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Investigation of the Opportunity of Heat Integration in a CDU in Egypt

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ABSTRACT

Energy integration is a critical solution for crude oil distillation units to reduce fuel consumption and environmental impacts. There are several proposed innovations for the networks of the heat exchanger, in addition to optimising operations in recent decades, for the purpose of enhancing the energy and economic efficiencies of the system of distillation of crude oil. However, retrofitting is often constrained by the constraints of maintenance, safety, and the topology of the process, but revamping managers prefer to exploit the existing equipment. Retrofit aims to reuse the existing equipment more efficiently to achieve various objectives, such as reducing utility consumption and minimising CO_2 emissions. This study introduces an optimal operation modification for the purpose of enhancing the energy savings of the existing unit of crude oil distillation. First, a detailed process operation is explained, simulated, and analysed with ASPEN HYSYS software, and then two modifications are proposed. The proposed modifications aim to achieve higher energy savings while retaining the economic gain than conventional approaches and without any additional investment in a retrofit in an industrial case, especially up to 28% of the reduction of energy. It is also equivalent to a 23.38% reduction of CO_2 emissions every year.

1. INTRODUCTION

Enhancing the efficiency of energy has been given great attention in the production management realm because of the significant importance of saving energy in addition to the reduced emission of greenhouse gases [1]. A crude distillation unit (CDU) that separates crude oil into various products in addition to distributing them to the units of the downstream could decrease the consumption of the energy of the plant by 24% [2]. Many techniques have been used to optimise CDUs while keeping economic performance and environmental effects into consideration [3]. They are primarily classified into two types; the first type focuses on the analysis of the integration of heat in addition to seeking to minimise the consumption of energy through the retrofitting of heat exchanger networks [4], while the other focuses on optimising process operations to maximise profit [5]. a CDU process, the feedstock of crude oil is preheated. This is achieved by heat exchange (HENs) between product streams and column pumps-around, i.e., the network of heat exchangers (HENs). Finding the most cost-effective HEN is the main purpose of the retrofit, considering any design and operational constraints. A retrofit procedure can include changing the topology of the network, adding area for heat transfer, upgrading heat exchangers, re-piping streams, and re-assigning the matches for heat recovery. For HEN retrofitting, there are some approaches such as mathematical programming, thermodynamic analysis, and heuristic integration of more than one of these approaches [6]. Heuristic knowledge and insight gained from the experiences of engineers could provide aid to finding a viable solution at an early stage for retrofitting modifications [7]. The methods of pinch analysis that involve numerical and graphical tools allow monitoring for a step-by-step user of the design of HEN to find retrofit feasible solutions in the industry [8]. Mixed Integer Nonlinear Programming (MINLP) problems are commonly used in mathematical programming methods to model HEN retrofitting. For the purpose of computational complexity of solving the problems of complex MINLP, linearization iterative efficient methods for finding the optimal solutions for HEN retrofit with the modifications of network structure have been developed [9]. In addition to the intensification of heat transfer [10] hybrid methods, which combine various methods with the individual advantages of each method, gain synergistic benefits. Pinch analysis has been stated to be a useful method for the purpose of proposing initial solutions for optimization at a large scale [11]. The HEN retrofit approaches could achieve significant energy savings with specified product numbers in CDUs, but only for configurational changes,

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unit upgrades in existing networks of heat exchangers, or configurational changes "investment is required" [12]. In recent literature, process optimum operations have also been extensively studied to increase the economic performance of CDUs. The real measurement of thermodynamic equilibrium, summation, pseudo components, light components, and heat balance are all part of the operation of modelling CDU. Reducing the CDU's organisational complexity is critical. In some commercial software, such as models that are non-linear and highly rigorous, PIMS (Aspen Tech) are transformed to simpler models with constraints that are bi-linear or linear, and are solved by using the process of sequential linear programming [13]. The data-driven models, which explain relationships between the variables of the input and the output, were introduced as an alternative to complex and deterministic models, involving simplified and robust models. As an atmospheric distillation column is represented by metamodels, a second-order polynomial regression was used [14]. For the purpose of generating samples for the operations of the CDU, the methods of Latin hypercube sampling are used, which could provide an accurate statistical estimation of the distribution of the probability [15]. In addition, regression of support vector was used in order to integrate the models of CDU with nonlinear regression into the system of the optimization [16]. Rigid and complete models are widely considered in many commercial simulators for the purpose of simulations such as UniSim, Pro-II, Plus, and Hysys. Basak *et al.* [17] developed an iterative method for on-line optimization for the device of the distillation of crude oil while taking commodity properties into account and implementing process simulators for the calculation of TBP back in addition to the tuning of stage efficiency. Ibrahim *et al.* [18] optimised a crude oil distillation system using a genetic algorithm, with the operational and structural values included in the distillation column superstructure of the crude oil determined in the setting of Aspen HYSYS. For the purpose of solving the optimising CDU operating revenue multi-objective optimization problem, AI-Mayyahi *et al.* [19] combined the composite grand curve with a genetic algorithm that is non-dominated by sorting. According to rigorous, reports, there is computational complexity in solving the nonlinear problems, leading to suboptimal solutions [20]. As previously mentioned, most current methods for the optimization of CDU either suffer from computational difficulty from obtaining optimal operations or retrofitting costs that necessitate substantial HEN. This thesis novelty is the first research that discusses the comprehensive operation of CDU with detailed unit outputs in order to increase energy savings by retrofitting of the existing heat exchanger network process main operating parameters without changes in the operating conditions and the product specifications. Two modifications for optimization are proposed to deal with challenges caused by the existing crude distillation unit process. The structure of the rest of this research is as follows: in the next section, the existing crude oil distillation description is introduced, and then there will be CDU modelling and thermodynamic analysis. Finally, the proposed modification results were compared with the base case data. The objectives of this article are: (1) Performance analysis of refinery existing plant energy. (2) Enhancing the efficiency of the refinery's existing plant energy. (3) CO_2 emissions reduction in the atmosphere Compared to the base case, the whole process has overall sustainability with the least operating costs, and thus it is applicable for implantation. Pinch technology or analysis is an excellent technique for energy savings in processes where cooling is predominant. Hohman [21], Huang and El shout [22], Linnhoff *et al.* [23], [24], and Umeda *et al.* [25] discovered and established the principle of pinch analysis in the late 1970s. Nonetheless, it is improving over time, in addition, there is an expansion in its applicability, primarily for decreasing the flows of materials [26] in addition to increasing the supplies of energy. Pinch analysis is a common and effective form of energy savings mathematical programming because the results obtained are very impressive, *i.e.*, 10–35% in the saving of energy, and that it is conceptually easy [8]. The principals of the pinch analysis are a group of rules that could be developed using some techniques of graphical representation, such as calculation-based methods such as the 'Problem Table Algorithm' [27], Grand Composite Curves, or Composite Curves. For a given operation, for constructing composite curves, Smith [28] and others [26], [23] and [25] provide a systematic procedure in addition to information about the graph’s key characteristics. By adhering to the pinch analysis principles for a given ΔT_{min}, it’s possible to achieve the maximum recovery of heat. Furthermore, it's possible to increase the quality of the energy, in a retrofit scenario, through the elimination of violations of these concepts. For the purpose of establishing the goals of energy, in addition to directing the new HENs design, composite curves are widely used. In addition, for the purpose of synthesising new HENs in different applications of energy, the method of Pinch design is used. For many chemical/refining processes, new exchanger networks are often planned. Existing HENs must be reused or upgraded. Several objectives, such as energy savings, capacity expansion, or changing feed or product requirements. It is possible to achieve these goals by adding new areas, installing new exchangers, heat transfer enhancement, and stream splitting, and thus the existing network will be changed. The effectiveness of the approaches to retrofit is heavily based on whether opportunities for change or inefficiencies in the energy in the network that was existing could or couldn't be found. This could imply:

