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A Case Study on the Design and Optimization of an Ethylbenzene Production Plant

by
William Peaster

A thesis submitted to the faculty of The University of Mississippi in partial fulfillment of
the requirements of the Sally McDonnell Barksdale Honors College.

Oxford
May 2018

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Abstract

This thesis will detail two goals of my capstone design project. The first goal is to simulate a given base case of a vapor phase reaction ethylbenzene production facility in PRO/II simulation software and discretely optimize the process in order to increase the profitability of the plant. The second goal is to develop a base case process flow diagram, stream tables, utility tables, and equipment tables for an ethylbenzene facility with liquid phase reactions. The base case simulation and optimization is a group task with my assigned group members in Ch E 451. The first part of this study looks at the plant during startup and the first twelve years of its operation. Both the vapor and liquid phase models of the ethylbenzene facility contain appropriate stream, equipment, and utility tables. The vapor phase model also includes an analysis of the plants profitability and a comparison of the optimized plant to the base case model. The base case for the liquid phase reaction ethylbenzene plant was modeled using Microsoft Excel.

Table of Contents:

1. Summary of Chemical Engineering Design and Optimization	1
2. Executive Summary: Vapor-Phase Reaction Ethylbenzene Process	4
a. Project Introduction	4
b. Project Optimization	5
c. Project Safety	14
d. Project Summary Conclusion	14
3. PFD Development	16
a. Project Introduction	16
b. Project Method.....	16
c. Calculation of Stream Tables, Equipment Tables, and Utility Tables	19
d. Conclusion	23
4. References.....	24
5. Appendix.....	25
a. Appendix A: Liquid Phase Reaction Ethylbenzene Plant.....	25
b. Appendix B: Vapor Phase Reaction Ethylbenzene Plant	29

Summary of Chemical Engineering Design and Optimization

There are two essential steps to process design in chemical engineering. These two steps are process design and process optimization. Process design is generally broken down into five steps when designing a new process. One must decide whether the process will be batch or continuous, identify the input/output structure of the process, define the process recycle structure, design the process separation system, and design the heat exchanger or energy recovery system [1]. The process optimization is divided into two segments: topological optimization and parametric optimization. Topological optimization is focused on the actual physical arrangement of process equipment in the plant, and it is usually considered first because of a larger impact on the plant profitability. Parametric optimization is involved with the operating variables, such as temperatures, pressure, and concentration, for each piece of equipment. Though topological optimization is usually considered first, it is important to remember that sometimes an optimization of a parameter may lead to a change in plant topography.

The most effective and easy way to describe a plant is through the use of flow diagrams. There are three major types of flow diagrams: block flow diagrams (BFD), process flow diagrams (PFD), and piping and instrumentation diagrams (P&ID). BFDs are the simplest of these diagrams. They simply show a major outline of the process with minimum equipment detail. P&IDs are extremely detailed and show every mechanical aspect of the plant allowing enough information for the construction of the plant. These diagrams are too detailed to fit the entire plant on one sheet of paper. One plant usually has hundreds or thousands of P&ID diagrams. For a more detailed but simple overview, PFDs are generally the most commonly used diagram for a process engineer. For this reason, PFDs will be examined at more closely in this paper.

According to Turton's *Analysis, Synthesis, and Design of Chemical Processes* [1], a good process flow diagram should contain the following: all major equipment in the process with a description, unique equipment number, and a name; all process streams identified by a number with an accompanying stream table showing the process conditions and chemical composition of

each stream; all utility streams supplied to equipment that serves a function in the process; and finally, basic control loops. All PFDs should be clear and easy to follow. Generally, the information on a PFD can be divided into process topology, stream information, and equipment information. To formulate the PFD, one generally follows the hierarchy of process design mentioned in the opening paragraph of this paper.

Parametric and topological optimization are critical tools for a chemical engineer. Parametric focuses on parameters such as temperatures and pressures, where topological focuses on equipment location. To optimize a plant there must be some variable that needs to be optimized. This is the objective function. For example, if the goal is to maximize the net present value of a plant, then the objective function is the net present value. On a smaller scale the objective could be to minimize the amount of unwanted by-products in the plant. In order to get the objective function to its desired value, decision variables, or design variables, are manipulated. These are variables such as temperature, pressure, reactor size, and tower design. The optimum is the point at which either the maximum or minimum of the objective function is reached. It is important to note that a global optimum for an objective function will most likely never be reached. There will always be ways to improve a process. Generally, local optimums are sought after in the industry.

A process can be discretely or continuously optimized. Discrete optimization involves looking at the process step by step and finding local optimums. The process will be repeated several times through. For example, if one is trying to optimize the reactor temperature, then temperature may be adjusted to find a better value for the objective function. Next, one will look at reactor volume. An even better value for the objective function may be found. If the temperature is adjusted now, an even higher value for the objective function can be achieved. Due to the time constraints for the fall design class, after one parameter was optimized we did not go back and optimize again later. Continuous optimization solves for the global optimum all at

one time by changing every optimizable parameter at once. For the nature of this project, discrete optimization is the method used.

Executive Summary: Vapor-Phase Reaction Ethylbenzene Process

Project Introduction

The constraints of the production facility are to produce 80,000 tonnes per year of ethylbenzene at 99.8 percent by mole. As shown below in Figure 2.1, diethylbenzene is produced as an unwanted side product. The maximum allowed diethylbenzene allowed in the ethylbenzene product stream is 2 parts per million.

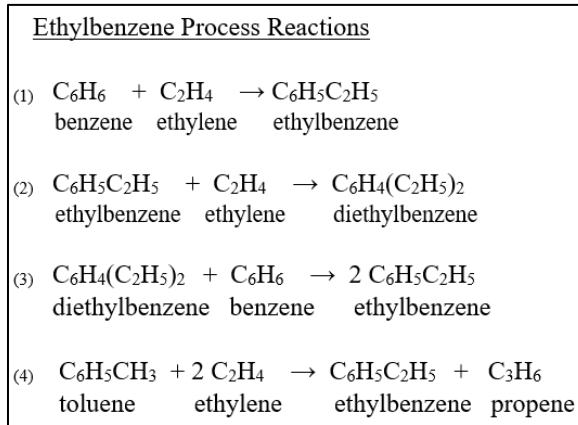


Figure 2.1: Ethylbenzene Process Reactions

The goal of our team was to determine the profitability of the ethylbenzene plant base case. The next step was to determine if the plant could be optimized to achieve greater economic value. The objective function in the optimization of the project was the net present value. In this case, the goal was to maximize the objective function. In order to determine the profitability of the plant, a sensitivity analysis was created in order to figure out which variables had the greatest effect on the net present value of the plant. The sensitivity analysis can be seen below in Figure 2.2. The cost of raw materials and the selling price of our products have the greatest effect on the net present value of the plant. The profit margin for the production of ethylbenzene based on the current raw materials cost is profitable. If the prices are predicted to fluctuate significantly or the market for ethylbenzene decreases, then this project will have a greater risk. The breakeven price for ethylbenzene in the base case was \$1,504.99/kg and the breakeven price in our current optimized project was \$1,294.73/kg.

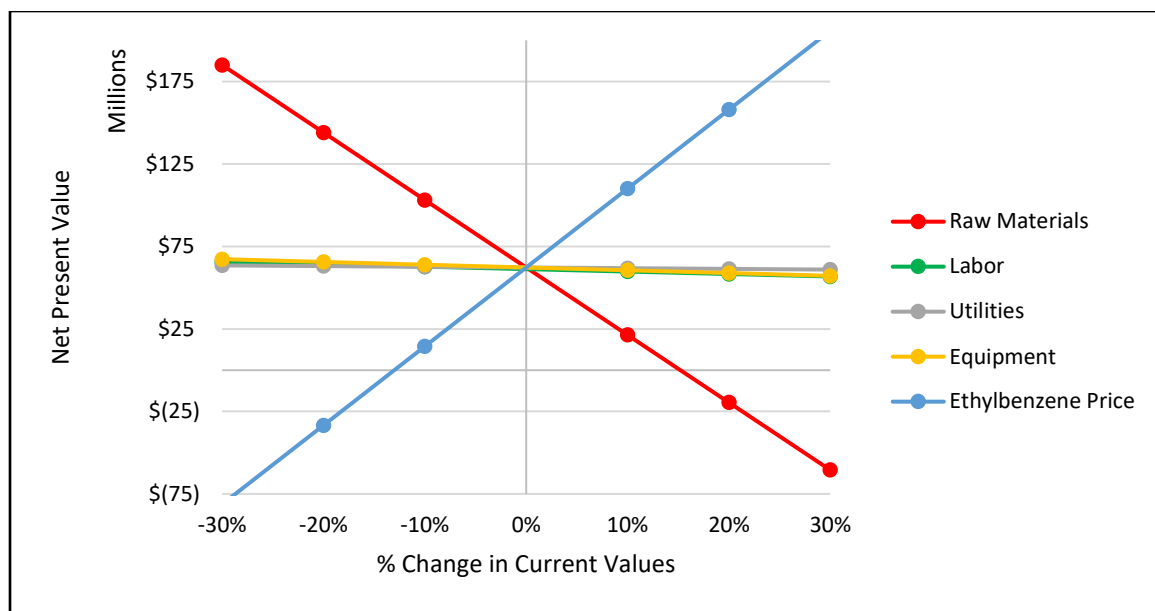


Figure 2.2: Ethylbenzene production plant sensitivity analysis

Project Optimization

The team began to determine whether the project should be undertaken with by evaluating changes to the process by using discrete optimization. There were several recommended changes given in the case study. They include using a different catalyst in our reactor system, using a lower quality feed, and installing a distillation tower to separate toluene from our raw materials in order to sell pure toluene as a product.

