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New Retrofit Approach for Optimisation and Modification for a Crude Oil Distillation System

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Crude oil atmospheric distillation units consume substantial amounts of energy, as equivalent to a 2 % of the total crude oil processed (Errico, 2009). An existing crude distillation unit is costly to modify due its complex configuration and existing limitations of structure, space area, matches, bottlenecked equipments, etc. Thus, a few new crude distillation units are built and most projects are directed to revamping existing equipments. Modifying an existing plant is a tedious task, more complex than a new process. While revamping, many parameters must be considered and structure limitations need to be met.
This paper develops a new revamping method based on rigorous simulation and optimisation procedures. This method accounts for both the distillation column and the associated heat exchanger network at the same time to maximise the use of existing equipments. The methodology considers process changes and structural modifications together with the interactions between the existing distillation process and heat recovery system. The new method is valid for multiple objective functions, i.e. saving energy, reducing emissions, enhancing production capacity, and profit improvement.
The new presented methodology is applied to a local atmospheric plant for MIDOR, as an Egyptian refinery case study. Many revamping options were obtained, including no structural modifications, simple additional exchanger areas, and additional units or equipments.

1. Introduction

Crude distillation units (CDUs) are major energy-consuming units and therefore require extensive energy management. There are many ways to increase energy efficiency, and heat exchanger network (HEN) design and process heat integration are widely used methods. Heat transfer from hot products and pump-around streams to the crude feed by the applications of HEN reduces the energy demands of both coolers and furnaces. This reduction of energy demands diminishes the operating cost while increases the capital cost for exchanger area installation, therefore, the retrofit design is more preferable than the grass-roots design for oil refineries (Pejpichestakul, 2013). Standard objectives of revamping include increasing the plant throughput, reducing the energy demands, utilising more efficiently the raw materials, reducing the atmospheric emissions and waste generation. All these objectives preferably be fulfilled without modifying much the physical constraints of the unit, such as column actual diameter, pump-arounds and side-columns locations, exchanger matches and areas, maximum heat loads (fired heaters), etc. The interactions between the existing distillation process and heat recovery system have a critical impact on the revamping of the overall process. These interactions are the operating conditions of the distillation column, including feed preheating temperature, steam flow rate, pump-around duties and flow rates and reflux ratio, in addition to the existing exchanger matches and areas of the heat recovery system (Gadalla, 2003).Many researchers worked on revamping crude distillation units by sequential approaches i.e. column revamping then HEN or vice versa, or in simultaneous approaches with targets of Pinch Analysis (Gadalla et al., 2003). In these research works, existing heat exchanger networks were considered through their targets only and not via their matches or physical constraints.
The main objective of this work is to develop a new methodology for revamping and simulation framework for heat-integrated crude oil distillation systems. This approach is based on rigorous simulation and

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considers the existing distillation process simultaneously with the existing heat recovery system, meeting their physical constraints. The new methodology provides a system for considering and exploring structural modifications to the existing flow sheet and heat exchanger network. The trade-offs between the capital investment required by retrofit modifications and the cost savings are to be accounted for the new retrofit objectives, as well as the conventional goals, are tackled. Retrofit design options may be explored to result in better performances and achieve objectives with minimum fixed cost and larger cost savings.

2. A New revamping methodology

The new revamping approach focuses on the efficient reuse of existing equipment without major modifications; however, it also accounts for changing the structure of the distillation column and the exchanger network. Figure 1 presents the new methodology for revamping of crude distillation systems.
The retrofit approach presented in Figure 1 treats an existing column design and details of the associated heat exchanger network. The new approach as shown is based on a rigorous simulation modelling of both units, *i.e.* column and HEN, as all physical details of these units are fixed. In the simulation step, existing column with side strippers, pump-arounds streams, reboilers and condenser are simulated for the given products' specifications. On the other hand, all heat exchanger matches of the existing exchanger network are simulated simultaneously with the distillation unit. Interactions between the two units (columns, HEN) are considered in the simulation through all product streams, top vapours stream, bottom liquid streams, and pump-around recycled streams. The simulation obtained in this step is a robust model and can be used for any revamping study, sensitivity analysis, and development projects. All these streams are routed into their exchanger matches to preheat the crude oil feed before the fired heater. Then the two units are optimised simultaneously by varying process conditions and further by modifying the system structure for a given objective. During optimisation, existing physical constraints and product specifications are maintained. The objective of the retrofit optimisation can be varied, including energy, cost and atmospheric emissions. Optimisation results will be set of optimum operating conditions, optimum distribution of heating and cooling loads within the HEN and in other cases some structural modifications to existing hardware