1. Calculating the current network's relative output in relation to energy goals,
2. Quantifying the existing network's deviation from the principles pinch analysis
3. Finding the network pinch that restricts the potential recovery of energy.

2. RESEARCH METHODOLOGY

In this section, the description of a real refining plant is discussed. Two modifications are proposed to minimise the energy consumption and CO_2 emissions, considering the modifications' cost and payback.

2.1 Description of the Existing CDU

It has been considered to apply the pinch analysis on a real plant for refining the incoming oil distilling into different fractions with different ranges for boiling. The capacity of the existing plant is 1.8 M ton/year of crude oil feed. In addition, it contains tower of an atmospheric crude distillation, pre-flash tower, heat exchangers and a fired heater. In this article, CDU is fed by the Kuwaiti crude oil at 25°C in order to generate five products namely gas oil, gasoline, naphtha, kerosene, and fuel oil. The refinery existing data has been collected from the data that are available in the EGPC textbook. Figure 1 shows the configuration in addition to the equipment of the basic process of the existing plant. The data for processes of the cold streams as well as hot streams involving heat loads and supply temperature are provided in Tables 1 and 2. It has been observed that 10°C (*i.e.*, ΔT_{min}=10°C) is the minimum temperature for all exchanger units driving force. The layout of the existing network of heat exchangers focusing on the feed stream preheats train of the crude oil which is shown in Figure 1. The details of the process of exchanging heat in hot streams with a cold stream that is the main process that is the feed of the crude oil are shown in Figure 1. As presented in the figure, the feed temperature of crude oil just prior to the fired heater is 165°C. The crude oil temperature that is leaving towards the atmospheric tower from the heater is 300°C.

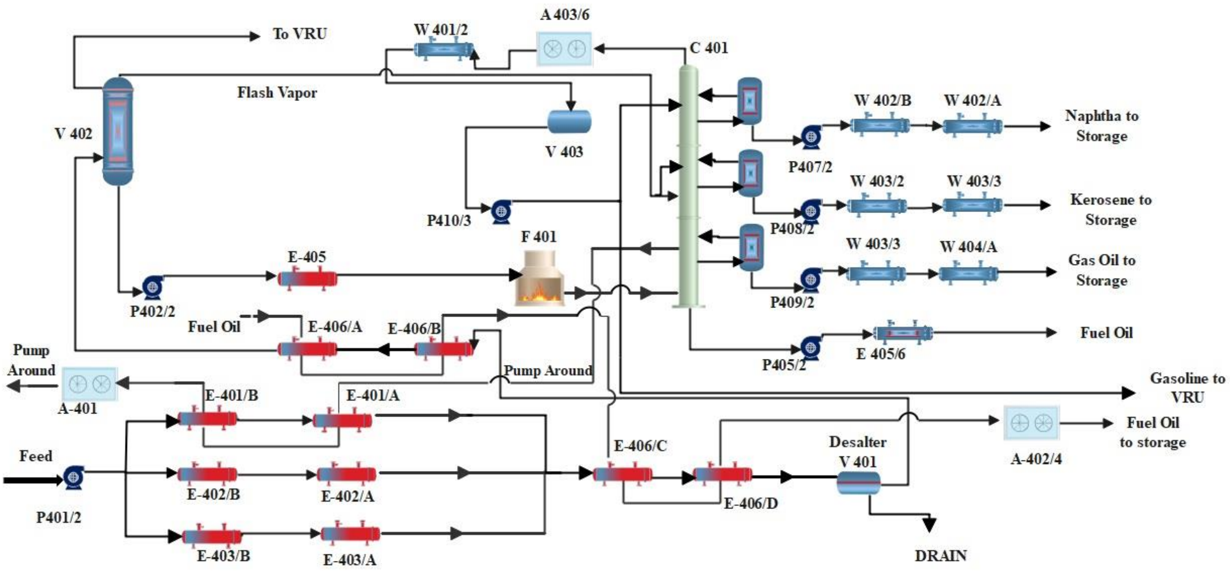

Fig. 1. Existing plant configuration.

Table 1. Heat exchangers data for the existing plant.

Heat EX.	T_c^{in}(°C)	T_c^{out}(°C)	T_h^{in}(°C)	T_h^{out}°C)	Load (kJ/hr)	Area (m^2)
E401/A	70.03	115.1	163.6	121.9	$1.084*10^7$	140.4
E401/B	25.79	70.03	121.9	81.94	$9.67*10^6$	195.9
E402/A	45.97	65.95	116.5	97.64	$1.114*10^6$	18.58
E402/B	25.79	45.97	97.64	78.72	$1.076*10^6$	24.311
E403/A	50.95	76.91	198.9	158.3	$4.913*10^6$	32.06
E403/B	25.79	50.95	158.3	118.5	$4.498*10^6$	50.50
E405	126.2	164.3	277	218	$1.973*10^7$	93.09
E406/A	124.9	135	218	201	$5.289*10^6$	35.71
E406/B	115	124.9	201	184.2	$5.196*10^6$	38.71
E406/C	105	115	184.2	167.5	$5.150*10^6$	45.94
E406/D	94.39	105	167.5	149.1	$5.438*10^6$	58.52

Table 2. The existing refinery targets of energy for ΔT_{min}=10ºC.

