Rigorous Procedure for the Design of Conventional Atmospheric Crude Fractionation Units. Part II: Heat Exchanger Network

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This paper is the second of a two-part series. In the first part, the optimal condenser and pump-around duties were determined for two types of crude, one light and one heavy. These crudes constitute the targets for the design of a multipurpose heat exchanger network. A model for the design of such a network is presented in this paper. The conjecture that this network can be used as a design that can handle crudes of intermediate density is also tested.

Introduction

The relevance, historical background, and previous work on the subject of this paper were presented in Part I.[1] In the first part, targets for duties of the condenser and the pump-around circuits for a crude fractionation unit were provided. These targets correspond to an optimal arrangement of these duties, so that the overall energy consumption is minimized. This second part attempts to design a multipurpose heat exchanger network capable of handling both crudes.

A heat exchanger network that is optimal for one type of crude is not necessarily optimal or even feasible for another crude. Thus, the problem consists of determining the best heat exchanger network that accommodates the needs of maximum efficiency for a variety of crudes at the same time. To design such a network, targets corresponding to two crudes in the extremes of a range of density are used. The conjecture tested in this paper is that a design that satisfies the maximum efficiency needs of these two extreme crudes, that is, the very light and the very heavy, can also be efficient in processing all of the crudes with densities in between.

Extensions of well-known heat exchanger network synthesis models (transshipment and others) to address this problem will be presented first. The energy targets that will be used to design the heat exchanger network are determined by the procedure proposed in the first paper of this series.

Multipurpose/Multiperiod HEN Model

As mentioned previously, refineries or chemical plants must have the flexibility to process different feedstock, so the heat exchanger network (HEN) must also be designed to absorb such changes and maintain the operating costs close to the optimum. Then, the problem reduces to the design of a HEN that has the capability of handling the same sets of hot and cold streams with different flow rates and different inlet and outlet temperatures in different periods of time. Floudas and Grossmann[2] presented an extension of the transshipment model to multiperiod operation. Their work was based on the following three assumptions: (1) Each heat exchanger has the same pair of hot/cold streams. (2) Each heat exchanger can handle variable heat loads. (3) Those pairs of hot/cold streams that exchange heat above and below the pinch points in different periods will have different exchangers in each subnetwork.

The first assumption is related to cost containment. For the second assumption to hold, a bypass is necessary. The third assumption is fully related to the duplication of exchangers when matches occur above and below the pinch. This assumption is not really required because bypass arrangements are possible.[3] By using this model, the authors obtained matches and heat loads for each operation period so that the HEN design could be subsequently developed by hand.

In a subsequent paper, Floudas and Grossmann[4] presented the automatic generation of the HEN by using the superstructure model. Papalexandri and Pistikopoulos[5] also solved the problem using "hyperstructures" in which all possible heat exchange matches are considered. The resulting model is a mixed integer nonlinear programming problem (MINLP).

It is well-known that the design of crude fractionation units presents the engineer with a problem in which pinch analysis exhibits some difficulties. As we shall see, many crudes exhibit a region of temperature in which the temperature approaches between the hot and cold composite curves are fairly similar. This region has been called a "pinch region", or a "continuous pinch", instead of a pinch point. To overcome these difficulties, it is necessary to use models in which the heat recovery is fixed by picking a value of the heat recovery minimum approximation temperature (HRAT) and the design is performed by allowing a smaller temperature approach in the exchangers. This smaller approach has been named the exchanger minimum approach temperature (EMAT).

We present here a model based on the transshipment model proposed by Papoulias and Grossmann[6] and the global vertical heat transfer with the dual temperature approach (DTA) by Gundersen and Grossmann.[7] Our focus is to concentrate on the determination of what could be a universal design of crude units to handle design flexibility.

We first define the different sets:

$$\text{Cold streams: } C = \{\text{C1, C2, CW}\}$$

Stream C1 represents the raw crude with the water added. Stream C2 is the crude feed to the column.

