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Exceeding Pinch limits by process configuration of an existing modern crude oil distillation unit – A case study from refining industry

Omar S. Bayomie^{a,b,*}, Omar Y. Abdelaziz^{c,d}, Mamdouh A. Gadalla^{e,f,*}

^a*Department of Energy and Processes, MINES ParisTech, PSL Research University, 75006 Paris, France*

^b*Department of Chemistry, The American University in Cairo, New Cairo 11835, Egypt*

^c*Department of Chemical Engineering, Cairo University, Giza 12613, Egypt*

^d*Department of Chemical Engineering, Lund University, 22100 Lund, Sweden*

^e*Department of Chemical Engineering, Port Said University, 42523 Port Said, Egypt*

^f*Department of Chemical Engineering, The British University in Egypt, Cairo 11837, Egypt*

*omar.bayomie@univ-psl.fr

*mamdouh.gadalla@bue.edu.eg

Abstract

Crude Distillation Unit (CDU) represents significant challenge for retrofitting and energy optimisation as the most energy intensive consumer in a conventional crude oil refinery. Pinch Technology and its based-methodologies are found primary keys for decades to energy savings in refining industries for a range of common economic-based and environmental objectives or applications. Typical benefits in energy savings are reported within 20 to 40% of original designs. However, such savings are limited and questioned when modern refiners are dealt with. The current paper addresses the revamping of a modern refinery exhibiting an existing high energy efficiency ($\approx 93\%$). This implies the maximum potential energy savings would only be 7% at current process conditions. The present research proposes an algorithm that tackles energy recovery of modern refiners, enabling additional savings beyond the energy targets set by the existing process. The algorithm starts by process simulation and validation against real plant data, followed by a network optimisation, e.g. stream splitting, to reach the energy targets set by Pinch Analysis. The energy targets are then moved to

another lower level by performing potential process modifications to reduce the energy consumption further. Results showed that the current modern refinery unit could reach its energy targets by stream splitting modifications with hot energy savings of 2.69 MW. Process modifications resulted in additional energy savings of 31.3% beyond the current level of the existing plant alongside less than a year of payback period for estimated capital investment. An environmental assessment is performed, and comparable reductions were obtained with respect to greenhouse gas, with reduction in CO₂ emissions by 45.1%. The proposed retrofit methodology is applicable to minimising energy consumptions of refiners including modern units to achieve energy levels beyond energy targets by new process modifications.

Keywords

Oil refinery; Process modelling; Energy savings; Emission reduction; Pinch Analysis; Retrofit

1. Introduction

The oil downstream industry is facing big challenges to reduce its CO₂ emissions while balancing its economic positive margins. Toward making renewable energies more affordable and energy efficiency techniques achieving big steps to CO₂ emissions reduction, the existing processes should reduce their energy consumptions. In reference to current refinery configurations and oil demand, petroleum refineries are already operating at their most efficient level ~ 94% (Abdul-Manan et al., 2017), making the energy saving and emission reduction activates a difficult task with such challenging profitability margins. Economic and environmental redesigns of existing crude distillation units (CDUs) appear to be of significant importance in order to achieve the

highest possible technical efficiencies simultaneously with a reasonable total investment cost (Waheed et al., 2014).

One of the main considerable technologies in the last decade is heat and process integration in the chemical and refining industries with significant advances in that domain (Ulyev et al., 2018); many of techniques have been applied to reach more efficient performance in the case of maximising energy recovery (Mehdizadeh-Fard and Pourfayaz, 2019). Pinch Analysis is an effective methodology that proved its credibility in many process industries for retrofitting and designing heat exchanger networks (HENs) and utilities consumption (Čuček et al., 2019). Within the different tools developed, the application of such techniques for existing oil refinery case studies has gained importance (Kamel et al., 2017).

Various research efforts have been performed on the energy management of CDUs since they are considered the major energy consuming unit operations in refinery infrastructures. Thorough studies tackled the retrofit of existing CDU using sequential approaches, i.e. column optimisation then HEN or vice versa (Joe and Rabiou, 2013), and simultaneous approaches with Pinch Analysis targets (Gu et al., 2014). Retrofit designs are regarded to be favourable than the grassroots design when it comes to petroleum refineries (Enríquez et al., 2014), with process integration methods (Pejpichestakul and Siemanond, 2013). Other solutions involving distillation design modifications and different revamping procedures were also applied to enhance the crude fractionation distillation yield and raise the energy efficiency in existing CDUs (Bagajewicz et al., 2014). More research works were focusing on enhancing the economic viability for operation by analysing progressive distillation over conventional design (Nalinakshan et al., 2019), retrofit tracing grid diagram (Nemet et al., 2015), or minimising hot utility

demand by optimising operating variables to the preflash units associated with CDUs (Ledezma-Martínez et al., 2018).

Shortcut models were advanced to account for existing HEN details with respect to the optimisation of refinery CDUs (Ochoa-Estopier et al., 2014). Such models were applied to optimise existing CDUs with objectives of energy efficiency enhancement and CO₂ emission minimisation. Also other models of CDUs were developed for various design and optimisation objectives and purposes, including area-based (Ochoa-Estopier et al., 2018), rigorous simulation (Gadalla et al., 2015) and nonlinear programming models (López et al., 2013). Improvements were also extended to entail the process itself, and studies have reported investigating feed blending for desired products maximisation (Cerdá et al., 2017) and two-column process for energy demands reduction (Kim, 2017). Therefore, there is a regular concern to generate solutions to enhance the energy efficiency and minimise emissions within existing refinery CDUs.

It is worth mentioning that refineries and HEN units studied in most reported research outcomes on retrofit/analysis of HENs were relatively old. This implied large potential energy savings when these networks were retrofitted and thus results were significant. On the other hand, results of modern refineries would not be very promising providing that their actual operated energy efficiencies are relatively high. Retrofit of such refining units would be very challenging. In the present study, process simulation and performance evaluation were performed for an existing crude distillation unit present in a local modern refinery in Egypt with a high energy efficiency. A rigorous process model was developed using Aspen HYSYS commercial software, and the results were validated with real data, considering specific crude oil properties. Pinch Analysis technique was conducted for the existing HEN to define the minimum energy

requirements for hot utility and cold utility, with the aid of Aspen Energy Analyser. An algorithm involving the incorporation of process modifications was proposed. Process modifications were applied on the case studied, achieving additional savings beyond the energy targets identified. Environmental assessment was finally carried out to evaluate CO₂ emission reduction levels. The study provides tools addressing the energy aspects and environmental concerns in CDUs of industrial relevance, which can enhance the overall sustainability profile of modern refineries.

