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Comparison of the Isomerization Units from the Perspective of Process Design and Energy Studies for Producing High Quality Gasoline

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ABSTRACT

Optimizing energy consumption and gasoline quality in isomerization units has become very important, indicating the need to employ suitable process design methods and energy studies. Aspen HYSYS, Aspen Energy Analyzer, and Aspen Pinch software were utilized for simulation. Optimum heat integration opportunities were identified using composite curves and driving force plots. This study showed significant variations in the minimum approach temperature ($\Delta T_{min.}$) among the units. Unit 1 showed the lowest $\Delta T_{min.}$ at 8°C, followed by units 2 and 3 at 9 and 10°C, respectively. Consequently, unit 1, with 35.65 MW of hot and 41.24 MW of cold utilities consumption at $\Delta T_{min.}$= 8°C, demonstrated the highest potential for heat recovery when increasing $\Delta T_{min.}$ from 8 to 10°C. The first heat recovery scenario aimed to examine the Research Octane Number (RON) of the output stream from the Deisopentanizer (DIP) column, and the amount of isopentane recovery to improve the existing processes. Unit 2 has an octane number and isopentane recovery of 85 and 75.24%, respectively, which is the lowest amount among the three units. Economic analysis using Aspen Process Economic Analyzer revealed that unit 1 had the highest fixed ($7.6\times10^{+6}$ USD), operating ($7.48\times10^{+6}$ USD/year), and total costs ($13.75\times10^{+6}$ USD/year). The second scenario showed that 17% and 20% of the total required low pressure steam in Unit 1 and Unit 2 can be provided due to steam generation using process energy pockets, respectively. Results showed that the process design and operating conditions directly influence energy and RON.

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1. Introduction

The demand for high-octane gasoline has prompted refineries to implement isomerization processes to eliminate tetraethyl lead as an anti-knock compound due to environmental concerns [1]. The isomerization process converts low-octane, C_5, and C_6 paraffins into high-octane branched-chain paraffins (isomers), free of benzene and aromatics [2-3]. An innovative approach was developed to optimize heat recovery in isomerization units through strategic modifications of heat exchanger networks. In this study, the researchers reduced the number of heat exchangers from 28 to 25. This modification lowered capital and annual costs by 3.77% and 0.87%, respectively, using the Aspen HYSYS process simulator and analyzing the Heat Exchanger Network (HEN) with the Aspen Energy Analyzer [3]. Refineries have incorporated isomerization units using various processes, including Ipsorb, Penex, and isomerization with DIH (Deisohexanizer) recycle. Nowadays, the need for energy and cost reduction in industrial units has significantly emphasized the investigation of isomerization units and their energy analysis using the pinch technology developed in the 1970s [4].

Process modifications, such as heat recovery and optimization of heat exchanger networks, resulted in substantial energy savings. Specially, Aspen Energy Analyzer software using pinch technology has provided process optimization and the capability to achieve the best design in terms of heat exchanger networks [5]. CO_2 Emission Pinch Analysis (CEPA) is a technique used for planning and optimizing energy systems to minimize CO_2 emissions. The use of natural gas instead of fuel gas, and energy recovery from flue gas to preheat the incoming streams to the furnace are found to be the reasons for the efficient CO_2 emission reduction of the refinery unit [6]. Retrofitting isomerization units with distillation columns (e.g., de-iso-pentanizer and de-iso-hexanizer) could reducing of CO_2 emissions and increasing gasoline production by 16.6%. this study applied Aspen energy analyzer to optimize energy use, targeting minimum heating and cooling (37 MW, and 42 MW at $\Delta T_{min.}$=10°C) [5].

In 2024, Algahtani et al. applied the Aspen Energy Analyzer to optimize an industrial plant. They reported a reduction in heating and cooling requirements by 56.7% and 12.7%, respectively, the operating cost by 35.4%, and the capital cost increased by 9.6%. Overall, the savings in total annual costs was about 10.9% [7]. Researchers studying isomerization plant showed that the addition of fractionators (de-hexanizer and de-isopentanizer) before and after the reactor is more profitable with the highest product octane number compared with other modifications [8]. The energy efficiency of Light Naphta Isomrization process can be enhanced by applying the pinch method. The research showed that minimum hot and cold targets with $\Delta T_{min.}$=10 °C are 37 and 42 MW, respectively. An energy pocket was observed at 120-180°C, offering the potential for improving heat recovery. Generating new hot utilities in the energy pocket was infeasible due to inadequate driving forces.

The researchers proposed using the heat pump to enhance heat transfer and reduce the overall energy consumption of the Light Naphta Isomerization unit using Aspen Energy Analyzer V10 [9]. The results showed the reduction of hot and cold utilities by 0.63% and 0.33%, respectively, and an 11.3% decrease in cross-pinch heat transfer. Naqvi et al. discussed the process of Isomerization of Light Naphtha to increase the octane number [10]. In this study, different isomerization technologies are compared and evaluated by employing Aspen Pinch. They designed a heat exchanger network to minimize total annualized costs. The new HEN design with $\Delta T_{min.}$=15°C successfully reduced the annualized cost to \$579,295.66/y, demonstrating the effectiveness of pinch analysis in enhancing the energy efficiency of industrial processes [11]. In 2024, Hazazi et al. presented a systematic heat integration methodology for industrial facilities, focusing on thermal oil systems through pinch analysis. By optimizing heat exchanger networks and generating LP/MP steam from previously wasted heat sources, the retrofit achieved significant energy savings of 53.3 MW (181.8 MMBtu/h) [12].

