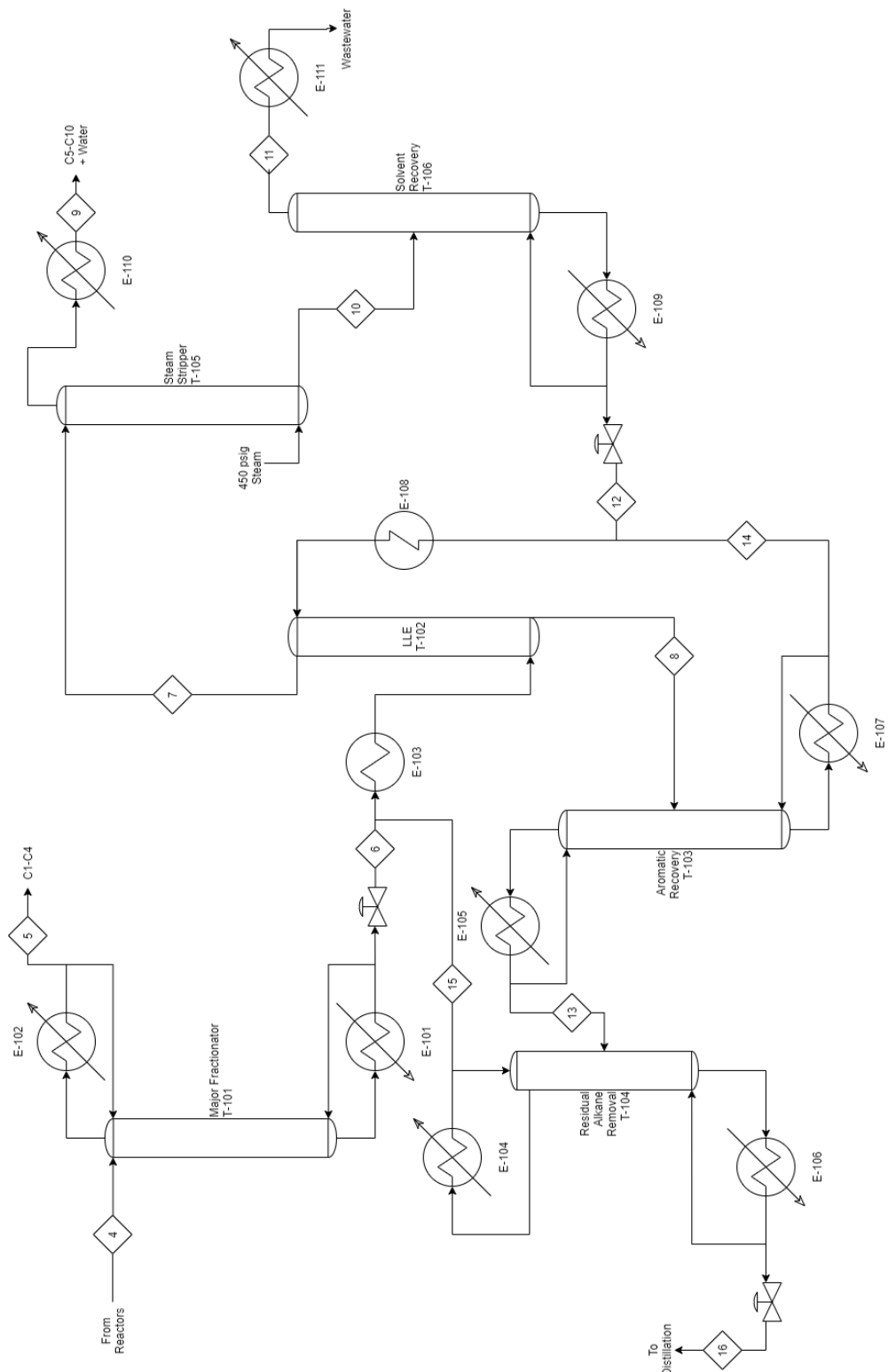


Figure 2: Extraction Section Process Flow Diagram

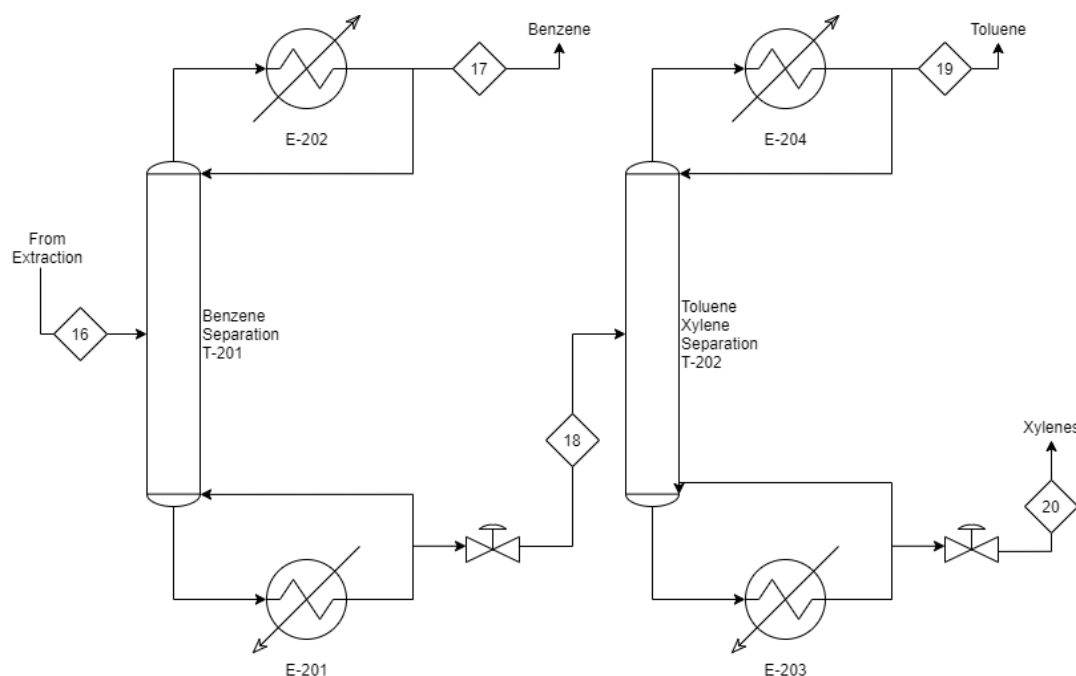


Figure 3: Distillation Section Process Flow Diagram

Table 1: Conditions of Major Process Inputs & Outputs

Stream	Temp (°C)	Flow (kmol/h)	Composition
1 (Input)	252	294	59.7% Decane, 31.3% Cyclohexane, 9% Benzene
3 (H ₂ -Rich)	38	258	92.3% Hydrogen, 2.6% Methane, 2.2% Ethane, 1.2% Propane, 0.9% Benzene, 0.5% Butane, 0.3% Pentane
5 (C ₁ -C ₄)	51.7	13.1	44.5% Butane, 30.6% Propane, 11.1% Ethane, 11.1% Hydrogen, 2.4% Methane, 0.1% Pentane
9 (C ₅ -C ₁₀) /H ₂ O	30	339	47.3% Water, 43.1% Decane, 1.9% Nonane, 1.9% Octane, 1.9% Pentane, 1.9% Heptane, 1.9% Hexane
17 (Benzene)	78.3	36.2	99.4% Benzene, 0.4% Heptane, 0.1% Cyclohexane
19 (Toluene)	110	49.4	98.5% Toluene, 1.2% Xylenes, 0.2% Octane, 0.1% Benzene
20 (Xylenes)	138	19.9	98.6% Xylenes, 1.3% Decane, 0.1% Nonane

Economic Analysis and Sensitivity Analysis

Any analysis of a profit-seeking venture must also count the cost, so to speak, in order to determine the viability of the project prior to construction. This document contains the anticipated Fixed Capital Investment, Utilities costs, Labor expectation, Taxes under the Kurdish and Iraqi regimes, and the projected Internal Rate of Return, and Return on Investment. In addition, a sensitivity analysis demonstrates the probability of achieving a profitable outcome and the factors to which the final profitability is most sensitive.

