Investigation of Heat Exchanger Network Flexibility of Distillation Unit for Processing Different Types of Crude Oil

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This paper summarizes the results of a study aimed to investigate the effect of the type of crude oil on the design of heat exchanger network of an atmospheric/vacuum distillation unit with maximum heat recovery, and to form a flexible heat exchanger network that is capable for processing all of the selected crude oils. Three different types (light – Crude A, medium – Crude B, heavy – Crude C) of crude oil were selected. Results showed that the heat recovery was the lowest (68.66 MW) and the external heat requirement was the highest (101.58 MW) in case of Crude A comparing to the other crude oils being investigated. It means that the processing of light crude oil is less energy efficient relative to medium and heavy ones. On the contrary the calculated heat exchanger area was the highest in Crude C (12,165 m^2). This crude oil contained the high boiling fractions of low overall heat transfer coefficient in relatively large amount increasing the required heat transfer area. Heat exchanger network that is applicable for processing each selected crude oil with maximum heat recovery also formed. Results showed that only 4 additional heat exchanger units is required to form a network that can be used in a flexible way for processing all of the selected crude oils. The total heat exchanger area of the flexible heat exchanger network is 15,238 m^2, which is higher than the value obtained for each selected crude oil separately. Economic viability of the flexible heat exchanger network should be calculated considering the investment and operating costs as well as the amount of the different types of crude oil to be processed annually.

1. Introduction

Petroleum refining industry faced several challenges in last years. The economic crisis forced companies to implement efficiency improving steps including reduction in the operating costs as Al-Mutairi (2010) presented for a FCC unit as well as Liang and Feng (2011) for a CCR unit. On the other hand, the political crisis that emerged in petroleum exporting countries necessitates the diversification of feed sources, which requires flexibility in the crude oil processing.

Kumar et al (2010) investigated the grass-root design of crude distillation unit using binary crude mixtures; however the heat exchanger network was not involved in the study. But the optimal work of heat exchanger network basically determines the energy efficiency of the crude distillation unit. Somjitt et al (2011) presented the efficiency improvement of an existing crude preheat train applying the relocation technique for single type crude oil. However, processing of different types of crude oil means the change in the optimal working point of heat exchanger network, too. Yimyam and Siemanond (2012) showed solutions for retrofit problem of crude preheat train with different types of crude oil. Varga et al (2010) investigated an existing crude oil distillation unit involving the heat exchanger network to determine the operation constrains for processing different types of crude oil. EL-Temtamy and Gabr (2011) presented the sensitivity table approach for flexibility problem of heat exchanger network. Verheyen and Zhang (2006) presented a superstructure-based MINLP model which minimizes the total annualized cost containing heat exchanger area cost and utility costs.

This paper summarizes the results of study to be performed to determine the optimal heat exchanger network of a grass-root atmospheric/vacuum distillation unit for processing three different types of crude oil, while ensuring the required product quality and applying the minimal utility consumption approach.

2. Methodology

The following steps were carried out during the study. At first the model of crude oil distillation system was prepared for each crude oil to determine the heat and material balances. Then the pinch technique was applied to prepare the heat exchanger network (HEN) system for maximum heat recovery in every case. Finally the HEN that can be applied in case of all selected crude oils for maximum heat recovery was prepared.

Figure 1 shows the simplified process flowsheet of the crude distillation system applied in the study, which contained the following parts: feed preheating line including the desalter, pre-flash tower for light product recovery, atmospheric furnace followed by the atmospheric tower, vacuum furnace followed by the vacuum tower and stabilizing tower for adjusting the initial boiling point of light naphtha. Crude oil was separated into fuel gas, LPG, light- middle- and heavy naphtha fractions, kerosene, atmospheric light and heavy gas oils, light and heavy vacuum gas oils, and vacuum residue.

Three different types of crude were selected in order to determine the effect of the crude quality on the heat exchanger network properties (area, pinch temperature, minimal cold and hot energy requirement etc.). The crudes were a light, a medium and a heavy type; their properties and yield of products obtained on processing them are summarized in Table 1.

Detailed model of the distillation system was prepared with PROII 8.3 process engineering tool. Crude stream was pre-heated before entering the pre-flash tower the temperature at inlet point of the tower was depended on the type of the crude oil. The pre-flash tower constitutes 16 theoretical stages and steam stripping is considered using steam at 250 °C and 11 bar. The pre-flash column produces light ends along with light naphtha product that is removed as the top product using a partial condenser and a side product that is drawn at stage 4 (stage numbering is from top to bottom).