2. Process description

The crude distillation unit studied typically consists of a topping section, fractionation section and heat exchange section. This fractionation unit belongs to a modern refinery located in Egypt with originally a high energy efficiency. The local refinery has four CDU units, one of which is more recent compared to the other three. Fig. 1 presents a schematic of the process flow diagram for the existing refinery CDU.

Fig. 1 Comes Here

The crude oil is fed through the first booster pump to the topping section; this section contains a preflash (or prefractionator) tower C-1 (20 trays – numbering is bottom-up), air cooler, trim coolers (water coolers), overhead drum (OVHD D-1), topped crude pumps, light naphtha product and reflux pumps, topped crude recirculation pumps and topped crude/fuel oil exchangers. The preflash tower C-1 feed is preheated in the heat exchange section to 213 °C then fed to the tower on the 16th tray. Overhead vapours are partially condensed in the air cooler and trim coolers from 110 °C to 40 °C. Vapour-liquid mixture enters OVHD where uncondensed vapours are routed to the vapour recovery unit. Condensed liquid is pumped by the light-naphtha product and reflux

pumps a cold reflux to the tower C-1 and light naphtha to the vapour recovery unit. To improve separation and strip the preflash bottom at its bubble-point, two heat exchangers are used to re-boil topped crude by fuel oil product from the fuel oil pump discharge directly as a heating medium.

The fractionation section contains fired heater F-1, main fractionator C-2 (32 trays), air cooler, trim coolers (water coolers), (OVHD D-2) and several pumps to discharge outlet streams for fuel oil, gas oil, bottom pump-around, kerosene, top pump-around, heavy naphtha and reflux. Topped crude is heated from 230 °C to a temperature not more than 360 °C in the fired heater operating conditions (total hot energy consumption of 37.16 MW in the existing refinery) with air-preheater in operation; the main fractionator C-2 is fed with topped crude at temperature not more than 360 °C from the fired heater below the 5th tray from the bottom. Overhead vapours are totally condensed in the air cooler E-121 and trim cooler E-124 from 130 °C to 40 °C. The condensate enters the drum OVHD D-2 where any uncondensed vapours are either routed to the vapour recovery unit or vented to flare. Condensed liquid is pumped by the heavy naphtha product and reflux pumps as a cold reflux pumped to the tower C-2 and heavy naphtha product pumped to storage tanks. Kerosene is drawn from the 21th and 23rd stages to feed kerosene side-stripper C-3. Top pump-around reflux is drawn from the kerosene draw-off to the heat exchange section then returned to the tower above trays 22 and 24. Gas oil is drawn from the 9th and 11th trays to feed kerosene side-stripper C-4. Bottom pump-around reflux is drawn from the gas oil draw-off to the heat exchange section then returned to the tower above stage number 12. Fuel oil is drawn from the bottom to the heat exchange section.

The heat exchange section contains a HEN complex, where the heat content of products (kerosene, gas oil and fuel oil) is used to preheat the crude oil. Not only the products heat content, but also that of intermediate circulating refluxes (top pump-around and bottom pump-around) is used to preheat the crude oil before desalting and preflash. In addition, the fuel oil heat content is utilised to re-boil the preflash bottom for better separation.

The heat exchange section is divided into three parts as following and depicted in Fig. 2 for the existing network. Furthermore, Table 1 shows the exchanger details for hot/cold streams exchanging heat, their temperatures, existing exchanger areas, and heat loads. The existing HEN includes 24 exchanger units.

1- *Preheating of the crude before desalting*, where the crude is preheated before desalting as follows:

- a- against kerosene in (E-100).
- b- against top pump-around in (E-101b, E-102).
- c- against gas oil in (E-103).
- d- against fuel oil in (E-105, E-106, E-107).

The crude temperature becomes about 139 °C while kerosene, gas oil and fuel oil temperatures become 55, 110 and 138 °C, respectively.

- kerosene is cooled to 40 °C in water cooler (E-118).
- gas oil is cooled to 55 °C in water cooler (E-104).
- fuel oil is cooled in water heater (E-117) then water cooler (E-119) to 80 °C.

2- *Preheating the crude before preflash tower*, such that the crude at temperature of 139 °C is heated to 213 °C against:

- a- bottom pump-around in (E-109).

b- fuel oil in (E-111, E-112, E-113).

3- *Re-boiling of preflash bottom*; in here, the bottom of preflash tower (topped crude) is re-boiled by the fuel oil from the main tower in exchangers (E-114a, E-114b).

Fig. 2 Comes Here

Table 1 Comes Here

3. Research methods

The research approach introduced in the current work has main aspects to focus on. Firstly, it concentrates on the process analysis through the simulation phase of the CDU studied. Existing data from the refinery are extracted and validated with the simulation results with respect to product specifications as hard constraints for the process. Then, energy targeting is identified using Pinch Analysis technique to ascertain Composite Curves (CC) and consequently determining minimum energy requirements for the existing CDU without major modifications, except for HEN structure. The HEN, the preflash and the main column, are hence simulated while the products specifications, yields, process conditions, energy consumption, and intermediate temperatures are used as validation constraints between the simulation model and existing actual data. Process modifications are finally proposed with the objective of achieving extra savings in terms of energy demands, beyond targets set by initial energy analysis. Fig. 3 shows the algorithm used in the present study.

Fig. 3 Comes Here

3.1. Simulation of CDU

Building a rigorous and robust simulation model for a real CDU is a challenging task and involves many iteration-based calculations, such as recycle and reflux streams,

integrated hot/cold streams, pump-arounds (PAs) liquid draws, etc. Aspen HYSYS V8.4 of an academic license is used for the simulation, modification, and validation purposes of the work. The model can be attributed to include all equipment details; a good strategy is to initially model the fractionation section to get all products that are necessary for pre-heat train network including the pump-arounds (side coolers). Further, the structure of the main column in the real plant should ideally match the simulated model unit (e.g., pump-arounds draws location, steam flow of the bottom of the column and side strippers, distillation number of stages, initial and final boiling points (IBP, FBP) of products, beside cut points and gap/overlap of products' draws from the main column). Process parameters employed during the simulation include crude oil feed composition, temperature and pressure, flow rate and intermediate temperatures between exchangers/units of flowsheet. As for the fractionation column, process parameters are the reflux ratio in the main tower, the steam flow and conditions in main tower and side strippers, the liquid flow rate and the temperature differences across PAs. These process parameters are adjusted in a way to validate the process simulation results with the existing refinery data. Product flow rates are, on the other hand, adjusted as those for the existing refinery data. The product specifications of the crude oil productions are checked/maintained while the simulation is executed.