The above-mentioned studies have generally concentrated on energy optimization or gasoline quality improvement within individual isomerization units, with limited attention to a comprehensive analysis involving multiple units simultaneously. Moreover, integrated studies focusing on the use of thermal pockets and heat exchanger network optimization across several isomerization units are scarce. This research employs a combined approach utilizing Aspen HYSYS for product quality enhancement (through the addition of DIP columns) and Aspen Energy Analyzer for comprehensive energy analysis.

In this research, three different isomerization processes were studied in both the process and energy. The base case of all three units was investigated and compared for the optimization of minimum approach temperatures, composite curves, driving force plots, and cross-pinches. In this study, DIP columns were added to units 2 and 3 and simulated using Aspen HYSYS to evaluate the increase in product quality in energy consumption. Additionally, Aspen Energy Analyzer was employed to compare and optimize three isomerization processes using thermal pockets, resulting in enhanced energy saving. This study offers novel insights by simultaneously investigating and comparing three different existing isomerization units with new scenarios, whereas most previous research has focused on a single unit or technology. The application of energy analysis using Aspen Energy Analyzer, specially utilizing thermal pockets for optimization and energy saving across these diverse units, represents a practical innovation. Furthermore, the inclusion of DIP columns in two of the units to achieve the highest gasoline quality and evaluate their impact on product quality using octane number and energy consumption adds a new dimension that barely anyone has already studied.

2. Concepts and methodology

Three types of isomerization processes were studied in the present research: Ipsorb, Penex, and isomerization with DIH. The feed composition and octane number of each type of process are shown in Table 1. All three isomerization units use a light naphtha feed containing C_5 and C_6 hydrocarbons. The flow rate, composition, and octane numbers of the feed streams are different in the three processes. Comparison of isomerization processes helps us to investigate the yield of each process since some processes are more effective in improving the octane number of gasoline; therefore, they have a better yield.

For all three isomerization processes, the isomerization reactions are carried out in two-stage, fixed-bed reactors namely, lead and lag reactors, respectively. Chloride is injected into the system to maintain the activity of the alumina catalyst and remove hydrogen chloride. The stabilization column separates hydrochloric acid (HCl) and light gases. The reactor effluent is directed into the stabilization column, and the overhead product from this column is sent to a scrubber for neutralization.

The three studied isomerization processes differ primarily in the design of the absorption sections. According to Fig. 1 the reference unit employs a molecular sieve adsorption (MSA), while the two other units utilize DIH towers for absorption. In the reference unit, unlike the other two process trains, the DIP adsorption tower is installed for purpose of isopentane separation. Unlike the other two units, in unit 3, the light naphtha first enters the hydrogenation purification section and undergoes desulfurization in the reactor. Additionally, unit 3 has a stripper tower downstream of the hydrogenation purification section to separate the purified feed from the overhead vapor stream. The purified feed is sent to dryers before being fed into the isomerization reactors.

Table 1. Data extraction

	Unit 1 (Reference)	Unit 2	Unit 3
Process type	Ipsorb	Penex	Isomerization with DIH Recycle
Licensor	Axens	UOP	Axens
Octane number (RON)	66.51	61.47	70.41
Component name	**Flowrate (kgmol/h)**		
Propane	-*	-	0.3
i-butane	0.33	0.01	6.3
n-butane$_4$	19.72	0.32	6.3
i-pentane	187.9	151.69	99.2
n-pentane	239.54	245.16	103.5
2,2-methylbutane	1.73	-	3.3
2,3-methylbutane	2.88	-	6.9
2,2-dimethylbutane	-	11.24	-
2,3-dimethylbutane	-	101.74	-
2-methylpentane	144.38	518.69	42.2
3-methylpentane	75.01	381.92	25
n-hexane	137.82	639.91	62.6
Cyclopentane	2.43	31.9	5.7
Methylcyclopentane	76.52	342.58	17.6
Benzene	30.37	70.21	13.9
Cyclohexane	22.04	241.93	12.6
2-methylhexane	-	15.74	0.4
3-methylhexane	-	-	0.1
2,2-dimethylpentane	-	-	1.1
2,3-dimethylpentane	-	-	0.1
2,4-dimethylpentane	-	15.4	1.6
3,3-dimethylpentane			1.9
1,1-dimethylcyclopentane	-	-	1.9
1trans 3MCP	-	-	0.1
2,2,3-methylbutane	-	-	0.3
Methylcyclohexane	-	23.8	-
Dimethylcyclohexane	-	3.42	-
n-heptane	1.14	-	-
Toluene	4.41	-	-
C_7^+	48.94	-	-
Total flowrate (kgmol/h)	995.15	2792.85	412.8

* “-“ refers to absent compounds.