Using a given preliminary process design, the base case net present value was \$(6.4) M. This base case design had a benzene inlet stream of 7,862 kg/hr composed of 98% benzene and 2% toluene. There was also an ethylene inlet stream of 2846 kg/hr composed of 93% ethylene and 7% ethane. The fired heater in this process had an outlet temperature requirement of 400 °C, cost \$2.5 M to build, and has a duty of 22,612 MJ/hr, leading to a series of three packed bed reactors with volumes of 20 m³, 25 m³, and 30 m³ respectively. A feed ratio to the first reactor of 8:1 benzene to ethylene was specified. The amount of diethylbenzene produced required the use of a second tower in this preliminary design due to the 2 ppm diethylbenzene specification, and a

reactor that converts recycled benzene and diethylbenzene to ethylbenzene. Based on our initial analysis we saw significant room for improvement to the preliminary process.

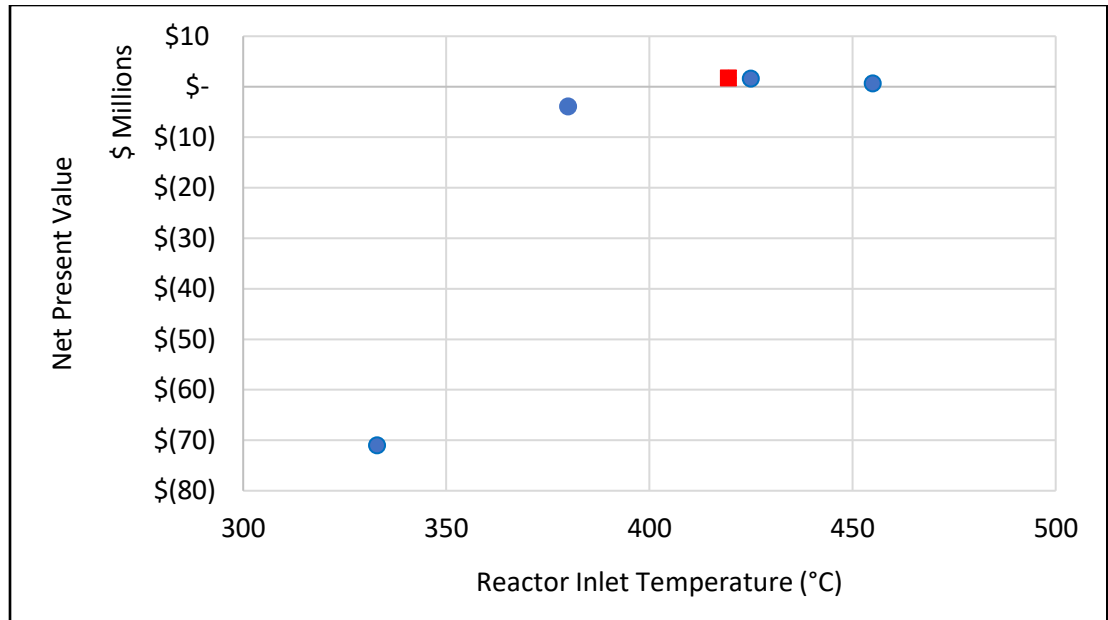
After studying the base case, our group saw many areas for improvement in the ethylbenzene plant through the evaluation of parameters such as reactor operating temperature, reactor volume, reactor feed ratio, and several other areas. We were confident in our ability to make these process optimizations. We chose to proceed with further development and optimization of the project because of the economic potential of this process.

The first process change is the first recommended change, a new catalyst advertised to suppress the production of diethylbenzene by changing the reaction rates. This new catalyst costs \$8/kg and had an expected lifetime of four years, compared to the catalyst used in the base case with a cost of \$5/kg and an expected lifetime of three years. With this change, we needed to buy catalyst one extra year during the twelve-year life of the process for approximately \$300,000, but we save nearly \$500,000 per year in raw materials and utility costs. We saved money with this catalyst because we lost less benzene out of our fuel gas stream from V-302. Using the base case catalyst, we had a net present value of \$(6.4) M. By changing to the proposed catalyst, we lowered the feed streams of benzene and ethylene to 7,828 kg/hr and 2,834 kg/hr, respectively. With this change of catalyst in our reactors, we saw an increase in net present value to \$(3.9) M. From here, we decided that we would use the new catalyst presented in change one for the remainder of the design process.

After completing the catalyst optimization, we chose to continue looking at the reactor section and its effect on the process. We chose to look at reactors before changing feed quality because we knew the reactors section would have the greatest effect on net present value. Looking back, we realize that changing the feed quality would have a larger effect on our NPV, since it directly affects raw material cost.

We started the optimization of reactors by looking at reactor feed temperature because we knew that the temperature would affect the reaction kinetics, granting higher selectivity to

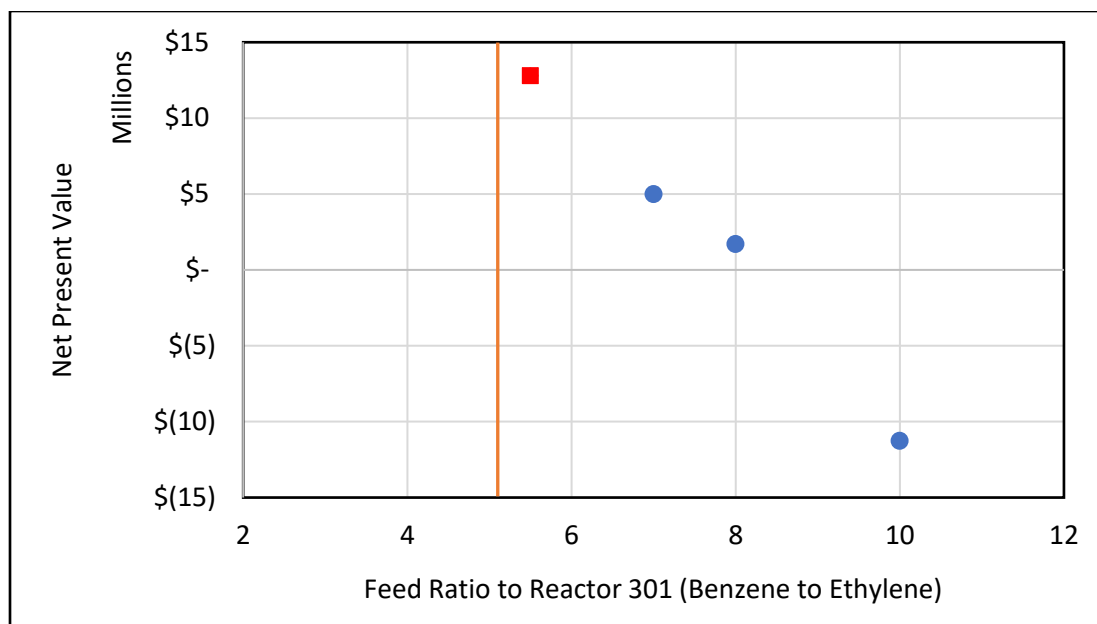
ethylbenzene. A decrease in the production of diethylbenzene allowed for the removal of T-302, R-304, V-304, E-308, E-309, and P-305, which reduced utility and equipment cost. We kept the temperature to each reactor constant. Using the conversion of benzene and selectivity vs. temperature graph that we created in the process justification, we were able to select a few temperature points for further investigation. The initial base case value for the inlet temperature was 380 °C. At 333 °C, the conversion was almost at a maximum, but the selectivity was lower. At 455 °C, the conversion reached its maximum, and the selectivity was essentially infinite (i.e. no diethylbenzene is being produced). We investigated a few points between those aforementioned and found an economic optimum at 420 °C. At this point, the conversion of benzene was at a maximum and selectivity was very high, and our utility cost was reduced. Operating at increased temperature required the purchase of stainless steel for several pieces of equipment, including all reactors R-301, R-302, R-303, the fired heater H-301, and the first three heat exchangers E-301, E-302, and E-303. This more expensive material was justified by the increase in net present value. The base case temperature into each reactor was 380 °C, and the net present value is \$(3.9) M. Our optimized temperature into each reactor was 419.5 °C, which gave a net present value of \$1.7 M. The points chosen for optimization and their effect on net present value can be seen below in Figure 2.3.