To define the crude oil feed stream to the simulation, a crude oil assay is needed. The design basis of the real plant is based on Shukair blend crude with assay given in Table 2 for true boiling point (TBP) with its products ASTM distillation (Table 3). Aspen HYSYS represents the assay data given to a set of hypothetical pseudo-components plus water and light ends (C5-). The thermodynamic property model PR (Peng Robinson) is

defined as fluid package for simulation basis which is commonly used for organic hydrocarbons (Yela, 2009).

Table 2 Comes Here

Table 3 Comes Here

3.2. Energy analysis

Process streams of the existing data are extracted from the real process, and energy targeting using Pinch Analysis principles is performed. Pinch Analysis of the heat exchanger network is used to get the optimal network design. The analysis and targeting are made at the minimum temperature approach difference of the existing HEN. The optimisation of the network is carried out based on minimising the energy consumption of the existing process. The energy consumption defined by the energy targets are achieved by modifying the network through adding extra heat exchangers areas with stream splitting. Aspen Energy Analyser is used to achieve that task after hand calculations. The existing heat exchanger network of CDU consists of seven hot streams and four cold streams in the form required for Pinch Analysis.

The data extraction is the first step in the heat integration task carried out. Process data streams are extracted for Pinch Analysis from material and heat balances of the operating CDU and used as input to the Aspen Energy Analyser software platform. The extracted data of the current process that include heat duty, heat capacity flow rate, and supply and target temperatures are given in Table 4.

Table 4 Comes Here

3.3. Environmental assessment

After applying energy analysis procedure, there is CO₂ emissions reduction to be taken into account with equivalence to fuel burnt in the fired heater. The equation used for calculating the carbon dioxide emission amount (kg/s) is given below (Gadalla et al., 2006).

$$\text{CO}_{2\text{emissions}} = \alpha \left(\frac{Q_{\text{fuel}}}{\text{NHV}} \right) \left(\frac{C\%}{100} \right) \quad (1)$$

Where α is the molar masses of CO₂ and C (3.67), Q_{fuel} is the amount of fuel burnt in furnace (kW), NHV (net heating value) with 50% natural gas (51,600 kJ/kg) and 50% heavy fuel oil (39,771 kJ/kg), and C% is the carbon mass percent in the fuel (75.4% for natural gas) and (86.5% for heavy fuel oil).

4. Results and discussion

By applying the strategy mentioned in the simulation of CDU section, a convergence of the simulation model was reached. Fig. 4 displays the complete simulation model developed for the existing CDU.

Fig. 4 Comes Here

Based on accurate simulation outputs, Table 5 gives the design yields of products for the real plant based on Shukair blend crude 33.7 API (American Petroleum Institute) which are used to compare with results obtained in the model validation section.

Table 5 Comes Here

4.1. Model validation and performance assessment

To validate the simulation model developed, a comparison between the simulation results and the equivalent real data is drawn and presented in Table 6 and Fig. 5 in terms of products flow rates, furnace duty, condensers duties and pump-arounds temperature differences. The developed model showed a virtuous agreement with the existing data,

as the maximum error obtained between model results and real data was less than 4.5%. Table 7 compares the gap/overlap points between the real data and simulation results. It showed good match in terms of model prediction for H.N–kerosene and kerosene–gas oil. It should be noted that an extra comparison of gap/overlap between L.N/H.N could have been calculated if L.N was extracted from the main column and not from the preflash.

Table 6 Comes Here

Fig. 5 Comes Here

Table 7 Comes Here

4.2. Retrofit and energy savings with economic improvements

The minimum temperature approach difference for the existing HEN is extracted from the data given in Table 1 and found to be approximately 10 °C. The corresponding Pinch temperatures are 240 and 230 °C for hot and cold Pinches, respectively. The hot and cold energy targets for the existing HEN are 34.47 and 21.5 MW, respectively. Despite that, this CDU in particular is regarded to be modern compared with the other three CDUs in the same refinery, its HEN entails two exchangers across the Pinch. The existing grid diagram of the CDU process consists of 16 heat exchanger units (process-to-process), eight cooling utility units (water coolers and air coolers) and one hot utility unit (furnace).

To improve the energy efficiency of the existing HEN, an optimum exchanger network was designed with the aid of systematic techniques reported in the field textbooks (Klemeš, 2013; Smith, 2005) to initially draw the CC and Grand Composite Curves (GCC), as can be seen in Figs. 6a and 6b.

Fig. 6a-6b Comes Here

As reported in Fig. 6a, the hot energy target for the existing refinery is 34.47 MW at ΔT_{\min} of 10 °C. This results in an energy efficiency of the existing refinery of almost 93% compared with energy targets. Therefore, this reported high efficiency of the existing plant will make the retrofit task challenging as the potential maximum energy savings would be limited to 7% in this case. After estimating the minimum heating and cooling requirements (energy targets) for the process, a certain algorithm should be respected in terms of designing the HEN diagram that achieves the maximum energy recovery (Kemp, 2007). During the design of HEN, existing structure is fixed and streams matching. When the process parameters inside the distillation column are changed, this change is reflected on the operation of the network so as changing the temperatures and flow rates of the streams in exchanger units. Also, changing temperatures and splitting streams would influence the operation of the network leading to changes in additional areas required, and heat loads. The considered changes in retrofit are represented by changes in temperatures of streams in HEN in addition to additional areas and heat loads. An optimum HEN has been successfully designed. Firstly, by splitting the crude before the desalter to be integrated with the light and heavy naphtha process streams with respect to their mass heat capacities. As the rule below the Pinch point should respect the principle $CP_h \geq CP_c$. Afterwards, by following the same rule a split stream was integrated with the fuel oil with ratio equals to $CP_{(\text{fuel oil})}/CP_{\text{crude}}$ and the other split integrated with the kerosene stream. The remainder of the crude stream was integrated with bottom pump-around (BPA) stream. Secondly, the crude stream after the desalter was split to four streams with the respect of CP_c of top

pump-around (TPA), gas oil, fuel oil, and (BPA) respectively. Then, the remainder was integrated with the Fuel Oil as shown in Fig. 7.