The bottom product of the pre-flash tower enters the atmospheric furnace and undergoes a further vaporization. The pressure of the furnace is maintained at 2.4 bar. The atmospheric distillation column constitutes 38 theoretical stages and consists of a total condenser, a coupled side column and two pump-around circuits. The feed to the atmospheric distillation column enters on stage 32. Steam stripping that aids enhancement of volatility via the reduction of partial pressure is facilitated by using steam at 266 °C and 4.2 bar. The average pressure drop of the atmospheric distillation tower is taken as 0.16 bar with the first stage at 0.98 bar, and the bottom stage pressure of 1.14 bar. Two pump-around circuits are facilitated to provide internal reflux at various sections of the tower.
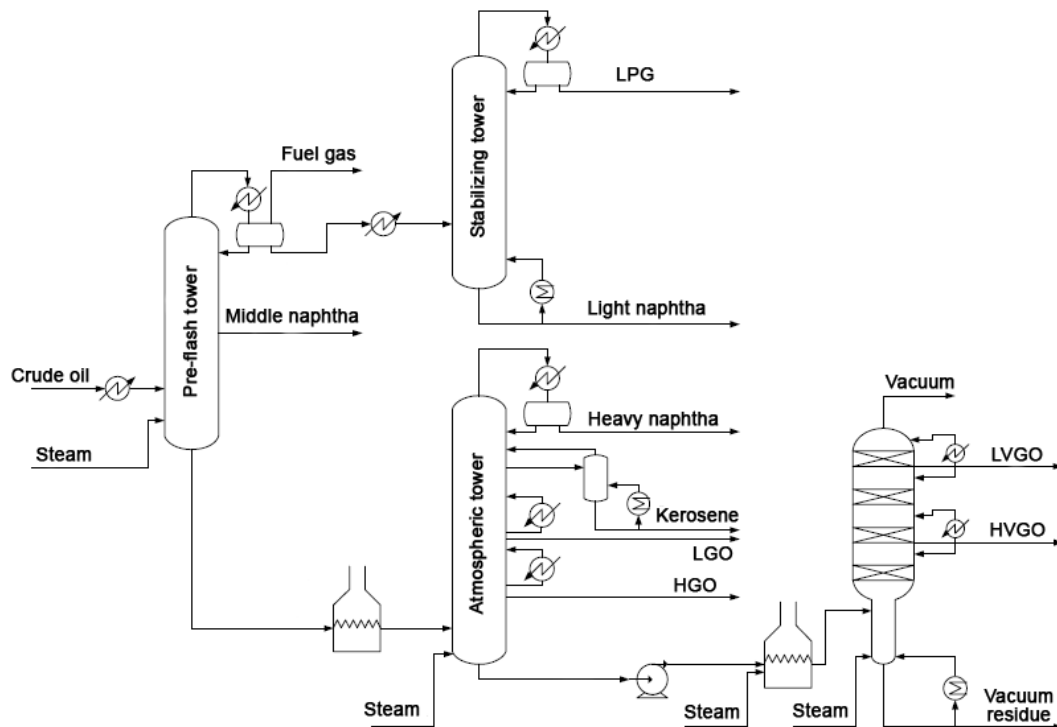

Figure 1: Simplified flowsheet of the crude oil distillation unit

Table 1: Properties of selected crude oils

Properties	Crude A	Crude B	Crude C
Density @15 °C, kg/m^3	819.0	864.3	874.0
Kinematic viscosity @20 °C, mm^2/s	3.90	12.7	17.4
Sulphur content, %	0.58	1.39	2.10
Nitrogen content, mg/kg	700	1631	1645
Ni + V content, mg/kg	8.5	42	95
Yield of products on the crude, %			
Fuel gas	0.02	0.04	0.02
LPG	2.08	1.30	0.67
Light naphtha	7.97	4.41	2.27
Middle naphtha	3.67	3.45	4.64
Heavy naphtha	13.69	7.28	7.88
Kerosene	9.54	7.67	8.22
Light atmospheric gas oil	18.04	14.24	17.11
Heavy atmospheric gas oil	9.04	6.67	7.91
Light vacuum gas oil	9.91	7.99	6.81
Heavy vacuum gas oil	15.62	28.83	22.3
Vacuum residue	10.42	18.13	22.16

The location of the first pump around is from stage 16 to stage 15, the second pump around is located from stage 24 to stage 23.

The side column consisting of 4 equilibrium stages is fed with liquid drawn from stage 11 of the main column. The lighter product produced in the side column is fed to stage 10 of the main column and the bottom product is the kerosene stream.