Red = chosen optimum; Blue = each point investigated

Figure 2.3: Reactor 301 inlet temperature vs. process net present value

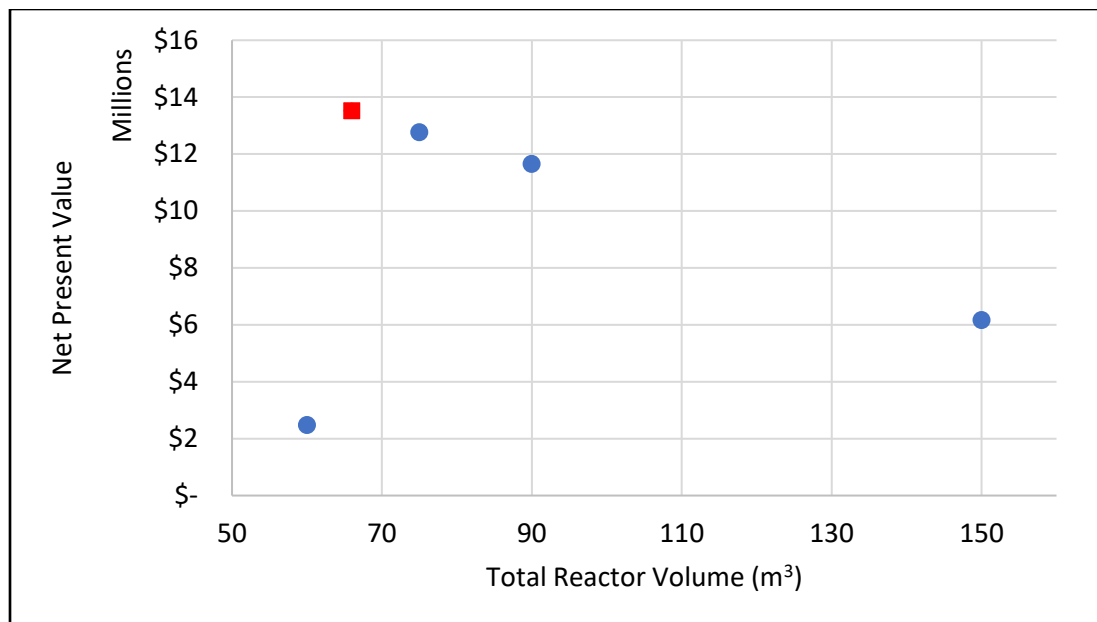
Continuing to look for ways to decrease our raw material cost, we investigated the feed ratio of benzene to ethylene to the first reactor (R-301). In the base case, the required feed ratio is 8:1 to suppress the production of diethylbenzene. The new catalyst, along with a higher temperature, allows us to lower this ratio. We decided that our optimized ratio of benzene to ethylene is 5.5:1. A lower feed ratio would have allowed for a higher net present value; however, the lower ratio would have burned up our catalyst, which cannot operate at a greater temperature than 525 °C. The net present value of the base case with optimized temperature and new catalyst was \$1.7 M. After optimization at a ratio of 5.5:1, the net present value of our plant was \$12.8 M. The points chosen for optimization and their effect on net present value can be seen below in Figure 2.4.



Red = chosen optimum; Orange = catalyst burn out barrier; Blue = each point investigated

Figure 2.4: Feed ratio to reactor 301 vs. process net present value

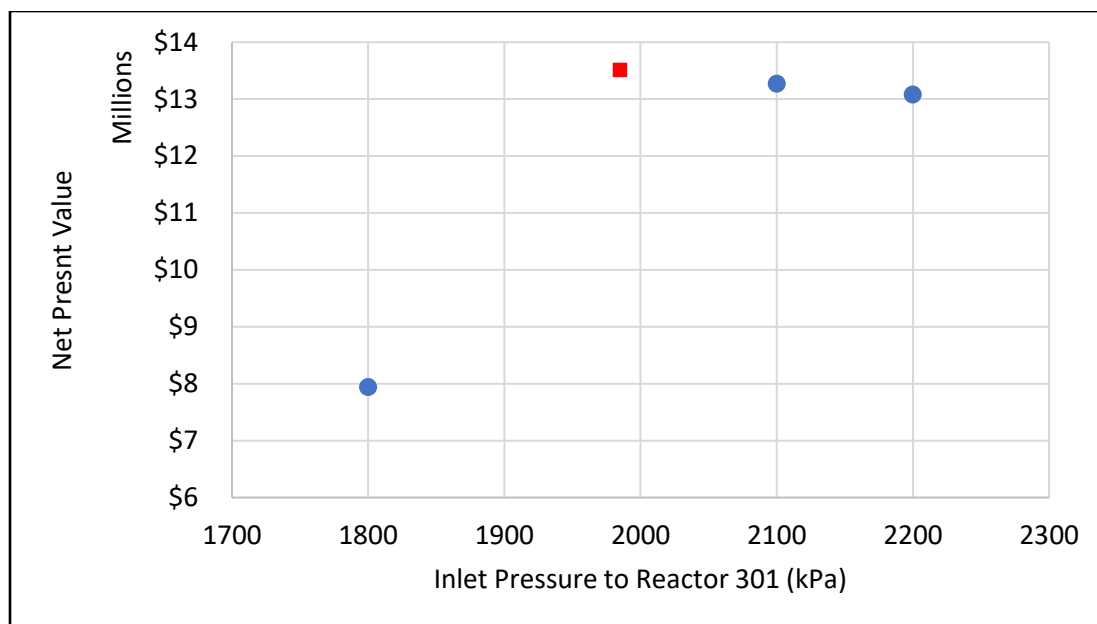
The next parameter we chose to look at in the optimization of our reactor system was reactor volume. To look at reactor volume, we made the volume of each of our reactors equal. We did this because we wanted to focus on the total reactor volume and decided that it would be easiest to make all reactors an equal volume. In further optimization, we will optimize the volume of each reactor individually. Volume of reactors affects the conversion which, in turn, affects the amount of raw materials in the recycle. In order to correctly model the pressure drop in our new reactors, we calculated a change in pressure using the Ergun equation. Our total unoptimized reactor volume for the three reactors was 75 m³, and the net present value for this size is \$12.8 M. After optimization we used three reactors at 22 m³ (a total volume of 66 m³) giving a net present value of \$13.5 M. A lower volume than this greatly decreased the net present value, since it lowered the conversion of the reactor series and increased the production of diethylbenzene calling for more equipment. The points chosen for optimization and their effect on net present value can be seen below in Figure 2.5.



Red = chosen optimum; Blue = each point investigated

Figure 2.5: Total reactor volume vs. process net present value

Pressure was the last variable examined in the reactor optimization. The inlet pressure to the first reactor was varied to study the effects on the net present value. The rate laws for our reactions were based on concentrations, which are affected by pressure since the reactions are occurring in the vapor phase, leading to our decision that pressure would be a good parameter for optimization. We noted that an increase in reactor inlet pressure from our base case inlet pressure of 1985 kPa slightly decreased our net present value. A decrease in pressure decreased our net present value because the decrease in pressure caused a rise in diethylbenzene production calling for extra equipment and utilities. Due to these instances, we kept our inlet pressure to our reactor at 1985 kPa because the slightest decrease produced too much diethylbenzene, and a small change would have made an insignificant difference in our net present value. The points chosen for optimization and their effect on net present value can be seen below in Figure 2.6.



Red = chosen optimum; Blue = each point investigated

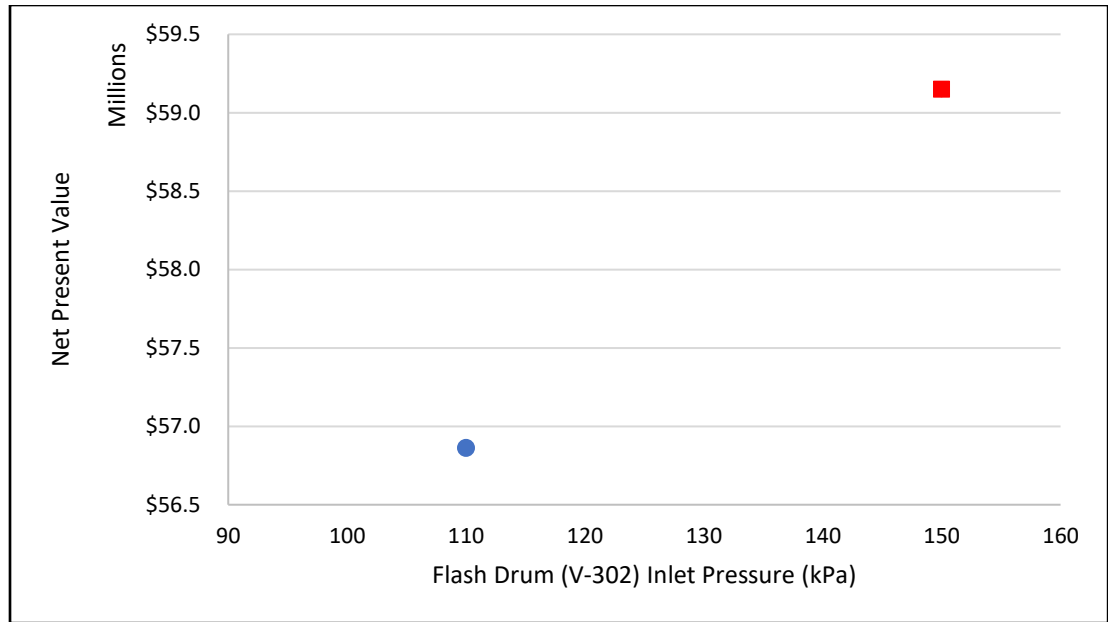
Figure 2.6: Reactor 301 inlet pressure vs. process net present value

The second recommended change in the project statement allowed a lower grade of benzene feed to be purchased. The original benzene feed was 98% benzene and 2% toluene at a price of \$1.014/kg. The new feed was 90% benzene and 10% toluene at \$0.85/kg. Knowing that raw material costs have a huge effect on the value of a plant, we experimented with this change. After installing this change, we created more diethylbenzene and had to reinstall the equipment that we removed after our temperature optimization (T-302, R-304, V-304, E-308, E-309, and P-305); however, our net present value increased from \$13.5 M to \$49.1 M. This was mostly due to the new raw material cost which decreased from \$86.3 M/yr to \$81.6 M/yr. The cheaper feed was the most profitable change in the optimization of the plant.