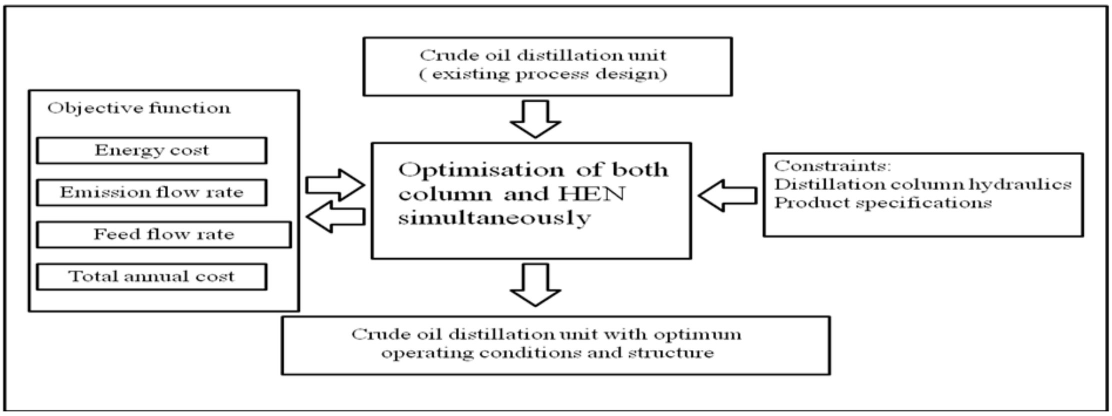

Figure 1: The new retrofit approach methodology

3. Case study Application

The new developed revamping method is applied to a real refinery plant for minimising the energy consumption and carbon dioxide emissions considering low modification costs. Case study data are industrial actual data obtained from MIDOR (Middle East-oil refinery, Egypt).

3.1. Existing refinery crude unit

The existing column configuration is given in Figure 2; it uses three side-strippers and two pump-arounds. Steam at 6.1 barg and 340 ^oC is used for stripping at the bottom of the main column to strip the light components dragged in the liquid, in the bottom light diesel side-stripper and the bottom of heavy diesel side stripper, while reboiling is employed in the kerosene side-stripper. The existing distillation tower processes 100,000 barrels/d (2811 kmol/h) of crude oil blend of 50 % Arabian light and 50 % Arabian heavy. The blending of different stocks is normally done to obtain the required product yields and also to meet the process constraints.

The atmospheric distillation tower produces five products: vapours which are further processed to produce naphtha, kerosene, light diesel (LD), heavy diesel (HD), and residue (RES), the flow rates of the products are presented in Table 1.

Table 1: The flow rates of the products of the atmospheric distillation tower

Product streams	Flow rate (kg/h)
Over head vapour	125,000
Kerosene	49,240
LD	106,349
HD	11,369
Residue	293,460

The crude oil is heated in a heat exchanger network from 25 °C to 264 °C by exchanging heat with process hot streams in two trains. The first train consists of five heat exchangers after which the temperature of the crude reaches 130 ^{o}C, and then the crude passes through a desalter to remove inorganic salts, impurities and soluble metals. Then the desalted crude enters the second train which consists of six heat exchangers. The crude oil can reach a maximum temperature of about 264 °C. This temperature is still too low to achieve the grade of crude vaporisation necessary for the separation in the main column and thus a furnace is always necessary. The temperature of the exiting stream from the furnace is about 365 °C and fuel oil or fuel gas, depending on the refinery availability, is used as energy source. All the heat needed for the separation is given in the furnace, so no reboiler is present in the main column.

