Benzene-Toluene-Xylene Production Process from Liquefied Petroleum Gas Using Aspen HYSYS and Aspen Energy Analyzer

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Abstract: - **In this work, Aspen HYSYS simulation software was used to develop a process for Benzene-toluene-xylene BTX synthesis; a leading feedstock in the petrochemical industries for the synthesis of several chemicals, from liquefied petroleum gas LPG. The pinch and retrofit analysis as applicable to the developed model was carried out using Energy Analyser V10 in order to develop a network of heat exchanger, possible energy savings and payback period of the process. The result of the simulation showed that benzene, toluene and xylene have compositions of 89.3%, 95.5%, and 100% respectively suggesting that the process is highly favoured by low pressure and high reflux ratio. The energy analysis revealed that fifteen (15) heat exchanger network (HEN) were developed comprising of seven (7) heaters and eight (8) coolers for the base case simulation and about 0.6MWof wasted heat was recovered by incorporating three (3) additional heat exchangers to the base HEN leading to 9.6% total energy savings of the entire system. The retrofit design suggested three case scenarios yielding payback periods of 2.062 years, 1.802 years and 6.579 years respectively among which the one the second case scenario with 1.802 years was selected to be the preferred case scenario.**

Keywords: **Aspen HYSYS, Pressure, Reflux, Aspen Energy Analyzer, Heat Exchanger Network (HEN), retrofit.**

I. INTRODUCTION

Aromatization of liquefied petroleum gas (LPG) to benzene, toluene and xylene (BTX) has drawn attentions to process engineers since it was proven to produce high purity BTX using Aspen HYSYS[1]. BTX have several applications in the synthesis of other chemicals especially petrochemical products like nylon, detergents, paints, drugs, glues etc. and also to the extraction of benzene and blending of gasoline [2]. The process follows the cycler process for synthesis using naphtha feedstock with high demand of energy and low product purity[3][4].Because of sulfur treatment prior to naptha reforming, the cost of production of BTX becomes unreasonably expensive [5]. The introduction of low chain alkanes produces higher purity BTX in comparism to naphtha feedstock [3]. The reaction is strongly endothermic and requires high temperature of about 400-540^{0}C to activate the reaction and yield better products[6].Following the recent innovation for the design of a BTX plant from LPG feedstock, it is relevant to investigate the effect of several parameters to enable quality products. Factors such as reaction temperature, pressure and type of catalyst have significant effect on the conversion of reactants to product [7]. The type of fractionating columns used for industrial applications and its mode of operation contributes to how well products are formed [8]. In fractionating towers, column pressures, reflux ratios, boil-up ratios, feed rate, distillate rate, number of plates etc. are all very significant to determining products purity [9]. Employing process integration for the development and modification of new design strategies to optimise plant operations also determines how detailed a designed process is[10][11].

As a new engineering trend in energy efficiency, pinch technology is applied to design an energy efficient systems [12]. It is commonly used in developing heat exchanger network(HEN) and utilities of process(es) to minimize cost, reduce energy consumptions and emission of harmful substances to the environment[13][14][15]. The recoverable energy in pinch analysis depends on the minimum change in temperature (ΔTmin) which is considered as the pinch temperature [10]. Pinch analysis is best described in terms of composite curves, grand composite curves, grid diagram and grid diagram for retrofit analysis. Hot streams requiring cooling and cold streams requiring heating need to be identified and marched below and above the pinch separately to utilize wasted heat [16]. Secondly, zero enthalpy is required across the pinch [13]. Thirdly, Excess flow of energy in heat transfer equipment add up to the capital cost by increasing heat transfer area[17]. The retrofit is used to perform modification on the existing HEN and essential to balance the problem between energy savings and capital investment [18]. Retrofitting of process can reduce the cost of utilities and the pay back period of process.

This research work uses Aspen HYSYS V10 to develop a detailed BTX plant and Aspen energy analyzer V10 to perform pinch analysis on the simulated BTX plant from LPG.

II. METHODOLOGY

The process consists majorly of one reactor, one heater, two coolers, three pumps and six distillation columns. Figure 1

shows the process flow diagram. The process stream data was obtained from the base case Aspen HYSYS V10 and transferred to Aspen Energy Analyzer V10. Tables 1, 2 and 3 shows the extracted process stream data.
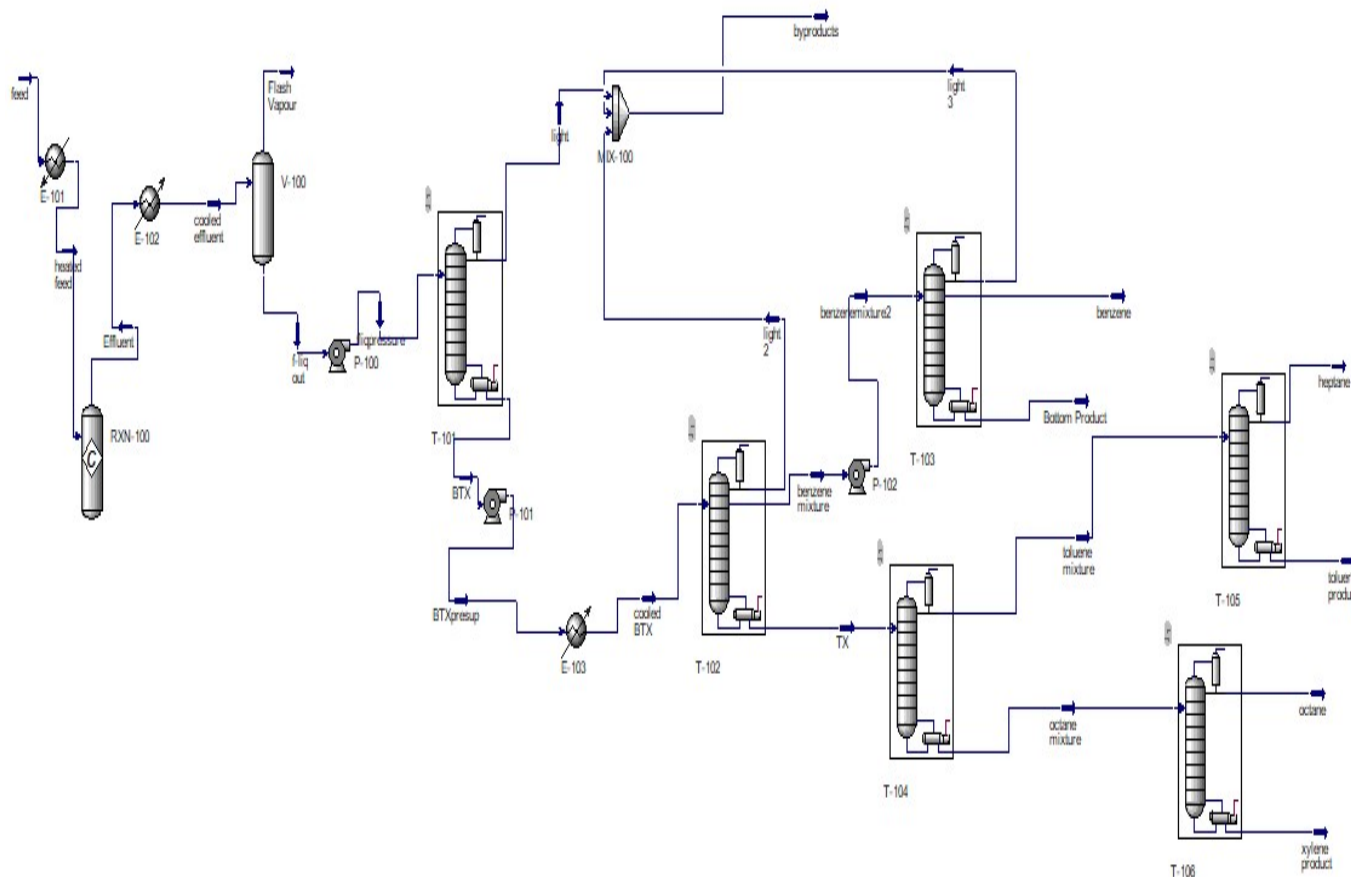

Fig. 1: Process flow diagram for aromatization of BTX from LPG

TABLE I

FEED OPERATING CONDITION

Component name	Value
Propane (kmol)	80
Butane (kmol)	20
Temperature (^{0}C)	150
Pressure (kPa)	120
Mass flow (kg/hr)	4690

III. RESULT AND DISCUSSION OF RESULT

A. Effect of feed operating conditions on the fractional formation of BTX in the reactor

1) Pressure Effect

Figure2 shows the effects of feed pressure on the reaction of LPG to BTX. No product gradient was observed as the pressure increases from 0-500 kPa. This shows that unaffected yield by pressure indicates that the operation can be carried out at either atmosperic, higher or vacuum pressure. Therefore, safe pressure can be suitably selected for the reactor operations.
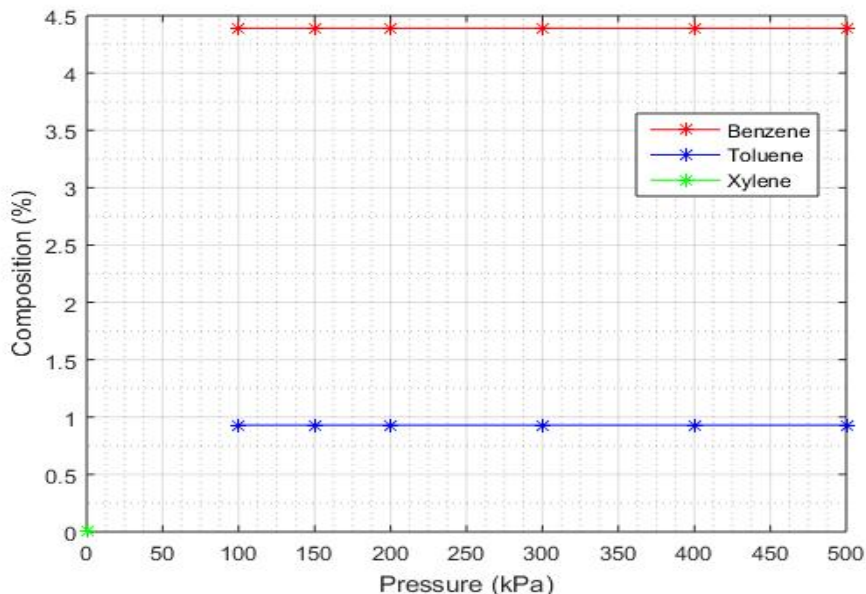
Fig.2: Effect of feed pressure on the yield of BTX