The third recommended change allowed toluene to be separated and sold for \$0.91/kg at a purity greater than 99.5 percent by mole. Our group saw this change as an opportunity to eliminate all the equipment that was added after changing to a lower feed quality and an opportunity to have another product sold in our plant. We installed a distillation column in our plant directly after our benzene feed stream to separate the toluene. In this new tower (T-303), we

specified the distillate to be 98% benzene and 2% toluene by mole and the bottoms stream to be 99.5% toluene by mole. This allowed us to buy cheaper raw materials and keep the composition of the feed the same as before we changed our feed quality, allowing the removal of T-302, R-304, V-304, E-308, E-309, and P-305 once again. This change also gave us another profitable product stream of toluene. After installing this change to our process our net present value increased to \$56.9 M.

The first parameter in the separations phase of our optimization is the valve outlet pressure, which in turn affects the pressure and temperature entering our flash drum. At the current pressure, 110 kPa, we were losing a large amount of benzene (almost 12% of the benzene being fed to the process) in the flash drum vapor stream. By increasing the outlet pressure of the valve, we decreased the amount of benzene being lost due to a change of flash point temperature in the stream. We experimented with several different pressures and found our optimum pressure at 150 kPa. This pressure change increased our net present value to \$59.1 M from \$56.9 M when the outlet pressure was 110 kPa. This increase was due to the lower cost of raw materials. There was a third data point removed from this study due to an error. The points chosen for optimization and their effect on net present value can be seen below in Figure 2.7.



Red = chosen optimum; Blue = each point investigated

Figure 2.7: Flash drum inlet pressure vs. process net present value

Next in our separation process was the optimization of the feed tray locations in our two towers using an optimizer in PRO/II. Originally, T-303 was fed at tray 5 and T-301 was fed at tray 6. After optimization, both towers were fed at tray 8. This did not have a significant effect on the net present value of our project. We decided to proceed with this change anyway because it balanced the duty required of our condensers and reboilers.

Our goal throughout the heat integration process was to decrease our yearly utility cost while keeping a safe operating process. We simulated several different methods of heat integration. In one method, we heated our fired heater feed stream with high pressure steam created from E-303 to lower the fired heater duty. Then we use reactor effluent from the third reactor to take the place of the steam used for our reboiler on T-303. This integration method raised our net present value to \$60.2 M. We further investigated a method that would almost entirely integrate the heat in our process. We used the effluent from each reactor to heat the feed stream to the fired heater H-301. This heater feed stream was fed through E-301, E-302, and E-303 cooling the reactor feeds to the desired temperatures. The heater feed stream then heats the

reboiler in T-303 and finally returns to the fired heater, reducing that duty significantly. The reactor effluent from E-303 cools to 260 °C, and that subsequently is used to heat the reboiler on T-301. This process eliminated the need to buy boiler feed water and steam for our reboilers and reduced the duty on the fired heater, which raised our net present value to \$61.3 M. We realize that this runs the risk of a runaway reaction if the reactor effluent gets too hot, therefore it would be good to go back and look into safer designs for the final heat integration process.

The optimized design can be seen in the appendices in Figure B.1.

Process Safety

Although the process does not produce extremely toxic and poisonous materials, the materials used or produced are inherently flammable and have a potential to be vigorously explosive. Moreover, ignition sources should be eliminated if possible. We must have a highly capable controls system to keep our catalyst from overheating since the reactor temperature gets within 10 degrees of the burnout temperature.

The EPA has a list of Toxics Release Inventory (TRI) chemicals that, if exposed to humans, can cause cancer or other chronic human health effects, significant adverse acute human health effects, or adverse environmental effects. The following chemicals used in this plant are listed: ethylene, benzene, toluene, ethylbenzene, diethylbenzene, and propylene. These chemicals must be safely stored and exposed to the environment as little as possible. Significant spills from any of these chemicals can cause major environmental and health issues. Some of the dangers that can come from these chemicals can be seen in Figure B.5 in the appendices.

Executive Summary Conclusion

According to our discretely optimized design project, our group recommended further investigation into this project. We concluded it is worth further investigation as the current net present value is at \$62.3 M. Our results after discrete optimization confirm our initial recommendations of performing further optimizations. The next steps in this process would be to explore new optimums for each of our parameters and further explore other parameters through

continuous optimization. An economic analysis and comparison of base case to optimized case values are shown below in Table 2.1 and Table 2.2.

Table 2.1: Operating costs of base case vs. optimized design

	Base Case	Current Design
Utilities (\$M/yr)	\$2.36	\$0.86
Raw Materials Cost (\$M/yr)	\$89.51	\$80.60
Product Revenue (\$M/yr)	\$123.10	\$127.03
EAOC (\$M/yr)	\$(119.35)	\$(106.47)
Net Present Value	\$(6.44)	\$62.27

Table 2.2: Economic parameters of base case vs. optimized design

	Base Case	Current Design
Discounted Cash Flow Rate of Return	8.55%	40.22%
Annual Equivalence (\$)	\$(946,000)	\$9,143,000
Conventional Payback Period (yr)	7.81	1.02
Breakeven Ethylbenzene Price (\$/tonne)	\$1,504.99	\$1,291.67

PFD Development for Liquid-Phase Ethylbenzene Process

Project Introduction

After completing the optimization and design of the base case plant, my advisor and I thought it would be a good idea to create an ethylbenzene process that operated on a liquid-phase reaction scheme. There can be several benefits that come from a liquid phase reaction such as decreased utility costs, the removal of the fired heater, a safer plant, and a decrease in raw materials required. I wanted to see if this was possible. The plant was to have two continuous-stirred tank reactors, or CSTR, and two distillation columns. I set out to create a base case process flow diagram equipped with stream tables, equipment tables, and utility tables. Instead of a process simulator, Microsoft Excel was used to perform both the mass and energy balances in the plant as well as the other required calculations. The PFD's are modeled using Microsoft Visio. The constraints given for this plant were 80,000 tonnes per year of ethylbenzene at a purity of 99.5% and no more than 500 parts per million of diethylbenzene allowed in the product stream. To create this PFD and accompanying tables, I followed the hierarchy of process design from Turton's *Analysis Synthesis and Design of Chemical Processes* mentioned earlier in this paper.

Project Method

The first step to the creation of this process flow diagram was to determine whether the process will be batch or continuous. A very important factor in determining whether a process should be batch or continuous is the amount of product needed in a year. Batch production is generally only used when the production is less than or equal to 500 tonnes per year, and continuous production is used when the production is greater than or equal to 5,000 tonnes per year. Since the constraints for the ethylbenzene process calls for 80,000 tonnes per year, it was easy to choose continuous as the favorable process model.

The second step in the PFD design process was to create the input/output structure for the process. Generally, the first step for an input/output structure is a process concept diagram. This

is an easy task in this project since the raw materials, reactions, and products were all given. The process concept diagram for this process is shown below in Figure 3.1.

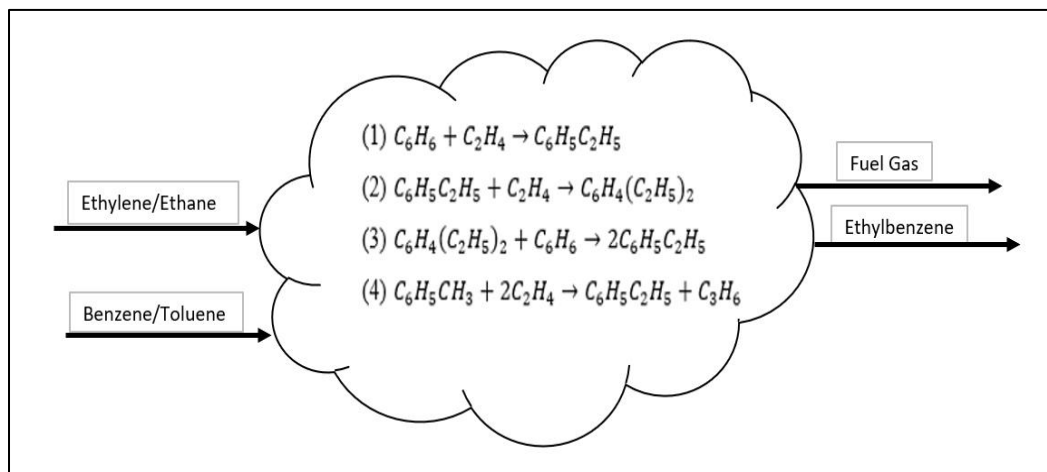


Figure 3.1: Process concept diagram for the ethylbenzene process

Once a process concept diagram was constructed, I formulated a generic BFD. BFD's are briefly mentioned at the beginning of this paper. They are helpful when developing a process and can start to give a better picture of what the entire process flow diagram will look like upon completion. I created a basic block flow diagram based on the recommended two reactors and two distillation columns. This first block flow diagram can be seen in the appendix in Figure A.1.