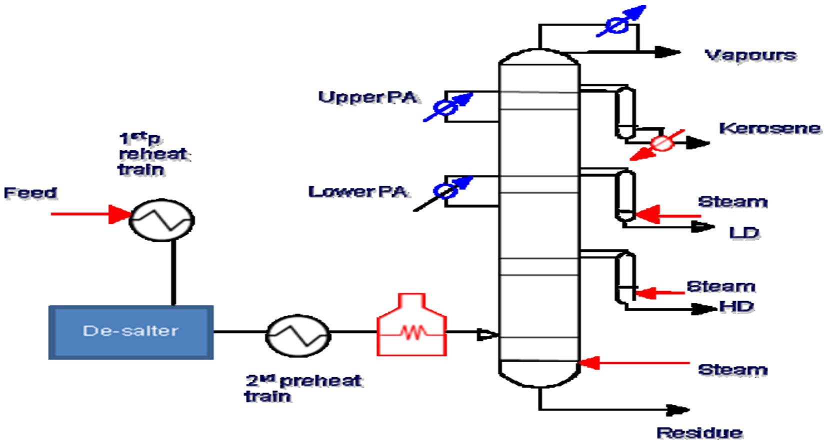

Figure 2: Crude distillation unit configuration

The large temperature difference between the inlet and the outlet streams of the furnace and the high flow rate of the crude processed make the furnace as one of the highest energy consumer of the whole refinery, the duty of the furnace is 61.96 Mkcal/h (71.9 MW) with an efficiency of 91 %, so the net duty of the furnace is 56.3 Mkcal/h (65.4 MW). It follows that also the cost of this unit is a meaningful part of the overall production costs. The cost of energy used in the furnace can be calculated from the data available from the refinery that each 10^6 kcal/h (1.163 MW) cost 4,982 $/month, the total cost of the energy used in the furnace is then 3,330,245 $/y, and this value will be used later in calculations of energy savings. The CO_2 emissions from burning the fuel in the furnace is calculated by the equation developed in (Delaby and Smith, 1995) where the NHV of fuel oil is 39,771 kJ/kg, C is 86.5 % and α is 3.57 to be 18,679.8 kg/h. Figure 3 presents a schematic diagram of the existing heat exchanger network with its eleven heat exchangers; the blue circles represent the coolers.

The objectives of this case study are developing a rigorous process design model for the existing refinery distillation plant, increasing the energy efficiency, optimising the current operating conditions of the existing refinery distillation unit and modifying the existing HEN to reduce the energy consumption and increase the profits

4- Results

4.1. Optimising the process conditions of the distillation unit

During this optimisation, the existing distillation column and the associated heat exchanger network (HEN) are considered; optimisation is done by changing operating condition without adding any new equipment in order to minimise the total annualised cost of energy consumption and operating cost. During optimisation, the existing distillation column configuration, number of stages, locations of condenser, pump-arounds and side strippers are kept fixed. Also the details of the existing heat exchanger network including connections, existing number of exchangers are fixed.

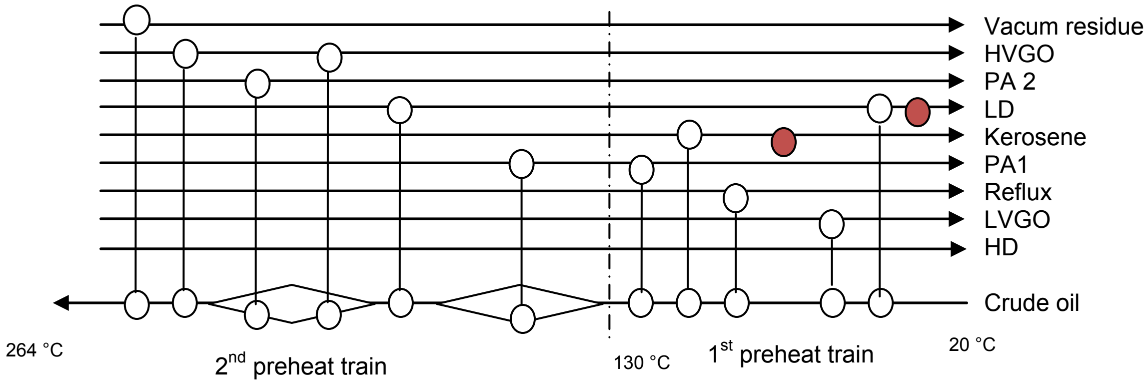

Figure 3: The existing HEN

In this part, the optimisation is divided into two sections; the first section is concerned with the optimisation done on the distillation tower itself. The optimisation is done by changing the column operating conditions like pump-arounds flow rates, reflux ratio, stripping steam rate and temperature across the pump-arounds to reduce the energy consumption. The second section is concerned with optimisation done to the column and at the same time the redistribution of heat loads between all exchanger matches by the usage of utility paths and loops or by changing the splitting ratio of different streams. This method aims to increase the temperature of the crude oil before entering the furnace as much as possible by maximising the heat recovered from hot streams to cold streams. As a result, the duty of the furnace is reduced and consequently the energy consumption of the overall system is decreased.