Fig. 7 Comes Here

Above the Pinch region, is for the segment reflecting the considerable economic and energy savings for heating requirements of the network. An integration with the second heat exchanger of the existing HEN was applied to exploit the highest temperature of the fuel oil to get around 9 °C extra to the topped crude before the fired heater. In addition, a new heat exchanger of bottom pump-around was added to be integrated with the bottom C-1 stream that acts as re-boiler; with this effective replacement, more energy savings were reached. A better heat integration was achieved by raising temperature before the fired heater to 239.4 °C instead of 230 °C. The potential energy saving includes the heating utility saving of about 7% and the cooling utility saving of 11% over the current design. Consequently, the amount of CO₂ emissions for the new network is estimated to be 52,705 t/y using natural gas as fuel gas to save 23.5% comparing with the existing network which emits 68,898 t/y using natural gas and heavy fuel oil as combined source of fuel for the furnace.

4.3. Process improvements

In order to enhance the process performance and minimise energy consumption, a new process cycle is projected, with ensuring the column and products specifications are fixed without violation. A preflash pump-around cycle could be installed to use excess energy in HEN (Errico et al., 2009; Waheed and Oni, 2015). However, this modification will increase the vaporisation rate in the preflash and diversify its temperature profile, which lets heavier components to amend the distillate product composition. So, this

modification was not implemented due to design limitations of the column, for avoiding high pressure, entrainment or flooding during operation. However, there are some solutions in operation to reduce vaporization rate i.e. reducing reboiler duty, increasing reflux by the air coolers rotor speed or decreasing feed rate to the column. Nevertheless, feasible operation changes with zero topology modifications, like changing intermediate temperature of CDU top vapours or increasing temperature difference of CDU pump-arounds has proven to result into considerably minor savings (Gadalla et al., 2016). Thus, a novel modification is proposed through exchanging a total of 8 MW of energy via heater E-101 between (two material streams) topped crude stream and new pump-around stream with amount of 112 t/h drawn from stage four at 337 °C. In real plant revamping, this installation will be more than one heat exchanger due to the large area to be installed.

Since it is the highest stage temperature of the column profile, the new installation of pump-around at that stage with temperature difference of 95 °C from return temperature to stage five acts as 2nd bottom pump-around. E-101 is simulated as a heater and tied with stream set SET-1 to the new pump-around cooler inside C-2 column environment with the same energy quantity for both the heater and the cooler, as shown in Fig. 9.

Fig. 9 Comes Here

The outlet steam of the heater stream 17 enters a flash drum V-101 to separate 18.6% of the vapour fractions resulted from the extra heat gained by the topped crude and introduced to the main column directly to decrease the head load of the fired heater F-1 by decreasing the total mass flow rate inside the fired heater. The remainder liquid stream 18 is pumped by P-8 to F-1 at 262 °C instead of 231 °C to achieve a total of

24.76 MW hot utility, resulting in 31.3% reduction of fuel consumption of the fired heater and 45.1% CO₂ emissions reduction over the existing design.

4.4. Energy and economic comparisons for retrofit scenarios

A complete comparison of the existing case, first retrofit solution and second retrofit solution is presented in Table 8, showing a summary of energy savings and emissions savings with economic consideration for each scenario and their supplementary additional capital investment.

Table 8 Comes Here

The utility cost for fuel and cooling water are obtained from actual plant as 5 \$/MMBTU for fuel gas and 13.44 \$/MMBTU for fuel oil. Additional capital investment was related to the additional area installed for each scenario, providing that the existing column and the flowsheet structure are fixed. Thus, additional capital costs are obtained from the economic equations in manual of process economic evaluation (Chauvel et al., 2003) with additional cost index modifications. For retrofit scenarios and calculations, existing exchanger areas are fixed, while additional areas for retrofit solution are estimated for new heat loads. For these solutions, additional capital investments were inflicted whereas energy and utility savings were obtained. The payback time for first scenario was almost one year and a half for energy savings equivalent to 670,247 \$/y. Second retrofit had a higher and more substantial energy savings equivalent to 3,089,618 \$/y and less than a year as a payback period. CO₂ releases were significantly minimised by 45.1% regarding to 8,699 kg/h of the existing plant releases.

5. Conclusions

A process simulation model has been developed for a modern crude distillation unit with preflash existed in an oil refinery. The model has been validated with real operation parameters and intermediate details of the CDU, showing a good agreement in the comparative results. A heat integration retrofit analysis has been carried out to identify energy targets and potential savings for the existing HEN. Pinch Analysis Technique was adopted for designing a heat integration network that improves energy efficiency of the process. New modifications of the process design showed that stream splitting of the main crude stream before and after the desalter was located below the Pinch point, which subsequently decreased the cooling utility and electricity consumption. Above the Pinch, better heat integration has been achieved with raising the temperature before the fired heater to 239.4 °C instead of 230 °C. The potential energy saving included heating utility saving of 7.2% and cooling utility saving of 11.1%, with 23.5% CO₂ emissions reduction over the current design. Furthermore, the developed simulation model has been used to propose process modifications, which are then applied to achieve less energy consumption than defined by the energy targets. This led to substantial energy savings beyond the Pinch limitations by 31.3% and 45.1% CO₂ emissions reduction.

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Nomenclature

CDU crude distillation unit

ΔT_{\min} minimum temperature approach difference

HEN heat exchangers network

L.N Light Naphtha

H.N Heavy Naphtha

OVHD overhead drum

TPA top pump around

BPA bottom pump around

CP heat capacity

CC Composite Curves

GCC Grand Composite Curve

IBP initial boiling point

FBP final boiling point

α molar masses

Q_{fuel} fuel burnt in furnace, MW

NHV net heating value

C carbon mass percent in the fuel

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Figure captions

Fig. 1. Simplified process flow diagram of existing CDU.

Fig. 2. Existing heat exchanger network.

Fig. 3. The approach methodology for simulation and energy optimisation of modern CDUs.

Fig. 4. The developed simulation model for CDU case study.

Fig. 5. Products flowrates comparison of real data and simulation model values.

Fig. 6. Composite Curves with ΔT_{\min} 10 °C (a) Hot and Cold CC (b) Grand Composite Curves GCC.

Fig. 7. The optimum HEN grid diagram.

Fig. 8. Illustration of new process modification into the simulation model.

Table captions

Table 1. Existing heat exchanger network design data and specifications.

Table 2. Crude Assay for TBP distillation of Shukair blend crude oil.

Table 3. ASTM D86 distillation for light naphtha, heavy naphtha, kerosene and gas oil.

Table 4. Design yields for Shukair blend crude oil.

Table 5. Data extraction (process streams).