The third step in the creation of the PFD is determining the recycle structure of the process. As mentioned earlier, the largest cost for every chemical plant is the cost of raw materials. Therefore, the separation and recycling of unreacted raw materials is critical for a plant to stay profitable. The recycle structure of the plant can take a small single-pass reactor conversion and make it economical through a large overall conversion. There are three ways to recycle raw materials in a continuous process: separate raw material from products and recycle, recycle feed and product together then use a purge stream, recycle feed and product together and do not use a purge stream. In this process, the use of a separation system for the raw materials makes the most sense. It is easy to separate benzene from ethylbenzene and ethylbenzene from diethylbenzene, so separation makes sense. The recycles with the product only make sense when

separation is difficult and when the product can react further in the system. Generally, a purge stream will be needed for this unless the reaction can reach an equilibrium, so steady-state can be reached instead of accumulation. There are two recycles in this process. The benzene is separated in the first column and recycled to raw materials, and the diethylbenzene is separated in the second column and recycled to the inlet stream of the second reactor. It is important to note that an increase in the recycle size of a plant is going to cost more money in the form of equipment size and utilities, but usually the money saved on raw materials outweighs these costs. However, there is a balancing act between the two that would be further investigated in the optimization of a plant.

The fourth step in the process flow diagram development is formulating the structure of separation system. Once the recycles are determined, the actual equipment responsible for causing these recycles must be put in place. There are many heuristics for the development of the separation system in a plant. First, one needs to determine the type of separation equipment. If the purity of both products is required, distillation columns should be the first choice. Generally, the largest stream is removed first so that the second separation system is smaller. Also, do the easy separation first. If these conflict, it is usually the largest stream that is removed first. Following these heuristics, it is easy to see that distillation columns are the preferred choice for this process, and the benzene should be separated first as this is the easiest separation and largest stream. I also had to consider the fact that the ethylene would need to be released in a fuel gas stream in order to prevent accumulation in the plant. Since the conversion of ethylene is so high, a recycle stream would not be profitable, and it was more cost effective to let the ethylene be burned off as fuel gas. I decided that the most efficient way to do this would be to install a partial condenser on the first tower instead of using a flash drum like the one in the vapor phase process.

The final step in the development of the PFD is the heat-exchanger network. The strategy behind the heat-exchanger network is the idea that energy from the process can be saved to reduce utilities by transferring that energy to part of the process that is lacking energy. Most

plants will use boiler feed water to cool hot process streams, creating steam that can be used in other parts of the plant to heat cold streams. Process streams can also be used to transfer heat between each other. Too much integration can make the process difficult to control. For the sake of the development of this PFD, I did not perform heat integration. Naturally, there is a reboiler and condenser for each distillation column. I then simply found the areas in the plant that needed heat transfer and placed exchangers. The reactor feed needed to be cooled after the compression of the ethylene. The reactor effluent needed cooling to allow better separation in the first tower. The product stream needed to be cooled to the specified temperature, and the bottoms from the second column needed cooling before entering the reactor.

I followed the five general steps in the hierarchy of process design and created the process flow diagram shown in the appendix in Figure A.2. This process flow diagram is in no way optimized, but it provides a working base case for an operational liquid phase reaction ethylbenzene production plant. The accompanying stream tables, equipment tables, and utility tables are shown with the PFD in the appendix.

Calculation of Stream Tables, Equipment Tables, and Utility Tables

A process flow diagram is not complete without the accompanying stream tables, equipment tables, and utility tables. For the PFD development part of this capstone to be complete, the mass and energy balances for process need to be completed. All the calculations done were done in an Excel workbook using the solver function. The completed stream tables can be found in the appendix in Figure A.5. While reading this section of the paper, it may be helpful to have the process flow diagram in Figure A.2 on hand.

Before concerning myself with temperatures of mixing streams, tower reboilers, etc., I wanted to complete the overall mass balance. The first two variables are the inlet flows of the two raw materials in streams one and two, benzene and ethylene respectively. The two constraints for these streams are setting the molar flow rate of the product stream to its specified value and setting the benzene to ethylene ratio to the first reactor to three to one. The purpose of

the product constraint is obvious, and the purpose of the benzene to ethylene ratio is to suppress diethylbenzene production. The feed ratio is an optimizable parameter that can be adjusted during optimization. Stream 3 is simply the addition of stream 1 and the recycle stream 14. The calculation of stream 14 will be discussed during the discussion of T-101. Stream 5 is equal in mass and composition to stream 3 at a higher pressure calculated using the equation given for the calculation of reactor pressure seen in the appendix in the problem statement plus the 0.2 atmosphere pressure drop that will occur in the first heat exchanger. Stream 4 is equal to stream two in mass and composition, however, the ethylene is compressed to the pressure of stream 5 in order to liquify the ethylene. Stream 6 is stream 5 plus stream 4. Stream 7 is equal mass and composition to stream 6 with a 0.2 atmosphere pressure drop and heated to the reactor temperature of 150°C.

To calculate stream 8, the effluent from the R-101, there need to be five variables and five constraints. The five variables are the reactor volume and the outlet concentrations of the four components. The five constraints are the four component mole balances and the conversion of ethylene at ninety-nine percent. The molar volume was assumed as constant in these calculations. The generated term in the mole balance was calculated using the reaction rate of the component multiplied by the variable volume. The inlet and outlet terms were calculated using the inlet and outlet concentrations, respectively, multiplied by the volumetric flow rate. The process is steady-state so each balance must be equal to zero. Stream 8 will be equal to the solved outlet concentration times the volumetric flow rate.

Stream 9 is the addition of stream 8 and stream 20. Stream 20 is the recycle of diethylbenzene from the bottoms of T-102. The calculation of stream 20 will be discussed at the same time as T-102. Stream 9 is at the same temperature of 150°C and is the inlet to the second reactor, R-102.

The effluent of R-102 is stream 10. It was calculated in the same way as the calculations for the effluent of R-101 were except that the reactor volume was no longer a variable. In turn,

the conversion was no longer a constraint. The reactor volume was equal to the volume of the first reactor. This was a given in the problem statement. The four variables were now only the outlet concentrations of the reactor, and the four constraints were the component mole balances. Stream 10 will exit the reactor at 150°C.

The function of streams 11 and 12 were to prepare the reactor effluent to enter the separation section of the plant. Since the condenser on Tower 101 is a partial condenser, and fuel gas is leaving the from V-102, the temperature of stream 12 needs to be at the bubble temperature. Feeding the column at the bubble point temperature will provide the most efficient separation possible as the light component will immediately go to the vapor phase and the heavier component will stay liquid when the feed stream enters the column. To solve for the compositions, flows, and temperatures of streams 11 and 12 there were six variables and six constraints. The temperature of stream 11, the total liquid flow rate in 11, the liquid compositions of stream 11, and the bubble temperature of stream 12 were the variables. The constraints were the four vapor-liquid mole balances around stream 11, the energy balance around the valve, and the sum of the vapor fractions of stream 12. The vapor-liquid equilibrium in this project was modeled using Raoult's Law that states that the partial vapor pressure of each component in a mixture was equal to the vapor pressure of each component multiplied by its mole fraction. Using Antoine's equation to find the vapor pressure, the guess variables for the liquid mole fractions in stream 11, and the guess variable for the total liquid flow rate, the mole balance around the vapor and liquid component of stream 11 was modeled. Then using the heat capacities of each component, the energy balance around the valve was modeled knowing that the enthalpy of stream 12 must equal the enthalpy of the liquid in 11 plus the enthalpy of the vapor in 11. Finally, the bubble temperature of the stream was found using the Antoine's equation for vapor pressure, the liquid composition of stream 12, and the fact that the vapor fraction at the bubble point equals the mole fraction times the vapor pressure at the bubble point divided by the total

pressure. The sum of these vapor fractions must equal one which provided the necessary constraint.

The next streams to determine are streams 13, 14, and 15 from T-101. There were four variables and four constraints to determine these streams: the condenser temperature, liquid fraction of benzene and ethylene in the distillate, and the total molar flow rate of the distillate stream 14. The four constraints were the sum of the vapor fractions of each component calculated from Raoult's Law equal to one, and the three mole balances for ethylene, benzene, and ethylbenzene. It was assumed that zero diethylbenzene will be in the distillate stream since it is the heavy component. It was assumed that anything lighter than the light key was completely recovered in the distillate and anything heavier than the heavy key was completely recovered in the bottoms. The mole balance around the tower was determined using the column specifications give on 99.95% benzene recovery and a 99.9% ethylbenzene bottoms recovery. The stream 12 feed stream is known. The bottoms stream, 15, was determined solely based on the specifications. Streams 13 and 14 were determined using the liquid mole fraction times the total liquid rate and the vapor fraction times the total vapor rate found using Raoult's Law.

The balance around the second column was much simpler. There was no ethylene in the feed to T-102 and all benzene will be in the distillate stream 16. There were two variables: the molar flowrate of the bottoms and the fraction of ethylbenzene in the bottoms. The two constraints were the ethylbenzene and diethylbenzene mole balances. Using the column specifications of 99.9% recovery of ethylbenzene and diethylbenzene in the distillate and bottoms respectively, the compositions of the bottoms and distillate were easily found.

Stream 17 is the ethylbenzene product stream and it is simply the same molar flow and composition of stream sixteen reduced to a specified temperature of 50°C. Stream 19 was stream 18 pumped to 0.2 atmospheres higher than the reactor temperature. Stream 20 was stream 19 cooled to the reactor temperature. Solver was then used on all twenty-three variables and constraints.