The Major Equipment will include the Reactor Train, the Furnaces, 6 Still columns named in this section as Stills 1-6, and two extractors. The estimation of the fixed equipment costs is outlined below:

- Reactor Train: \$ 7,830,000
- Extractor Train: \$ 5,234,000
- Distillation Train: \$ 772,000
- Total: \$ 13,836,000

Depreciation is performed under a 16-year MACRS schedule. As the majority of capital expenditures fall within this life cycle, the analysis of the facility was determined according to this timeframe. MACRS was chosen in order to more quickly recapture equipment cost and improve cash flow in the early years of the facility.

Labor costs were estimated according to US standards for wages. According to Glass Door, 2021 Chemical Operators make between \$59,000-\$63,000 each year. The high end of this metric was utilized for our analysis. In order to determine the number of operators required to operate the facility, the number of each manned piece of equipment was counted and multiplied

by the proportion of each shift that piece of equipment would require. This information can be found in **Table 2** below.

Table 2: Equipment Operator Information

Equipment	Proportion of Shift	Pieces	Total Operator Time Required
Furnace	.5	3	1.5
Heat Exchanger	.1	1	.1
Reactor	.5	4	2
Tower	.35	8	2.8
Flash Separator	.3	1	.3
Cooler	.5	3	1.5
		Total	8.2

As 4.5 operators are required to maintain each required shift over the course of the year, this results in a required 36.9 operators needed. To meet this, 39 operators are requested, including 3 working supervisors to supervise each daily shift.

The utilities expenditures during normal operation were estimated to cost \$836,4773 per year. Heat integration has been investigated in order to improve the total utilities cost, though much of the utilities used in separations are required, limiting the impact such projects have had.

A full Free Cash Flow analysis is appended at the end of this document containing a detailed breakdown of the costs and revenues expected for our facility. From this analysis, under the Iraqi tax scheme our annual ROI was determined to be 84.11%, our IRR was determined to be 39.35%, and our NPV according to discount rates comparable to rates of return for similar facilities was \$359,764,871. Under Kurdish taxation, our ROI was determined to be 112.31%, our IRR was determined to be 44.45%, and our NPV according to discount rates comparable to

rates of return for similar facilities was \$488,938,022. This demonstrates that the facility has great potential to be profitable given existing market conditions persist. A plot of the NPV with respect to time is shown in **Figure 4** below. It is worth noting that profits increase substantially under the Kurdish tax system.

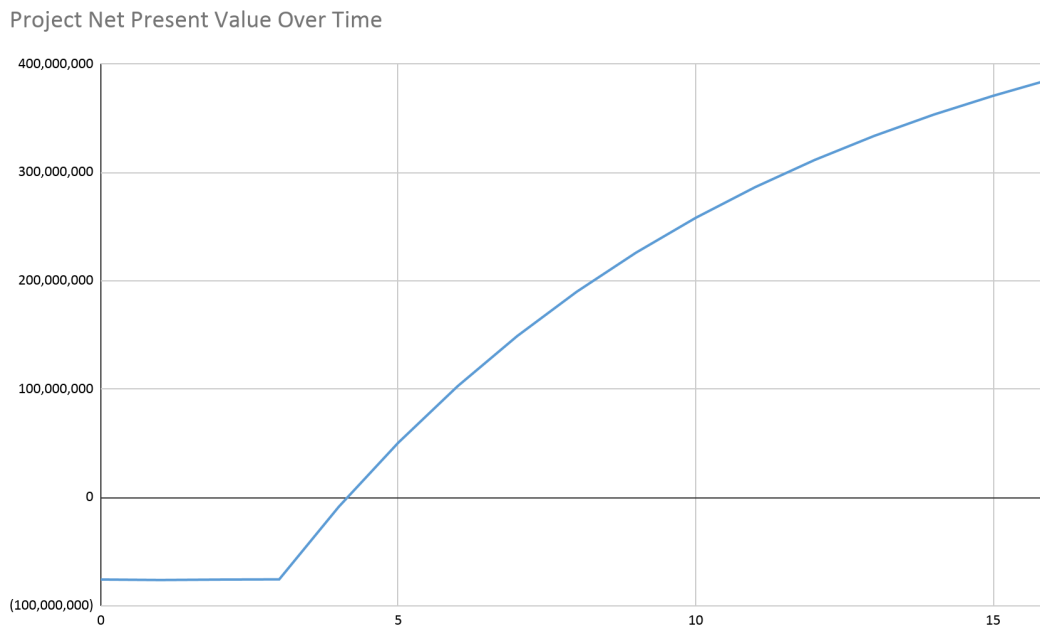


Figure 4: Net Present Value Over Time

All proposed improvements upon the original design were analyzed using an INPV analysis, using the IRR for the base project as the discount rate. In this manner, cost-saving modifications were tested to determine that the savings improved overall process profitability. Such improvements proposed include the installation of an economizer on the reactors with an INPV of \$3,263,043. This indicates that these proposals would result in a significantly more profitable facility overall and are thus recommended for inclusion in the final proposal.

In order to determine the sensitivity of the project's profitability to external changes in market conditions, a sensitivity analysis was performed using a 10% absolute deviation while holding all else constant. A graphical representation of this analysis is shown in **Figure 5**.

Sensitivity Analysis

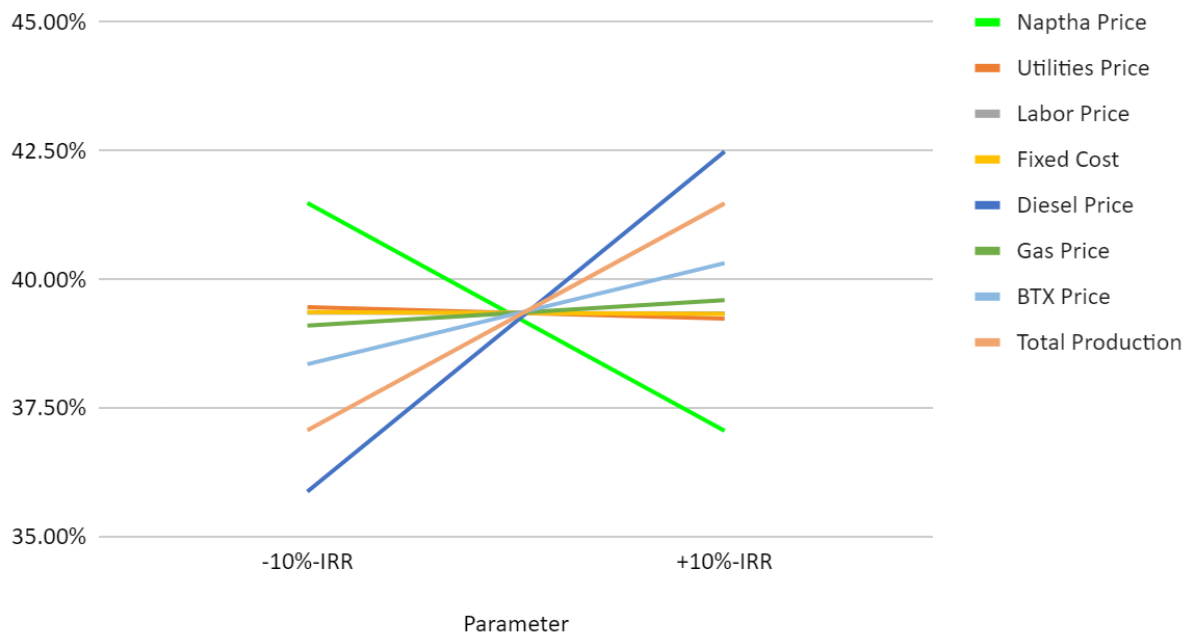


Figure 5: Cost Sensitivity Analysis

A Monte Carlo simulation was also performed using a 10% maximum absolute deviation in each substantial parameter in order to approximate the probability of deviation from expected profitability. The internal rate of return is plotted in a bell curve in **Figure 6** below.