Hot energy (fuel) kJ/hr	66,590,000
Cold energy (water) kJ/hr	49,170,000
Hot pinch temperature °C	136.2
Cold pinch temperature °C	126.2

The actual true boiling point (TBP) data (crude assay) of the crude oil is shown in Table 3 and Figure 2; this information is based on the available data from the EGPC textbook. Pseudo-components are used for the purpose of representing the crude assay, which could be achieved by using the technique of oil characterization included in a simulation rigorous package. The thermodynamics, in addition to the physical properties of each pseudo-component (*e.g.*, critical properties, boiling temperature, molecular weight, *etc.*), are calculated using the model of Peng Robinson, and are then extracted from HYSYS. Table 4 shows the composition of the feed. The rigorous model was used for the purpose of stimulating the existing atmospheric unit. In the calculations, practical and physical equipment constraints, such as hydraulic limitations of distillation columns, allowable pressure drop, and maximum heat loads of particular exchangers, are taken into account during simulation and are kept fixed, in addition to the operating conditions, including the rates of steam flow. The reflux ratio, the temperature drops along the pump-arounds, and the pump-around duties are then specified. The model calculates the flow rate of the product in addition to its temperature, the liquid flow rate recycled through the pump-around, the cooler and the heat exchanger duties, and the flow rate for each component.

The results obtained from the simulation are summarised in Table 3 and are then used to initialise the computation of the rigorous simulation (HYSYS), for the same feed specifications, column and the operating conditions. The simulation results are then put in Table 3 and compared to the real processing data.

After comparison between the simulation and actual values of the crude oil specifications (TBP curve), it is clear from Table 3 and Figure 2 that the simulated model predicts the results with good agreement with those of the actual case. Furthermore, in Figure 2, the true boiling curves (TBP curves) of the crude oil for the rigorous simulation are shown. The true boiling curve is obtained from HYSYS by inputting the product stream data (temperature, pressure, and component mole fractions) that are predicted by the model. It's obvious that between the results, there is a good agreement. The maximum deviation of the results from the simulated models is 8%. This good agreement between the results validates the simulated model successfully. The good agreement illustrates the adequacy of the retrofit shortcut model and supports the application of the model for retrofit studies.

In the products of refinery distillation, the curves of true boiling, as well as the composition of the products, have great importance as they determine the specifications of the products and hence their qualities to meet the requirements of the market. Tables 5 to 9 and Figures 3 to 6 show the product specifications obtained from the actual data and compare them with results from rigorous simulation.

The figures are plotted as the specifications of the product in the actual process against specifications of the same product in the simulation. As shown, the rigorous simulation model predicts the product compositions in good agreement with the actual composition.

Table *3.* Comparison between assay and simulated properties of crude oil.

Property	Actual	Simulation
Specific gravity (60/60)	0.8596	0.8603
Kinematic viscosity @ 50°C	5.21	5.225
Reid vapor pressure	0.44	0.4193
Characterization factor	10.79	10.79
IBP (°C)	10	12.81
ASTM D-86 distillation 5% (°C)	59.7	109.4
ASTM D-86 distillation 10% (°C)	79.5	132.3
ASTM D-86 distillation 15% (°C)	109	164.3
ASTM D-86 distillation 20% (°C)	139.2	186
ASTM D-86 distillation 25% (°C)	169.3	211.7
ASTM D-86 distillation 30% (°C)	209	237.2
ASTM D-86 distillation 35% (°C)	239.5	260.5
ASTM D-86 distillation 40% (°C)	269.9	287.4
ASTM D-86 distillation 45% (°C)	305.4	314.8
ASTM D-86 distillation 50% (°C)	339.5	342.6
ASTM D-86 distillation 55% (°C)	370.4	371.5

A conclusion can be drawn from this example that the retrofit shortcut models developed for design of crude oil distillation columns are reliable for applications with good accuracy compared with existing data. The model predicts the product separations with comparable specifications and qualities with those obtained by actual data. The predicted results include flow stream rates, compositions, duties and true boiling curves. In addition, the models can initialize rigorous calculations.

Table 4. Comparison between actual and simulated material balance (composition of the feed).

Property	Unit	Actual	Simulation
Gases	TPH	5	5
Gasoline	TPH	13.2	11.4
Naphtha	TPH	12.11	12
Kerosene	TPH	21.99	23.99
Gas oil	TPH	40.5	46
Fuel oil	TPH	126.3	121.6
Losses	TPH	0.9	0.01
Crude oil	TPH	220	220

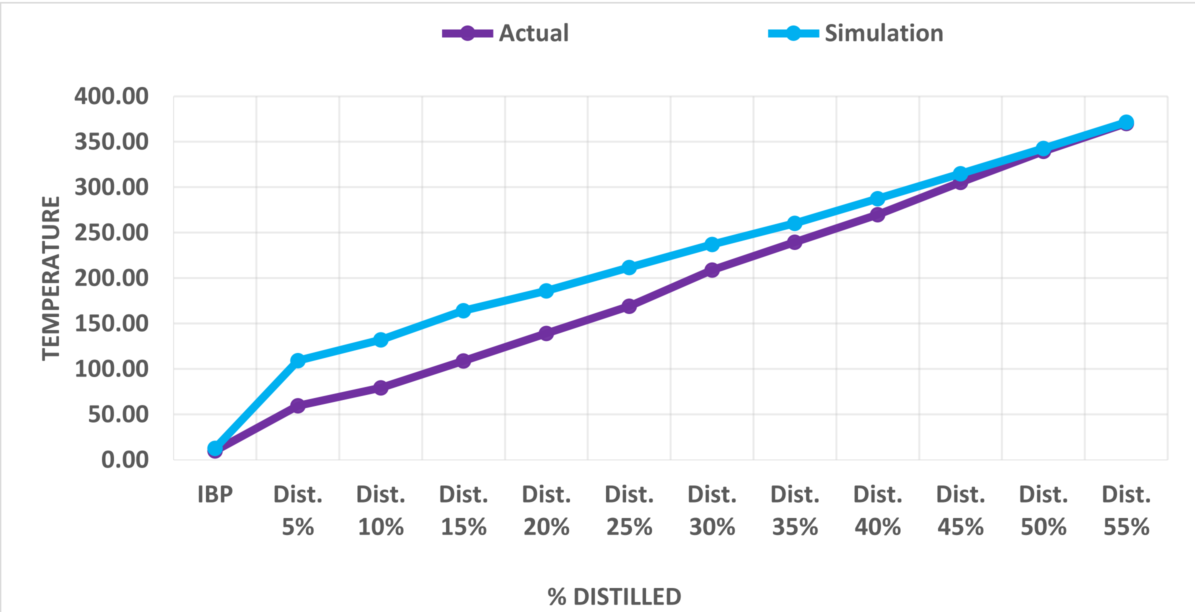

Fig. 2. Comparison between assay and simulated properties of crude oil.

Table 5. Comparison between actual and simulated properties of gasoline product.