After the mass balance was solved, the temperatures of streams 15, 16, and 18 were determined using Antoine's equation to solve for the temperature as the variable and Raoult's equation using the sum of the vapor fractions as a constraint. The temperature of stream 3 was solved by using the heat capacities of the streams and creating an energy balance around stream 1, 3, and 14. The temperature of the compressed ethylene was determined using an equation from thermodynamics that relates the temperature and pressure of a system. The temperature of stream 6 was determined the same way as stream 3 by creating an energy balance around streams 4, 5, and 6. The PFD, stream tables, equipment tables, and utility tables can be found in Appendix A in Figure A.2, Table A.3, Table A.1, and Table A.2, respectively.

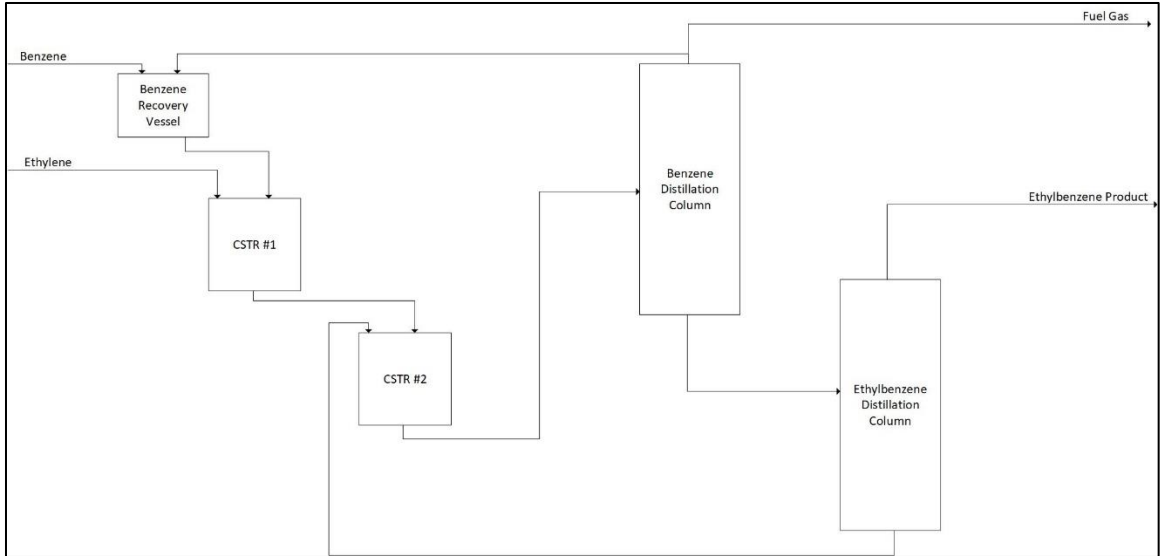
Conclusion

The purpose of the study of a liquid-phase reaction ethylbenzene was to compare the value of this plant to the vapor-phase reaction plant. Given further time, a complete economic evaluation would be done on this process, as performed for the vapor-phase plant in the first part of this study. The plant is attainable, and a base case is now ready to be studied and optimized. In the vapor reaction part of the study, a simulation software was used to model the system and optimize. In the liquid reaction part of the study, Excel was used to model the system. Both modeling techniques require a deep knowledge of chemical engineering but modeling a system in Excel seems to require more attention to detail and a more extensive knowledge of more difficult concepts of thermodynamics and vapor-liquid equilibrium. It would be much harder to optimize a base case using Excel due to the ease of errors, but it helps solidify a good knowledge of the process.

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- [1] Turton, Richard. *Analysis, Synthesis, and Design of Chemical Processes*. Upper Saddle River, NJ: Prentice Hall PTR, 2012. Print.

Appendix



Appendix A: Liquid Phase Reaction Ethylbenzene Plant

Figure A.1: Ethylbenzene production facility block flow diagram

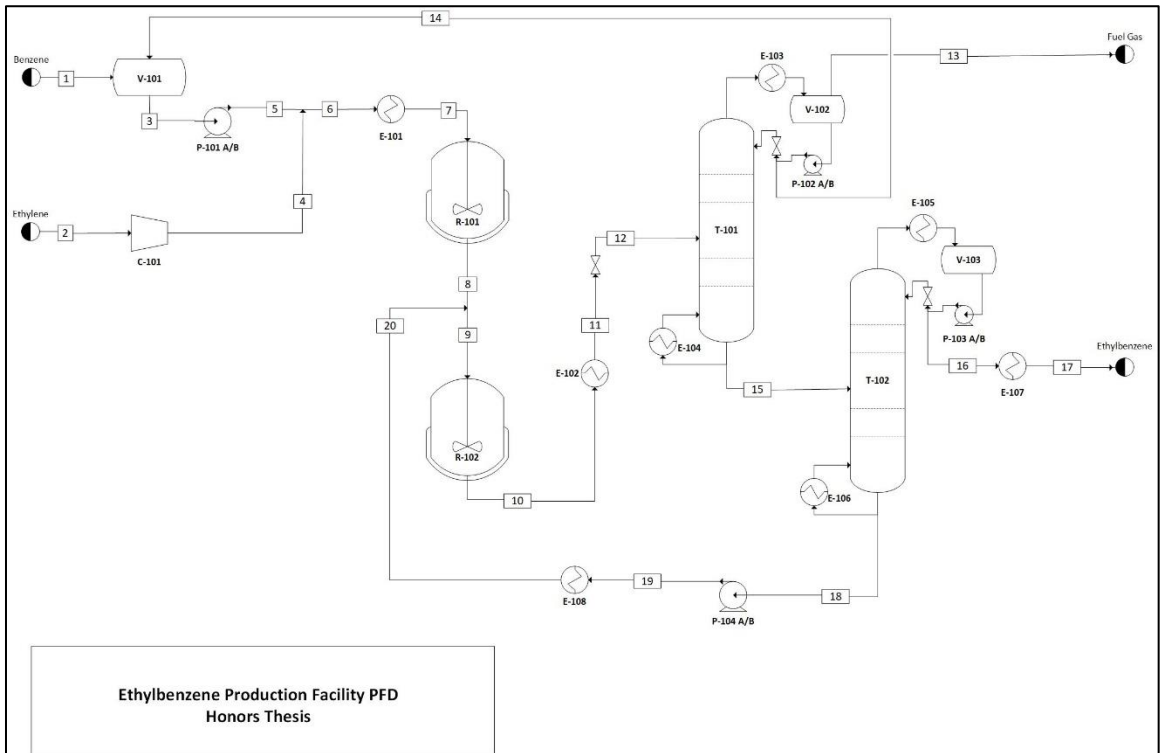


Figure A.2: Ethylbenzene production facility process flow diagram

Table A.1: Ethylbenzene production facility equipment tables

Reactors

R-101 Agitated Jacketed $V_{\text{reactor}} = 56.4 \text{ m}^3$ $P_{\text{Max}} = 68.4 \text{ atm}$	R-102 Agitated Jacketed $V_{\text{reactor}} = 56.4 \text{ m}^3$ $P_{\text{Max}} = 68.4 \text{ atm}$
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Pumps

P-101 A/B Carbon Steel - Positive Displacement Efficiency 75%	P-103 A/B Carbon Steel - Centrifugal Efficiency 75%
P-102 A/B Carbon Steel - Centrifugal Efficiency 75%	P-104 A/B Carbon Steel - Centrifugal Efficiency 75%

Heat Exchangers

E-101 1-2 exchanger, floating head, carbon steel $P_{\text{max}} = 68.6 \text{ atm}$ $Q = 169.9 \text{ kW}$	E-105 1-2 exchanger, floating head, carbon steel $P_{\text{max}} = 1.8 \text{ atm}$ $Q = 979.8 \text{ kW}$
E-102 1-2 exchanger, floating head, carbon steel $P_{\text{max}} = 68.4 \text{ atm}$ $Q = 998.1 \text{ kW}$	E-106 1-2 exchanger, kettle reboiler, carbon steel $P_{\text{max}} = 2.6 \text{ atm}$ $Q = 311.7 \text{ kW}$
E-103 1-2 exchanger, floating head, carbon steel $P_{\text{max}} = 2.02 \text{ atm}$ $Q = 1618.1 \text{ kW}$	E-107 1-2 exchanger, floating head, carbon steel $P_{\text{max}} = 2.9 \text{ atm}$ $Q = 410.1 \text{ kW}$
E-104 1-2 exchanger, kettle reboiler, carbon steel $P_{\text{max}} = 2.8 \text{ atm}$ $Q = 1291.6 \text{ kW}$	E-108 1-2 exchanger, floating head, carbon steel $P_{\text{max}} = 68.6 \text{ atm}$ $Q = 51.26 \text{ kW}$

Compressors

C-101 A/B Carbon Steel - Centrifugal Work = 545.7 kW
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Vessels

V-101 V _{vessel} = 7.61 m ³ Carbon Steel P _{max} = 2.72 atm Horizontal	V-102 V _{vessel} = 7.72 m ³ Carbon Steel P _{max} = 2.72 atm Horizontal	V-103 V _{vessel} = 6.02 m ³ Carbon Steel P _{max} = 1.82 atm Horizontal
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Towers

T-101 Carbon Steel 60% Efficient Trays .5 m tray spacing V _{tower} = 25.94 m ³ P _{max} = 2.32 atm	T-102 Carbon Steel 60% Efficient Trays .5 m tray spacing V _{tower} = 16.09 m ³ P _{max} = 2.62 atm
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Table A.2: Ethylbenzene production facility utility tables