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$$\text{Hot streams: } H = \text{H1} \cup \{\text{Furnace}\}$$

H1 = {kerosene, diesel, AGO, residue, naphtha, sour water, condenser, PA1, PA2, PA3}

$$\text{Intervals: } N = \{T_0, T_1, ..., T_N\}$$

$$\text{Crudes: } P = \{\text{light, heavy}\}$$

Objective Function. As the targeting already provided the minimum heating utility, the objective function is to maximize the vertical heat transfer. This has been shown to produce the lowest overall network area.

$$\mathbf{(P1)} = \min\{\alpha^{\mathrm{L}}\sum_{i\in H}\sum_{j\in C}S_{i,j}^{\mathrm{L}} + \alpha^{\mathrm{H}}\sum_{i\in H}\sum_{j\in C}S_{i,j}^{\mathrm{H}}\} \quad (1)$$

The parameters α correspond to the fraction of the year that each crude is processed. The superscripts L and H identify the light and heavy crude sets, respectively. The variable S represents the amount of heat vertically transferred between hot and cold streams. The number of exchangers is controlled through a constraint.

Constraints. We first start with the transshipment constraints. The reader is referred to the original paper that presents the transshipment model[7] for diagrams and thorough explanations of the variables.

The following corresponds to the heat balances in hot streams without the furnace:

$$R_{i,THs_i}^k + \sum_{j\in C} V_{i,j,THs_i}^k = WH_{i,THs_i}^k \quad i \in \text{H1},\ k \in P \quad (2)$$

$$R_{i,T}^k - R_{i,T-1}^k + \sum_{j\in C} V_{i,j,T}^k = WH_{i,T}^k \qquad i \in \text{H1},\ T \in N,\ k \in P,\ T \neq THs_i \quad (3)$$

The furnace needs a separate set of constraints. It is important to note that the load of the furnace is fixed by targeting. In addition, it is confined to the intervals with the highest temperatures.

$$R_{F,THs_F}^k + \sum_{j\in C} V_{F,j,THs_F}^k = Q_F^k \quad k \in P \quad (4)$$

$$R_{F,T}^k - R_{F,T-1}^k + \sum_{j\in C} V_{F,j,T}^k = 0 \qquad T \in N,\ k \in P,\ T \neq THs_F,\ T \neq THe_F \quad (5)$$

$$-R_{F,THe_F-1}^k + \sum_{j\in C} V_{F,j,The_F}^k = 0 \quad k \in P \quad (6)$$

The following constraints correspond to the transshipment heat balances of the cold streams. One important feature is that the heat capacity of the crude is determined as a function of temperature intervals, which allows a more rigorous way of representing the heating of the crude through the heat exchanger network.

$$\sum_{i\in H} V_{i,j,TCs_j}^k = FC_j^k Cp_{j,TCs_j}^k \Delta T_{TCs_j} \quad j \in \{\text{C1,C2}\},\ k \in P \quad (7)$$

$$\sum_{i\in H} V_{i,j,T}^k = FC_j^k Cp_{j,TCs_j}^k \Delta T_{TCs_j} \qquad j \in \{\text{C1,C2}\},\ T \neq TCs_j,\ k \in P \quad (8)$$

The cooling utility is limited to the last temperature interval:

$$\sum_{i\in H} V_{i,CW,TCs_j}^k = Q_{CW}^k \quad k \in P \quad (9)$$

$$\sum_{i\in H} V_{i,CW,T}^k = 0 \quad T \in N,\ k \in P,\ T \neq TCs_j \quad (10)$$

The existence of a match between hot and cold streams is defined by the following well-known constraint:

$$\sum_{T\in N} V_{i,j,T}^k - \Gamma_{i,j}^k Y_{i,j} \leq 0 \quad i \in H,\ j \in C,\ T \in N,\ k \in P \quad (11)$$

where $\Gamma_{i,j}^k$ is larger than the maximum of the heat loads of hot stream i and cold stream j for crude k. The number of units is limited to reduce the number of alternative solutions and also to reduce cost.

$$\sum_{i\in H}\sum_{j\in C} Y_{i,j} \leq N^* \quad i \in H,\ j \in C \quad (12)$$

The constraint for vertical heat transfer is

$$\sum_{T\in N} V_{i,j,T}^k - S_{i,j}^k = Qv_{i,j}^k \quad i \in H,\ j \in C,\ k \in P \quad (13)$$