2) Temperature Effect

Figure 3 shows the temperature effect on the composition of BTX. The analysis was carried out at a temperature between 100^0C and 550^0C and it was observed that there was no product formation between 0-300C. However, benzene and toluene were formed at a reaction temperature of 320^0C which form the activated reaction temperature for benzene and toluene while the reactants were inert at same temperature. As temperature increases, there was corresponding increase in the composition of both benzene and toluene from 0.89% and 0.05% at 320^0C to 4.39% for benzene and 0.93% for toluene at 395^0C respectively while xylene formation with composition of 0.01% was observed at 385C which remained constant with increasing temperature. Complete product (BTX) formation from LPG was achieved at 395C which gave a better result as compared to a process using naphtha were product formation can only be achieved at 500C.
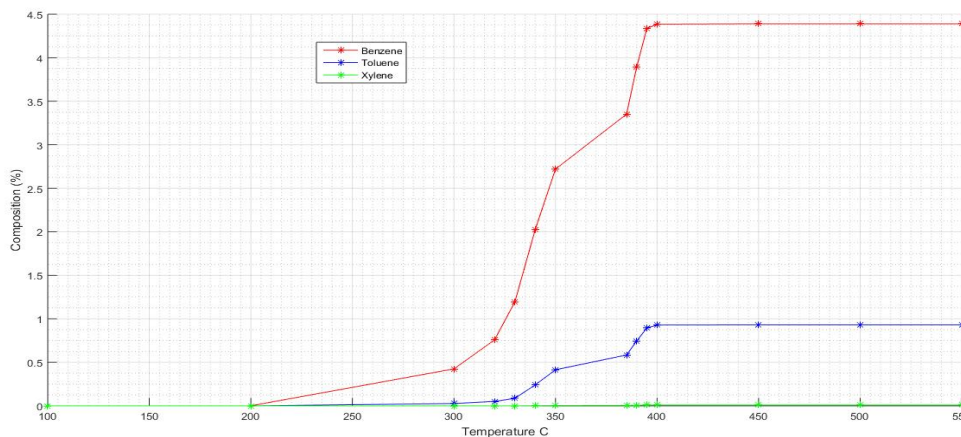

Figure3: Feed temperature effect on the formation of benzene-toluene-xylene

B. Effect of Distillation Column Operating Conditions on the Purification of BTX

1) Effect of Condenser and Reboiler Pressures on the purification of BTX in BTX Stripper (COL1)

Figure4 shows the variation in condenser and reboiler pressures from 100 kPa to 170 kPa in COL1. At all pressures, BTX composition in the column remains constant. Since no concentration increase was observed, higher pressures will inevitably increase the temperature and so much energy will be wasted in doing no work. It is more convenient to operate at a safe working pressures for this system.
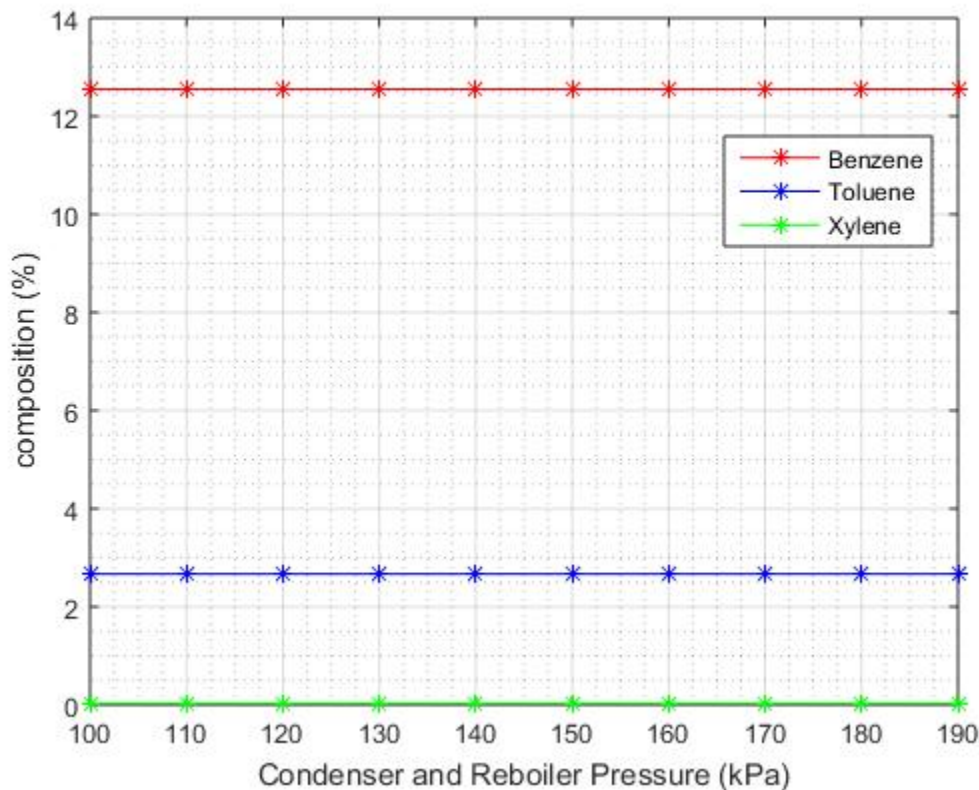
Fig. 4: COL1 (BTX Stripper from lighter hydrocarbons)

2) Effect of Reflux Ratio on BTX formation in Column One (COL1)

Table 7 shows the effect of reflux ratio to percentage purity of BTX. No increase in composition of the products was observed with increasing reflux ratio.
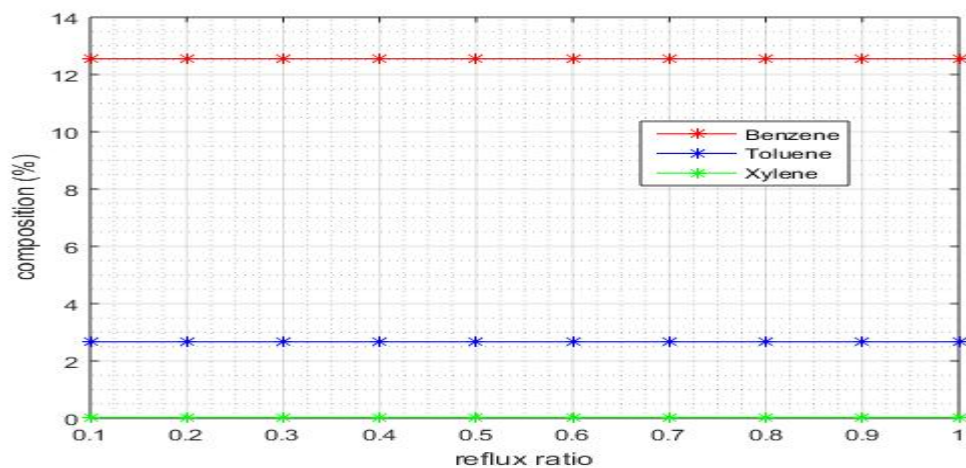

Fig. 5: Effect of reflux ratio

3) Stage Conversion of BTX in COL1

COL1 is a stripper that removes light components such as methane, ethane and ethylene. The column contains 9 theoretical trays in addition to one condenser (denoted by zero in figure 6) and a reboiler (denoted by ten in figure 6). With the lighter product going to the top condenser and BTX exiting the reboiler. As shown in table 8, there is no increase in composition from stage 1-4 for all the products. However, at tray 5, the products purity increases. BTX has higher

composition in the reboiler compared to the other stages. Therefore, no side draws are required as the reboiler draw
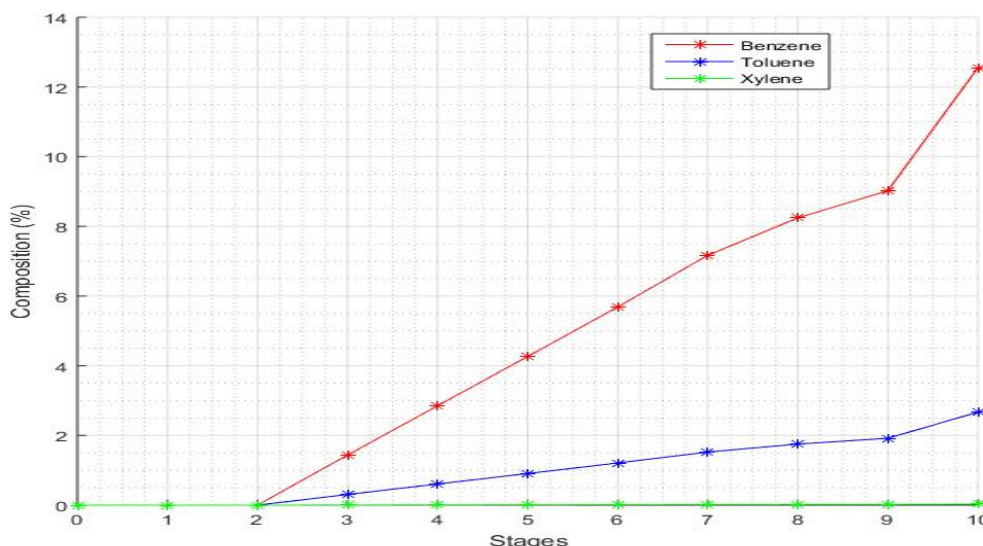
contains more of BTX.