Monte Carlo Histogram

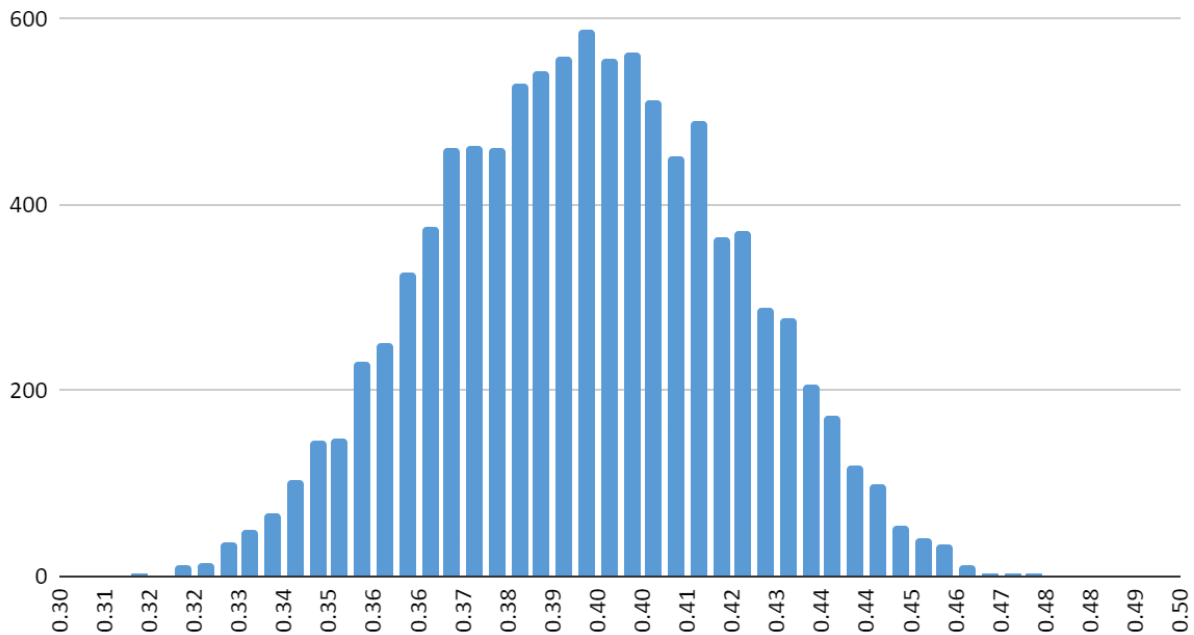


Figure 6: Monte Carlo Method

The Monte Carlo provides a Standard Deviation of 2.63%, indicating that there is approximately a 95% chance that the realized rate of return will be in excess of 34.08%. These Analyses suggest that this project is likely to turn a profit, though there is major risk of losses associated with changes in the price of naphtha and the price of diesel. Unfortunately, this predominant risk is not practical to mitigate.

Process Safety

When designing a plant, it is crucial to evaluate the environmental, health, and safety aspects to ensure a practical, maintainable operation. Process safety centers on the prevention of fires and explosions, hazardous chemical release, and workplace injury. Inherently safer design addresses safety and health concerns that can be prevented by changing the design of a system through substitution of less hazardous chemicals, minimization of chemical amounts or equipment sizes, moderation through dilution of chemicals, and simplification of plant facilities. Another important tool for designing a sustainable and responsible plant is the application of process safety management programs as regulated by the Occupational Safety and Health Administration (OSHA). Many of the decision-making processes during the design of a plant facility are governed by inherently safer design and process safety management as will be discussed further in the following sections.

Inherently Safer Design

As previously mentioned, inherently safer design is a method of eliminating hazards through substitution, minimization, moderation, and simplification within a chemical plant facility. It is important to note that while inherently safer design may produce a much safer work environment, it does not eliminate all hazards and it is necessary to employ a process safety management program to identify remaining hazards in the plant. The first strategy for eliminating hazards in a plant is the substitution of less hazardous materials, processes, and equipment. Some prime substitution examples for this process include using welded pipe over flanged pipe, less toxic solvents and chemicals (in our case, using sulfolane over DMSO), mechanical gauges over mercury, water over hot oil for heat transfer, etc. Through the utilization

of this engineering control, select process hazards can be permanently eliminated before they become a threat to workers and property.

The second strategy to produce an inherently safer process design involves minimization of hazards associated with chemicals and equipment through employment of chemical dilution or equipment downsizing. Examples of minimization include switching from a large batch reactor to a small continuous reactor, reducing inventory of hazardous raw materials, and improving control systems to prevent holdup of hazardous intermediate chemicals. By employing a continuous process instead of a batch process, the holdup of potentially hazardous chemicals within the refinery is minimized. Additionally, storage of hazardous chemicals should be limited to avoid potential fires or explosions within the plant; the storage vessels should also be grounded and bonded. Maintaining control systems in a clear and concise manner also proves to be an effective method for developing a safe and efficient process.

Moderation is the third technique for producing an inherently safer process design, and it involves using process conditions that are less hazardous and reducing the concentration of hazardous chemicals. This strategy of engineering control involves placing control rooms far away from operation, refrigerating storage vessels rather than using high pressure, operating at conditions that prevent runaway reactions, and diluting hazardous chemicals to limit the effects.

Finally, the last strategy for eliminating inherent process hazards includes simplification of process details for a user-friendly facility. This technique includes clear labeling of piping, vessels, controls, and design panels as well as choosing equipment with a less frequent maintenance schedule and lower failure rate. HMI panels are designed to be clearly read with flashy colors used only for alarms and warnings. The goal is to place a lot of the responsibility on the control system rather than on the operators to prevent catastrophic events.

Process Safety Management

Process Safety Management (PSM) is a regulation developed by OSHA soon after the Bhopal accident in 1984 that defines the requirements for managing hazardous chemicals within a facility. There are fourteen major sections of the standard with categories including but not limited to process safety information, process hazard analysis, operating procedures, and emergency planning and response. The process safety information section focuses on providing documentation to all employees and operators to facilitate a plant-wide understanding of the hazards present. This documentation usually includes process flow diagrams, process chemistry, and process temperatures, pressures, flows, and compositions. Once this documentation is developed, a process hazard analysis (PHA) is conducted by a team of experts - engineers, chemists, industrial hygienists, operators, etc. - based on the complexity of the process. For more complex processes a HAZOP study is conducted, whereas for less complex processes a what-if analysis or failure mode and effects analysis may be conducted. It is necessary for a PHA to be conducted at least every five years. With the use of the information obtained from the process hazard analysis, emergency planning and response documentation can be developed so employees and operators have the knowledge and training in the event of a process upset or other emergencies. In **Table 3**, a list of all chemicals present throughout the refinery with associated safety and hazard information is included. For this process, a what-if analysis was conducted to identify and provide solutions for potential hazards and process upsets for the major fractionator. This information can be found in **Table 5**.