Utilities					
	E-101	E-102	E-103	E-104	E-105
Utility	CW	CW	CW	LPS	CW
m _{flow} (kg/hr)	14,619	85,876	139,223	2,230	84,308
	E-106	E-107	E-108	R-101	R-102
Utility	HPS	CW	CW	CW	CW
m _{flow} (kg/hr)	661	35,289	4,411	258,242	42,744

Table A.3: Ethylbenzene production facility stream tables

Stream Number	1	2	3	4	5	6	7
Temperature (°C)	25.0	25.0	61.5	516.8	61.5	165.0	150.0
Pressure (atm)	1.0	1.0	1.0	66.9	66.9	66.9	66.7
Total (kmol/hr)	86.1	85.9	257.9	85.9	257.9	343.8	343.8
Total (kg/hr)	6722.3	2409.9	20145.1	2409.9	20145.1	22555.0	22555.0
Molar Flowrates (kmol/hr)							
Ethylene	0.0	85.9	0.0	85.9	0.0	85.9	85.9
Benzene	86.1	0.0	257.8	0.0	257.8	257.8	257.8
Ethylbenzene	0.0	0.0	0.1	0.0	0.1	0.1	0.1
Diethylbenzene	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Stream Number	8	9	10	11	12	13	14
Temperature (°C)	150.0	150.0	150.0	71.1	71.1	79.8	79.8
Pressure (atm)	66.7	66.7	66.7	66.5	0.5	0.3	1.0
Total (kmol/hr)	258.7	283.2	282.4	282.4	282.4	0.0	171.8
Total (kg/hr)	22538.4	25777.0	25794.4	25794.4	25794.4	0.0	13422.7
Molar Flowrates (kmol/hr)							
Ethylene	0.9	0.9	0.0	0.0	0.0	0.0	0.0
Benzene	181.4	181.4	171.9	171.9	171.9	0.0	171.7
Ethylbenzene	67.9	68.0	86.0	86.0	86.0	0.0	0.1
Diethylbenzene	8.6	33.1	24.5	24.5	24.5	0.0	0.0

Stream Number	15	16	17	18	19	20
Temperature (°C)	124.1	142.9	50.0	179.2	179.2	150.0
Pressure (atm)	0.6	1.2	1.0	0.9	66.9	66.7
Total (kmol/hr)	110.6	86.1	86.1	24.5	24.5	24.5
Total (kg/hr)	12371.7	9133.1	9133.1	3238.7	3238.7	3238.7
Molar Flowrates (kmol/hr)						
Ethylene	0.0	0.0	0.0	0.0	0.0	0.0
Benzene	0.2	0.2	0.2	0.0	0.0	0.0
Ethylbenzene	86.0	85.9	85.9	0.1	0.1	0.1
Diethylbenzene	24.5	0.0	0.0	24.4	24.4	24.4

Appendix B: Vapor Phase Reaction Ethylbenzene Plant

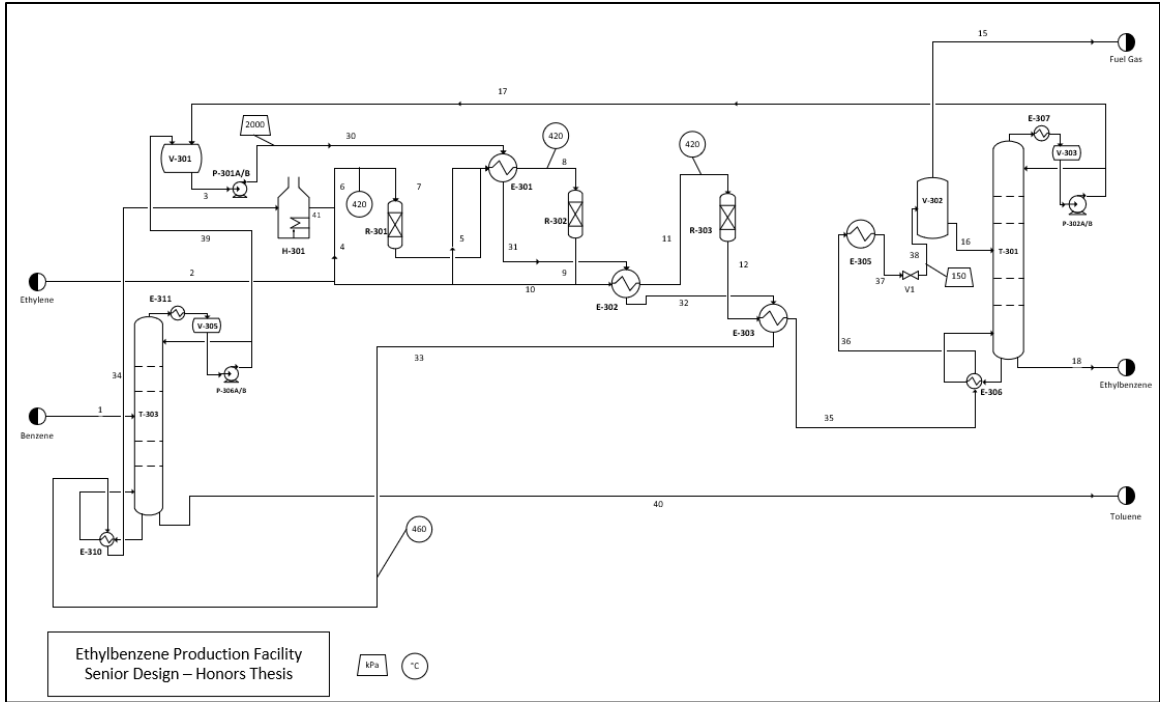


Figure B.1: Ethylbenzene production facility process flow diagram

Table B.1: Stream tables

Stream Number		1	2	3	4	5	6
Temperature	°C	25	25	56	25	25	420
Pressure	kPa	110	2000	110	2000	2000	1985
Vapor Mole Fraction		0.0	1.0	0.0	1.0	1.0	1.0
Flowrate	kmol/hr	102	100	159	30	35	188
Total Mass Rate	kg/hr	8150	2820	12308	845	988	13153
Total Molar Comp. Rates	kmol/hr						
ETHYLENE		0.00	92.98	0.00	27.86	32.56	27.86
ETHANE		0.00	7.04	1.61	2.11	2.46	3.72
PROPENE		0.00	0.00	1.18	0.00	0.00	1.18
BENZENE		92.25	0.00	153.12	0.00	0.00	153.12
TOLUENE		10.25	0.00	1.88	0.00	0.00	1.88
EBENZENE		0.00	0.00	0.71	0.00	0.00	0.71
14EZ		0.00	0.00	0.00	0.00	0.00	0.00
WATER		0.00	0.00	0.00	0.00	0.00	0.00

Stream Number		7	8	9	10	11	12
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Temperature	°C	500	420	519	25	420	509
Pressure	kPa	1979	1979	1972.5	2000	1972.5	1965.8
Vapor Mole Fraction		1.0	1.0	1.0	1.0	1.0	1.0
Flowrate	kmol/hr	164	199	165	35	200	167
Total Mass Rate	kg/hr	13153	14140	14140	988	15128	15128
Total Molar Comp. Rates	kmol/hr						
ETHYLENE		1.02	33.58	0.00	32.56	32.56	0.00
ETHANE		3.72	6.18	6.18	2.46	8.65	8.65
PROPENE		3.06	3.06	3.06	0.00	3.06	3.06
BENZENE		130.07	130.07	96.46	0.00	96.46	63.90
TOLUENE		0.00	0.00	0.00	0.00	0.00	0.00
EBENZENE		25.62	25.62	59.25	0.00	59.25	91.81
14EZ		0.03	0.03	0.00	0.00	0.00	0.00
WATER		0.00	0.00	0.00	0.00	0.00	0.00

Stream Number		15	16	17	18	30	31
Temperature	°C	73	73	15	153	61	131
Pressure	kPa	150	150	150	160	2000	2000
Vapor Mole Fraction		1.0	0.0	0.0	0.0	0.0	0.0
Flowrate	kmol/hr	12	155	64	91	159	159
Total Mass Rate	kg/hr	583	14545	4932	9613	12308	12308
Total Molar Comp. Rates	kmol/hr						
ETHYLENE		0.00	0.00	0.00	0.00	0.00	0.00
ETHANE		7.04	1.61	1.61	0.00	1.61	1.61
PROPENE		1.88	1.18	1.18	0.00	1.18	1.18
BENZENE		2.82	61.08	60.91	0.17	153.12	153.12
TOLUENE		0.00	0.00	0.00	0.00	1.88	1.88
EBENZENE		0.68	91.13	0.71	90.42	0.71	0.71
14EZ		0.00	0.00	0.00	0.00	0.00	0.00
WATER		0.00	0.00	0.00	0.00	0.00	0.00

Stream Number		32	33	34	35	36	37
Temperature	°C	214	460	285	260	199	80
Pressure	kPa	2000	2000	2000	1965.8	1965.8	1965.8
Vapor Mole Fraction		0.1	1.0	1.0	0.9	0.0	0.0
Flowrate	kmol/hr	159	159	159	167	167	167
Total Mass Rate	kg/hr	12308	12308	12308	15128	15128	15128
Total Molar Comp. Rates	kmol/hr						
ETHYLENE		0.00	0.00	0.00	0.00	0.00	0.00
ETHANE		1.61	1.61	1.61	8.65	8.65	8.65
PROPENE		1.18	1.18	1.18	3.06	3.06	3.06
BENZENE		153.12	153.12	153.12	63.90	63.90	63.90
TOLUENE		1.88	1.88	1.88	0.00	0.00	0.00
EBENZENE		0.71	0.71	0.71	91.81	91.81	91.81
14EZ		0.00	0.00	0.00	0.00	0.00	0.00
WATER		0.00	0.00	0.00	0.00	0.00	0.00