Fig.6: Tray composition of BTX along COL1

4) Effect of Condenser and Reboiler Pressure on Column Two (COL2)

Figure 7 shows the effect of condenser and reboiler pressures to the composition of BTX. The lighter product being benzene is the top product drawn from stage 4 close to the condenser, while toluene-xylene are both bottom product drawn from the reboiler. when the condenser pressure is increased from 10kPa to 130kPa, there was a noticeable decrease in the benzene purity from 29.53% to 28.40% respectively. At the reboiler, Toluene was observed to slightly decrease in composition from 53.37% at 20 kPa reboiler pressure to 53.16% at 120

kPa, but increases again to 53.39% at 130 kPa then drops to 53.36% at 140 kPa displaying irregular response to pressure increase. Increase in pressure leads to corresponding increase in temperature thereby leading none separation of the reactants at the increased operating conditions does not favour separation of the product. However, there was a slight increase in the composition of xylene from 0.73% at 10kPa to 0.74% at 40kPa beyond which pressure has no significant role in composition of xylene. or maximum yield the condenser pressure should be operated between 10kPa to 60kPa (preferably 10 kPa) while the reboiler pressure should be operated between 30 kPa to 50 kPa.
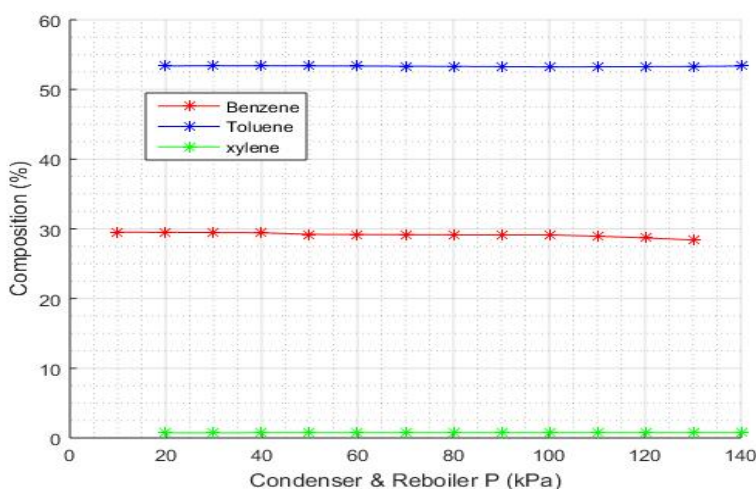

Fig.7: Benzene Stripper from Toluene-Xylene (TX) (COL2)

5) Stage Composition of BTX in COL2

Figure8 shows the stage conversions of BTX in COL2. Seventeen (17) theoretical stages were required for the separation the condenser was denoted as stage zero (0) while the reboiler was denoted as the stage eighteen (18), the rest of the stages retained one to seventeen (1-17). Benzene was seen to be very low at stages closed to the reboiler. High composition of benzene was seen at stages between 3 to 6. These stages qualified for desired draw stage of benzene, but tray 4 yields better conversion than the rest of the trays. However, TX composition was seen to approach a maximum composition of 53.12% and 0.74% on getting to the reboiler. Therefore, tray 4 was considered as the required draw stage for benzene while the reboiler was considered the draw stage for TX.
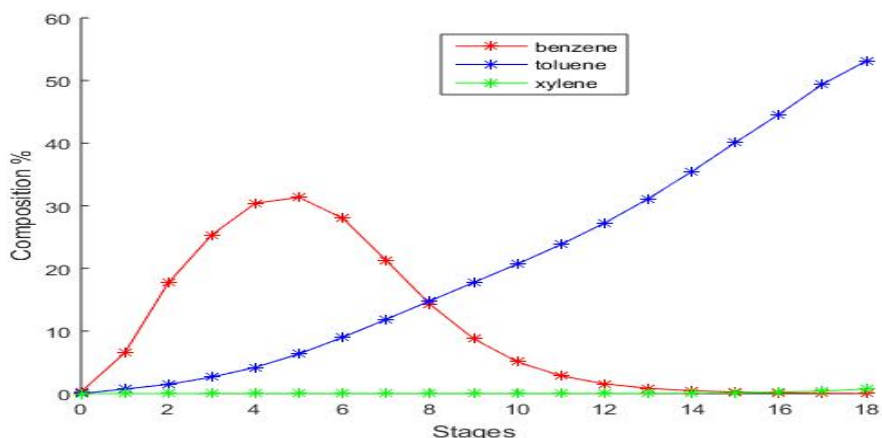

Fig.8: Stage composition of BTX in COL2

6) Effect of Condenser and Reboiler Pressure on Column Three (COL3)

Figure 11 shows the effect of condenser and reboiler pressures (kPa) on the separation efficiency of benzene from n-hexane, n-pentane, 1-hexene and 1-pentene. The column has 20 trays. The top of the column at the condenser draw contains n-hexane, n-pentane, 1-hexene and 1-pentene, while the bottom draw is rich in n-heptane and benzene is drawn from side at stage 13. At the condenser, n-hexane, n-pentane, 1-hexene and 1-pentene shows no significant changes with changes in pressures. At the reboiler, the composition of n-heptane shows a significant decrease as the pressure increase from 5 kPa to 130 kPa. The peak pressure for maximum composition of 67% for the bye-product n-heptane was stable at 5 kPa. At 130 kPa, the composition drops to 59%. Meanwhile, benzene shows significant increase with decrease in both condenser and reboiler pressures attaining a composition high of 89.29% at 5 kPa condenser and reboiler pressure. Low pressure tends to be more significant in the separation columns than higher pressures.
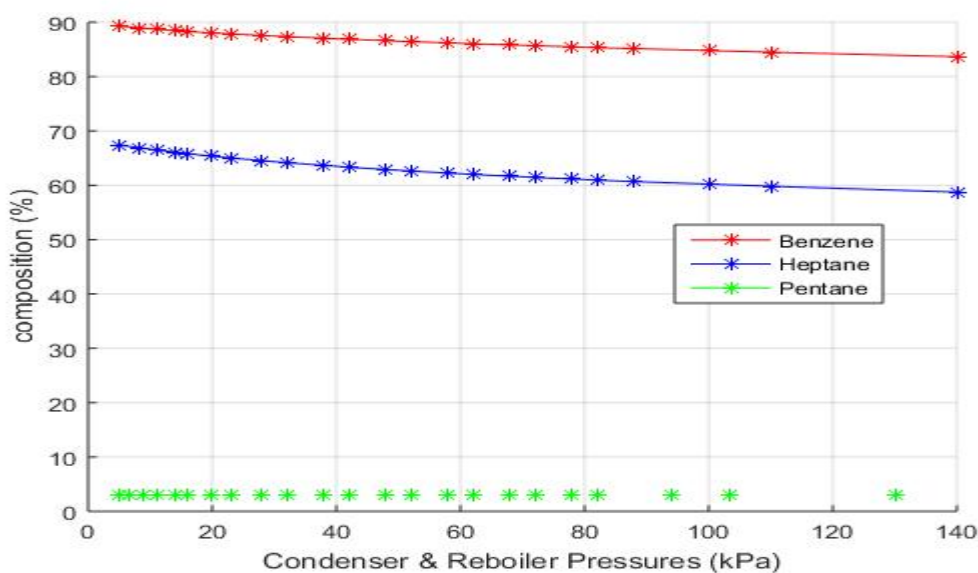

Fig.9: Condenser and reboiler pressure effects on the composition of benzene

7) Stage Composition of Benzene COL3

Figure 12 shows the estimated stage composition from top to bottom of benzene across COL3. 20 stages were required for the seperation, 0 represent the condenser and 21 represent the stage in the reboiler. At the condenser, the composition was 6.87%steadily increases down the column attaining the highest composition of 92% then gradually drops to 89% at tray 13 then drops steadily to 2.207% in getting to the reboiler. This implied that benzene does not require high temperatures and pressures to attain high composition as mid column temperature are moderate compared to high temperature reboiler and low temperature condenser. Therefore, pure benzene can be harnessed from stage 7 to 14 with compositions ranging between 86% to 90%.
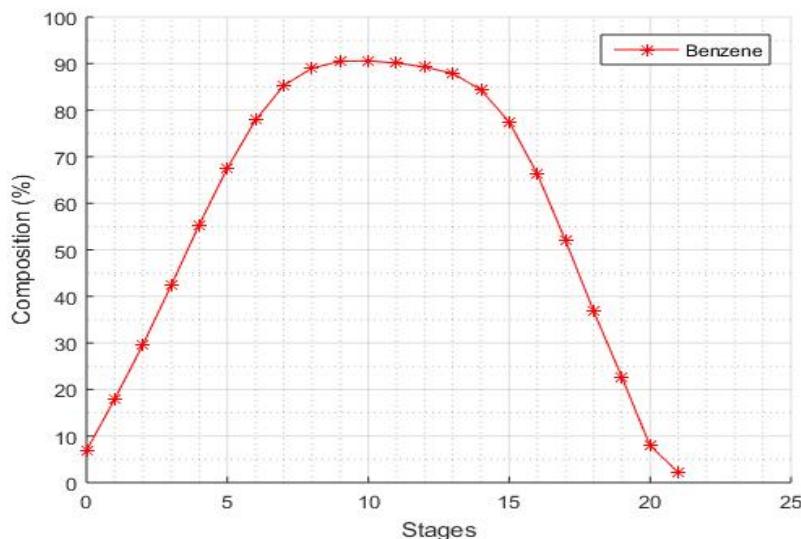