Table 3: Chemicals and Associated Hazards

Chemical	OSHA PEL	NIOSH REL	NFPA Diamond				LD ₅₀ Limits	GHS Signal Word
			Health	Fire	Instability	Special		
iso-butane	-	800 ppm TWA	1	4	0	-	Oral: > 5,000 mg/kg Dermal: > 5,000 mg/kg	Danger
benzene	1 ppm TWA	0.1 ppm TWA	2	3	0	-	Oral: > 2,000 mg/kg Dermal: > 8,260 mg/kg	Danger
cyclohexane	300 ppm TWA	300 ppm TWA	1	3	0	-	Oral: > 5,000 mg/kg	Danger
hydrogen	-	-	0	4	0	-	No data available	Danger
methane	-	1000 ppm TWA	2	4	0	-	No data available	Danger
n-decane	500 ppm TWA	-	1	2	0	-	Oral: > 5,000 mg/kg Dermal: > 5,000 mg/kg	Danger
para-xylene	100 ppm TWA	100 ppm TWA	2	3	0	-	Oral: > 3,426 mg/kg Dermal: > 5,000 mg/kg	Danger
pentane	1000 ppm TWA	120 ppm TWA	1	4	0	-	No data available	Danger
sulfolane	-	-	1	1	0	-	Oral: > 2,143 mg/kg Dermal: > 2,000 mg/kg	Danger
toluene	200 ppm TWA	100 ppm TWA	3	3	0	-	Oral: > 5,000 mg/kg Dermal: > 5,000 mg/kg	Danger
water	-	-	0	0	0	-	Oral: > 90,000 mg/kg	NA
ethane	-	-	1	4	0	-	-	Danger
propane	1000 ppm TWA	1000 ppm TWA	2	4	0	-	-	Danger
butane	800 ppm TWA	800 ppm TWA	1	4	0	-	-	Danger
hexane	500 ppm TWA	50 ppm TWA	2	3	0	-	Oral: > 25 g/kg Dermal: > 3,000 mg/kg	Danger
heptane	-	85 ppm TWA	1	3	0	-	Oral: > 5,000 mg/kg Dermal: > 3,000 mg/kg	Danger
octane	300 ppm TWA	75 ppm TWA	3	3	0	-	Oral: > 5 g/kg Dermal: > 2 g/kg	Danger
nonane	200 ppm TWA	200 ppm TWA	3	3	0	-	-	Danger

Best Available Control Technology (BACT) is a standard mandated by the Clean Air Act for pollution control. It requires facilities to use the most stringent measures they are capable of to reduce their emissions. A distillation column, such as the “Major Fractionator” in the process shown in this paper, naturally requires the application of BACT in order to reduce the chance of an accidental release. While “last line of defense” safety measures such as pressure relief valves are in place to prevent incidents, these measures also result in the partial release of chemicals. With the implementation of BACT, abnormalities can be resolved far before accidental release conditions are met.

The Major Fractionator receives the post-flash mix of hydrocarbons from the catalytic reforming process, and it removes methane, ethane, propane, and butane. With the separation of butane and pentane being as difficult as it is, the Major Fractionator must be significantly large to compensate. It is therefore the most critical unit operation in the system to develop safety controls for. From the perspective of BACT, this column has the potential for the largest amount of pollution out of nearly any operation in the system, and from the perspective of safety at large, it holds the potential for the most devastating incident in the form of deflagration. Because of this, a controls system must be in place to maintain design pressure, fluid level, and temperature, as is shown in **Figure 7**.

A pressure sensor is installed at the top of the column, as this is the most important location for pressure control with the main relief valve located here. A pressure transmitter sends pressure data to a pressure-indicating controller, which in turn manipulates an actuator on the condenser’s cooling water control valve. Increasing the flow of cooling water increases the rate of condensation, lowering the pressure at the top of the column. Alarms for abnormally low and abnormally high pressures are integrated directly into the pressure control system.

Temperature regulation occurs at the bottom of the column, where the largest temperatures will be present. It is important to regulate the temperature in this section of the column, as abnormal temperatures may not distribute up the length of the column within the time frame in which abnormalities must be addressed. Utilizing a similar system to the pressure control system, temperature control is achieved by directly manipulating the heat applied to the system. Utilizing temperature data from the bottom of the column, the actuator in the temperature control system adjusts the flow rate of 450 psig steam into the reboiler, allowing heat input to be minimized in the event that temperature rises above normal conditions. This system also contains alarms for abnormally high and abnormally low temperatures.

Level regulation also occurs at the bottom of the column, as measuring fluid pressure at the bottom of the column is the most convenient method of determining fluid height. Column fluid level is conveniently controlled by manipulating the flow rate out of the bottom of the column. A similar system exists for the fluid level in the condenser drum. Drum fluid level is regulated by manipulating the flow of distillate back into the column. Alarms for abnormally high and abnormally low fluid levels are installed in both of these control systems.

A control system for pressure regulation in the condenser drum is also present, but the principle used for control is different. The condenser drum is positioned after the heat exchanger, so controlling pressure by condensation is not an option. The drum pressure control instead manipulates the outlet flow of vapor to keep pressure stable. In the event that this system fails to adequately control pressure, the drum's pressure relief valve will serve as a final measure against overpressurization. As with all of the control systems, alarmed indicators are in place for abnormally high and abnormally low conditions.

Depending on the condition controlled and the variable manipulated to achieve said control, control valves should be set to fail into either the open or closed position to ensure that the least hazardous conditions will be presented. Level control valves should fail open, as low fluid levels are significantly less hazardous than high fluid levels. The same can be said for the pressure control valves, as the open position yields a reduction in pressure for both systems. The only valve that should fail into the closed position is the temperature control valve, as this will cut off the addition of heat into the process entirely.

A final consideration for safety instrumentation is the possibility of utility feed disruption. Hazardous conditions can result if the availability of steam or cooling water changes drastically, and these factors may well be outside the control of our control systems. To compensate for this, two alarmed flow indicators are present. A high flow alarm is installed at the steam inlet, as the over addition of heat to the system is the most hazardous scenario. Conversely, a low flow alarm is fitted to the cooling water inlet, as a lack of cooling is also especially hazardous.

The pressure relief valve on the column can be sized with the following equation

$$W = K_d A P_1 \sqrt{\frac{\gamma g_c M}{R_g T} \left(\frac{2}{\gamma+1}\right)^{(\gamma+1)(\gamma-1)}}$$

where W is the mass flow rate, K_d is the discharge coefficient (which can be assumed to be 1 since choked flow is assumed for gases), P_1 is the upstream relieving pressure, γ is the heat capacity ratio, M is the molecular weight, R_g is the ideal gas constant, and T is the absolute temperature. With an upstream relieving pressure of 175 psig, the area of the orifice of the pressure relief valve was calculated to be 56.0 mm².

Deflagration is the phenomenon that occurs when an explosion occurs where the reaction front moves slower than the speed of sound in the unreacted medium. Most of the time,

deflagration occurs when flammable gases and vapors do not find an ignition source and rather begin to develop overpressure in the system. Sometimes this may lead to detonation, but this does not always occur. An effective method for understanding the energy associated with a vapor cloud deflagration is the TNT equivalency method. This method assumes that an exploding fuel mass will behave similarly to TNT on an equivalent energy basis. Additionally, the mathematical relationship is based on empirical overpressure curves. The TNT equivalency relationship is shown below with m_{TNT} being the equivalent mass of TNT, η being the empirical explosion efficiency, m is mass of hydrocarbon, ΔH_c is the energy of the explosion of flammable gas which can be estimated to be the heat of combustion, and E_{TNT} is the energy of explosion of TNT.

$$m_{TNT} = \frac{\eta m \Delta H_c}{E_{TNT}}$$

Using the fraction of components being fed into the major fractionator, masses for each component can be calculated to be used in the TNT equivalency formula. This value as well as the upper and lower explosive limits of each component are included in **Table 4**. The TNT equivalency values assume an explosion efficiency of 2% and a TNT explosion energy of about 4686 kJ/kg.

Table 4: TNT Equivalency and Upper/Lower Explosive Limits

Chemical	Equivalent Mass of TNT (g)	Lower Explosive Limit	Upper Explosive Limit
		%vol Fuel in Air	%vol Fuel in Air
n-Decane	5.06E-01	0.8	5.4
Cyclohexane	5.32E-04	1.3	8
Benzene	2.53E-01	1.4	7.1
Methane	4.89E-04	5	15
Ethane	3.81E-03	3	12.5
Propane	1.32E-02	2.1	9.5
n-Butane	2.48E-02	1.8	8.5
n-Pentane	3.39E-02	1.4	7.8
Hydrogen	2.32E-04	4	75
n-Hexane	3.81E-02	1.2	7.5
n-Heptane	4.51E-02	1	7
n-Octane	5.15E-02	0.8	6.5
n-Nonane	5.77E-02	0.7	5.6

Based on the scaling law, the side-on overpressure approximately 100 ft away from the major fractionator will be about 5.47 kPa which is enough to cause minor damage to house structures along with shattering of windows. At approximately 200 ft away from the major fractionator, a blast is expected to cause limited minor structural damage. With these values in mind, the safety of the nearby operators must be considered. A what-if analysis can be found in **Table 5** to detail hazards, consequences, and potential control methods to limit the hazards present for operators near the major fractionator.