Property	Actual	Simulation	Specifications
Specific gravity (60/60)	0.7221	0.7236	0.69-0.73
IBP	25	22	Recorded
ASTM distillation 10% (°C)	90	113.5	
ASTM distillation 30% (°C)	110	117.5	
ASTM distillation 50% (°C)	115	120.4	
ASTM distillation 70% (°C)	120	124.6	
ASTM distillation 90% (°C)	125	129.8	
FBP (°C)	140	148.2	Max. 180

Table 6. Comparison between actual and simulated properties of naphtha product.

Property	Actual	Simulation	Specifications
Specific gravity (60/60)	0.756	0.7531	0.69-0.73
IBP (°C)	70	57.3	35-65
ASTM distillation 10% (°C)	150	139.9	
ASTM distillation 30% (°C)	152	141.5	
ASTM distillation 50% (°C)	155	147.6	
ASTM distillation 70% (°C)	161	150.3	
ASTM distillation 90% (°C)	172	156.9	
FBP (°C)	185	176.9	Max. 180

Table 7. Comparison between actual and simulated properties of gas oil product.

Property	Actual	Simulation	Specifications
Specific gravity (60/60)	0.825	0.8301	0.82-0.87
IBP (°C)	176	174.6	Recorded
ASTM distillation 10% (°C)	230	250.7	
ASTM distillation 30% (°C)	250	265	
ASTM distillation 50% (°C)	270	284.1	
ASTM distillation 70% (°C)	310	306	
ASTM distillation 90% (°C)	350	342.8	
FBP (°C)	380	384.5	
Distilled @ 350˚C	90%	92%	Min. 85%
Flash point	75	100.3	Min. 60

Table 8. Comparison between actual and simulated properties of fuel oil product.

Property	Actual	Simulation	Specifications
Specific gravity (60/60)	0.9409	0.9394	Max. 0.995
Kinematic viscosity @ 50°C	345.3	321.8	Max. 388
Redwood 1 viscosity @ 100°F	1350	1200	Max. 1500
Flash point	98	152.7	Min. 65

Table 9. Comparison between actual and simulated properties of kerosene product.

Property	Actual	Simulation	Specifications
Specific gravity (60/60)	0.7891	0.7803	Max. 0.82
IBP	144	145.2	Recorded
ASTM distillation 10%	180	176.2	
ASTM distillation 30%	195	180.9	
ASTM distillation 50%	205	188.8	
ASTM distillation 70%	215	198.4	
ASTM distillation 90%	235	214.6	
FBP	272	254.2	Max. 300
Flash point		45.3	Min. 44

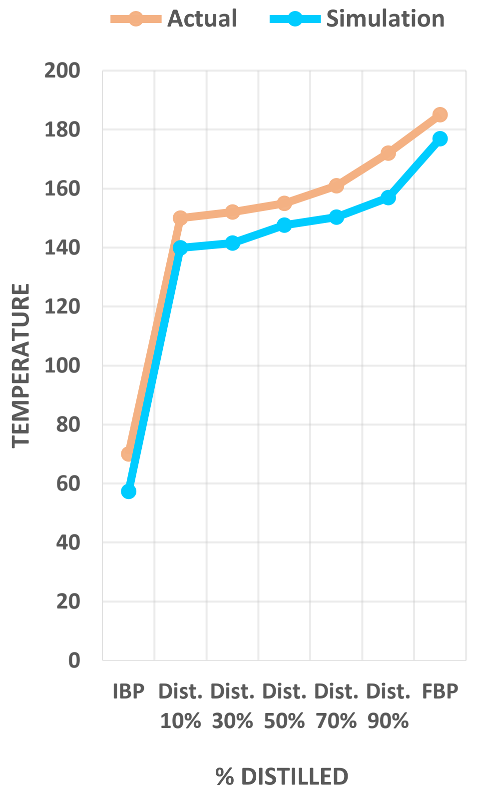

Fig. 3. Comparison between actual and simulated properties of naphtha product.

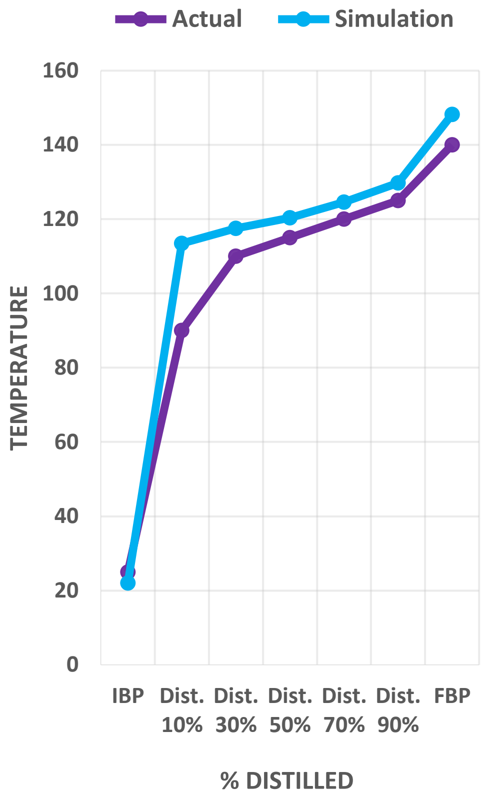

Fig. 4. Comparison between actual and simulated properties of gasoline product.

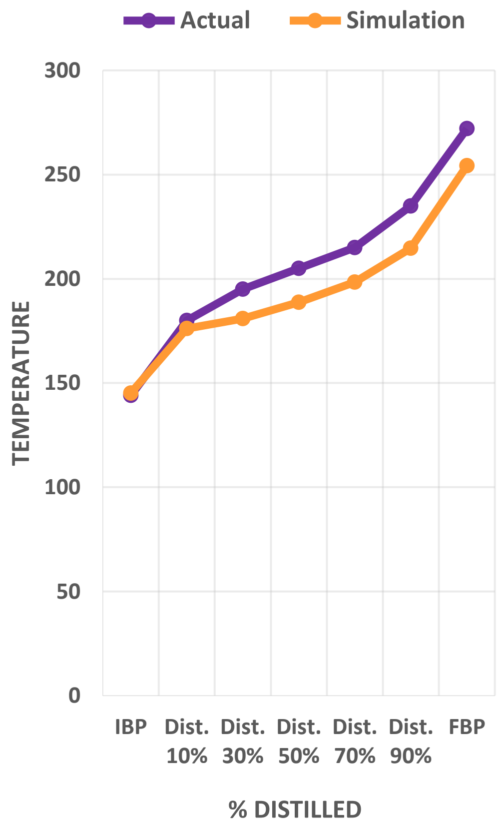

Fig. 5. Comparison between actual and simulated properties of gas oil product.