This research is based on the following assumptions: a) both Start of Run (SOR) and End of Run (EOR) conditions are considered in the Process Flow Diagrams (PDFs) of the three selected units; however, this study focuses entirely on the EOR condition, b) the high and low pressures of the DIP columns in units 2 and 3 are set equal to the pressure of the DIP column in unit 1 to ensure comparability of the design results for the DIP columns in the three selected units, c) the cooling air temperatures for the air coolers in the three selected units are based on the average temperatures of summer air over three months. For units 1 and 2, the inlet and outlet air temperatures are 28 and 38°C, respectively. For unit 3, the inlet and outlet air temperatures are considered 48 and 58°C, respectively.

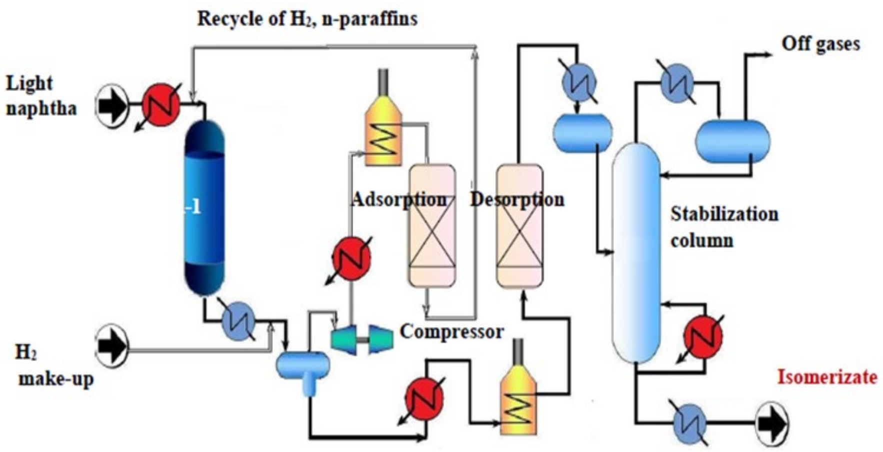


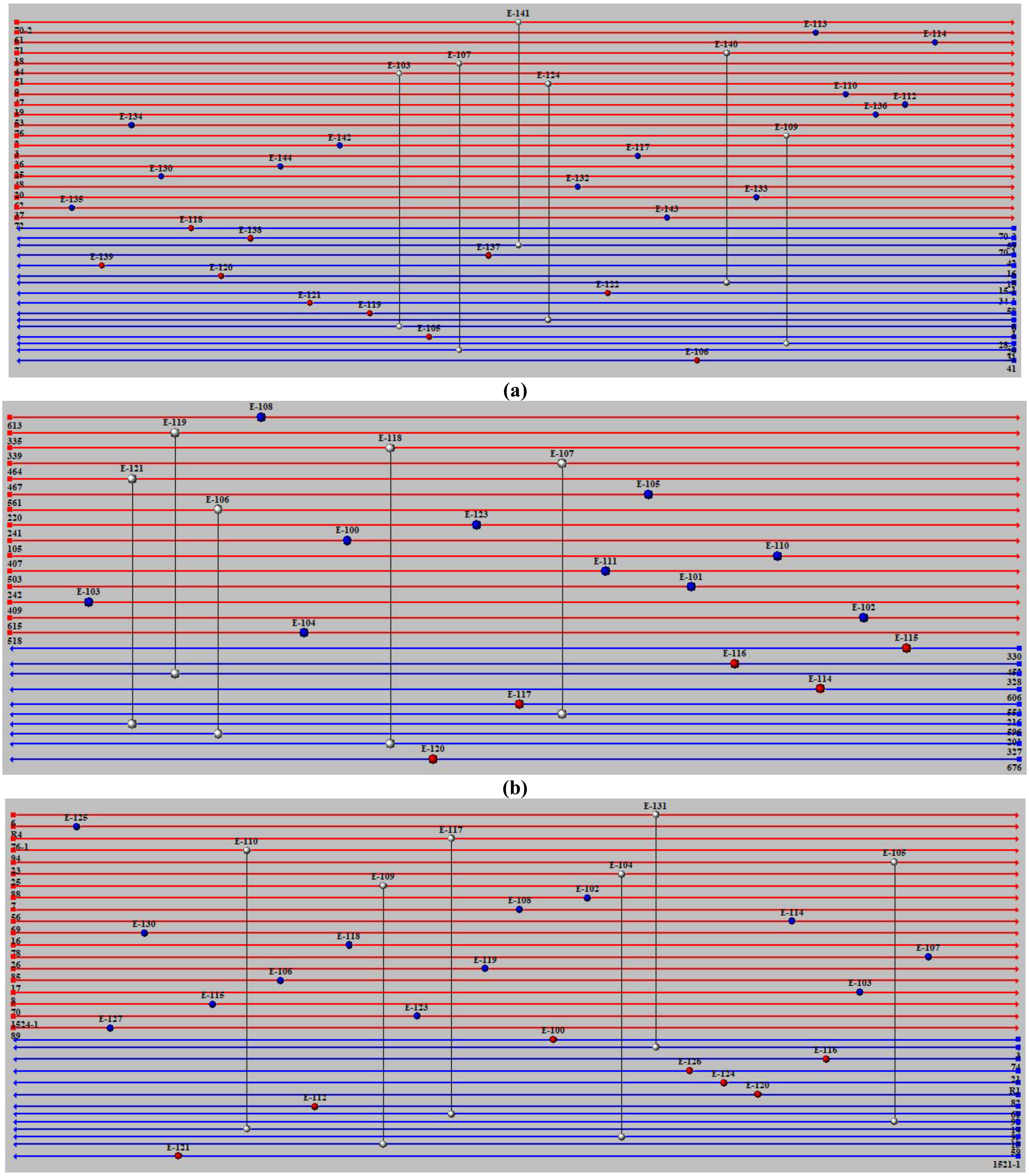
Fig. 1. Schematic diagram of Ipsorb process [11]