Table 5: What-If Analysis of Major Fractionator

What-If Question	Hazardous Scenario	Consequence	Control Method
What if excessive pressure builds up in the major fractionator?	Plug in the system with continued normal operation results in pressure buildup.	Pressure vessel burst or boiling liquid vapor expansion explosion.	Pressure sensor installed with alarms for abnormal pressure conditions. Relief valve installed.
What if the vent valve is left open?	Failure to close vent valve during start-up.	Process upset.	Pressure indication throughout the system. Regular pressure test procedure during start-up.
What if air is introduced to the system?	Air is introduced to the system during routine vessel cleaning.	Energetic event. Potential deflagration.	Pressure testing procedure with purge.
What if non-Class 1 Division 2 equipment is left near the major fractionator during operation?	Potential for flammable materials to reach ignition sources with this equipment present.	Potential ignition of a flammable vapor if a leak is present.	Pressure testing to identify leaks. Alarms for flammable vapors.
What if pressure builds up in the condenser drum?	Outlet flow of vapor is obstructed resulting in overpressurization.	Pressure vessel burst. Process upset.	Pressure relief valve. Alarms for abnormal pressure conditions.
What if the temperature at the bottom of the column increases?	Increasing temperature at the bottom of the column is not able to redistribute through the column causing heat increase in the system.	Process upset. Potential energetic event.	Alarms for abnormal temperature conditions. Temperature control system lowers steam flow rate.
What if the reflux level is too low?	Reflux pump is started despite the level being too low.	Process upset.	Low level alarms. Level interlock on reflux pump.
What if there is a leak downstream of the distillation column?	Failed connection, release of hazardous chemicals.	Harmful atmosphere. Inhalation toxicity to nearby operators.	Pressure check procedure along with pressure indication throughout the system. Emergency respirators for nearby operators. Proper PPE.
What if the column is overfilled?	Pump fails to stop filling column.	Process upset. Overflow.	Level indicators and alarms. Pressure indicators and alarms.
What if there is a power outage?	Plant facility power failure.	Forced shutdown. Loss of monitoring and process control.	Initiation of shut-down procedure.

Safety Summary

With the presence of hazardous chemicals in a plant of this capacity, it is of utmost importance to train and educate operators in the event of a chemical release or plant accident. As shown in the What-If analysis, there are many hazardous scenarios in a chemical plant and even for one unit operation. Each operator should be provided with SDS information for all chemicals and byproducts relevant to a process so that the proper response can take place in the event of an accident or exposure. The SDS also provides information about proper personal protective equipment such as hard hats, safety glasses, gloves, steel-toed boots, and in the event of an emergency release, NIOSH-approved respirators should be made available. Operators should be made aware of all relevant documentation such as flow diagrams, chemical information, standard operating procedures, emergency procedures, and safety review information. It is the operator's right to know where to find this information at all times, so that they may respond in an appropriate manner to process upsets to reduce loss of life and property. The main hazards associated with this toppings refinery include loss of pressure control within the distillation columns resulting in deflagration to detonation as well as the release of toxic chemicals known to affect individuals through inhalation. Safety review methods are never completely comprehensive, and operators and technicians should not become complacent with the idea that all potential hazards have been identified.

Conclusion

The proposed naphtha reforming process successfully purifies the given naphtha fraction, meeting all regulatory requirements. The catalytic reforming reactions successfully convert cyclohexane to benzene and other alkanes, greatly increasing the ease of separation. The extraction system effectively removes aromatics from the hydrocarbon mixture, and the alkanes are separated into a fraction that the existing facility should easily be able to process. These alkanes now have an aromatics content low enough to meet EPA guidelines. The distillation system then separates the BTX to 99% purity, making all components immediately salable.

The financial analysis demonstrates that, even under reasonably adverse market conditions, the proposed facility has the potential to be remarkably profitable. Under the least favorable expected conditions, the internal rate of return estimated for this facility is in excess of 38%, offering significant returns in relatively short order. The most important costs to manage are the cost of Naphtha and the price of Diesel, as these constitute the two factors to which the overall profitability is most sensitive to. While the fixed capital cost is near \$90,000,000, the full cost is recouped within a year of operation. While there stands the possibility of politics influencing the bottom line, this proposal should be resilient to changes in tax policy as the result of moving borders in the region.

The proposed system is also remarkably safe. Through the lens of inherently safer design, this process is impressive due to its simplification via the safety control systems resulting from BACT application. Pressure, level, and temperature controls on the major fractionator coupled with alarmed indicators for these factors greatly reduce the risk of accidental release. The deflagration calculations and what-if analysis performed greatly emphasize the importance of this control scheme.

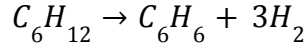
Appendix

Reactor Train Detail

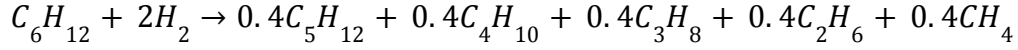
In this section, naphtha from crude oil goes through a series of 3 reactors to reduce the cyclohexane content. The main reaction that initially takes place converts cyclohexane into benzene and hydrogen. These hydrogens can then react with cyclohexane or n-decane to produce C1-C9 alkanes. Since these reactions are endothermic, the feed entering each reactor must be heated up to the optimum temperature of the reactor for the reaction to take place, which should be over 400°C. The product from the last reactor is flashed to remove hydrogen, and the bottom stream moves on to the extraction section.

Catalyst deactivation is an important issue to address when refining naphtha. Over time, catalysts deactivate and must be regenerated in order to be used again. One way to address this issue is by implementing a swing reactor. A swing reactor is another reactor that is not always in use, meaning that there are now 4 total reactors and 3 reactors in operation. When the catalyst in one reactor needs to be regenerated, the feed to it is turned off and instead directed toward the swing reactor. This cycle can occur in periods of 24 or 48 hours. This constant cycling allows the catalyst to have periods of regeneration so that it does not deactivate as soon.

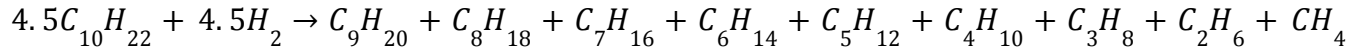
The pump at the start of the system was chosen to supply 10 bar of head. The operating pressure must be high throughout the reaction system in order to avoid catalyst deactivation. The reactor conditions were tuned in Aspen to optimize cyclohexane conversion. Cyclohexane dehydrogenates to the more profitable benzene, and cyclohexane is difficult to remove during downstream processing because its relative volatility to benzene is very close to one. The kinetic equations in the problem statement were given as follows:



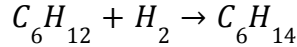
$$rate = 9.4928 * 10^{13} e^{-160506.4/(8.314T)} P_{C_6H_{12}} - 8.2728 * 10^{-4} e^{-52170.4/(8.314T)} P_{C_6H_6} P_{H_2}^3$$



$$rate = 3.6704 * 10^{21} e^{-287756.8/(8.314T)} P_{C_6H_{12}}$$



$$rate = 3.6704 * 10^{21} e^{-287756.8/(8.314T)} P_{C_{10}H_{22}}$$



$$rate = 3.33674 * 10^{19} e^{-275285.8/(8.314T)} P_{C_6H_{12}} P_{H_2} - 4.19816 * 10^{21} e^{-312237.9/(8.314T)} P_{C_6H_{14}}$$

where the rate is in kmol/(m³ hr), the pressure is in MPa, and the temperature is in K. The kinetic equations and literature indicate that reaction rates are negligible below 400°C. Cyclohexane conversion starts with reaction 1. Reactions 2 and 3 have a higher pre-exponential factor and activation energy than equation 1. So, the reaction temperature must be high enough to react cyclohexane, but low enough to prevent a significant amount of reactions 2 and 3. The optimizer in Aspen was used to vary the temperatures of the three reactors to minimize the amount of cyclohexane coming out of the third reactor, leading to operating temperatures of 470°C for each reactor. From there, the temperature of the last reactor was modified to be 467°C to allow for the best downstream separation, according to Aspen.