Figure 10: Stage composition of benzene at COL3

7) Effect of Reflux Ratio on the Composition of Benzene in COL3

Figure 11 shows the effect of reflux ratio on the composition of benzene. At no reflux ratio, a composition of 45% was observed in the COL3. Steady increase in reflux ratio was seen with progressive reflux. 80% composition was recorded at a reflux ratio of unity which implied same volume of distillate returned to the column as the quantity leaving. However, benzene composition steadily increases to 90% at a reflux ratio of 11 and remained constant at larger reflux ratio implying more of the distillate returned to the column to increase benzene composition, maximum reflux ratio of 11 is recorded for COL3.
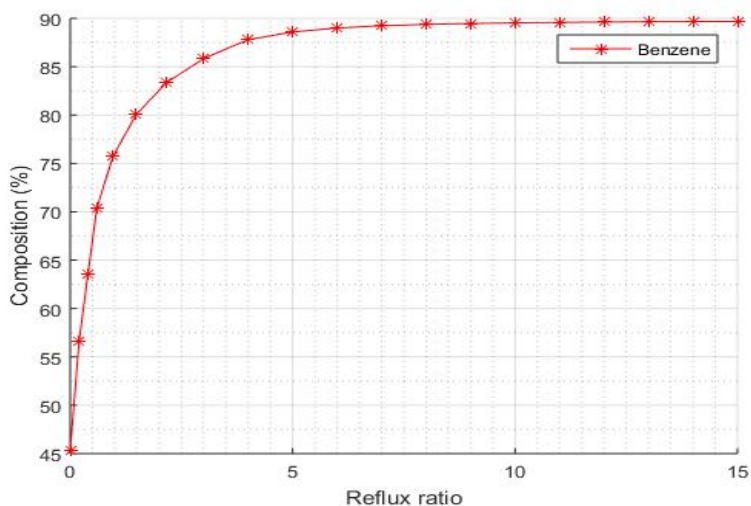

Fig.11: Effect of reflux ratio to the composition of benzene

7) Effect of Condenser and Reboiler Pressures on Column Four (COL4)

Figure 12 shows the composition effect of toluene and xylene with condenser and reboiler pressures in COL4. The figure depicts that condenser and reboiler pressure increase has no significant effect to the composition of toluene and xylene. Therefore, low and safe pressure that will minimize energy consumption is more suitable for this column.
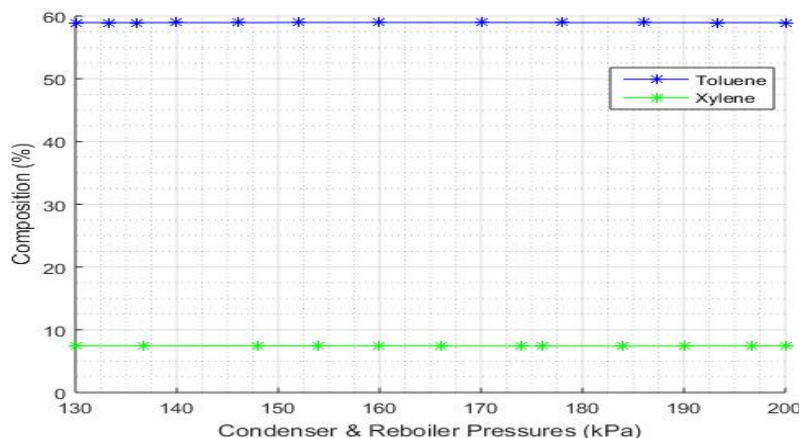

Fig. 12: Condenser and reboiler pressure effects on the composition of toluene and xylene

8) Stage Composition of Toluene and Xylene in COL4

Figure 13 shows the composition of toluene and xylene in COL4. The tower contains 30 theoretical stages, a condenser and a reboiler. Xylene being the heaviest component, little traces of xylene was found in the condenser until the seven stage. Maximum composition of about 7 % was found in the reboiler for xylene. However, toluene concentration was 60% in the condenser but gradually increases as it approaches tray 4. At tray 4, the maximum composition of 66% was recorded after which it steadily drops to zero at the reboiler. With this regard, to obtain highly rich toluene, it should be drawn between tray 1 and tray 4 as the composition range is between 63-66% respectively. The reboiler gives high composition of xylene with 7% and therefore qualified as the draw stage for xylene in COL4.
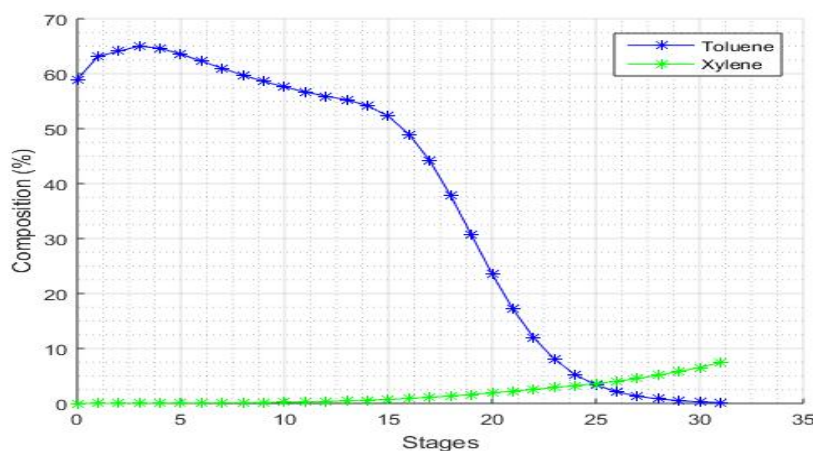

Fig.13: Stage Compositions of Toluene and Xylene in COL4

9) Effect of Condenser and Reboiler Pressures on Column Five (COL5)

COL5 seperates toluene from heptane. The lighter product is heptane and the heavier product is toluene. At a pressure of 5 kPa, toluene has a composition of about 89.7% while haptane has a composition of 83.5%. With increase in presssure from 5 kPa to 180 kPa, a slight decrease from 89.7% to 88.4 % was observed while heptane drops to 81.60% at 160 kPa. This implied that higher pressures have little significant decreases in composition of both toluene and heptane. Operations should be handled at cheap and safe pressures in COL5.
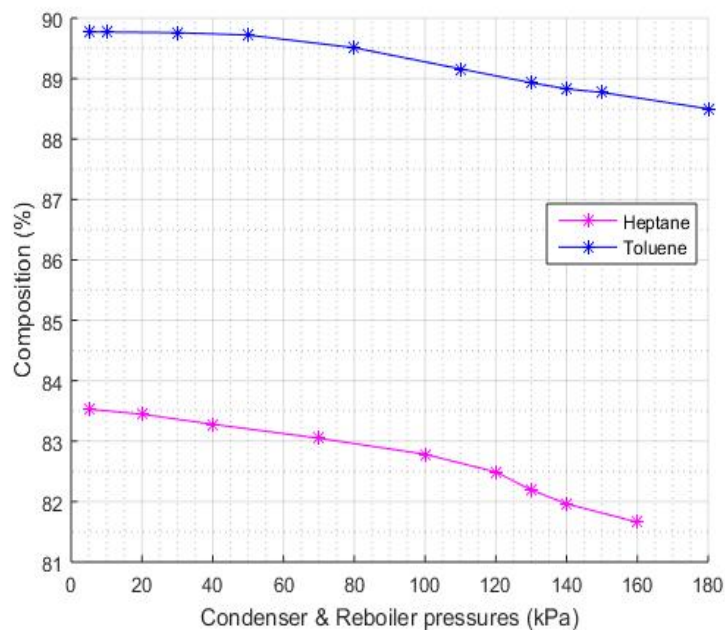
Fig. 14: Effect of condenser and reboiler pressures on COL5

10) Effect of Reflux Ratio on the Composition of Toluene in COL5

Figure 15 shows the effect of reflux ratio on the composition of toluene in COL5. At no reflux ratio, composition of 60% was recorded. About 70% composition was recorded at unity implying same volume returned to the top of the column as they left while the composition of toluene significantly increased. 90% composition was attained at a reflux ratio of 10 after which it the composition became constant at infinite reflux ratio. This signifies that higher composition will be attained if more of the distillate returned to the column than the leave. Therefore, maximum reflux ratio is attained at 10.
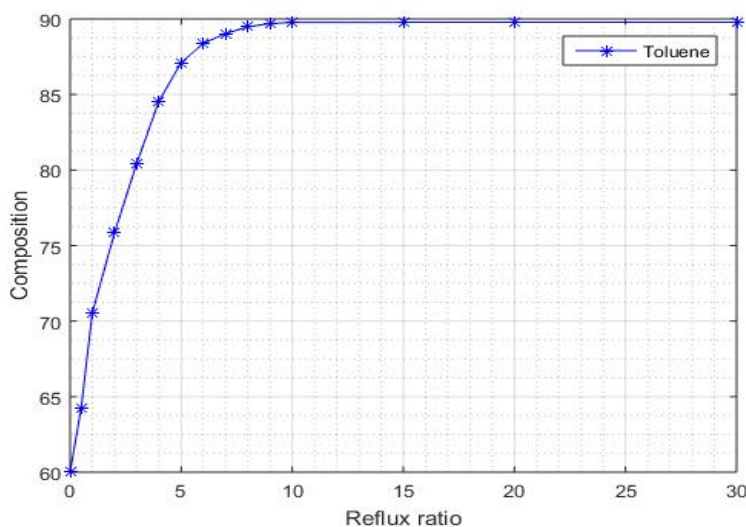