2.1. Problem statement

The enhancement of product octane number is required in isomerization units while minimizing energy consumption and operational costs through comparative analysis, preliminary design, and redesign of the DIP column. Producing gasoline with a high octane number that complies with environmental standards has always been a significant challenge for the refining industry. Various factors such as complex refining processes, feedstock quality, environmental issues, and economic costs make this challenge more difficult. Three distinct gasoline production units were investigated to employ varying processes to identify the relationship between these processes and energy consumption, and to assess their effects on gasoline production. A comparison of different processes will lead to suggesting suitable retrofit designs. The DIP column is a crucial component in the refining process, and its efficiency directly affects the quality and cost of the final product. The main challenges of retrofit designs are to improve isopentane recovery, reduce steam consumption, and compare the production rate of LP steam using heat pockets in all three units. This study focuses on three isomerization units with distinct configurations to optimize gasoline production with varying octane numbers.

3. Results and discussion

Three separate isomerization processes were selected for process evaluation and energy analysis. In comparison with previous studies, such as those by Ghazizahedi et al. [9] and Naqvi et al. [10], which focused on single-unit optimization or individual retrofit strategies, this study presents a comprehensive evaluation across multiple units. The HEN of the isomerization units is shown in Fig. 2. The reference unit (unit 1) contains 30 heat exchangers, while units 2 and 3 have 20 and 26 heat exchangers, respectively. In the reference unit, the number of process exchangers, heaters, air coolers, and water-cooled heat exchangers are 6, 10, 7, and 7, respectively. Four of the heat exchangers are furnaces. Unit 2 has 5 process exchangers, 5 heaters, 4 air coolers, and 6 water-cooled heat exchangers, but it does not have a furnace. Unit 3 has 3 process exchangers, 7 heaters, 5 air coolers, and 8 water-cooled heat exchangers, and 1 furnace. The reference unit has 18 cold and 23 hot streams. Unit 2 has 12 cold and 17 hot streams and unit 3 has 15 cold and 23 hot streams. The $\Delta T_{min.}$ in units 1-3 are 8, 9, and 10°C, respectively. Aspen HYSYS V10, Aspen Energy Analyzer V10, and Aspen Pinch were selected for the process simulation, analysis, and optimization of the heat exchanger network of the isomerization units.



(a)

(b)

(c)

Fig. 2. Base case HEN, (a) unit 1, (b) unit 2, (c) unit 3

The selection of an appropriate $\Delta T_{min.}$ is essential for defining objectives and network design. Therefore, the $\Delta T_{min.}$ for all three units was obtained based on the total cost vs. delta T plot. The Composite Curve (CC), and the Driving Force Plot (DFP) for each of the three units were analyzed for the base case. Additionally, a design scenario was developed to save energy. A utility was produced using thermal pockets in the scenario. The temperature range of the thermal pockets and the steam temperature in each unit were considered when finding the new utility.

The base case design aims to evaluate energy consumption, capital, operating, and total costs for all three units to identify the most energy-efficient and cost-effective processes. The positions of hot and cold pinch temperatures with hot and cold utility consumptions are shown in Fig. 3 and the values are reported in Table 2.