Since the products of the reactions will be at an elevated temperature, they need to be cooled down before they enter the flash separator. The inlet naphtha feed also needs to be heated from 70°C to 470°C for the reaction. To save on utilities costs, these two streams were fed into a heat exchanger, saving \$227/hr in natural gas costs with an \$18,000 capital investment.

Hydrogen recycle was not used in the model, since an increase in hydrogen promotes reactions 2 and 3, which have a negative effect on overall profit. A recycle loop was tested in Aspen, which yielded results that supported this statement. **Figure 8** shows the compositions of the bottoms from the flash with no hydrogen recycle and a hydrogen recycle loop with a 20% purge. To maintain a low amount of cyclohexane, the system with the recycle loop had to be increased to 495°C in the final reactor. These figures show that the system with the recycle loop had less n-decane and benzene, which are the primary sources of profit, and more alkanes. This outlet composition also had a much more difficult downstream separation, which led to the choice of not having a recycle loop for the system.

	Mole Fractions		Mole Fractions
n-Decane	0.4776	n-Decane	0.3244
Cyclohexane	0.0003	Cyclohexane	0.0002
Benzene	0.3743	Benzene	0.3490
Oxygen	0.0000	Oxygen	0.0000
Nitrogen	0.0000	Nitrogen	0.0000
H2O	0.0000	H2O	0.0000
CO	0.0000	CO	0.0000
CO2	0.0000	CO2	0.0000
Methane	0.0010	Methane	0.0028
Ethane	0.0047	Ethane	0.0120
Propane	0.0132	Propane	0.0286
n-Butane	0.0191	n-Butane	0.0419
n-Pentane	0.0210	n-Pentane	0.0474
Hydrogen	0.0047	Hydrogen	0.0040
n-Hexane	0.0206	n-Hexane	0.0466
i-Butane	0.0000	i-Butane	0.0000
i-Pentane	0.0000	i-Pentane	0.0000
n-Heptane	0.0211	n-Heptane	0.0476
n-Octane	0.0211	n-Octane	0.0477
n-Nonane	0.0212	n-Nonane	0.0478
SULFOLANE	0.0000	SULFOLANE	0.0000
p-Xylene	0.0000	p-Xylene	0.0000
Toluene	0.0000	Toluene	0.0000

Figure 8: Composition of feed after flash without (left) and with (right) hydrogen recycle

Estimates for the cost of the equipment and catalyst are provided in **Table 6**. Some estimates were provided by Aspen, while the fired heaters and heat exchanger were researched in literature and scaled using the six-tenths factor rule.

Table 6: Capital Costs for Equipment in Reactor Train Section

Equipment	Cost
Pump	\$64,400
Heat Exchanger	\$43,000
Fired Heater 1	\$1,120,000
Fired Heater 2	\$572,760
Fired Heater 3	\$501,200
Reactor 1	\$640,000
Reactor 2	\$640,000
Reactor 3	\$640,000
Swing Reactor	\$750,000
Catalyst	\$500,000/year
Cooler	\$135,000
Flash Separator	\$127,700

The yearly operating costs are summarized in **Table 7**.

Table 7: Operating Costs for Reactor Train Section

Type of Cost	Cost/Year
Labor	\$ 302,000
Utilities	\$ 4,781,500

Extractor Section Detail

In this section, the products from the flash separator go through liquid-liquid extraction, where sulfolane is used to extract benzene from the feed. The remaining alkanes go through a steam stripper to be recovered, and the remaining sulfolane is recycled back to the extractor. The

benzene rich stream goes through a series of two distillation columns to yield pure enough benzene to be separated into its components of benzene, toluene, and xylene in the next section. The distillate from the final distillation column is then recycled back into the liquid extractor. The sulfolane in the extract is removed after the first distillation column and recycled back to the liquid extractor. The 2nd liquid-liquid extractor in the provisional PFD was replaced with a steam stripper. In the system, steam strippers can selectively vaporize the alkanes, while the sulfolane is heated and can eventually be removed by distilling the water.

In the liquid-liquid extractor, 130,000 kg/h of sulfolane is used to extract out benzene from the rest of the alkanes. All 129,700 kg/h of the sulfolane in the extract is able to be recovered because the relative volatility between benzene and sulfolane is very high. However, of the 300 kg/h of sulfolane that gets into the raffinate, only 200 kg/h is able to be recovered. This calculation does not take into account that the decomposition rate of sulfolane increases at temperatures above 200°C, which is the temperature at the bottom of both distillation columns. Assuming that sulfolane decomposes at a rate of 0.02% per hour, another 26 kg/h of sulfolane cannot be recycled. So, around 125 kg/h of sulfolane is unable to be recycled, and so that much needs to be continuously added to the first liquid-liquid extractor. Since sulfolane sells for \$5/kg, recycling the sulfolane saves around \$650,000/h.

The temperature profiles for the distillation columns in this section are presented in the figures below. **Figure 9** shows the profile for the major fractionator, which requires 16 stages and a reflux ratio of 8 to adequately separate butane from pentane. With these conditions, a small amount of pentane is lost in the vapor, but the majority of the pentane remains in the distillate, and all the butane leaves in the vapor. **Figures 10 and 11** show the profiles for the sulfolane recovery columns, which shows that very few stages were required to adequately recover the

sulfolane. **Figure 12** shows the profile for the final distillation column in the extraction section, which aims to remove as much cyclohexane and alkanes to yield benzene that is as pure as possible. This separation is difficult because most of the remaining components have similar volatilities, but the benzene must still be separated to a bit above 99% purity to ensure that the ensuing benzene, toluene, and xylene in the next section can be separated to at least 99% purity. As a result, 29 stages and a reflux ratio of 10 are required to get adequate separation for the distillation section.

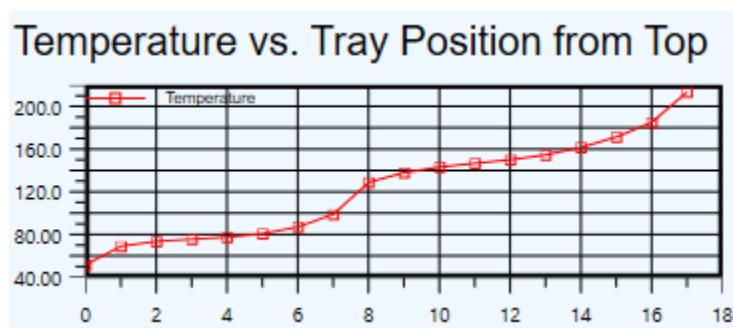


Figure 9: Temperature profile of the Major Fractionator

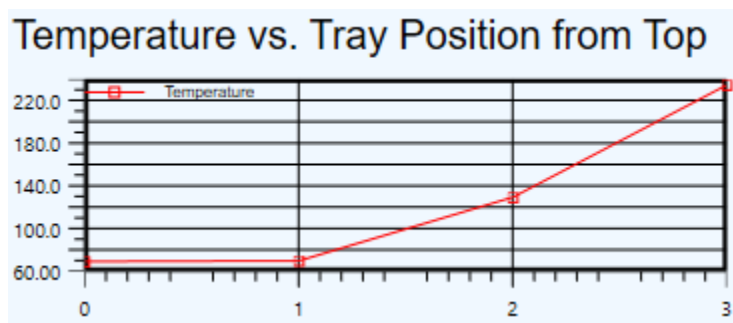


Figure 10: Temperature profile of the distillation column where sulfolane is recovered from water

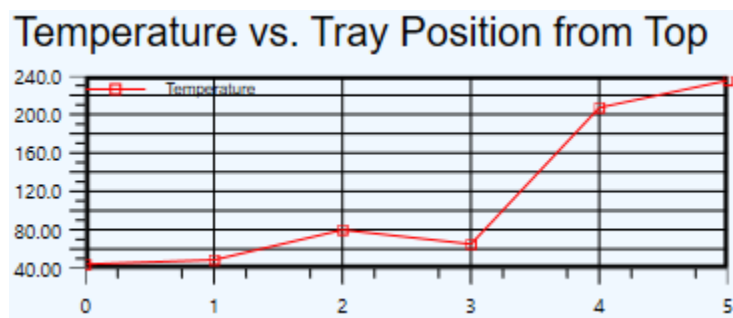


Figure 11: Temperature profile of distillation column where sulfolane is recovered from the aromatics