Fig. 15: Effect of reflux ratio on toluene purity in COL5

11) Stage Composition of Toluene in COL5

Figure 16 shows the composition of toluene at COL5. 30 trays were required for this seperation with the condenser denoted as stage zero and the reboiler denoted 31. The composition of toluene increases with the number of stages. Maximum composition of 95% was seen between trays 18-20 before gradually declining to 30% at the reboiler. Therefore, tray 20 is selected as the draw stage for toluene.
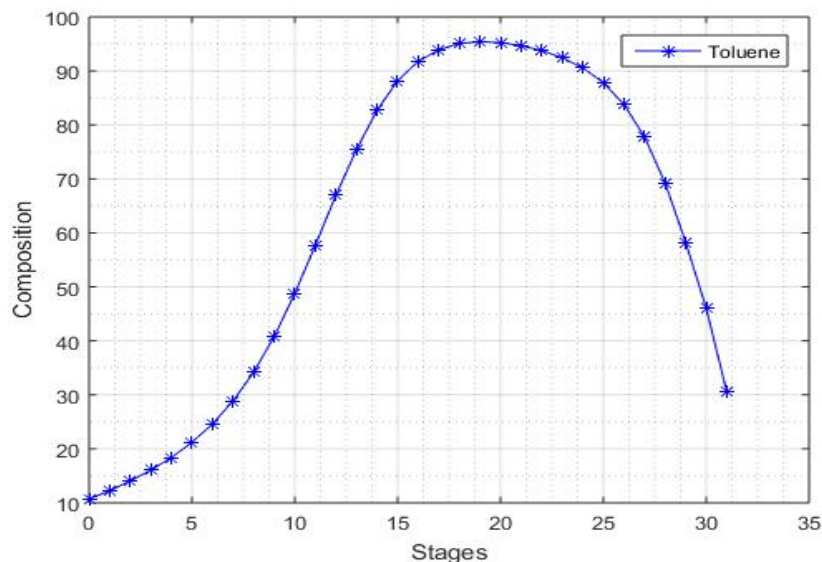
Fig.16: Stage composition of Toluene inCOL5

12) Effect of Condenser and Reboiler Pressures on Column Six (COL6)

Figure 17 shows the effect of condenser and reboiler pressure on the composition of xylene. Octane is drawn from the condenser and toluene is drawn from the reboiler. At 5 kPa reboiler pressure, the composition of xylene is 100%. Composition droped to 89% as the pressure increases from 5 kPa to 150 kPa. In general, xylene is not deeply affected by pressure changes. Composition of octane shows no significant decrease with increasing pressure as it droped from 97% at 5 kPa to 96.5% at 150 kPa.
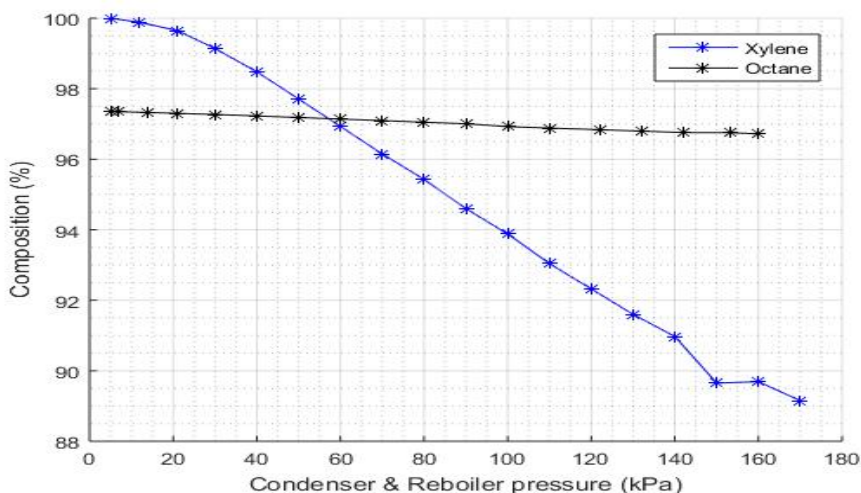

Fig. 17: Effect of Condenser and reboiler pressures on the composition of xylene in COL6

13) Effect of Reflux Ratio on the Composition of Xylene in COL6

Figure 18 shows the effect of reflux ratio on the composition of toluene and octane in COL6. At no reflux ratio, a composition of 20% was recorded for xylene. However, the composition increases steadily with an effect in reflux ratio. At unity, a composition of 44% was seen, implying that even when same volume of distillate returns to the column than they leave, a low reflux ratio was recorded. From the figure, it can be depicted that the purity of xylene solemnly depends on the reflux ratio. A composition of 100% was recorded at reflux ratio of 20 implying a large volume of distillate will return to the column. However, octane composition is slightly affected by increases in reflux ratios but a steady increase was observed from 92% at zero to 98% at 9 reflux ratios. Therefore, more of the distillate should be returned to the column for higher composition.
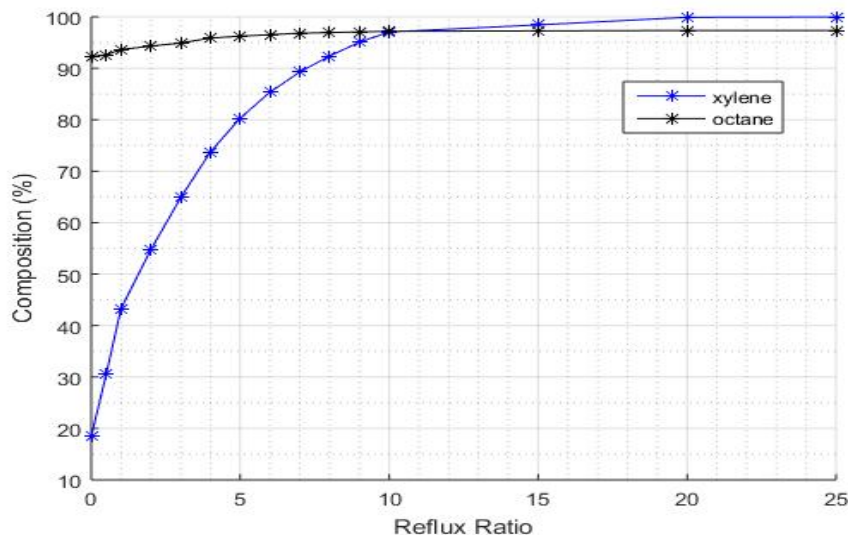
Fig.18: Effect of reflux ratio on the composition of xylene in COL6

14) Stage Conversion of Xylene at COL6

Figure 19 shows the stage conversion of toluene in COL6. Twenty-six (26) theoretical stages were required for the seperation of toluene from octane. The condenser is represented as 0 and the reboiler is represented as 26.Being the heavier component, the composition of xylene is low as 2% at the condenser but essentially increases down the column. Composition of 100% was seen at the reboiler. With this regard, the drawn stage for the purest composition is at the reboiler after tray 25.
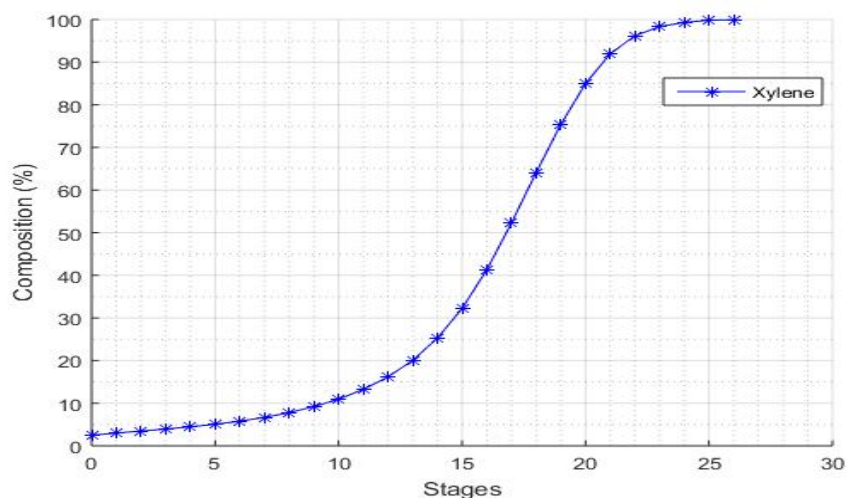

Fig. 19: Composition of xylene along COL6 to determine the draw stage

C. Energy Analysis using Aspen Energy Analyzer V10

1) Process Data Extraction

The data extracted from the process as shown in figure 4 and 5 contains seven cold streams and eight hot streams respectively. The utilities are low pressure (LP) steam generation, air, cooling water, four (4) refrigerants, very low temperature, fired heater, hot oil, high pressure (HP) steam, medium pressure (MP) steam and LP steam.

After the process data is extracted, pinch analysis is applied to obtain the heat exchanger network (HEN) design. Retrofit analysis is performed on the HEN case to obtain the possible energy savings and heat exchanger repositioning. Table II and III shows the cold and hot stream data respectively.

The process flow diagram in figure 1 shows that heat was exchanged in heaters and coolers which combination forms the grid diagrams representing the HEN of process streams, utility streams, split mixers and heat exchangers in figure 4. The blue colours in figure 2 represent cooling utility heat

exchangers while the red represent heating utility exchangers respectively. The silverish colour represent a normal network exchanger for heat recovery.