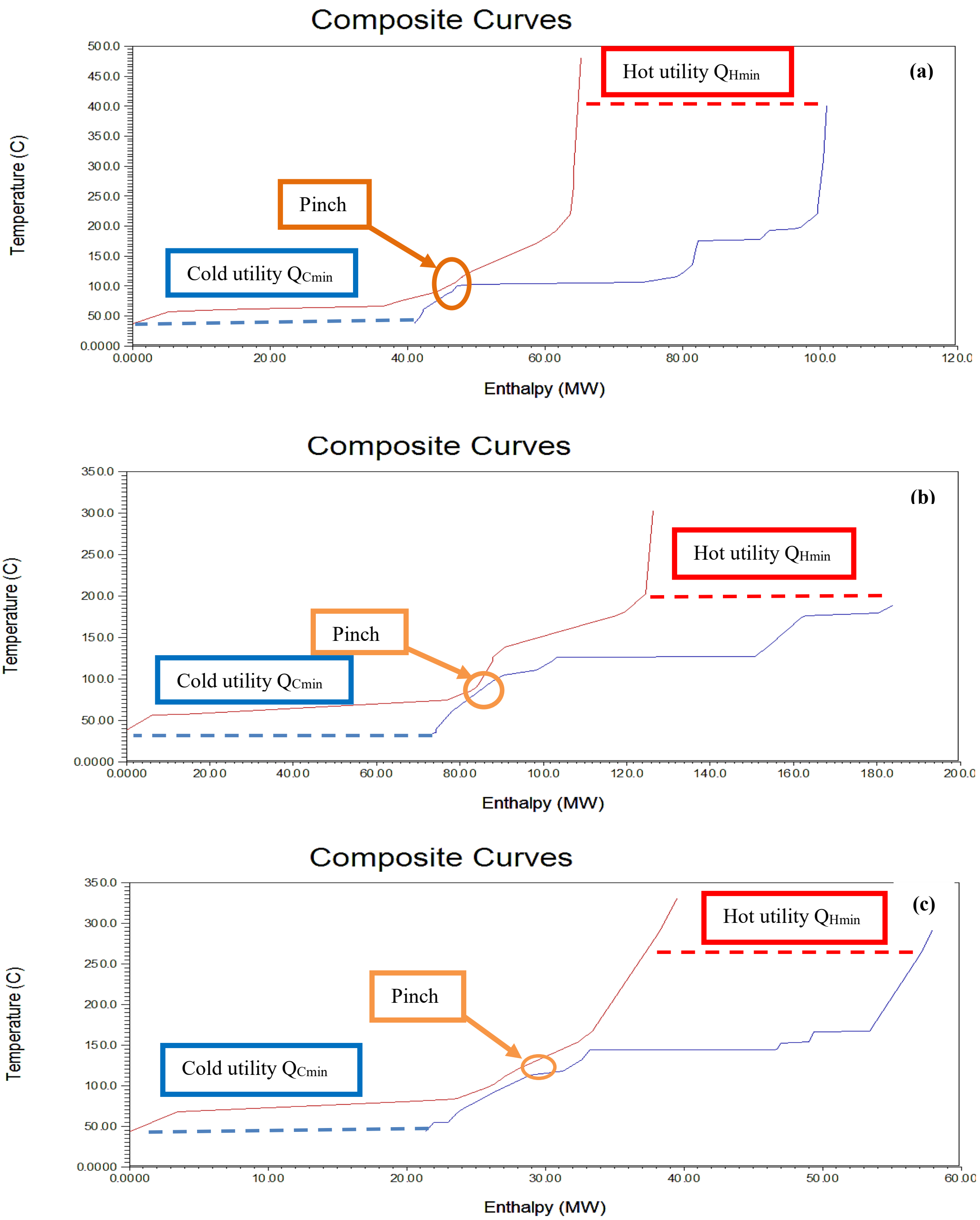


Fig. 3. Base case composite curve, (a) unit 1, (b) unit 2, (c) unit 3

Figs. 3(a),(b),(c) show the optimum $\Delta T_{min.}$ values are 8, 9, and 10°C for the units 1-3, respectively. The comparison of the amounts of hot and cold utilities shows that these amounts will increase from $\Delta T_{min.}$= 8°C to $\Delta T_{min.}$= 10°C for the base case design of unit 1 [5]. The results of the base case design indicate that unit 2 exhibits a higher consumption of hot and utilities compared to the reference unit and unit 3.

Table 2. Hot and cold pinch temperatures with hot and cold utility consumptions values

	Hot pinch temperature (°C)	Cold pinch temperature (°C)	Hot utility (MW)	Cold utility (MW)
Unit 1	108	100	35.65	41.24
Unit 2	90	81	57.18	73.64
Unit 3	98	88	19.19	21.4

Fig. 4 shows the DFP plot indicating the red lines of the driving force and black lines of heat exchangers deviating from each other. The deviant heat exchangers should be modified to approach the red line [6]. Modifying heat exchangers moves driving force and heat exchanger lines closer to each other to enhance energy recovery. The DFP plots of isomerization unit has not been studied in our previous works [3, 5].