4.1.1. Pump-arounds flow rates

The pump-around in our case is not only used as a cold reflux to the column to enhance the separation but also used as a heat exchanger in preheat train. Thus, any change in the pump-around flow rate will affect the HEN either by increasing or decreasing the area and duty required for heat transfer. Two procedures are considered, first focusing on the pump-around flow rates, second when the redistribution of heat loads on exchangers are accounted for. In the first part, the optimisation is focused on the column environment only, considering the interaction between column and HEN. As mentioned before the distillation column has two pump-arounds, so the optimum pump-around flow rate is obtained for both pump-arounds with respect to the minimum energy requirements. So by changing the upper pump-around flow rate, the minimum heating energy target tends to be constant after a certain value. Thus, the upper pump-around flow rate does not affect the energy consumption of the system. By applying the above procedure on the lower pump-around, the optimum value of the lower pump-around flow rate is 540 m^3/h which corresponds to the lowest minimum energy consumption value. By applying this value in the rigorous model simulation, the temperature of the crude oil before the furnace reaches 285.3 °C instead of 264 °C which consequently decreases the duty of the furnace from 56.7 Mkcal/h (65.8 MW) to 50 Mkcal/h (58.1 MW). The reduction in energy consumption is equal to 10.7 %; this corresponds to an operating cost saving of 366,327 $/y. Hence, this reduction of energy will reduce the CO_2 emission flow rate to 16,695 kg/h, with a reduction of 11.1 %. After getting the optimum value for the lower pump-around flow rate, the impact of this optimisation results on the heat exchanger network is obtained in the form of additional area required for heat transfer. The total additional area of heat transfer is found to be 3,052 m^2. This additional area is distributed between three heat exchangers, exchanger 1, exchanger 6 and exchanger 10. The cost of the additional area is calculated from the following equation (Gadalla, 2003):

$$\text{Heat exchanger area cost (\$)} = 1530*(\text{additional areas})^{0.63} \quad (1)$$

The cost calculated from this equation should be multiplied by the Marshall and Swift cost index of 2011 which equals 1.29 (Chemical Engineering magazine, 2011). So the cost of the additional area is 309,435 $ and the payback time is determined to be 0.85 y. It should be noticed that during the above optimisation, both the product flow rates and specifications are kept constant. In the second part, the optimisation will be based on the column (changing the lower pump-around flow rate) and maximising the heat recovered from all hot streams available in the process to preheat the feed crude to a maximum temperature before entering the furnace. This optimisation is done through interchanging the heat loads between the heat exchangers by utility paths and loops (Linnhoff, 1982), and by using all the energy from HVGO (heavy vacuum gas oil) in exchanger 10 instead of using part of it in exchanger 8. Also the splitting ratio of the crude after the first train and before the eights heat exchanger is optimised. The total additional area of heat transfer is calculated and is found to be only 485 m^2. This additional area is distributed between three heat exchangers, exchanger 4, exchanger 6 and exchanger 8. The resulting cost of the associated additional area is 97,117.7 $. Figure 5 shows the modified heat exchanger network with the additional area required for heat transfer on every exchanger unit and is presented by the dotted circles.

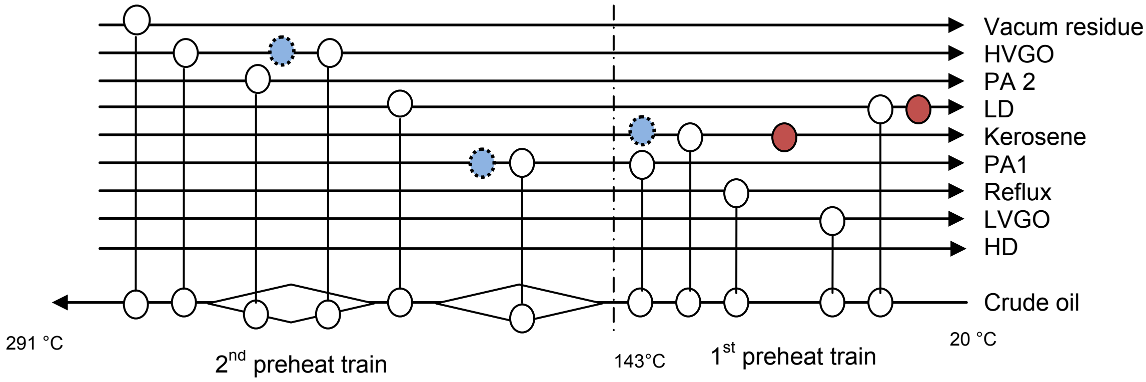

Figure 5: The modified HEN

As a result of applying the optimisation, the temperature of crude oil before the furnace increases 291 °C instead of 264 °C for the base case, which consequently decreases the duty of the furnace from 56.7 Mkcal/h (65.8 MW) to 46.98 Mkcal/h (54.98 MW), with a reduction in energy consumption of 16 %. This corresponds to an energy cost saving of 532,839 $/y.The payback time equals 0.18 y - less than that in the former case as the optimisation in the latter case is for both column and HEN. The CO_2 emissions also decrease to 15,686 kg/h, which corresponds to a 16.5 % reduction over the base case level of emissions.