The bottom product of the atmospheric distillation column is fed to the vacuum distillation. The vacuum unit separates the atmospheric column bottom product into off-gas, light vacuum gas oil, heavy vacuum gas oil and residual oil. The vacuum column constitutes 13 theoretical stages with pressure maintained at 43 mbar (stage 1) and 73 mbar (stage 13). The process furnace associated to the vacuum column operates at a pressure of 0.31 bar. The vacuum tower is enabled with two pump-around circuits, with the first located from stages 3 to 1 and the second located from stage 9 to 7. Light vacuum gas oil is taken out from stage 3 as a total draw. The heavy vacuum gas oil is withdrawn from stage 9 and vacuum residue is taken as bottom product of the vacuum distillation unit.

The liquid product of the pre-flash tower is fed to the stabilizing tower, which separates the feed stream into LPG and light naphtha. The stabilizing tower constitutes 25 theoretical stages and consists of a total condenser and a reboiler. The first stage pressure is 9.5 bar and the bottom stage pressure is 9.94 bar and the top stage temperature is 66 °C.

Data of product and pump-around streams was used as input for the pinch analysis. The minimum temperature approach was selected to 20 °C. The preparation of composite curves, calculation of minimum heating and cooling requirements as well as design of the HEN was done on the work of Linnhoff and presented in everywhere (Kemp, 2007).

3. Results and discussion

Main steps of the calculation is presented on processing Crude B. Figure 2 shows the composite curves and Figure 3 the grand composite curve, respectively. Grand composite curve shows that the system has pinch point which divides the system into heat source and heat sink parts. The pinch temperature is 139.7 °C. The calculated minimum cooling requirement is 38.345 MW and the minimum heating requirement is 50.191 MW. Design of the HEN was performed by applying the basic rules established by Linnhoff and Hindmarsh (Kemp, 2007) for maximum energy recovery. Additionally, the operability of crude oil distillation system was considered, too. For example, we formed the HEN such a way to heat the reboiler of the stabilizing tower with only two adequate hot streams to avoid the use of complicated heat exchanger construction and to ensure easy temperature control of bottom section of the column. Figure 4 displays the HEN to be designed by following the before mentioned criteria.
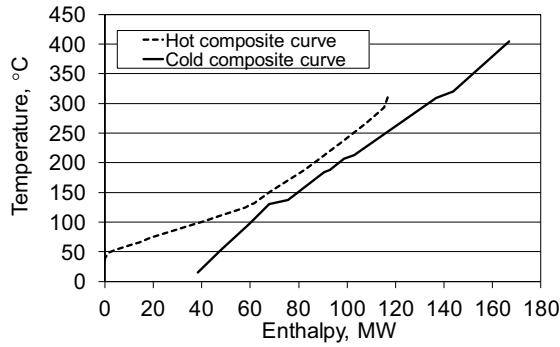
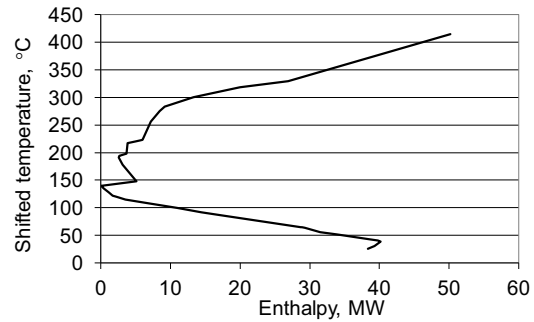*Figure 2: Composite curve obtained for Crude B*

Figure 3: Grand composite curve obtained for Crude B

HEN consists 35 heat exchanger units from which 25 process/process heat exchanger. The heat exchanger area also calculated. We applied different overall heat transfer coefficients depending on the type of hot and cold side streams. The calculated total heat exchanger area was 11,988 m^2 from which the process/process heat exchanger area was 9,865 m^2.

We applied the same algorithm in case of Crude A and Crude C for design the HEN. Most important results of these calculations are summarised in Table 2.