Because $Qv_{i,j}^k$ is the maximum possible vertical heat transfer between streams i and j, $S_{i,j}^k$ represents the unrealized vertical heat transfer. Finally, nonnegativity constraints and others are

$$R_{i,T}^k = 0 \quad i \in H,\ T = T_N \quad (14)$$

$$V_{i,j,T}^k \geq 0,\ Y_{i,j} = \{0,1\},\ R_{i,T}^k \geq 0 \qquad i \in H,\ j \in C,\ T \in N,\ k \in P \quad (15)$$

The above model is a multiperiod extension of the model presented by Gundersen and Grossmann,[6] which maximizes the vertical heat transfer constrained by a prespecified maximum number of heat exchangers. If this maximum number of heat exchangers is too small, the problem can become infeasible. To relax this condition, and only for numerical purposes, the furnace utility is added to the objective function, so that infeasible problems will be recognized easily for exhibiting larger utility usage. The model is solved for a value of the EMAT smaller than the HRAT used for targeting. To avoid areas growing too large, the EMAT was chosen to be 50% of the HRAT for the small HRAT (11.1 °C) and to be 75% of the HRAT for larger values (22.2 and 44.4 °C). In practice, it is suggested that an HRAT of 40−50 °C with a rather smaller EMAT (10 °C) should be used.

Gundersen and Grossmann[7] pointed out some of the deficiencies of the global vertical model. In particular, they pointed out that "multiple matches can be required when a match expands across an enthalpy interval temperature", of which the most significant interval temperature is the pinch. In our results, multiple matches are not present, so this is not a problem. They also warned that this global model "cheats", because "the same amount of heat may be assumed transferred vertically to more than one cold stream". Because we have only one cold stream in each interval here, this does not constitute a problem in our case either. In addition, they showed that the EMAT is not an optimization variable and pointed out that, although an EMAT of zero could be used, in industrial practice, a value larger than 5 °C should be used. Finally, they pointed

Table 1. Stream Data Sets for Light (L) and Heavy (H) Crude

stream	CP (MW/°C) L	CP (MW/°C) H	T_{in} (°C) L	T_{in} (°C) H	T_{out} (°C) L	T_{out} (°C) H
H1 (kerosene)	0.0720	0.0240	182.8	214.4	21.1	21.1
H2 (diesel)	0.0391	0.0388	262.2	264.4	21.1	21.1
H3 (AGO)	0.0720	0.0170	323.9	281.7	21.1	21.1
H4 (condenser)	0.4335	0.1594	143.3	120.6	43.3	21.1
H5 (PA1)	0.3427	0.1347	169.4	163.3	104.4	104.4
H6 (PA2)	0.4396	0.0897	254.4	254.4	177.8	171.1
H7 (PA3)	0.1123	0.0959	310.6	308.3	232.2	232.2
H8 (residue)	0.1539	0.4433	348.3	349.4	260.0	260.0
H9 (naphtha)	0.1002	0	43.3	–	21.1	–
H10 (sour w.)	0.0939	0.0943	104.4	104.4	21.1	21.1
C1	0.4668	0.5011	21.1	21.1	104.4	104.4
C2	*a*	*a*	104.4	104.4	359.4	354.4

[a] See Table 2.

Table 2. CP = $f(\Delta T)$ for Light and Heavy Crude

C2 (light) T (°C)	C2 (light) CP[a]	C2 (heavy) T (°C)	C2 (heavy) CP[a]
104.4	0	104.4	0
121.1	0.4038	121.1	0.4289
148.9	0.4264	148.9	0.4568
176.7	0.4710	176.7	0.4821
204.4	0.5247	204.4	0.5006
232.2	0.5712	232.2	0.5211
260.0	0.6063	260.0	0.5428
287.8	0.6281	287.8	0.5663
315.6	0.6440	315.6	0.5927
359.4	0.7528	354.4	0.6954

[a] CP = MW/°C

Table 3. Minimum Utilities and Pinch Points (HRAT = 11.1 °C)

feedstock	QH (MW)	QW (MW)	pinch (°C)
light	53.3	41.6	254.4−243.3
heavy	76.8	3.8	32.2−21.1

out that this model works well when the same heat transfer coefficient can be assumed for all streams, and they offered suggestions regarding how to handle the problem otherwise. Although there are some disparities between the film transfer coefficients for some streams, like the condenser and the disposed water, one can consider that the rest of the streams, being so close in nature (hydrocarbons) and phase (liquid), have the same film coefficient. Fortunately, these streams have an impact in temperature regions that are away from the pinch, where the influence of criss-cross heat transfer on area is not that significant. We consider this issue as a refinement that will be tackled in future work.