TABLE II

PROCESS COLD STREAM DATA

Name	Cold Tin ^{0}C	Cold Tout ^{0}C
Feed_To_Heated Feed	25.0	500.0
ToReboiler@COL3_To_Xylene product@COL3	164.3	185.8
ToReboiler@COL4_To_xylene@COL4	147.5	147.6
ToReboiler@COL6_To_toluene@COL6	131.9	132.1
ToReboiler@COL2_To_boilup@COL2	115.1	117.1
ToReboiler@COL1_To_boilup@COL1	110.4	113.6
ToReboiler@COL5_To_btx2@COL5	26.9	72.0

TABLE III

PROCESS HOT STREAM DATA

Name	Hot Tin ^{0}C	Hot Tout ^{0}C
Effluent_To_cooled effluent	193.1	-120
To Condenser@COL3_To_Octane@COL3	151.7	151.6
To Condenser@COL4_To_Reflux@COL4	117.1	115.4
To Condenser@COL6_To_Reflux@COL6	115.1	115.0
BTXPresup_To_cooled BTX	72.0	25.0
To Condenser@COL1_To_ogt@COL1	57.8	38.6
To Condenser@COL2_To_Reflux@COL2	10.7	-54.9
To Condenser@COL5_To_light 2@COL5	-47.3	-89.3

The total energy requirement for the process as calculated by Aspen Energy Analyzer was 6.243 MW (3.413 MW for heating and 2.830 MW for cooling).
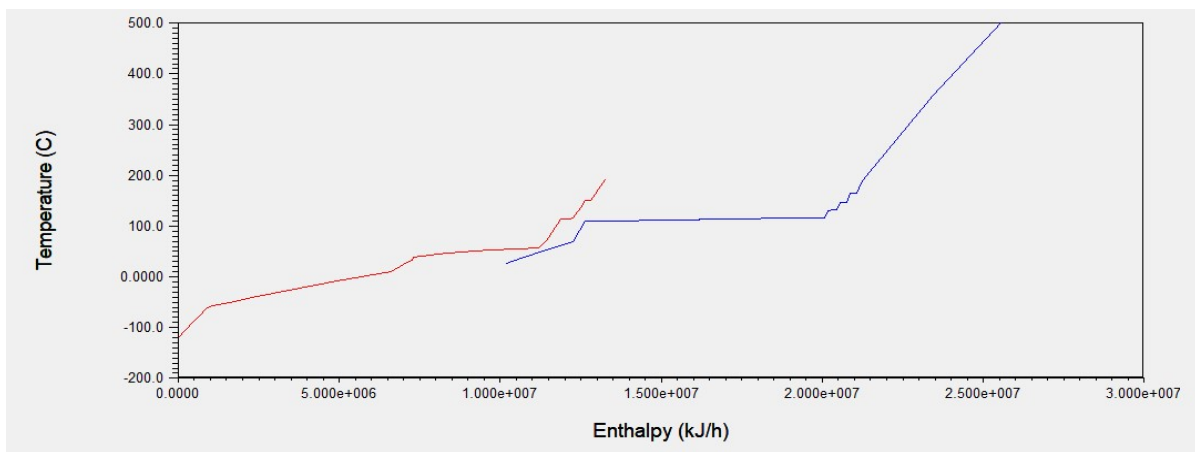

Fig.20: Grid diagram for base case

Figure 20 shows the composite curve for the design prior to optimization. In the figure, there exist an excess un-utilized energy of about 1.29MW from 193.1°C to -120 °C. The difference existing between the hot and cold temperature in the composite curve shows the possibility of recovery within that range of enthalpy and temperature.
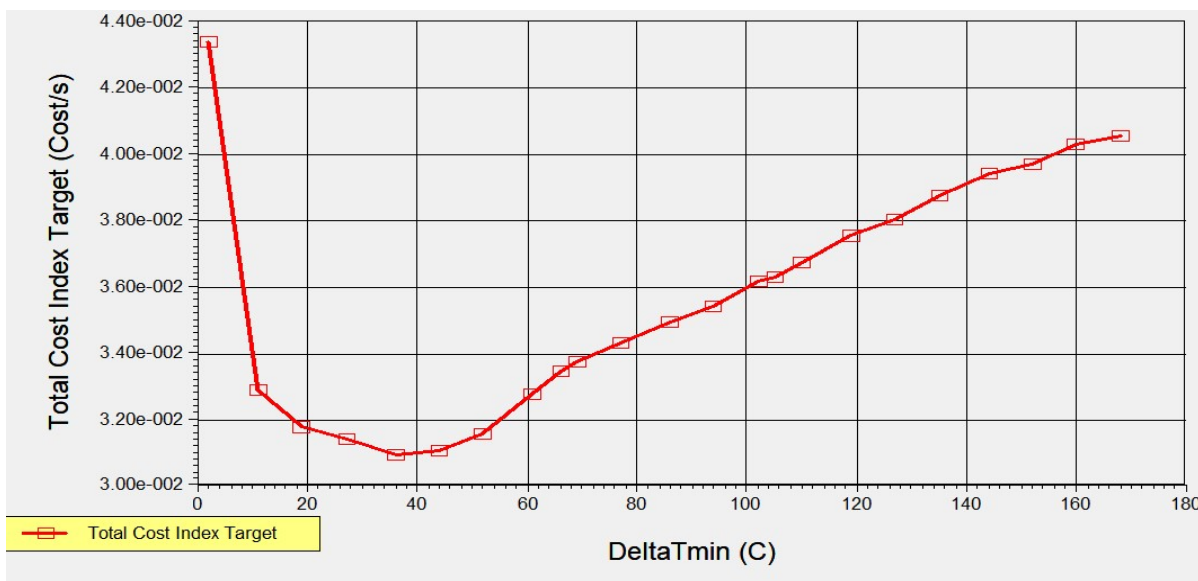
Fig. 21: Range target plot

Figure 3 shows the range target plot for optimization of the minimum approach temperature ΔTmin obtained by the minimization of the total annual cost which imply the best compromise between utility requirement, heat exchanger area and unit shell number. The ΔTmin varies with the total annual cost. In this process, the ΔTmin is calculated as 11^0C. The ΔTmin shows the minimum number of heat exchanger required for maximum recovery.

2) Performance Evaluation

Figure 4 shows the payback-investment curve from various scenarios of the retrofit case generated from Aspen Energy Analyzer. These scenarios give the best performance design yielding low packback period and less operating cost
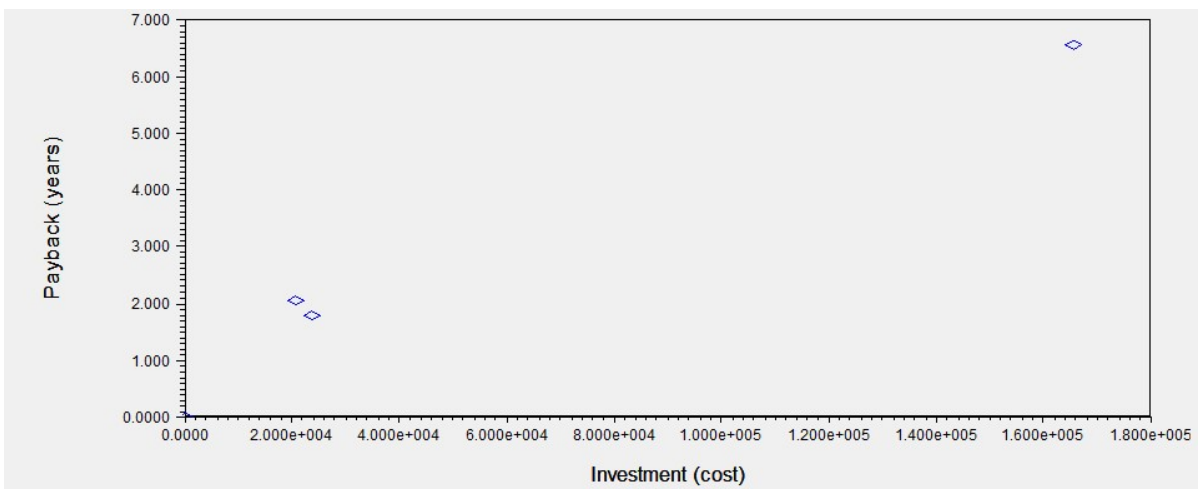

Fig. 22: Payback-Investment curve

Three scenario cases were generated for the design. The first case scenario as shown in figure 4generated a design that yielded a payback period of 2.062 years with an investment cost index of 20630. To achieve this, the retrofit design modified the reboiler utility heat exchanger at COL2 by shifting heating utility from MP steam to LP steam. This design saves about 2.87% of the total heating cost index required for the system but increase the surface area for heat transfer. The second case scenario generated a design yielding a payback period of 1.802 years with an investment cost index

of 23630. The retrofit design modified the condenser utility heat exchanger at COL6 from air cooler to cooling water. This increased the investment cost because of the increase in heat exchanger, but minimize the power requirement and operating cost of air coolers. The cooling load of the process was increased by 26% but saves about 0.614% of the total cooling cost index in addition to the 2.4% of the heating utility developed for the first case. In this case, the cost of using LP steam is small compared to MP steam. However, it is sufficient to use LP steam with larger area than MP with

smaller area because of cost of MP compared to LP. The third case scenario generated a design yielding a payback period of 6.579 years with an investment cost index of 165600. The retrofit design in addition to the first and second scenarios modified the utility condenser heat exchanger at COL1 from cooling water to air cooler. This scenario enormously raises the cost and maintenance of the air cooler. This scenario does not change the heating and cooling conditions but increases the heat transfer area. Table IV shows the performance summary of the case scenarios. looking at the three case scenarios, the second scenario yielded the best retrofit design.