Pinch temperature being calculated for the selected crude oils showed an interesting view. Despite significant difference was in the product yield between Crude A (light) and Crude B (medium) the calculated pinch temperature was the same (139.7 °C). However, pinch temperature was considerable higher in case of Crude C comparing to former ones. This is probably due to the different yield structure of crude oils. It can be stated that the yield of vacuum residue is significantly higher in case of Crude C than that is in case of Crude B. The higher flow rate of the vacuum residue of high heat content may cause the shift of the pinch temperature into the higher temperature range.
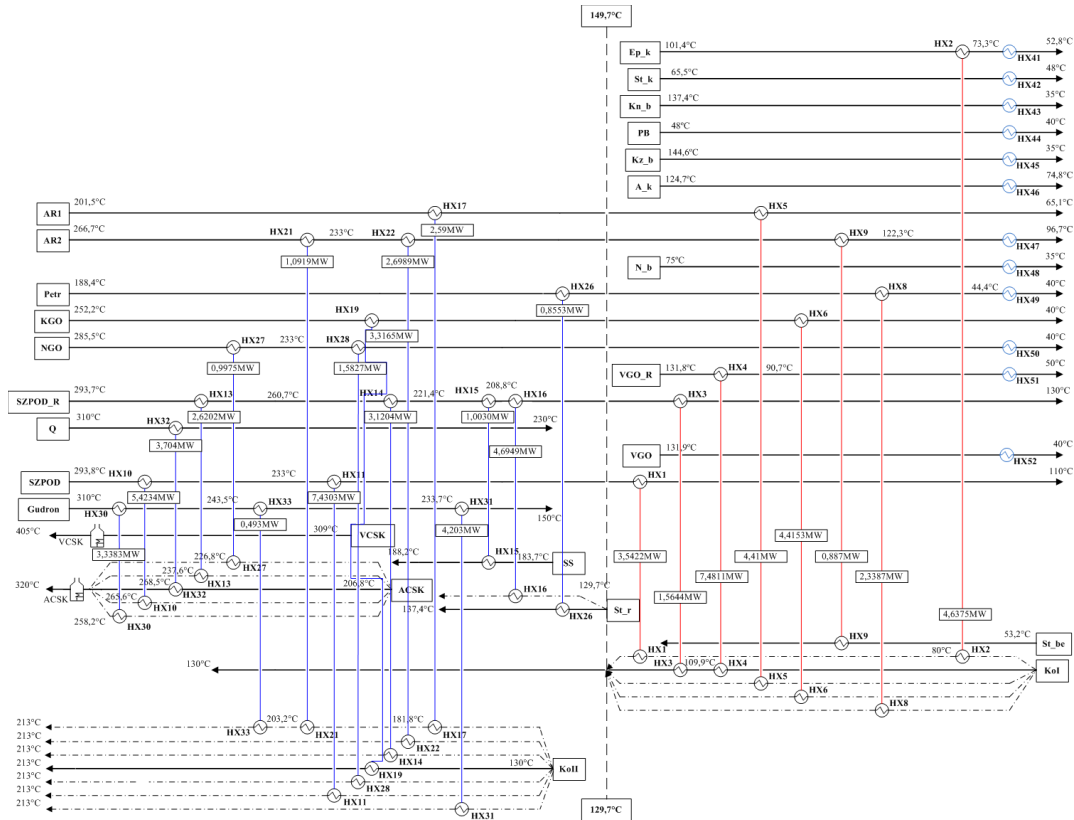

Figure 4: Heat exchanger network prepared in case of Crude B

Table 2: *Main results of heat exchanger network design for the selected crude oils*

Parameter	Crude A	Crude B	Crude C
Pinch temperature, °C	139.7	139.7	191.9
Minimum heating requirement Q_{Hmin}, MW	55.015	50.191	47.414
Minimum cooling requirement Q_{Cmin}, MW	46.564	38.345	36.603
Process/Process heat exchange, MW			
Below pinch point	30.215	29.274	47.077
Above pinch point	38.445	49.142	29.814
Heat duty of atmospheric furnace, MW	36.771	23.785	25.201
Heat duty of vacuum furnace, MW	18.320	26.325	22.208
Total number of heat exchangers	33	35	40
Process/process	24	25	29
Below pinch point	10	8	15
Above pinch point	14	17	14
Total heat exchanger area, m^2	9,003	11,988	12,165
Process/process	6,805	9,865	10,411
Below pinch point	3,310	3,018	5,933
Above pinch point	3,495	6,846	4,478
Area of coolers	2,199	2,133	1,753

Required minimum heating duty (Q_{Hmin}) decreased in the following order Crude A, Crude B and Crude C, respectively. This shows that the yield of the high boiling fractions is higher less external heat duty is required due to the increased heat recovery level. In case of Crude B and Crude C the high flow rate of high boiling fractions ensure the temperature difference between the hot streams (products and pump-arounds) and cold streams (crude oil) of the distillation unit.