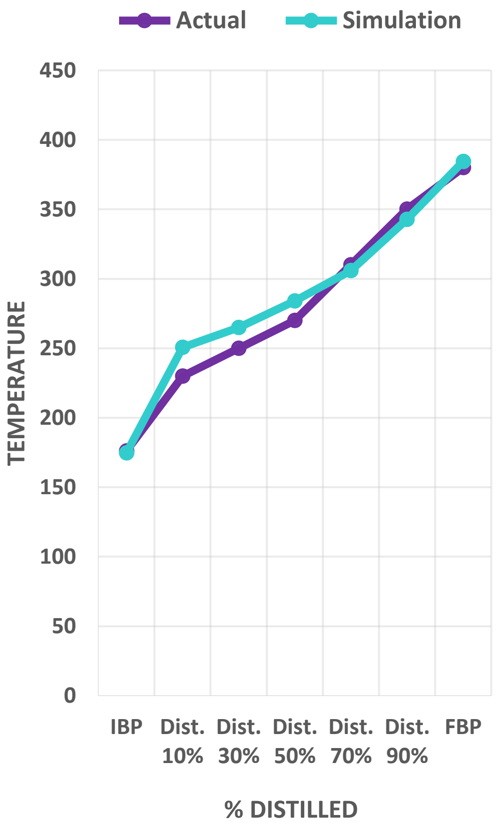

Fig. 6. Comparison between actual and simulated properties of kerosene product.

2.2 Proposed Modifications of the Existing HEN

It has been observed based on the previous discussion that the performance of the refinery plant is poor in terms of the energy performance is concerned; the actual calculation of the hot utility is $1.013*10^8$ kJ/hr compared with $6.659*10^7$ kJ/hr as a target, this led to increasing the fuel consumption by approximately 40%. This confirms that the existing train of the preheat wasn't designed in a manner which is efficient for energy. For the purpose of analysing the performance of the energy of the refinery, Aspen HYSYS was used to implement the pinch analysis. Note that the temperature of the cold pinch and the temperature of the hot pinch were obtained at this condition are 136.2℃ and 126.2℃, respectively at ΔT_{min} of 10℃ using Aspen Energy Analyzer. The poor energetic performance is caused by each of the principals violations mentioned above of pinch analysis in the existing HEN. It also shows that why the energy consumption is greater by 40% more than necessary. A range of the modifications of potential HEN could be identified from the observation reported above in order to enhance the performance of the energy of HEN in addition to reducing the consumption of fuel and cooling water. Within the existing HEN, the possible modifications that can be realized are:

Modification #1: Heat exchangers sizing modification.

Modification #2: Heat exchanger location modification.

Grid diagrams, like the one shown in Figure 1, are often used to represent existing HENs. These diagrams have the ability to identify the existing networks with the connection details. However, they couldn't quantify energy inefficiencies or qualitatively interpret the pinch analysis concept, especially for very large HENs. Furthermore, it's difficult to detect the network position and pinched match using HEN grid diagrams, especially when the network is very complex such as those used in the processes of refining. It's not possible to describe the existing exchanger using the composite curves until such time as such curves can completely characterize new HEN designs. In reality, for a complex HEN, its inconvenient and time-consuming to superimpose one exchanger match on these composite curves. If it's possible to represent the current network by stream matches details, temperatures, and duties, it's therefore advantageous. For better understanding and improved energy efficiency. It is also essential to recognize network pinch and limiting exchanger matches. After representing an existing HEN with all related information using pinch analysis simulator, the pinch analysis principles and recommendations can be understood more easily, resulting in energy savings. Such analysis will be extremely useful in conducting revamping studies. The current study aims to resolve these previous findings by creating a retrofit with new visualization capabilities that will address certain problems in existing methods.

3. APPLYING THE PROPOSED MODIFICATIONS

It's possible to apply the identified recommendations separately to the existing network. First, Modification #1 is applied to the existing HEN, to maximize the benefits from hot streams. After analyzing the whole HEN of the unit, it was found that some of the unit products temperature still relatively high after passing through the feed/product heat exchangers (E-401, E-402 and E-403) which means more relaying on the coolers (cold utilities). Pinch analysis is used to analyze how to decrease these temperatures by some design modifications. Modification #1 depends on some factors like the pressure drop and LMTD of each heat exchanger. This design modification may include adding more surface area (extra shell). It is observed that the fuel oil stream that exits from the fractionation tower and before passing through the air cooler (A-402) has high temperature (140℃) which means more relying on the air cooler. The air cooler (A-402) reduces the fuel oil temperature to 100℃ before entering the storage tanks, thus there is a chance to recover this load by adding extra area for the heat exchanger (E-405). Modification #2 is applied to the existing HEN, in order to exploit the benefits from hot stream (pump around stream). It is observed that pump around stream has high temperature before passing through air cooler (A-401). Lowering this temperature means saving more energy. Switching the hot stream of the heat exchanger (E-406/B) from fuel oil to pump around reduces the pump around temperature before passing through the air cooler (A-402) from 81℃ to 63℃. It is suggested to relocate heat exchanger (E-406/B), there is an option for relocation explained as shown in (Figure 7).

4. RESULTS AND DISCUSSION

4.1. Heat Integration

It is observed that stream data is collected and classified as hot streams, which release heat, and cold streams, which are heated up in the process. The principle of pinch analysis is to lump individual process streams together and regard them as two flows with common energy content, one composite hot stream and one composite cold stream. In Figure 8, the collected data is shown as hot and cold composite curves (CC). The vertical axis represents the temperature, while the horizontal axis represents the change in enthalpy load over the temperature intervals. The curves can be moved horizontally as the values on the x-axis show an enthalpy change, not an absolute value. The hot stream curve is usually placed with the lowest value at x=0. The cold stream curve is shifted horizontally so as to achieve maximum overlap (limited by the minimum allowable temperature difference T_{min}). T_{min} was set at 10°C in this study. A larger T_{min} would push the curves further apart, thus decreasing the overlap and causing an increasing demand for heating and cooling media ($Q_{h, min}$, $Q_{c, min}$). On the other hand, a higher driving force could facilitate an increased production rate as well as lower the cost of heat exchanger area.

In Figure 9 the curves have been converted into a grand composite curve diagram (GCC). This was made by moving the hot and cold composite curves respectively by 1/2 ΔT_{min} downwards and upwards until they come into close contact and then plotting the horizontal difference vs. temperature. The figure illustrates constant temperature utility streams. Note that when multiple utility levels are available for the hot utility, the ones with temperatures closest to the pinch point are often the cheapest one. The GCC curve can be used to determine whether there is any heat surplus (below the pinch) at useful levels, which could be used for heating. The concept of distillation targeting provides refinery modifications relevant to reduction in energy demand. The refinery targets (Heat loads and temperature) are based on the "Grand Composite Curve" (GCC), which determines the scope for appropriate modifications and sets the temperature targets for these modifications. The GCC is obtained from converged simulations. This GCC is then used to identify beneficial modifications; appropriate HEN modification for reducing the utility consumption further. Process modifications are those which result in reducing energy demand and improving the temperature-quality of heat sources. These modifications include appropriate feed location, reflux ratio modifications, feed conditioning and using side condensing or reboiling. Based on the analysis of the process flow diagram and flow data in Table 1 in the manuscript. Both hot and cold streams for the existing CDU were analyzed and ΔT_{min} was set at (10℃). Both hot and cold utilities targets were defined as ($6.659*10^7$ kj/kg and $4.917*10^{10}$ kj/kg) respectively. From Figures 9 and 10, it is observed that composite curve and grid diagram show that hot pinch point =136.2℃ and cold pinch point = 126.2℃.