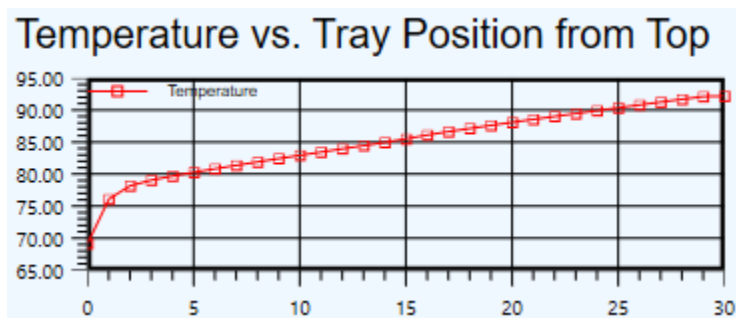


Figure 12: Temperature profile of final distillation column before Distillation Section

Estimates for the cost of the equipment and catalyst are provided in **Table 8**. Estimates were provided by Aspen.

Table 8: Capital Costs for Equipment in Extraction Section

	Cost
Distillation Column 1	\$576,900
Distillation Column 2	\$1,336,900
Distillation Column 3	\$727,700
Distillation Column 4	\$2,246,200
Cooler 1	\$134,000
Cooler 2	\$249,700
Liquid Extractor	\$1,726,600
Steam Stripper	\$832,200

The yearly operating costs are summarized in **Table 9**.

Table 9: Operating Costs for Extraction Section

Type of Cost	Cost/Year
Labor	\$ 210,000

Utilities	\$ 3,138,500
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Distillation Section Detail

In this section, the aromatics that passed through the extraction section are separated through distillation. Since the benzene earlier in the process was a proxy component for all aromatics, the equation in the problem statement was used to calculate the composition of the feed entering the distillation section while still including the impurities from the extraction section.

The relative volatilities of benzene, toluene, and para-xylene can be calculated using the following equations, assuming their solution is ideal:

$$\alpha_{1,2} = \frac{P_1^{vap}}{P_2^{vap}}$$

$$\ln P^{vap} = A - \frac{B}{T+C}$$

where P^{vap} is in kPa, T is in $^{\circ}\text{C}$, and the constants A , B , and C can be found in literature. Using the Antoine equation constants from Smith et al. and a temperature of 80°C yields relative volatilities that are summarized in **Table 10**. The table shows that the main separations have relative volatilities of around 2.5, making separation feasible. Previous commercial distillation operations that separated benzene and toluene have required 34 trays for adequate separation.

Table 10: Relative Volatilities of the Components

Binary Mixture	Relative Volatility
Benzene/Toluene	2.61

Benzene/p-Xylene	6.47
Toluene/p-Xylene	2.48

The temperature profiles for the distillation columns in this section are presented in the figures below. Both columns use 32 stages, which highlights the difficulty in separating the three aromatics. Benzene was able to be separated with 99.4% purity, but toluene and xylene were only able to be separated with 98.5% and 98.6% purity respectively.

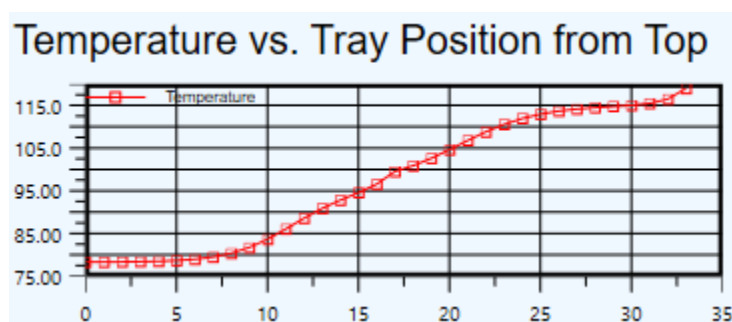


Figure 13: Temperature profile of the first distillation column

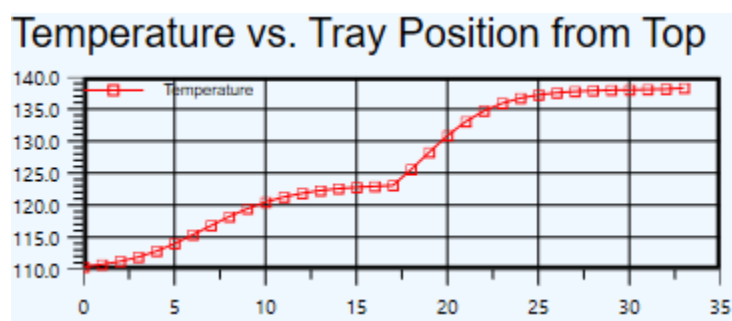


Figure 14: Temperature profile of the second distillation column

Estimates for the cost of the equipment and catalyst are provided in **Table 11**. Estimates were provided by Aspen.

Table 11: Capital Costs for Equipment in Distillation Section

	Cost
Distillation Column 1	\$501,800

Distillation Column 2	\$270,000
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The yearly operating costs are summarized in Table 12.

Table 12: Operating Costs for Distillation Section

	Cost
Labor	\$48,000
Utilities	\$87,000

Budget Sheets

AICHE Capital Budget																
Part 1: Key Input Data																
Equipment cost	\$89,943,248	Equip. market value at end of 4 yrs														
Increase in current	\$25,000,000	MOVC Recovery Rate														
Increase in payed	\$14,820,000	Tax rate														
Annual Unit sales see below		Required Return														
Annual Sales price see below																
Envision	\$0															
Variable cost per	\$0.00															
Fixed costs	\$15,913,872															
Opportunity Cost	\$0															
Part 2: Depreciation and Amort																
Initial Cost	\$9,943,248	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Year 14	Year 15
Equipment Depreciation Rate	5.0%	9.25%	8.6%	7.7%	6.9%	6.2%	5.9%	5.9%	5.9%	5.9%	5.9%	5.9%	5.9%	5.9%	5.9%	3.0%
Equipment Depreciation, Dollars	\$4,497,162	\$8,544,609	\$7,690,148	\$6,925,630	\$6,233,067	\$5,603,464	\$5,306,652	\$5,306,652	\$5,306,652	\$5,306,652	\$5,306,652	\$5,306,652	\$5,306,652	\$5,306,652	\$5,306,652	\$2,653,326
Ending Book Value, Cost - Accum Depn																\$0
Part 3: Net Salvage Values at End of Year 16																
Estimated Market Value at End of Year 16	\$2,000,000															
Book Value at End of Year 16																
Expected Gain or Loss	2,000,000															
Taxes paid or tax credit	700,000															
After-tax cash flow from salvage	\$1,300,000															
Part 4: Project Cash Flows																
Investment Outlays at Time = 0	(9,943,248)	Year 0	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Year 14
Equipment Depreciation	(9,943,248)															
Increase in MOVC	(10,180,000)															
Operating Cash Flows:																
Diesel	\$0	\$0	\$0	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709
Gasoline	\$0	\$0	\$0	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877
HTX	\$0	\$0	\$0	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412
Sales revenue	\$0	\$0	\$0	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998	\$314,463,998
Envision	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-
Netplus	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-
Utilities	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-
Variable costs	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-
Fixed operating costs	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73	\$3,595,811.73
Opportunity Cost	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-	\$-
Depreciation	\$4,497,162.42	\$8,544,608.59	\$7,690,147.73	\$6,925,630.12	\$6,233,067.11	\$5,603,464.37	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$2,653,325.83
EBIT	(68,448,274)	(312,406,420)	(311,641,959)	(316,563,178)	(317,596,546)	(318,135,948)	(318,423,767)	(318,423,767)	(318,423,767)	(318,423,767)	(318,423,767)	(318,423,767)	(318,423,767)	(318,423,767)	(318,423,767)	(318,423,767)
Taxes	(1,252,148,551)	(3,143,132,411)	(3,147,641,861)	(3,150,828,948)	(3,152,621,438)	(3,154,424,138)	(3,156,231,438)	(3,158,042,138)	(3,159,856,138)	(3,161,673,138)	(3,163,492,138)	(3,165,313,138)	(3,167,136,138)	(3,168,962,138)	(3,170,789,138)	(3,172,618,138)
EBIT (1 - T)	(69,699,825)	(315,549,531)	(314,783,820)	(319,724,336)	(320,217,989)	(320,558,086)	(320,575,905)	(320,594,630)	(320,614,271)	(320,634,826)	(320,655,295)	(320,675,669)	(320,695,948)	(320,716,122)	(320,736,296)	(320,756,470)
EBIT (1 - T) + Dep	\$4,497,162.42	\$8,544,608.59	\$7,690,147.73	\$6,925,630.12	\$6,233,067.11	\$5,603,464.37	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$2,653,325.83
EBIT (1 - T) + Dep	(69,694,671)	(324,1935)	(312,874)	(316,848,509)	(316,612,192)	(316,391,831)	(316,082,946)	(315,787,946)	(315,496,946)	(315,209,694)	(314,926,946)	(314,647,946)	(314,367,946)	(314,087,946)	(313,807,946)	(313,527,946)
Remove Cash Flows at Time = 16																
After-tax salvage value																\$1,300,000
Recovery of MOVC																2,036,000
Part 5: Key Output: Appraisal of the Project																
NPV (at the reqd. rate)	\$359,764,871															
IRR	39.55%															
ROI	84.11%															15,662,7966