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Table 8. Existing vs optimum HEN.

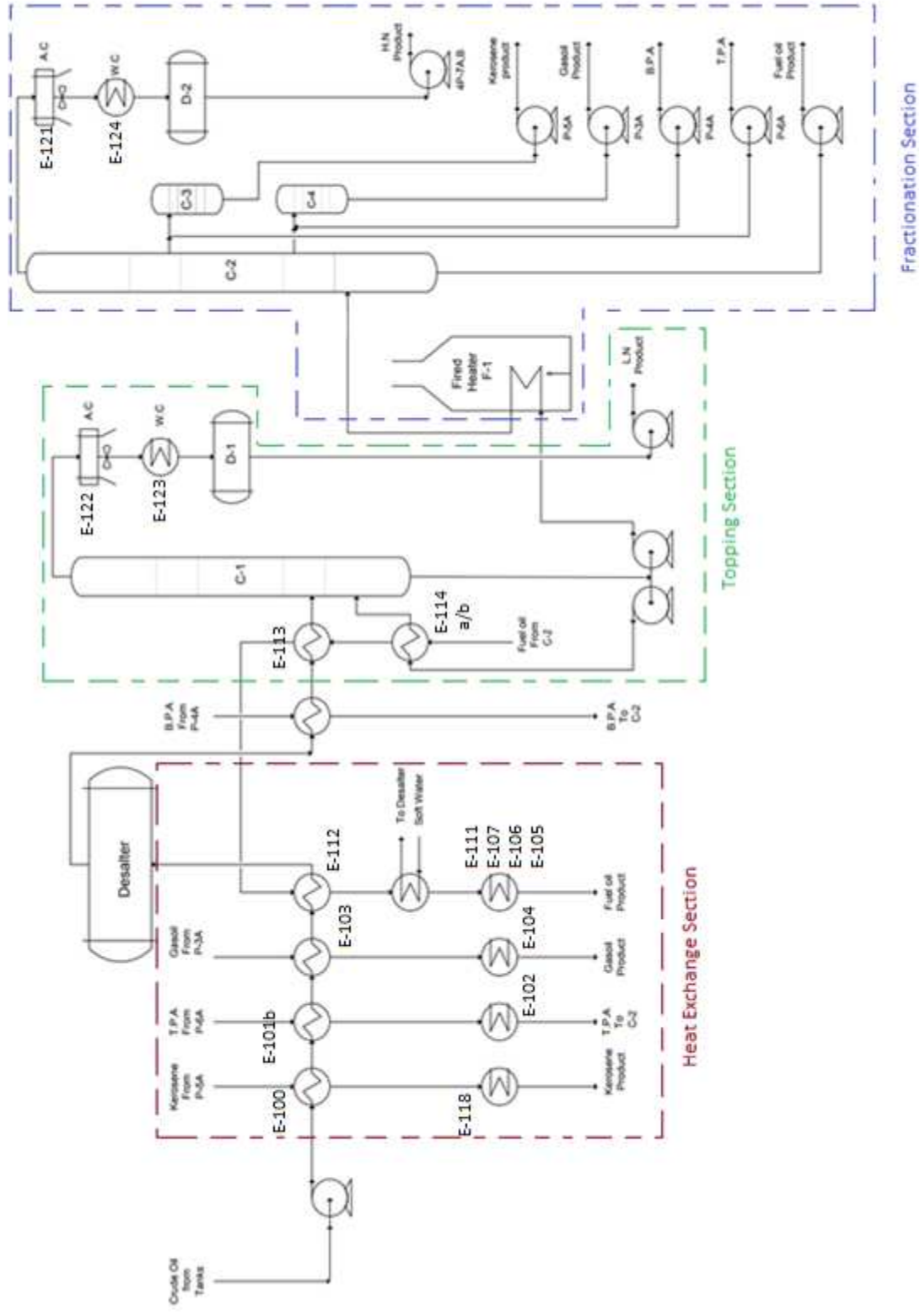


Fig. 1. Simplified process flow diagram of existing CDU.