Grand composite curves were performed indicating hot and cold target utilities. From grand composite curve, it is observed that shifted pinch point=131.2℃.

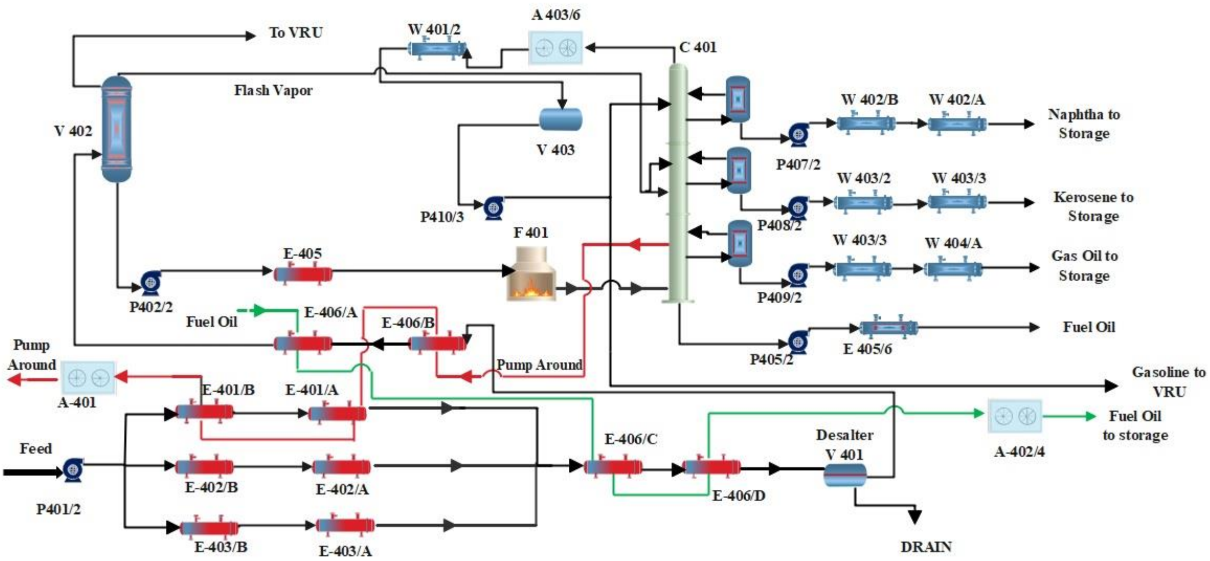

Fig. 7. Modified HEN (relocating E-406/B).

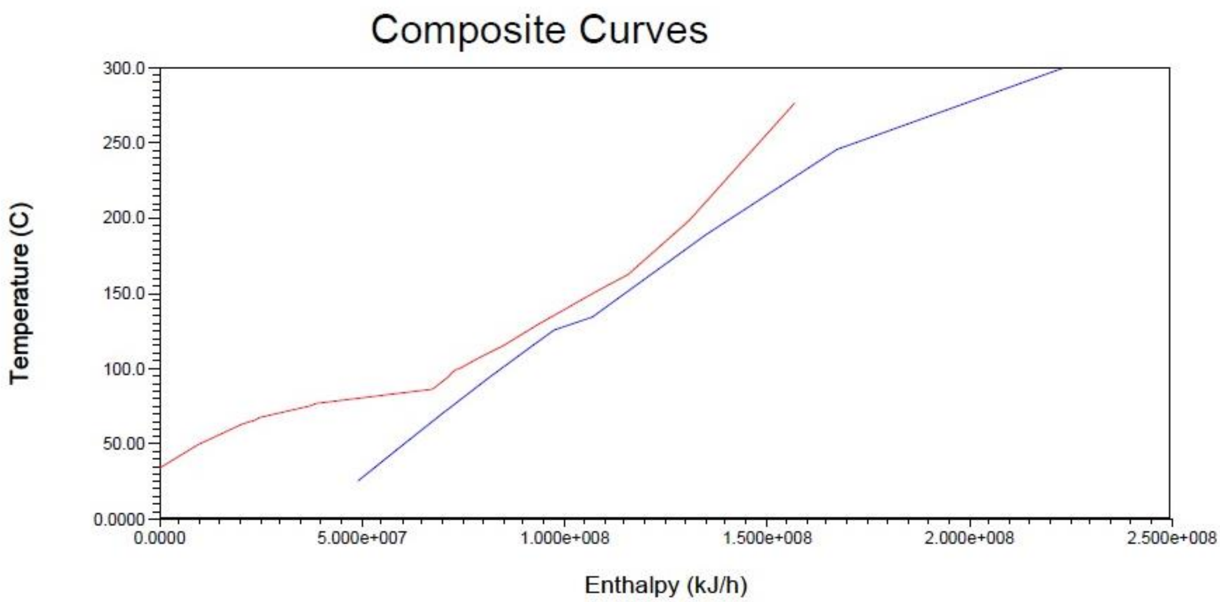

Fig. 8. Composite curve.

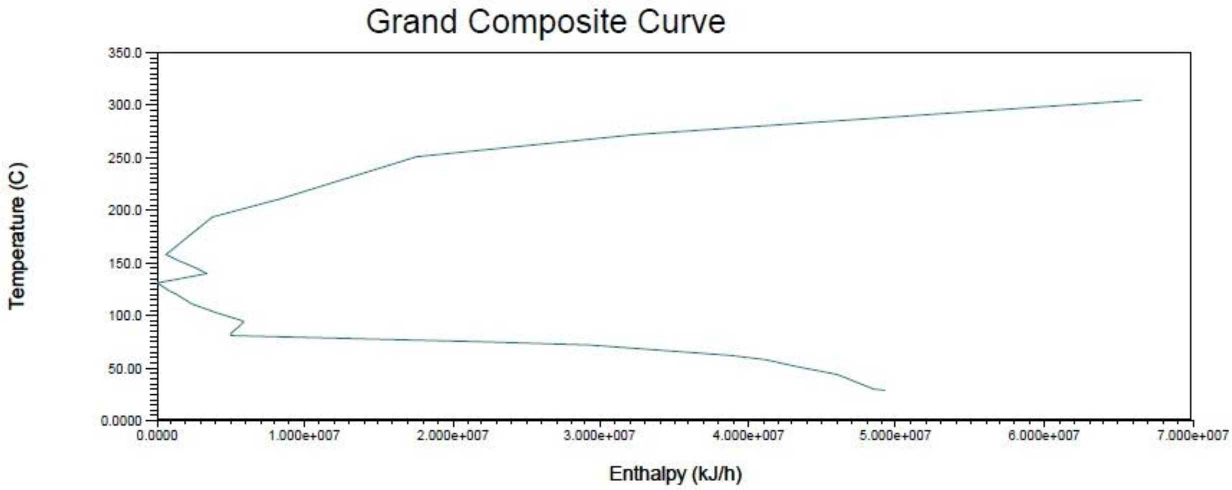

Fig. 9. Grand composite curve.