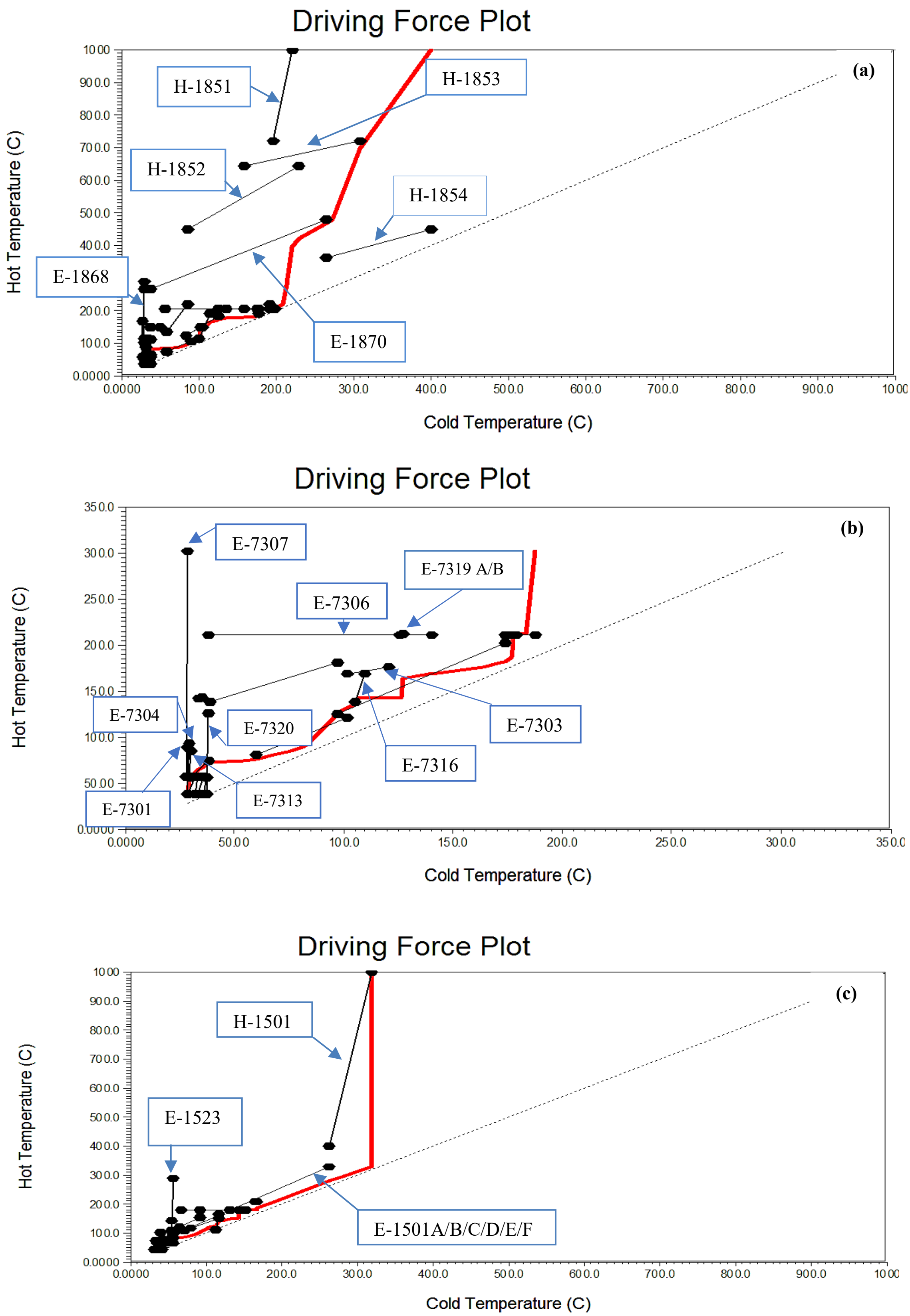


Fig. 4. Base case driving force plot, (a) unit 1, (b) unit 2, (c) unit 3

The deviant heat exchangers from the driving force line in the reference unit are four furnaces, namely, H-1851, H-1852, H-1853, and H-1854, process-to-process heat exchanger E-1870, and air cooler E-1868. In unit 2, the deviant heat exchangers from red line of driving force are process-to-process heat exchangers E-7303, and E-7316; water-cooled heat exchangers E-7307, E-7304, E-7313; heaters E-7306, E-7319 A/B; and air coolers E-7301, E-7320. In unit 3, the deviant heat exchangers are furnace H-1501, due to high fuel consumption, process exchanger E-1501A/B/C/D/E/F, and water-cooled heat exchanger E-1523, due to energy loss. The deviant heat exchangers should be modified to approach the red line. A detailed analysis of the force plot diagrams reveals that the deviation of the heat exchangers in unit 3 is less compared to the other two units, which could indicate a better design of the process in unit 3.

The comparison of HEN of Light Naphtha Isomerization units with Light Naphtha Hydrotreating unit [6] shows deviant heat exchangers in Isomerization units are more than that of Hydrotreating unit. Therefore, the heat loss in Isomerization units are more than Hydrotreating unit.

One of the crucial golden rules in pinch analysis is that there should be no heat transfer around the pinch point. The cross-pinch heat transfer of the reference unit is 8.805 MW and the highest amount of cross-pinch heat transfer is for air cooler E-1861. The air cooler E-1861 is designed to cool the recycle stream from the absorption section to the DIP column. The cross-pinch heat transfer in Unit 2 is 14 MW. The highest amount of cross-pinch heat transfer is 7.241 MW for E-7309 preheating the feed entering the reactor in Unit 2. In unit 3, the highest amount of cross-pinch heat transfer belongs to air cooler E-1523, located in the dryer regeneration section, which is 1.076 MW. Unit 2 has the highest amount of cross-pinch heat transfer, and unit 3 has the lowest the highest amount of cross-pinch heat transfer among the three units.

3.1. DIP column "Scenario 1"

Unit 1 includes a DIP column positioned before the reactors and the other two units have a DIH column after the reactors without a DIP column. Then, the DIP columns were added to units 2 and 3. The primary function of the DIP columns is to isolate i-C5 from other components. Simulations were conducted using Aspen HYSYS software to assess the performance and energy consumption of the modified units. In this scenario, the overhead and bottom pressures of DIP columns were considered 2.5 and 3.2 bar-g, respectively, to ensure consistent conditions for the DIP columns in all three units.

The percentage recovery of isoparaffins for all three units is presented in Table 3. Unit 1 has the highest recovery with 95.96 followed by unit 3 with 92.68%. However, unit 2 had a lower recovery of 75.24%, which can be explained by the lower concentration of isoparaffins in the feedstock to the DIP column in unit 2. The molar percentage of isoparaffins in the feedstock to unit1, 2, and 3 was 0.5996, 0.1031, and 0.2489%, respectively, which includes the isoparaffins in the recycle stream.