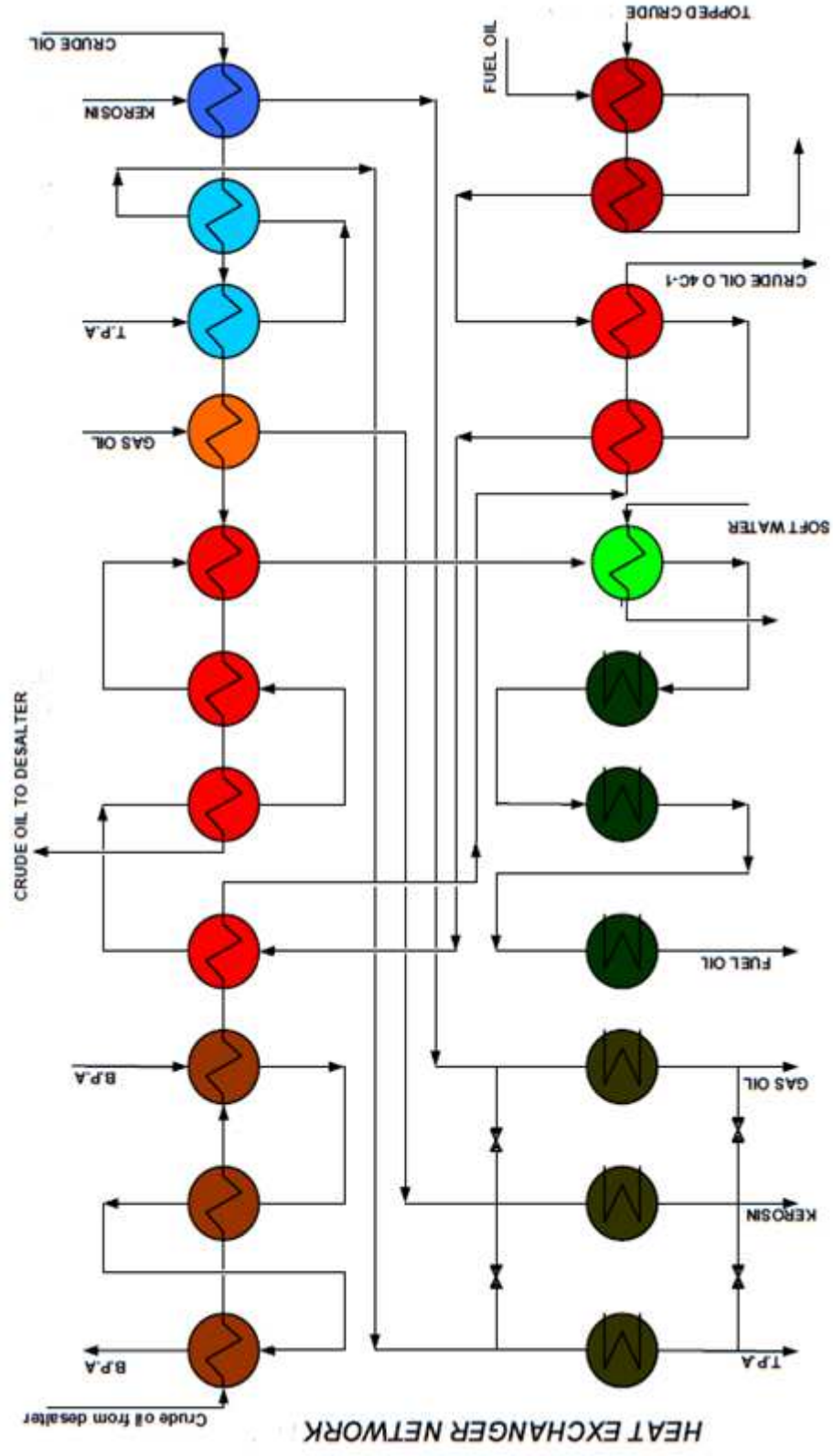


Fig. 2. Existing heat exchanger network (design data given in Table1).

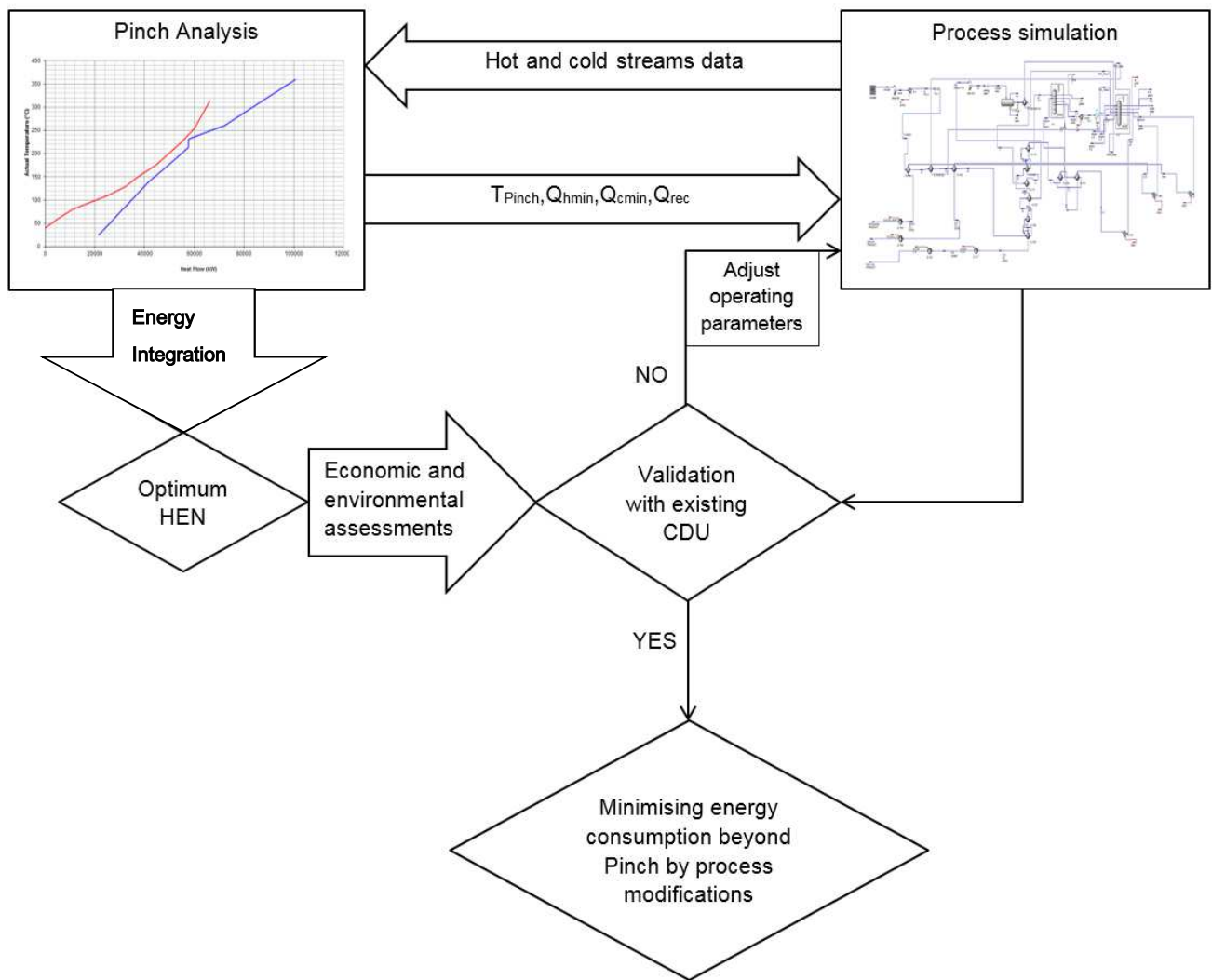


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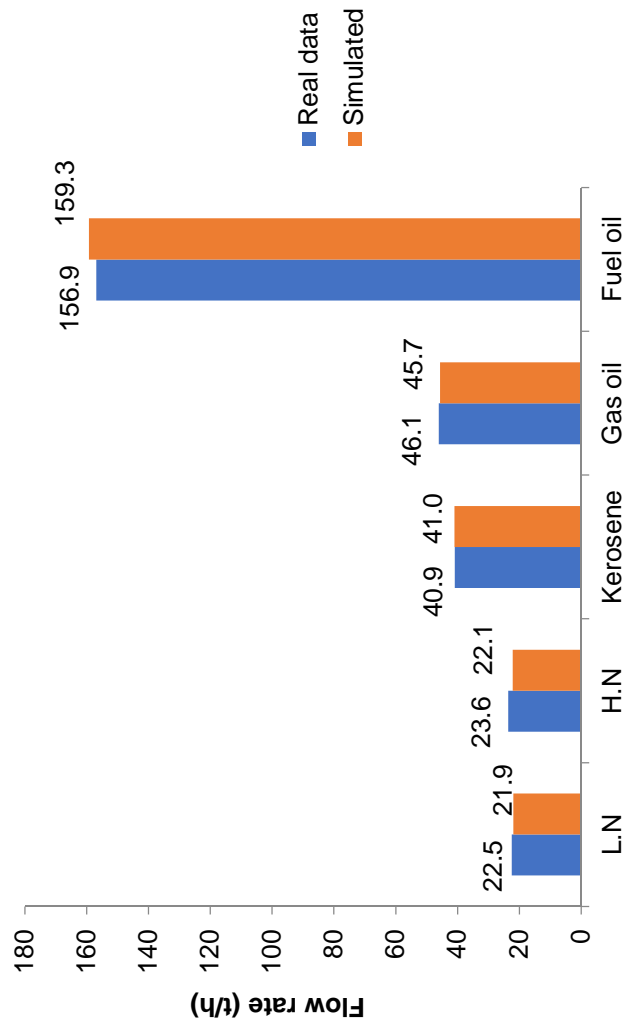