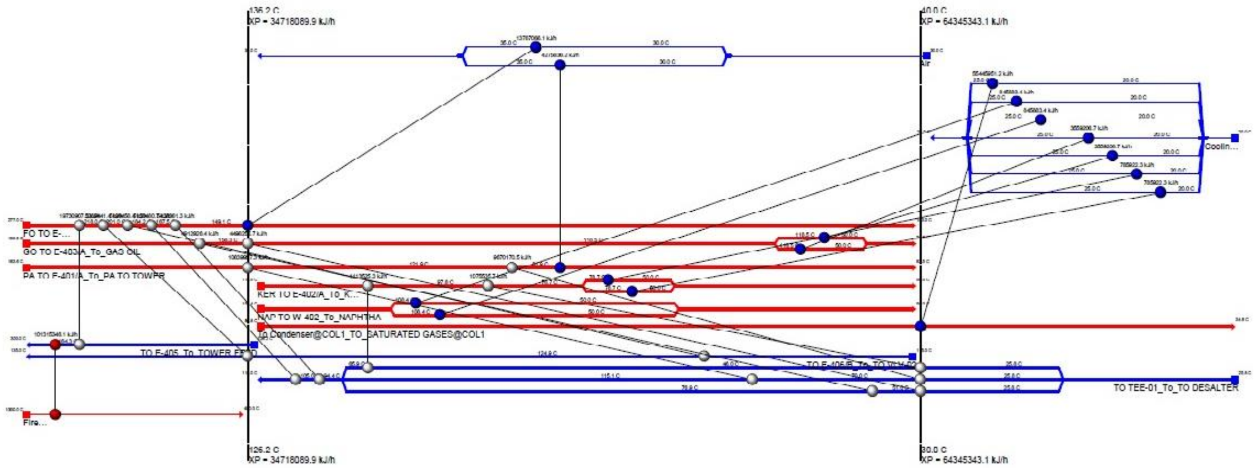

Fig. 10. Grid diagram.

For modification #1, it was discovered that increasing the area reduced the temperature of the fuel oil from 149.1°C to 124.9°C. chance to improve energy recovery. The same procedure is applied to the pump around the stream train. Before passing through the air cooler (A-401), the pump has a high temperature of (81°C). The air cooler (A-401) reduces the pump's operating temperature to 63.5°C.

Adding extra area to the exchanger (E-401/B) reduces the pump's operating temperature to 69°C, which means that the duty of the air cooler (A-401) is reduced. In addition, the gas oil stream train was studied, and it was noted that the gas oil product exits with a high temperature of 119°C before passing through the water cooler (W-404), which reduces the temperature of the gas oil to 50°C. Adding an extra area to the exchanger (E-403/A) reduces the gas oil temperature to 61°C, which means that the water cooler (W-404) can be eliminated. It is suggested to increase the area of some existing heat exchangers (E-405, E-403/A, E-401/B). The selection of the mentioned heat exchangers is due to the fact that they have the lowest LMTD value compared to the remaining heat exchangers. The existing fired heater of the CDU is designed to raise the crude oil temperature in the range of 110°C to 130°C, thus the optimum crude oil inlet temperature should be near 200°C as obtained from the fired heater data sheet. Reaching this temperature means that optimum heat transfer between the refined products and crude oil has been achieved. Practical and physical equipment constraints, such as hydraulic limitations of distillation columns, allowable pressure drop, and maximum heat loads of particular exchangers, are taken into account during optimization. Several objectives have been considered for retrofit applications, including energy reduction, modification cost, payback time, fixing product specifications, and emissions reduction. Structural modifications to the distillation process and the exchanger network have also been considered for retrofit. After applying the modifications, it has been found that, within the process, more heat is integrated and thus the load of the heat on the heaters decreases to $8.398*10^7$ kJ/hr compared to $1.013*10^8$ kJ/hr for the existing network (base case). There is also a 31.6°C increase in crude oil temperature (from 165°C to 196.6°C) over the base of HEN. At the same time, there is a decrease in the cold utility consumption to $6.656*10^7$ kJ/hr equivalence, and thus there will be a saving of 17.1% in fuel consumption. With the existing performance, 21.7% of the cooling water will be saved. The overall enhancement and savings for the HEN, which is modified, are summarised in Table 10.

It is shown in the table the significant savings in the demand for energy, consumption of oil fuel and the emissions of cooling water flows and atmospheric cooling water flows. It's clear that a great enhancement has been achieved by the modifications by the pinch analysis. The effect of adding areas is indicated by specified heat exchangers, and the effect of the heat is shown in Table 13.

For modification #2, the crude oil temperature after implementing the modification to HEN becomes 208°C with an increase in the value of the temperature of 43°C. The corresponding load of the heat on the furnace has been decreased after implementing the modification from $1.013*10^8$ kJ/hr to $7.761*10^7$ kJ/hr. Saving energy was achieved by approximately 28%. This shows that there will be a saving in the fuel flow rate in addition to the operating cost which reduced the annual cost from $3,896,519 to $2,997,322 for a savings of 23.1%. The consumption of the cooling water consumption is $8.389*10^7$ kJ/hr compared to $6.018*10^7$ kJ/hr, a saving of 13.1%. Data extracted for the modified HEN is summarised in Table 13, involving the duties for each unit of the exchanger and terminal temperatures. As a result of energy-saving, there is a decrease in the consumption of the fuel for the modified refinery. The overall enhancement and savings of the new modified HEN are shown in Table 11. The table shows the significant savings in the demands for energy, consumption of the fuel, atmospheric emissions, and cooling water flows. It's obvious that the modification proposed in this study by the pinch analysis enhances the integration of the existing refinery. The relocation modification is combined with the modified piping for the purpose of changing the process configuration. The cost of the piping work modification is neglected. Besides being a cheap modification, the advantage of this modification is that the network pinch approach does not consider changes to the distillation process operating conditions; it only considers modifying the existing exchanger network to reduce energy consumption. The overall enhancement and savings for the HEN, which is modified, are summarised in Table 12.