Table 3. Comparison of simulation results with unit 1 results

	Isopentane recovery (%)	Bottom product molar flow (kgmole/h)	Overhead product molar flow (kgmole/h)	Reboiler heat duty (MW)	Condenser heat duty (MW)	Reboiler temperature (°C)	Condenser temperature (°C)
Unit 1	95.96	1114	485.25	21.51	25.28	105.9	66.4
Unit 2	75.24	1317	154.4	13.61	11.07	115.2	71.65
Unit 3	92.68	610	102.9	11.92	9.252	116.7	67.43

The octane number of the overhead streams from the DIP column is one of the main criteria for determining the quality of those streams. The comparison is made using three different processes and the overhead streams of the DIP columns in this research. The isomerate streams of the three units are stored without any further processing. Therefore, the comparison criterion of the octane number of the overhead streams of the towers is a decision-making parameter. The octane numbers of the overhead streams of units 1-3 are 91.5, 85, and 91.28, respectively.

Unit 1 and 3 have higher octane numbers than that of unit 2. The Mole fraction of isopentane (0.9268) increases the octane number in the final product of unit 3 is higher than that of unit 2 (0.7524) and unit 1 (0.8859). The mole fraction of pentane reducing the octane number in unit 2 (0. 2455) is higher than that of unit 3 (0.0246) and unit 1 (0.0295). The improved design of the DIP column in units 3 and 2 has led to an increase in isopentane and a decrease in pentane in the final product, resulting in a higher octane number and better gasoline quality.

3.2. Energy analysis

The results of energy analysis using Aspen Energy Analyzer after adding DIP column to units 2 and 3 are presented in Table 4. Adding DIP column to units 2 and 3 will add a reboiler and a condenser to the heat exchanger network of units. Therefore, the amount of utilities increased in units 2 and 3. Unit 2 exhibited the highest minimum utility consumption compared to units 1 and 3.

Table 4. Results of the first modification for all three selected units

	Q_{Cmin} (MW)	Q_{Hmin} (MW)	Q_C (MW)	Q_H (MW)
Unit 1	41.24	35.65	50.04	44.91
Unit 2	84.71	70.79	98.71	84.79
Unit 3	30.65	31.11	35.87	36.33

Although Adding a DIP column to the isomerization process will increase the utility consumption, the octane number and process efficiency will also increase. The addition of the DIP column led to the development of a new process named DIP-Penex-DIH. Generally, the UOP process produces gasoline with an octane 7 units higher than that of the conventional Penex process. The UOP process also has a higher yield than that of the Penex process [13].

3.3. Economic analysis

The Aspen Process Economic Analyzer was used to evaluate the DIP columns in three selected units, and the results are presented in Table 5. Unit 1 shows the highest Total cost among the three units due to the largest condenser and reboiler areas, resulting in the highest Fixed and Operating costs. Units 2 and 3 showed lower costs. All calculations are based on a 20-year service life and a constant 20% Annual Return on Investment (ROI). These findings highlight the crucial importance of optimizing equipment sizing and energy efficiency to minimize operating costs for industrial units.

Table 5. Costs of DIP columns for all three selected units using Aspen Process Economic Analyzer

	Fixed cost (USD)	Operating cost (USD/year)	Utility (USD/year)	Total cost (USD/year)
Unit 1	$7.6\times10^{+6}$	$7.48\times10^{+6}$	$5.89\times10^{+6}$	$13.75\times10^{+6}$
Unit 2	$5.98\times10^{+6}$	$4.94\times10^{+6}$	$3.64\times10^{+6}$	$8.88\times10^{+6}$
Unit 3	$5.97\times10^{+6}$	$4.75\times10^{+6}$	$3.47\times10^{+6}$	$8.51\times10^{+6}$

3.4. Energy pocket "Scenario 2"

Fig. 5 shows the Grand Composite Curve (GCC) of units 1 and 2. All three units have a thermal pocket above the pinch point, meaning they all have excess thermal energy. In this design, the range of enthalpy and the temperature of each thermal pocket are important for producing utility in a modified design. The thermal pockets of units 1 and 2 can be used for producing LP steam but not unit 3 since the thermal pocket of unit 3 has a low amount of thermal energy. Therefore, it is not suitable for producing LP steam. The results of steam production are presented in Table 6.