Fig. 5. Products flowrates comparison of real data and simulation model value.

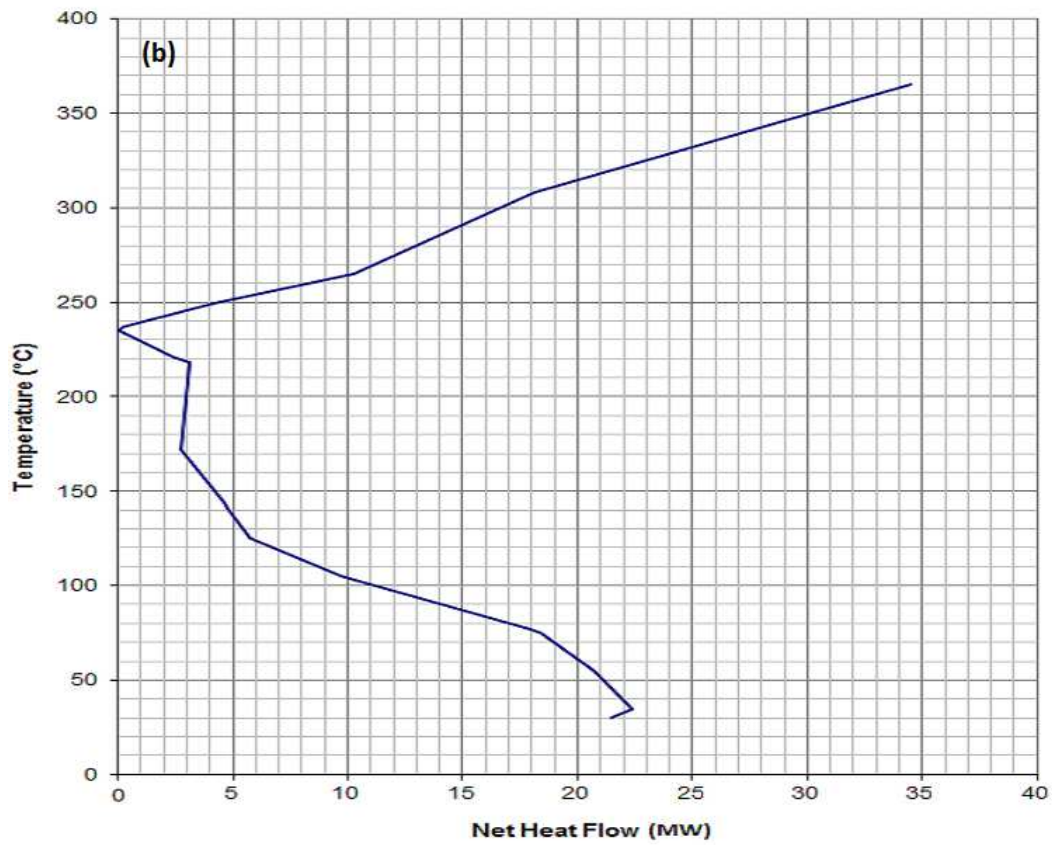
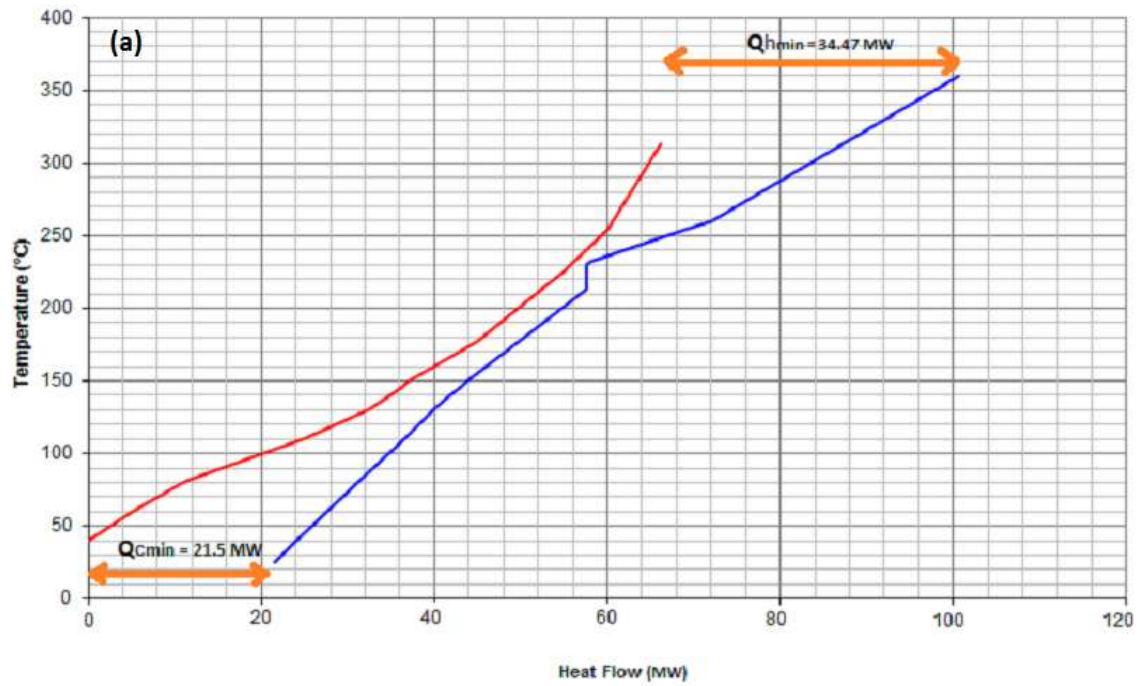


Fig. 6. Composite Curves with ΔT_{\min} 10 °C (a) Hot and Cold CC (b) Grand Composite Curves GCC.