Table 10. Savings for modified HEN.

Parameter	Existing unit	Modified unit	% Reduction
Hot utility demand (kJ/hr)	101,300,000	83,980,000	17.1
Cold utility demand (kJ/hr)	83,890,000	66,560,000	21.7
Crude oil temperature before entering furnace ℃	165	196.6	-
Pay back (year)	-	0.52	-
Green house reduction	-	-	17.18
Utility operating cost $/year	3,896,519	3,028,953	22.3

Table 11. The data of heat exchangers for HEN-based refinery plant which is modified.

Heat Exchanger	T_c^{in}(°C)	T_c^{out}(°C)	T_h^{in}(°C)	T_h^{out}(°C)	Load (kJ/hr)
E401/A	53	92.8	126.9	88.9	9,178,000
E401/B	25	53	88.9	63.6	5,864,000
E402/A	60	92.2	116.5	84.4	1,869,000
E402/B	25.8	60	84.4	50.7	1,855,000
E403/A	31.8	112.1	198.9	60.9	15,420,000
E403/B	25.8	31.8	60.9	50.4	1,073,000
E406/A	115	116.4	139.7	137	43,440,000
E406/B	116.4	135	136.6	126.4	7,602,000
E406/C	106	115	137	120.9	9,726,000
E406/D	100.5	106.2	120.9	110.5	4,533,000
E405	126.2	208.1	277	139.7	2,906,000

Table 1. Savings for modified HEN.

Parameter	Existing unit	Modified unit	% Reduction
Hot utility demand (kJ/hr)	101,300,000	77,610,000	28.2
Cold utility demand (kJ/hr)	83,890,000	49,170,000	13.1
Crude oil temperature before entering furnace ℃	165	208	-
Pay back (year)	-	0	-
Green house reduction	-	-	23.38
Utility operating cost ($/year)	3,896,519	2,997,322	23.1

Table 13. heat exchangers data for modified HEN.

H. E	T_c^{in}(°C)	T_c^{out}(°C)	T_h^{in}(°C)	T_h^{out}°C)	Load (kJ/hr)	Existing Area (m^2)	Added area (m^2)
E401/A	90.7	126.4	163.6	130	8,872,000	140.43	--
E401/B	25.8	90.7	130	69.9	14,540,000	195.96	180.6
E402/A	49.5	77.7	116.5	89.1	1,599,000	18.586	--
E402/B	25.8	49.5	89.1	66.4	1,269,000	24.311	--
E403/A	32	112.1	198.9	61.4	15,380,000	32.062	258.8
E403/B	25.8	32	61.4	50.5	1,102,000	50.505	--
E405	126	196.9	277	162	37,060,000	93.095	296.8
E406/A	124.9	135	162	143.5	5,289,000	35.711	--
E406/B	115	124.9	143.5	125.1	5,196,000	38.715	--
E406/C	115	115	125.1	125	18030	45.943	--
E406/D	114.9	115	125	124.9	18030	58.525	--

4.2 CO_2 Reduction Emissions

Carbon dioxide is generated in CDU distillation units mainly from fired heaters. The utility burns fuel to produce heat, steam and power. The fuel is combusted with air producing CO_2. In fired heater (furnace), the amount of fuel burnt can be related to the heat duty required by the process and efficiency of the fired heater as follow:

$Q_{Fuel}=Q_{process}/\eta_{furnace}$. If the value of the thermal efficiency of the existing fired heater $\eta_{furnace}$ =60%. If the used fuel has net calorific value of 40400 kj/kg, then the necessary rate of the flow of fuel for the healing process is ($1.013*10^8$ kJ/hr)/ (0.6 × 40400 kJ/kg) which is equivalent to 4179.2 kg/hr. The fuel oil has emission factor of 3.127 kg of CO_2 for every kg of fuel. So, the atmospheric emissions of CO_2 of the existing refinery plant because of fuel combustion is (13,067) kg/hr of CO_2 per each 4,179.2 kg/hr of fuel oil.

For modification #1 the consumption of oil fuel for the refinery which is modified is reduced because of saving energy. It has been computed as previously shown as (83,980,000 kJ/hr)/ (0.6 × 40,400 kJ/kg) or 3460.8 kg/hr. On the other hand, there is a reduction in the value of the emission of the CO_2 as it reduced to 10821.9 kg/hr by savings of (17.18%) with respect to the base case which was (13067) kg/hr. For modification #2 the same procedure is applied, then it was found that CO_2 emissions reduced to 10011 kg/hr which represents 23.38% with respect to the base case.

5. CONCLUSION

The Aspen HYSYS simulator is used in this work to perform a pinch analysis on an existing network of heat exchangers, involving information on the matches of exchangers, coolers, and heaters. In terms of pinch analysis principles, the simulation will analyse and evaluate the output of existing HENs or preheat trains. Current HEN energy inefficiencies are reported because of the performance review. Therefore, by quantitatively analysing these inefficiencies, the future energy recovery is calculated. The Aspen HYSYS analyzer can recommend numerous possible changes for improved energy integration that can be implemented. The existence of the diagonal is critical in visualising the existing exchanger's temperature driving force in the energy analysis and modification implementation. A case analysis of current CDUs, including atmospheric units, revealed substantial energy savings and emissions reductions. HEN modifications resulted in 28.2 percent energy savings and a 13 percent reduction in cooling water demands, according to typical case study findings. The study advantages of using the Aspen HYSYS analyzer simulator, which offers an interactive simulation of the analysis of HEN and can be used instead of commercial simulators for the analysis of energy (*e.g.*, Aspen Energy analyzer). Furthermore, the pinch analysis approach aids in the understanding of the issue of the integration of energy and its analysis activities. It also allows for the graphic transfer of heat between loops of an exchanger and the reduction of utilities' heat using utility-paths.

ABBREVIATIONS

CDU: crude distillation unit
HEN(s): heat exchanger network(s)
LMTD: logarithmic mean temperature difference
TBP: true boiling point
EGPC: Egyptian General Petroleum Corporation

NOMENCLATURE

T_c: temperature of process cold stream (°C)
T_h: temperature of process hot stream (°C)
T_{inh}: temperature of inlet hot stream to exchangers (°C)
T_{inc}: temperature of inlet cold stream to exchangers (°C)
$T_{out\ c}$: temperature of outlet cold stream from exchangers (°C)
T_{inh}: temperature of inlet hot stream to exchangers (°C)
ΔT: temperature driving force of exchanger (°C)
ΔT_{min}: minimum temperature approach difference (°C)

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