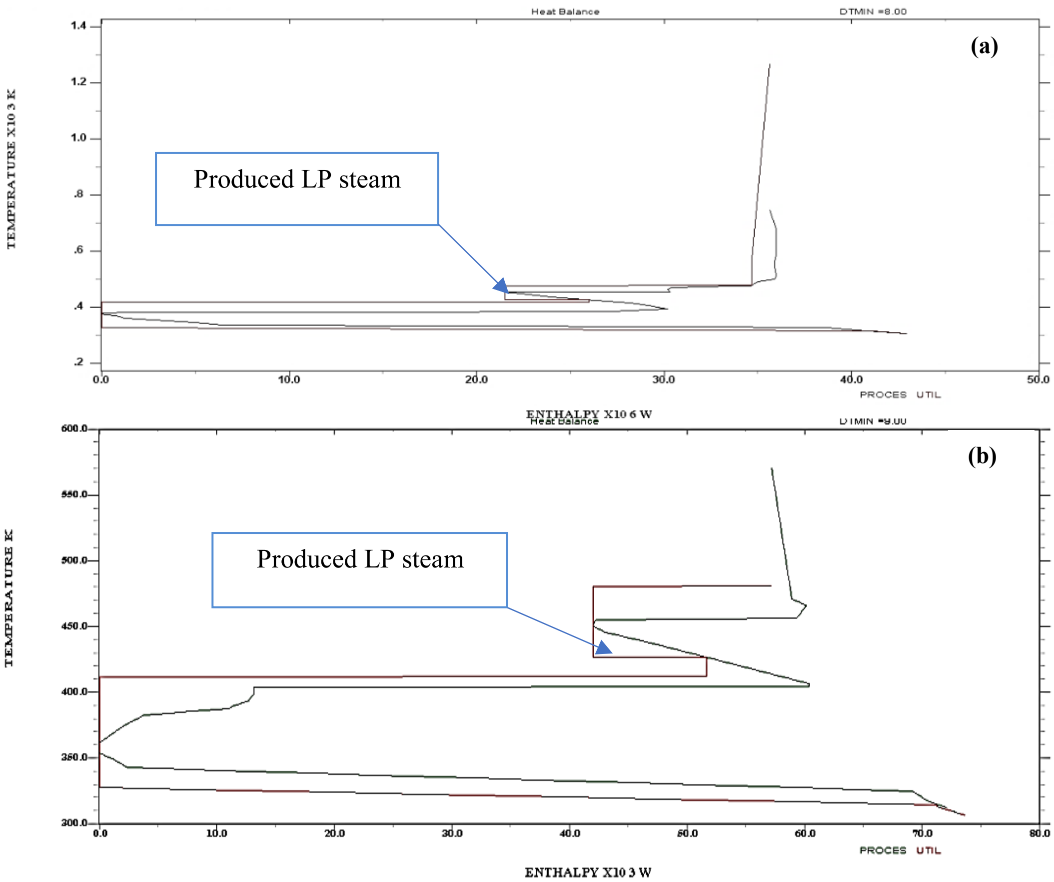


Fig. 5. Grand composite curves (GCC) (a) unit 1, (b) unit 2

Table 6. Analysis of the data collected from LP steam generation in the unit 1 and unit 2

	LP steam generation (MW)	LP steam of base-case (MW)	Cost saving (USD/year)
Unit 1	4.5	26	$35.75\times10^{+6}$
Unit 2	0.01	0.05	$77.47\times10^{+3}$

Tables 2 and 6 show that 17% (4.5/26=0.17) of the total required LP steam in Unit 1 can be provided by the energy pocket due to steam generation using process heat recovery. As a result of modifying unit 2, 20% (0.01/0.05=0.2) of the total required LP steam in unit 2 can be achieved through a significant improvement in energy efficiency by utilizing the energy pocket above the pinch point. Table 6 also shows the cost of steam production based on Aspen Pinch V.10 software, which is a useful tool for comparing the two unit costs. The highest annual economic savings is with $35.75\times10^{+6}$ USD/year for unit 1 with the highest steam production. This cost saving is with respect to the total LP steam requirement cost in unit 1.

4. Conclusion

A comprehensive energy analysis along with retrofits was conducted on three isomerization units to enhance energy efficiency after complete base-case analyses. The results revealed significant potential for heat recovery, particularly in unit 1. The highlight of this research lies in the simultaneous comparison and optimization of three distinct isomerization processes using Aspen HYSYS, Aspen Energy Analyzer, Aspen Pinch, and Aspen Process Economic Analyzer. Furthermore, the identification and utilization of energy pockets and analysis of ΔT_{min} for each unit provide novel insights into improving energy efficiency in industrial isomerization units. The first retrofit aimed at improving the octane number and isopentane recovery in three units by adding DIP columns. Adding distillation columns increased overall energy consumption. The results not only confirm the benefits of heat integration and DIP column implementation but also highlight unit-specific opportunities that have been less explored in previous research. The economic analysis of the designed DIP columns for all three units was conducted using the Aspen Process Economic Analyzer. The results showed that unit 1 had higher fixed, operating, utility, and total costs compared to the other two units. Heat exchanger network optimization in the second retrofit led to substantial energy savings with a 17% reduction in low-pressure steam consumption in unit 1. The investigation of the application of generated LP steam and designing a DIH tower for Unit 1 are very good and practical. The application of new LP steam generated should be studied in future projects. Since units 2 and 3 have DIH towers, we can design a DIH tower for unit 1 as part of the revamp, replacing its absorption section. We can evaluate the impact of this modification on paraffin thermal recovery, separation, and the passage of isoparaffins through the tower.

Nomenclature

Symbols

$\Delta T_{min.}$	Temperature (K)	GCC	Grand Composite Curve
Abbreviation		H	Heater
CC	Composite Curve	HEN	Heat Exchanger Network
CEPA	Carbon Emission Pinch Analysis	LP	Low Pressure
DFP	Driving Force Plot	MSA	Molecular Sieve Adsorption
DIH	DeIsoHexanizer	PFD	Process Flow Diagram
DIP	DeIsoHexanizer	ROI	Return on Investment
E	Exchanger	SOR	Start of Run
EOR	End of Run	USD	United States Dollar

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