Stream Number		38	39	40	41
Temperature	°C	73	84	116	447
Pressure	kPa	150	110	120	1985
Vapor Mole Fraction		0.1	0.0	0.0	1.0
Flowrate	kmol/hr	167	94	8	159
Total Mass Rate	kg/hr	15128	7376	774	12308
Total Molar Comp. Rates	kmol/hr				
ETHYLENE		0.00	0.00	0.00	0.00
ETHANE		8.65	0.00	0.00	1.61
PROPENE		3.06	0.00	0.00	1.18
BENZENE		63.90	92.21	0.04	153.12
TOLUENE		0.00	1.88	8.37	1.88
EBENZENE		91.81	0.00	0.00	0.71
14EZ		0.00	0.00	0.00	0.00
WATER		0.00	0.00	0.00	0.00

Table B.2: Utility table

Stream Name	cw to E-305	rhw to E-307	cw to E-311
Temp (°C)	30	-20	30
Pressure (kPa)	400	400	400
Flowrate 10 ³ kg/h	108	95	94
Duty (MJ/h)	-3948	-4172	-3933

Table B.3: Equipment table

Heat Exchangers

<p>E-301 1-2 exchanger, floating head, carbon steel process stream in tubes Q = 1664 MJ/hr maximum pressure rating of 4,200 kPa</p>	<p>E-302 1-2 exchanger, floating head, carbon steel process stream in tubes Q = 2604 MJ/hr maximum pressure rating of 4,200 kPa</p>
<p>E-303 1-2 exchanger, floating head, carbon steel process stream in tubes Q = 9601 MJ/hr maximum pressure rating of 2,200 kPa</p>	<p>E-305 1-2 exchanger, floating head, carbon steel process stream in shell Q = 3948 MJ/hr maximum pressure rating of 2,200 kPa</p>
<p>E-306 1-2 exchanger, kettle reboiler, carbon steel process stream in shell Q = 5239 MJ/hr maximum pressure rating of 200 kPa</p>	<p>E-307 1-2 exchanger, floating head, carbon steel process stream in shell Q = 4172 MJ/hr maximum pressure rating of 200 kPa</p>
<p>E-310 1-2 exchanger, kettle reboiler, carbon steel process stream in shell Q = 4810 MJ/hr maximum pressure rating of 200 kPa</p>	<p>E-311 1-2 exchanger, floating head, carbon steel process stream in shell Q = 3933 MJ/hr maximum pressure rating of 200 kPa</p>

Pumps

<p>P-301 A/B Carbon steel - positive displacement Efficiency 75%</p>	<p>P-302 A/B Carbon steel - centrifugal Efficiency 75%</p>
<p>P-301 A/B Carbon steel - centrifugal Efficiency 75%</p>	

Fired Heater

<p>H-301 75% thermal efficiency maximum pressure rating</p>	
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Reactors

R-301 stainless steel, packed bed, Adamantium catalyst $V_{bed} = 22 \text{ m}^3$ maximum pressure rating of 2000 kPa maximum catalyst temperature = 525 °C	R-302 stainless steel, packed bed, Adamantium catalyst $V_{bed} = 22 \text{ m}^3$ maximum pressure rating of 2000 kPa maximum catalyst temperature = 525 °C
R-303 stainless steel, packed bed, Adamantium catalyst $V_{bed} = 22 \text{ m}^3$ maximum pressure rating of 2000 kPa maximum catalyst temperature = 525 °C	

Vessels

V-301 $V = 14.85 \text{ m}^3$ carbon steel maximum operating pressure = 250 kPa horizontal	V-302 $V = 5.93 \text{ m}^3$ carbon steel maximum operating pressure = 250 kPa vertical
V-303 $V = 1.47 \text{ m}^3$ carbon steel maximum operating pressure = 300 kPa horizontal	V-305 $V = 2.04 \text{ m}^3$ carbon steel maximum operating pressure = 300 kPa horizontal

Towers

T-301 carbon steel 75% efficient trays .5 m tray spacing maximum pressure rating of 300 kPa	T-303 carbon steel 75% efficient trays .5 m tray spacing maximum pressure rating of 300 kPa
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Table B.4 Chemical Safety

Compound	Molecular weight	Melting Point °C	Boiling Point °C	Flash point/ Auto-Ignition °C	Flammability	Toxin	Irritating	Corrosive
C ₂ H ₄	28	-169.2	-104	/490	High	Suffocation Hazard	Yes	No
C ₃ H ₆	42.1	-185	-45	/460	High	Suffocation Hazard	No	No
C ₆ H ₆	78.1	5.5	80	-11.63/497.8	High	High	Yes	No
C ₆ H ₅ CH ₃	92.14	-95	110.6	4/480	High	Moderate	Moderate	No
C ₆ H ₅ C ₂ H ₅	106.2	-95	136.6	18/432	High	Yes carcinogenic	Moderate	No
C ₆ H ₄ (C ₂ H ₅) ₂	134.2	-42.8	184	55.6/430	High	Yes	Moderate	No

Ethylbenzene Process Description

Fresh 90% mole benzene, Stream 1, is fed into Tower 303 at 25 °C and 110 kPa. The bottoms stream, Stream 40, from this tower is a product stream at 99.5% toluene. The distillate stream, Stream 39, is fed into Vessel 301 at 98% mole benzene, where it is combined with a recycle, Stream 17, that contains mostly benzene at 15 °C and 150 kPa. The exit stream of V-301, Stream 3, is pumped to 2000 kPa by P-301. This stream is fed to the first heat exchanger, E-301, where it cools the reactor effluent to 420 °C. The stream leaving E-301, Stream 31, is fed to E-302 where it cools the second reactor effluent to 420 °C. Stream 32 leaving E-302 then cools the reactor effluent from R-303 to 260 °C. Stream 33 out of E-303 is at 460 °C. This stream is used to heat the reboiler for Tower 303, E-310. It leaves the reboiler as Stream 34 at 285 °C which is fed into the fired heater, H-301, where it is heated to 447 °C. 93% mole Ethylene at 25 °C and 2000 kPa is fed to the process through Stream 2. This stream branches into three streams, 4, 5, and 10. Stream 4 is combined with the outlet stream of the fired heater to make Stream 6 at 420 °C and 1985 kPa. This stream enters the first packed-bed reactor, R-301. This is the first reactor in a series of adiabatic reactors. The partially converted effluent from this reactor is Stream 7 at 500 °C and 1979 kPa and is combined with a fresh stream of Ethylene, Stream 5. The mixed stream is sent through a heat exchanger, E-301, to bring the temperature down to 420 °C. This cooled stream is Stream 8 and feeds the second reactor in series, R-302. The effluent from R-302 is Stream 9 at 1972 kPa, which is combined with the last fresh Ethylene feed stream, Stream 10. This mixture is sent through E-302, which brings down the temperature to 420 °C. The outlet stream of E-302 is Stream 11, which is fed to the final reactor in this series, R-303. The effluent stream of the final reactor in this series, R-303, is Stream 12 at 508 °C and 1966 kPa.

The Ethylene and Benzene react via catalyst to produce Ethylbenzene. There are several side reactions in this process. Ethylbenzene reacts with Ethylene to form Diethylbenzene. Diethylbenzene reacts with Benzene to form Ethylbenzene. Toluene reacts with Ethylene to make Ethylbenzene and Propylene. These exothermic reactions occur according to the following equations:

- (1) $C_6H_6 + C_2H_4 \rightarrow C_6H_5C_2H_5$
- (2) $C_6H_5C_2H_5 + C_2H_4 \rightarrow C_6H_4(C_2H_5)_2$
- (3) $C_6H_4(C_2H_5)_2 + C_6H_6 \rightarrow 2C_6H_5C_2H_5$
- (4) $C_6H_5CH_3 + 2C_2H_4 \rightarrow C_6H_5C_2H_5 + C_3H_6$

Stream 12 enters E-303 which cools the process stream to 260 °C. The exiting stream at 260 °C is used to heat the reboiler in Tower 301 and leaves the reboiler at 199 °C. This stream is Stream 36 which is fed to E-305 where it is cooled with cooling water to 80 °C. The process stream out of E-305 is expanded using a valve from 1966 kPa to 150 kPa where it is fed to V-302. Vessel 302 is a two-phase separator that separates the light gases from condensed liquid. The light gases in Stream 15 are sent to be fuel gas consumed in the fired heater, H-301. The condensed liquid in Stream 16 is sent to the first column, T-301 at 73 °C. The unreacted Benzene in the tower feed stream is separated overhead and recycled to the front of the process through Stream 17 at 15 °C. The bottoms from T-301 is Stream 18 which is the Ethylbenzene product stream at 154 °C. This Ethylbenzene has a purity of 99.8% mole and should contain less than 2 ppm Diethylbenzene. The Benzene out of the top of the first column T-301 in Stream 17 is recycled back to the process. Stream 17 is recycled into V-301.