Table 1. Existing heat exchanger network design data and specifications.

Heat exchanger	Cold stream	Cold T_{in} ($^{\circ}C$)	Cold T_{out} ($^{\circ}C$)	Hot stream	Hot T_{in} ($^{\circ}C$)	Hot T_{out} ($^{\circ}C$)	Area (m^2)	Duty (MW)
E-100	Crude Before 1	25	41	Kerosene 1	146	55	235	2.4
E-102	Crude Before 2	41	59	TPA 1	126.7	82.2	235	3.8
E-101b	Crude Before 3	59	81	TPA 2	176	126.7	235	3.6
E-103	Crude Before 4	81	103	Gas oil	226	110	235	3.7
E-108	Crude After 1	139	146	BPA 1	162.2	149	274	1.5
E-109	Crude After 2	146	154	BPA 2	196	162.2	274	1.9
E-110	Crude After 3	154	176.8	BPA 3	254	196	274	4.3
E-105	Crude Before 5	103	117.2	Fuel oil 8	155	137.78	235	2.8
E-106	Crude Before 6	117.2	126.6	Fuel oil 7	175	155	235	1.6
E-107	Crude Before 7	126.6	139	Fuel oil 6	199	175	235	2.3

E-111	Crude After 4	176.8	185	Fuel oil 5	215.5	199	274	2.2
E-112	Crude After 5	185	200.5	Fuel oil 4	240.5	215.5	274	3.2
E-113	Crude After 6	200.5	213	Fuel oil 3	265.5	240.5	274	2.8
E-114a	Bottom C1	232	244	Fuel oil 2	285	265.5	274	3.0
E-114b	Bottom 2 C1	244	260	Fuel oil 1	312	285	274	2.9
F-1	Topped Crude	230	360	Fired Heater				37.2
E-118	Cooling Water 1	30	37	Kerosene 2	54	40	69	0.4
E-104	Cooling Water 2	30	37	Gas oil 2	82	54	162	1.6
E-117	Cooling Water 3	30	37	Fuel oil 9	125	100	150	1.1
E-119	Cooling Water 4	30	37	Fuel oil 10	100	76	143	2.1
E-123	Cooling Water 5	30	40	L.N	60	40	228	1.9
E-124	Cooling Water 6	30	40	H.N	60	40	228	2.7
E-122	Air Cooling 1	33	35	L.N	110	60		4.8
E-121	Air Cooling 2	30	33.3	H.N	130	60		9.5

Table 2. Crude Assay for TBP distillation of Shukair blend crude oil.

Liquid mass percent, %	Temperature, °C
8	100
12	125
16	150
20	175
24	200
28	225
30	250
35	275
40	300
49	355
57	400
67	455
77	510

Table 3. ASTM D86 distillation for light naphtha, heavy naphtha, kerosene and gas oil.

Liquid volume percent, %	L.N, °C	H.N, °C	Kerosene, °C	Gas oil, °C
IBP	45	99	162	257
10	48	106	169	266
30	57	118	186	281
50	64	126	204	291
70	73	135	220	304
90	89	149	237	323
FBP	104	160	254	345

Table 4. Data extraction (process streams).

Stream name	Type	Inlet Temp. (°C)	Outlet Temp. (°C)	Specific Cp (kJ/kg °C)	Flow rate (kg/h)	Enthalpy (MW)
Kerosene	HOT	146	40	2.303	40950	2.7765
TPA	HOT	176.7	82.2	2.461	113578	7.3371
Gas oil	HOT	226	60	2.48	46165	5.2788
BPA	HOT	254.4	149	2.648	99879	7.7438
FO	HOT	312.8	80	2.358	156963	23.930
Crude Before Desalter	COLD	25	139	2.174	291667	20.083
Crude After Desalter	COLD	139	213	2.669	291667	16.003
Bottom C1	COLD	232	260	2.813	267703	5.8576
Topped Crude	COLD	230	360	3.844	267703	37.160
L.N (Reflux + Product)	HOT	110	40	6.627	52690	6.7893
H.N (Reflux + Product)	HOT	130	40	8.268	59376	12.273

Table 5. Design yields for Shukair blend crude oil.

Product	Yield (wt%)
Fuel gas	0.193
Propane	0.312
LPG	1.836
Stabilised naphtha	5.875
Heavy naphtha	8.10
Kerosene	14.04
Gas oil	15.828
Fuel oil	53.816

Table 6. Real data vs simulation model results.

Parameter	Real data	Simulation value
Feed, t/h	291.7	291.7
Furnace outlet temperature, °C	360.0	360.0
Furnace inlet temperature, °C	230.0	231.1
Furnace duty, MW	37.16	35.54
Top pump-around, t/h	113.6	114.1
Top pump-around draw temp., °C	176.7	173.4
Bottom pump-around, t/h	99.88	99.02
Bottom pump-around draw temp., °C	254.5	257.8
C-1 condenser duty, MW	6.74	6.67
C-2 condenser duty, MW	12.44	12.32

Table 7. ASTM D86 gap and cut points for certain products.

5–95 Gap/Overlap	Real plant (°C)	Simulation (°C)
H.N–kerosene	11	11.54
kerosene–gas oil	16	26.83

Table 8. Existing vs retrofitting scenarios.

Factor	Heating (MW)	Cooling (MW)	Reduction in CO ₂ emissions (%)	Energy reduction (%)	Additional area (m ²)	Additional capital investment (\$)	Payback (y)
Existing case	37.16	27.7	–	–	–	–	–
1 st retrofit	34.47	21.5	23.5	(7.21 Heating	5,850	1,056,937	1.57
(HEN optimisation)				+11.08 Cooling)			
2 nd retrofit	24.76	24.2	45.1	31.3 Heating	16,630	3,004,594	0.97
(novel process configuration)							

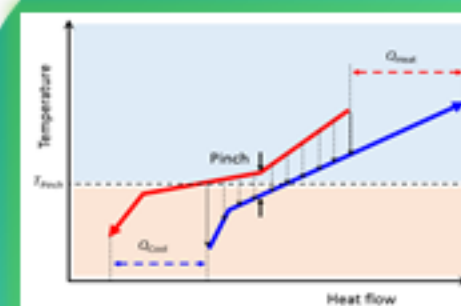
Modern CDU



Robust Simulation Model



Pinch Analysis



Process Modifications