4.2. Revamping with structural modifications

Structural modifications are done by the addition of new equipment like a preflash, a prefractionator or a new pump-around in order to enhance the capacity of the refinery and increase the energy efficiency of the system. This leads to an improved profit.

4.2.1. The installation of a preflash vessel

The addition of preflash removes the light vapour fractions of the crude oil feed before entering the furnace. The vapour fraction after the second train represents almost 44 % of the total feed mixture. So, the preflash is better situated at the outlet of this preheat train, i.e. train 2. Many trials of simulations are done with different liquid column-feed temperatures to adjust the gap/overlap temperature within the specified value. Each trial affects the percentage reduction of the furnace duty; these trials' results concluded that the liquid enters the tower at 370 °C. This increase in feed temperature will consequently increase temperature of the crude oil after the second train from 264 °C to 268.8 °C. After the crude oil exits from the second train of heat exchangers at 268.8 °C, it enters the preflash drum to separate between liquids and vapours where the liquids go through the bottom and the vapours leave through the top. The liquids are then directed to the furnace at 268.8 °C to be heated further till 370 °C and then enter the distillation tower. The duty of the furnace decreases from 56.7 Mkcal/h (65.8 MW) to 37.8 Mkcal/h (43.92 MW) and this corresponds to a 32 % reduction. The energy cost savings are of 1,065,678 $, and the CO_2 emissions are also decreased to 12,621.6 kg/h, with a corresponding reduction of 32.8 % reduction over the base case. The vapours on the other hand go directly to the distillation column; the position of vapours entering the tower is a very important parameter to be considered for efficient energy management. The best position for the vapours to enter the tower is at tray number 25 because its temperature is 264.8 °C which is the nearest temperature to the vapours. For this case, the optimisation is performed for the optimum distribution of heat exchangers loads; the total additional area is reasonably low and is equal to 485 m^2. The additional cost in this case is not only the cost of additional area required for heat transfer but also includes the cost of the preflash drum. The preflash drum is assumed to be vertical and its cost is obtained from the HYSYS model as 58,291$, while the cost of total additional area is calculated to be 97,117 $. The total modification cost required is 155,409 $; the payback time is 0.145 y.

4.2.2. The installation of a new pump around

The function of adding new pump-around is increasing the cold reflux to the column and thus to enhance the separation; it also works as a heat exchanger in the preheat train of the crude oil for extra energy integration. As result, the temperature of the crude oil before entering the furnace will increase, and the energy consumption in the furnace will decrease. After getting the optimum flow rate and the optimum duty for the new pump-around, the pump-around is installed into the preheat train after the 10th heat exchanger. The temperature of the crude oil entering the 11th heat exchanger changed to be 277.2 ^oC instead of 248.7 ^oC, which was the value of the temperature before adding the new pump-around. Adding a new pump around resulted in more energy recovery which in turn raised the temperature of the crude oil entering the furnace to 295.1 ^oC so the duty of the furnace decreased to be 45.5 Mkcal/h (52.8 MW) with 18. 7 % reduction in energy. The equivalent of energy cost savings is about 606,104.5 $/y. The total additional area required for heat exchangers is estimated to be 2,767 m^2, with a capital cost of 290,955 $; the payback time 0.48 y.

5. Conclusions

A retrofit design approach has been developed for existing crude oil distillation systems with their associated heat exchanger networks. The approach is a rigorous simulation and optimisation-based model that considers simultaneously the existing distillation column with its full details and the associated HEN. The optimisation procedure can be for single variable or multi-variables. The retrofit approach is applicable for several objectives, mainly improving the energy and hydraulic performances of an existing distillation system with its associated preheat train. The interactions between the two individual units have been optimally exploited. The developed retrofit approach can achieve various objectives for refinery crude distillation units, including: energy savings, atmospheric emissions reduction, capacity enhancement, efficient utilisation of the raw materials, feedstock changes, and profit improvement. An actual case study has been presented to show the applicability of the new retrofit method. Several retrofit solutions have been obtained, ranging from zero-modifications and simple additional exchanger areas to additional units or equipments. Simple optimisations of both the distillation operation and the HEN have lead to 17% savings in energy consumptions and emissions. The addition of a preflash to existing structure showed large energy savings of up to 32 % compared with base case and substantial utility cost savings per year of 1,065,678 $. The payback time for most retrofit solutions did not exceed a couple of months. Atmospheric CO_2 emissions have been reduced significantly with retrofit by up to 33% with respect to emissions of 18,679 kg/h for the base case.

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