TABLE IV

PERFORMANCE SUMMARY

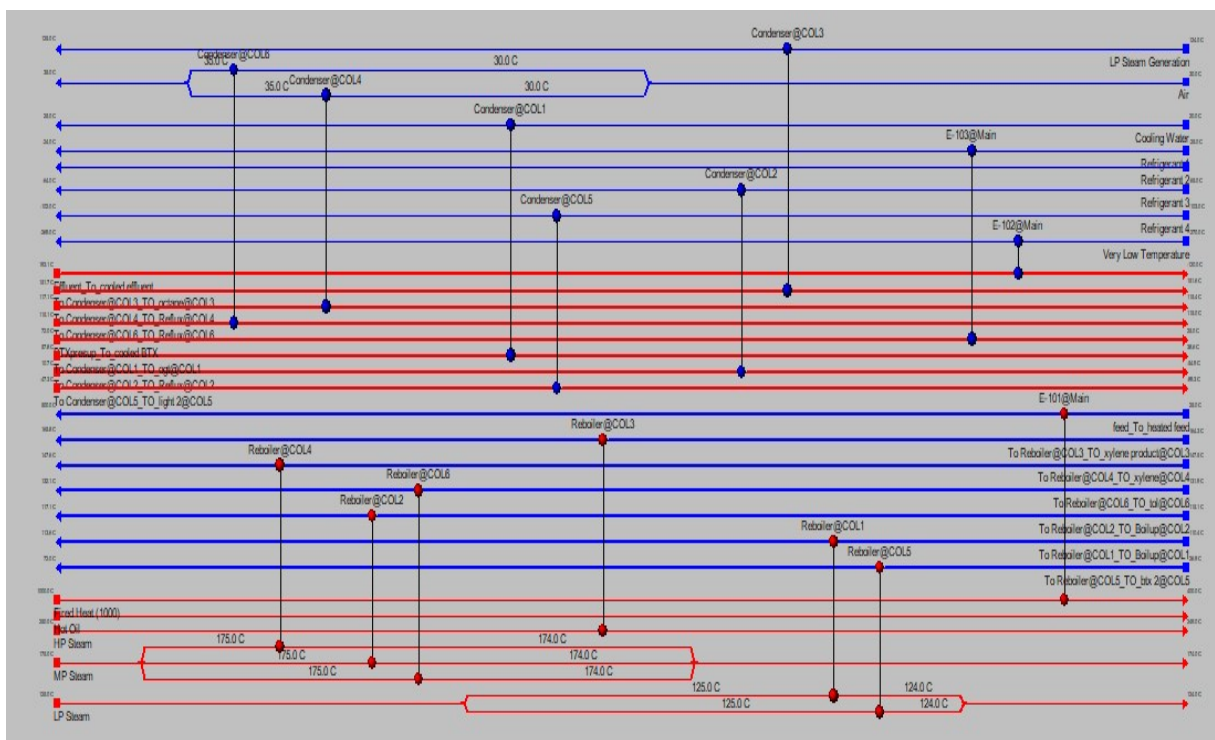| Scenarios	Base Scenario	Scenario 1	Scenario 2	Scenario 3
Heatingcost index (cost/s)	0.0114	0.01082	0.01082	0.01082
Heating load (MW)	3.782	3.782	3.782	3.782
Cooling cost index (cost/s)	0.01465	0.01465	0.01456	0.01417
Cooling load (MW)	3.200	3.200	4.033	3.28
Area (m^2)	887.90	953.70	956.7	1538
New area (m^2)	-	66.69	71.80	651.3

2) Heat Exchanger Network Analysis

Fig.23: HEN for the base design

The BTX process HEN is represented by grid diagrams in figure 23 The eight hot streams are represented by red lines while the seven cold streams are denoted by blue lines. As shown in figure 23, the base case HEN has fifteen (15) heat exchanger between process streams, Comprising of 7 heaters for heating and 8 coolers for cooling. The grid diagram was modified by addition of three (3) new heat exchangers represented with grey color to recover wasted heat energy. The incorporated heat exchanger recovered about 0.6 MW of heat energy saving about 9.6% of the total power required to run the process. The modified HEN for the base design showing the incorporated heat exchangers is shown in figure 24
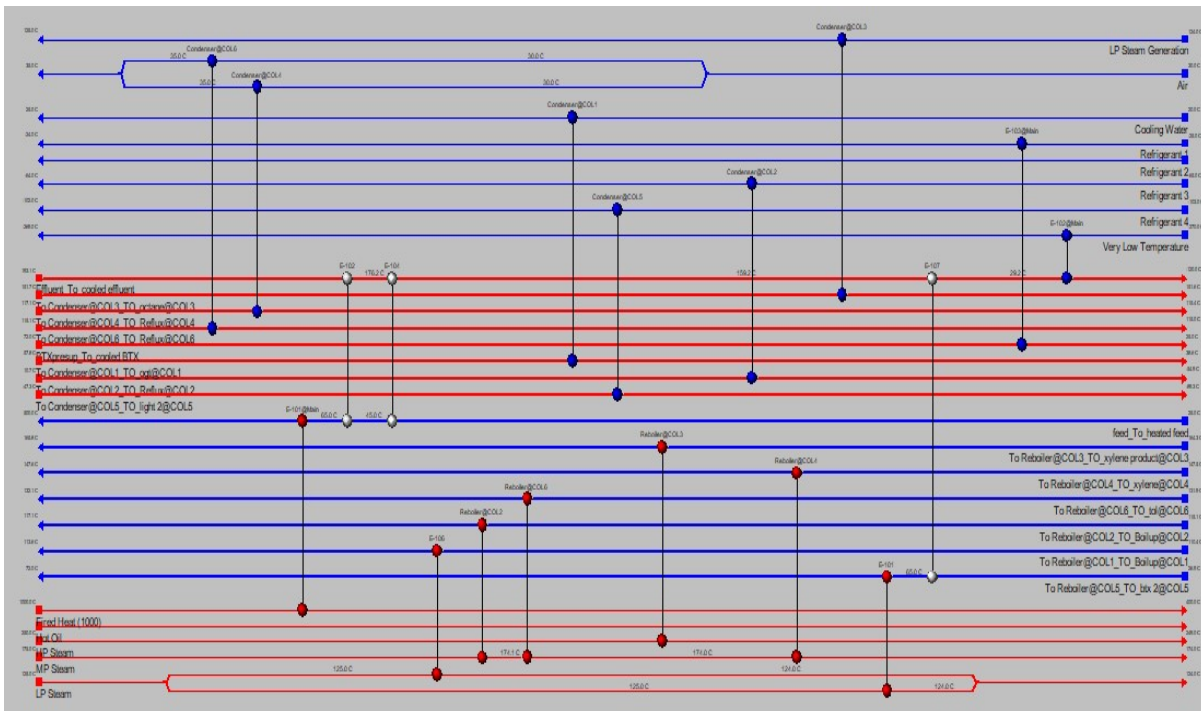
Fig.24: Modified HEN for the base design

Using Aspen Energy Analyzer, retrofit analysis was performed. No heat exchanger was added but rather modifications were made for each case scenarios in the retrofit environment, each modification is represented by a green color.
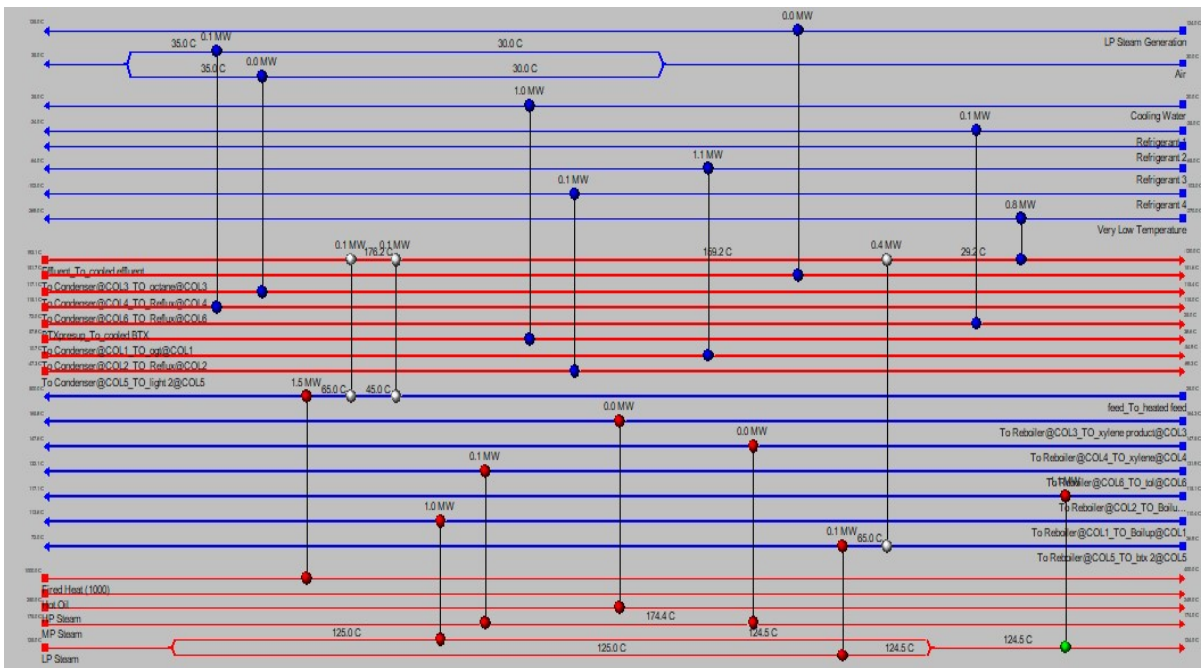