The required heat duty is added in the atmospheric and vacuum furnaces. In respect of atmospheric furnace, data of Table 2 show that the highest duty required in case of Crude A. This is in accordance with the yield structure of Crude A, because it contains the atmospheric products in the largest volume and considerable amount of heat is required to evaporate them. In respect of vacuum furnace it can be seen that the highest heat duty is required on processing Crude B due to it contains the vacuum side products in the largest volume.

On processing Crude A the external cooling duty (Q_{Cmin}) was the highest. This is because on one hand the largest volume of light products, which should be cool down to near ambient temperature, is obtained in Crude A. On the other hand the light products leave the distillation towers at relatively low temperatures resulting small temperature differences between cold and hot streams, therefore heat content of these hot streams can only be utilized in small extent.

Summing up the results in case of light crude oil (like Crude A) the energy efficiency of distillation unit is lower comparing to processing heavier crude oils due to the lower heat recovery.

Number of heat exchangers depended on the type of crude oil being processed. Total number of units varied in the range of 33 – 40. Number of coolers varied in the narrow range (e.g. 9-11) and it was practically independent from the type of crude oil. Number of process/process heat exchanger units increased from the light (Crude A) to heavy (Crude C) crude oils, which correspond to the heat to be recovered.

With respect to the calculated heat exchanger area the following statement can be made. The total heat exchanger area increased in order Crude A, Crude B and Crude C, respectively. Area of process/process heat exchangers relative to the total area increased in the same order corresponding to the attainable heat recovery. The heat exchanger area was also influenced by the amount of the heavier products, because the applied overall heat transfer coefficients were lower for the high boiling streams compared to those applied for low boiling ones. For example, we applied 170 W/m^2°C for crude oil/vacuum residue heat exchange and 340 W/m^2°C for crude oil/middle distillate (kerosene, light gas oil) heat exchange. Ratio of heavy products is significantly higher in case of Crude B and Crude C comparing to Crude A, which also contributes to the higher heat exchanger area to be calculated on processing the former crude oils.

The higher heat exchanger area means higher investment cost, too. This partially offsets the advantage of the higher heat recovery observed in case of Crude B and Crude C.

We designed the heat exchanger network that is applicable for processing each selected crude oil with maximum heat recovery. By comparing the heat exchanger networks to be established for processing the selected crude oils we determined that only 4 additional heat exchanger units is required to form a network that can be used in a flexible way for processing all of the selected crude oils. The total heat exchanger

area of the flexible heat exchanger network is 15,238 m^2, which is higher than the value obtained for each selected crude oil separately. However, the flexible network is applicable for processing crude oils of different composition with maximum heat recovery. Economic viability of the flexible heat exchanger network should be calculated considering the investment and operating costs as well as the amount of the different types of crude oil to be processed annually.

The potential CO_2 emission reduction being reached by the energy integration was also calculated for each selected crude oil. We considered a natural gas fuelled fired heaters and 90% furnace efficiency. The CO_2 emission changed in the following for Crude A: 24.2 t/h → 10.8 t/h, for Crude B: 25.2 t/h → 9.8 t/h, for Crude C: 24.4 t/h → 9.3 t/h, respectively. Results showed that more than 50 % reduction in the CO_2 emission can be achieved by applying energy integration.

4. Conclusions

Nowadays, the question of the feed flexibility and the energy efficiency comes to the front in the refining industry. Processing of different types of crude oil at maximum heat recovery which is also the lowest fuel cost is inevitable for the long term profitability. One of the possible solutions to achieve this goal is the design of flexible heat exchanger network which is applicable for processing different type of feeds.

Our object was to investigate the effect of the crude oil composition on the heat exchanger network of atmospheric and vacuum distillation unit. Heat exchanger network for maximum heat recovery was prepared for each selected crude oil (light – Crude A, medium - Crude B, heavy – Crude C). Results showed that the heat recovery was the lowest (68.66 MW) and the external heat requirement was the highest (101.58 MW) in case of Crude A comparing to the other crude oils being investigated. It means that the processing of light crude oil is less energy efficient relative to medium and heavy ones.

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Heat exchanger network that is applicable for processing each selected crude oil with maximum heat recovery also formed. Results showed that only 4 additional heat exchanger units is required to form a network that can be used in a flexible way for processing all of the selected crude oils. The total heat exchanger area of the flexible heat exchanger network is 15,238 m^2, which is higher than the value obtained for each selected crude oil separately. Economic viability of the flexible heat exchanger network should be calculated considering the investment and operating costs as well as the amount of the different types of crude oil to be processed annually.

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