Figure 15: Iraqi Taxation Budget Sheet

A/C III: Capital Budget																			
Part 1: Key Input Data																			
Equipment cost	\$89,943,248	Equip. market value at end of 4 yrs																	
Increase in current NOWC	\$25,000,000	NOWC Recovery Rate																	
Annual Unit sales price	\$14,820,000	Tax rate																	
Annual Sales price see below		Required Return																	
Annual Sales price see below																			
Variable cost per unit	\$0.00																		
Fixed costs	\$1,951,812																		
Opportunity Cost	\$0																		
Part 2: Depreciation and Amortization																			
Equipment Depreciation Rate	8%	Initial Cost	\$89,943,248	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Year 14	Year 15	Year 16
Equipment Depreciation, Dollars	\$4,497,162																		
Ending Book Value Cost - Accum'd Dep'n																			
Part 3: Net Salvage Values at End of Year 16																			
Estimated Market Value at End of Year 16	\$2,000,000																		
Book Value at End of Year 16	2,000,000																		
Expected Gain or Loss	300,000																		
Taxes paid or tax credit	\$1,700,000																		
After-tax cash flow from salvage																			
Part 4: Project Cash Flows																			
Investment Outlays at Time = 0:		Year 0	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Year 14	Year 15	Year 16	
Equipment	(\$89,943,248)																		
Increase in NOWC	(10,180,000)																		
Operating Cash Flows:																			
Diene	\$0	\$0	\$0	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	\$250,823,709	
Gaoline	\$0	\$0	\$0	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	\$18,847,877	
BTX	\$0	\$0	\$0	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	\$74,792,412	
Sales revenue	\$0	\$0	\$0	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	\$344,463,998	
Erosion	\$ -	\$ -	\$ -	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	\$168,408,000.0	
Utilities	\$ -	\$ -	\$ -	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	\$8,364,773.00	
Variable costs	\$ -	\$ -	\$ -	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	\$176,772,773.0	
Fixed operating costs	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	\$3,951,811.73	
Depreciation Cost	\$4,497,162.42	\$8,544,698.50	\$2,690,147.23	\$6,925,630.12	\$6,233,067.11	\$5,603,464.37	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	
EBIT	(\$8,448,974)	(\$12,496,420)	(\$11,641,959)	\$156,813,763	\$157,506,346	\$158,135,648	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	\$158,432,761	
Taxes	(\$1,207,346.12)	(\$1,374,463.05)	(\$1,746,293.92)	\$23,522,067.40	\$23,625,951.85	\$23,720,302.26	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	\$23,764,914.17	
EBIT(1 - T)	(\$7,181,628)	(\$10,621,957)	(\$9,895,666)	\$133,291,715	\$133,880,394	\$134,415,556	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	\$134,667,847	
Add back depreciation	\$4,497,162.42	\$8,544,698.50	\$2,690,147.23	\$6,925,630.12	\$6,233,067.11	\$5,603,464.37	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	\$5,306,651.65	
EBIT (1 - T) + DCP	(\$2,684,466)	(\$2,077,349)	(\$2,205,518)	\$140,217,345	\$140,113,461	\$140,019,621	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	
Terminal Cash Flows at Time = 16																			
After-tax salvage value																		\$1,700,000	
Recovery of NOWC																		2,006,000	
FCF: EBIT(1 - T) + DCP - CAPEX - Δ	(\$300,123,248)	(\$2,684,466)	(\$2,077,349)	(\$2,205,518)	\$140,217,345	\$140,113,461	\$140,019,621	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$139,974,499	\$144,312,500	
Part 5: Key Output: Appraisal of the Project																			
NPV (at the req'd rate)	\$488,935,022																		
ROI	114.45%																		
DO	112.31%																		20,278,380

Figure 16: Kurdish Taxation Budget Sheet

Stream Compositions

Table 13: Reactor System Stream Compositions (%)

Stream (#)	1	2	3	4
Flow (kmol/h)	29	565	258	307
n-Decane	59.70	25.94	0.02	47.76
Cyclohexane	31.30	0.02		0.03
Benzene	9.00	20.73	0.90	37.43
Methane		1.25	2.61	0.10
Ethane		1.25	2.17	0.47
Propane		1.25	1.16	1.32
Butane		1.25	0.47	1.91
Pentane		1.25	0.25	2.10
Hexane		1.15	0.08	2.06
Heptane		1.15	0.02	2.11
Octane		1.15	0.01	2.11
Nonane		1.15		2.12
Hydrogen		42.46	92.33	0.47
Temperature (°C)	252	38	38	38

Table 14: Extraction System Stream Compositions (%)

Stream (#)	5	6	7	8	9	10	11	12	13	14	15	16
Flow (kmol/h)	13	293	181	1202	339	286	284	2	123	1079	7	115
n-Decane		49.90	80.74	0.03	43.08				0.31			0.33
Cyclohexane		0.03	0.03		0.02				0.03		0.12	0.03
Benzene		39.11		10.10					98.88		90.73	99.39
H2O					47.28	99.24	99.98	0.01				
Methane	2.42											
Ethane	11.08											
Propane	30.85											
n-Butane	44.52										0.01	
n-Pentane	0.08	2.19	3.54	0.04	1.89				0.35		5.93	
Hydrogen	11.06											
n-Hexane	0.00	2.15	3.48	0.02	1.86				0.21		3.20	0.03
n-Heptane		2.20	3.49	0.01	1.86				0.11		0.02	0.12
n-Octane		2.21	3.54	0.01	1.89				0.07			0.07
n-Nonane		2.21	3.57		1.90				0.03			0.03
Sulfolane			1.61	89.78	0.22	0.76	0.02	99.99		100		
Temp (°C)	51.7	214	25.0	25.0	30.0	129	69.1	234	44.1	235	69.2	92.2

Table 15: Distillation System Stream Compositions (%)

Stream	17	18	19	20
Flow (kmol/h)	36	79	49	30
n-Decane		0.48		1.27
Cyclohexane	0.10			
Benzene	99.40	0.07	0.11	
n-Hexane	0.10			
n-Heptane	0.37	0.01	0.01	
n-Octane		0.10	0.16	
n-Nonane		0.04		0.12
p-Xylene		37.94	1.21	98.57
Toluene	0.04	61.36	98.51	
Temp (°C)	78.3	119	110	138

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