Fig. 25: Retrofit design for case scenerio 1

For the first case shown in figure 7, the LP steam was used to serve as boilup utility at column 2 reboiler in place of MP steam. This retrofit strategy increases the heat transfer area from 887.9 m^2 to 953.70 m^2 but the cost index reduces from 0.0114 to 0.0108yielding a payback period of 2.062 years. Steam consumption is increased but the cost of generating MP steam is minimized.
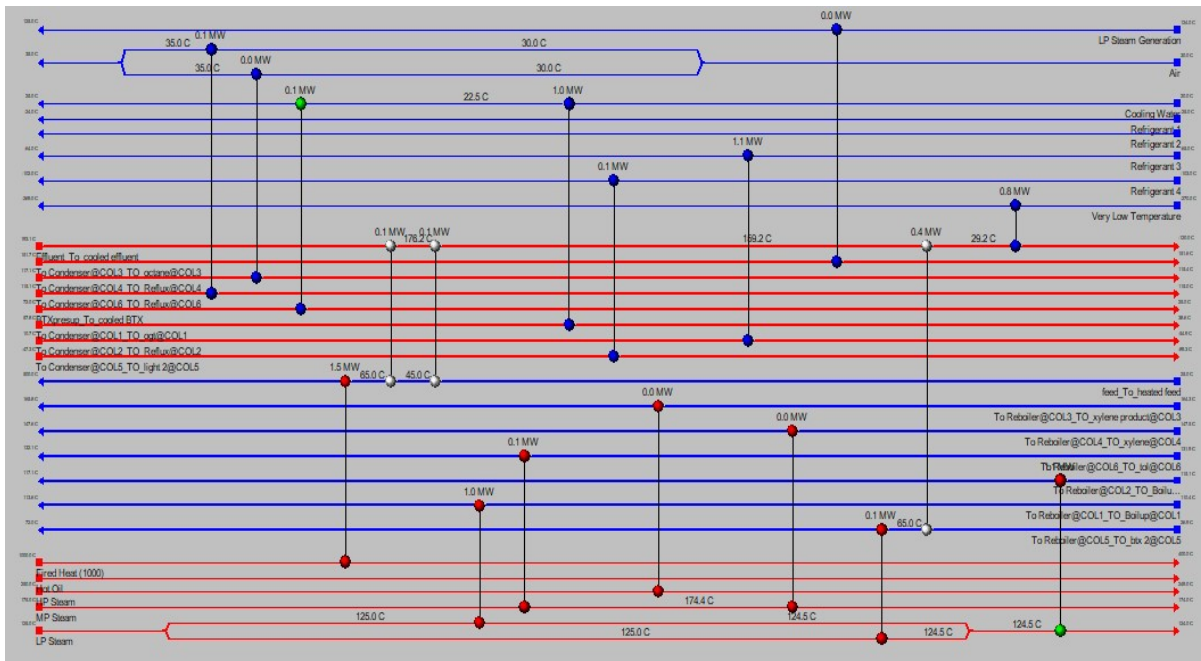
Fig.26: Retrofit design for case scenerio 2

The second case scenario shown in figure 27 in addition to scenario 1 modified the cooling exchange at Column 6 condenser by changing the cooling utility from air to cooling water utilizing a heat exchanger to replace the air cooler. This strategy increased the cooling load from 3.20 MW to 4.033 MW by increasing the heat transfer area from 887.9 m^2 to 956.7 m^2 but the cost index drops from 0.01465 to 0.01456 yielding a payback period of 1.802 years. This scenario considered maintenance in addition to the cost of the air cooler which is higher compared to cooling water using a heat exchanger. This case re-adjusted the cooling utility to produce a better payback period than the first case.
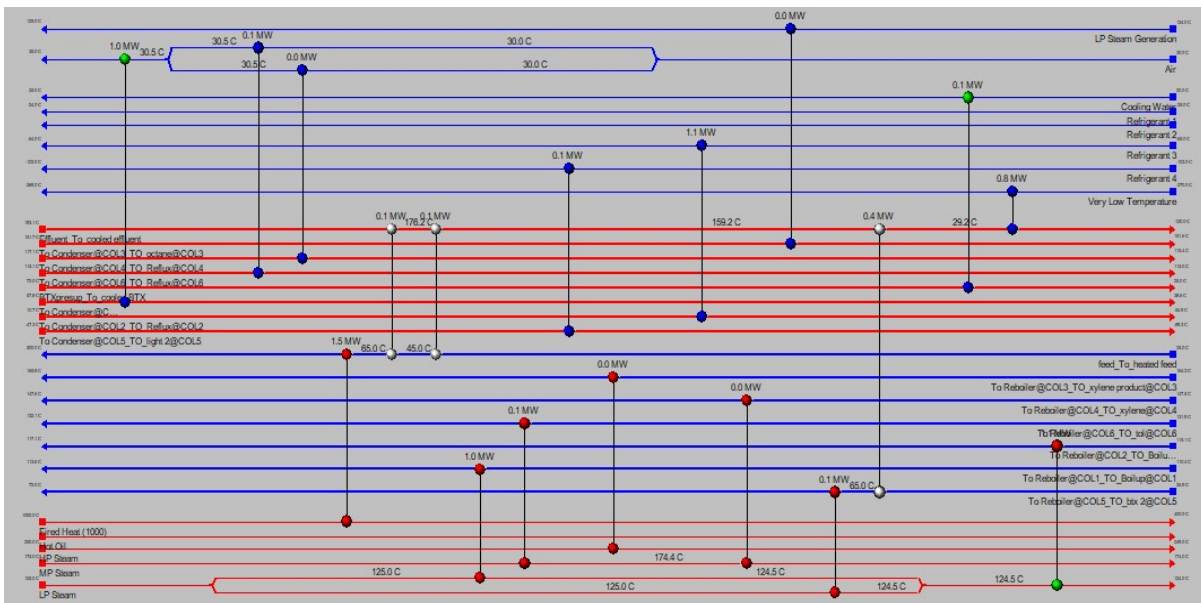

Fig. 27: Retrofit design for case scenario 3

In addition to the second case, the third case modified the cooling exchanger at Column 1 condenser by changing the cooling utility from cooling water to air. This strategy does not affect the cooling load but rather decreases the cost index from 0.0114 to 0.0417. However, the heat exchanger witnessed an increase from 887.90 m^2 to 1538 m^2 yielding a payback period of 6.579 years. The higher payback period is

due to the extensive increase in heat transfer area, cost and maintenance of the air cooler.

IV. CONCLUSION

The process for BTX production was developed with Product purity of 89.3% benzene, 95.5% toluene and 100% xylene. with an optimum reaction temperature of 395C. Feed pressure shows no effect in the conversion of BTX in the reactor. The separation process is favoured by low pressure and higher reflux ratio.Pinch and retrofit analysis were carried out using Aspen Energy Analyzer to develop base case for Heat exchanger network (HEN) consisting of 15 heat exchangers which comprise of eight (8) coolers and seven (7) heaters. Three (3) heat exchangers were added to modify the HEN base case recovering about 0.6 MW of heat, saving about 9.6% of the required energy. The retrofit design produced three case scenarios or energy saving by modifying utility exchangers Which yielded payback period of 2.062 years for scenario 1, 1.802 for scenario2 and 6.579yearsfor scenario 3 respectively. The retrofit case suggests that scenario 2 is ost effective design due to low payback period of 1.802 years.

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REFERENCES

[1]. U. S. Ubam, U. D. Amoka, and I. Shaibu, "Simulation of a Process Plant for the Production of Benzene, Toluene and Xylene from Liquified Petroleum Gas," *Niger. J. Eng. Sci. Technol. Res.*, vol. 4, no. 1, pp. 37–45, 2018.

[2]. Bignell, "The Petrochemicals Industry: Breaking Down BTX," *Fluenta*, 2018. .

[3]. A. Von Buettner, A. Destefano, T. Gunderson, and A. Mann, "Methane to Aromatics," University of Wyoming, 2016.

[4]. A. Chen, F. Crowley, J. Lym, and P. Sanchez, "Production of BTX from Ethane," 2015.

[5]. P. Jay, "Production of benzene, toluene and xylene (BTX) from FCC naphtha," 2007.

[6]. I. L. Egun and I. J. Otaraku, "Optimizing The Production Of Benzene , Toluene And Xylene (Btx) From Nigerian Treated Heavy Naphthene Samples," pp. 36–41, 2018.

[7]. O. Levenspiel, *Chemical Reaction Engineering*, 3rd ed. John Wiley & Sons, 1999.

[8]. H. P. McCabe W.C, Smith J.C, "Unit Operations Of Chemical Engineering, 5th Ed, McCabe And Smith.pdf." McGraw-Hill International, pp. 882–923, 1999.

[9]. J. H. H. and J. F. R. J. R. Backhurst, "Particle Technology and Separation Processes J.," in *Coulson and Richardon Volume 2*, 2002.

[10]. B. Linnhoff *et al.*, *Pinch Analysis amd Process Integration*, Second. Butterworth Heinemann, 2007.

[11]. W. I. Luyben, *Process Modelling, Simulation and Control for Chemical Engineers*, Second. McGraw-Hill Companies, 1996.

[12]. K. N. Rao, "Pinch Analysis of Acrylic Acid Process Plant," vol. 9, no. 06, pp. 432–439, 2016.

[13]. K. N. Rao, G. K. Reddy, P. R. Prasad, and V. Sujatha, "Pinch Analysis of Cumene Process using Aspen Energy Analyzer," vol. 1, no. 1, pp. 21–30, 2015.

[14]. L. C. Ebrada, M. D. G. De Luna, F. G. Manegdeg, and N. Grisdanurak, "The Effect of Minimum Temperature Difference in the Design and Optimization of Heat Exchanger Networks of a Brewery based on Pinch Methodology," in *World Congress on Engineering*, 2014, vol. II, pp. 0–5.

[15]. B. K. Piagbo and K. K. Dagde, "Heat Exchanger Network Retrofit Design by Eliminating Cross Pinch Heat Exchanger," *Am. J. Eng. Res.*, vol. 2, no. 5, pp. 11–18, 2016.

[16]. I. M. Idris, H. I. Mohammed, and B. K. Highina, "Pinch Analysis of Benzene Production Process via the Hydrodealkylation of Toluene," in *Faculty of Engineering Seminar Series*, 2015, vol. 6, no. december, pp. 73–78.

[17]. P. O. Kapustenko, L. M. Ulyev, M. V Ilchenko, and O. P. Arsenyeva, "Integration Processes of Benzene-toluene-xylene Fractionation , Hydrogenation , Hydrodesulphurization and Hydrothermoprocessing on Installation of Benzene Unit," *Chem. Eng. Trans.*, vol. 45, pp. 235–240, 2015.

[18]. S. vathsava Rajoli, "Retrofitting analysis on first generation ethanol production," University of Boras, 2015.