Illustration

We use the data presented in Part I[1] for a plant processing 795 m^3/h (120000 bpd). First of all, a value for the heat recovery approach temperature (HRAT) and the information about temperatures and heat capacity flow rates from Part I are used. Table 1 shows the data. Heat capacity flow rates as a function of temperature intervals are presented in Table 2. Minimum utility consumption and pinch temperatures for a HRAT value of 11.1 °C (20 °F) are shown in Table 3. Pinch diagrams for light and heavy crudes are shown in Figures 1 and 2, respectively.

The regular transshipment model (RTM)[6] was applied to both light and heavy crudes, above and below the pinch for the same HRAT (11.1 °C). The results show

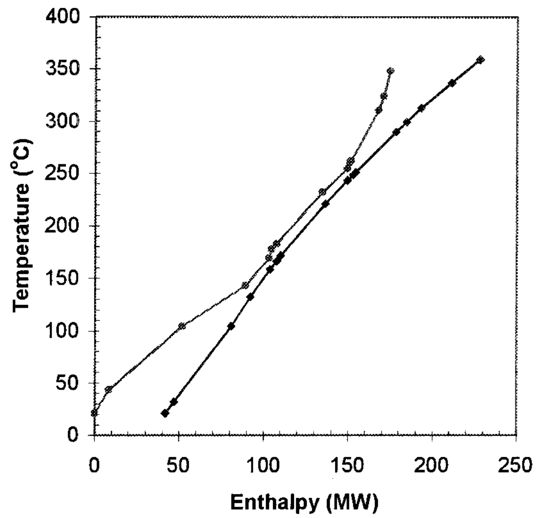

Figure 1. Pinch diagram for light crude.

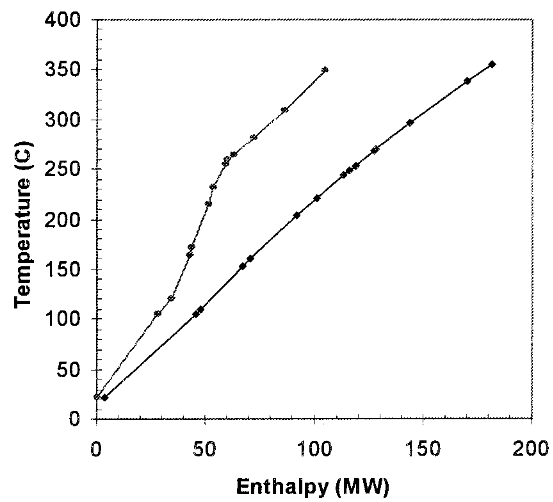

Figure 2. Pinch diagram for heavy crude.

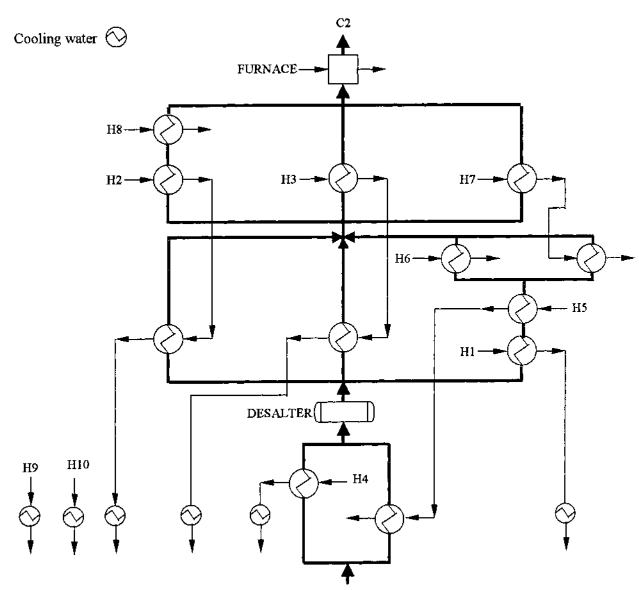

Figure 3. RTM applied to light crude.

that, for the light crude, a network with 18 exchangers is required (Figure 3), whereas for the heavy crude, 15 exchangers are needed (Figure 4). The first observation is that the network for the light crude can perform the

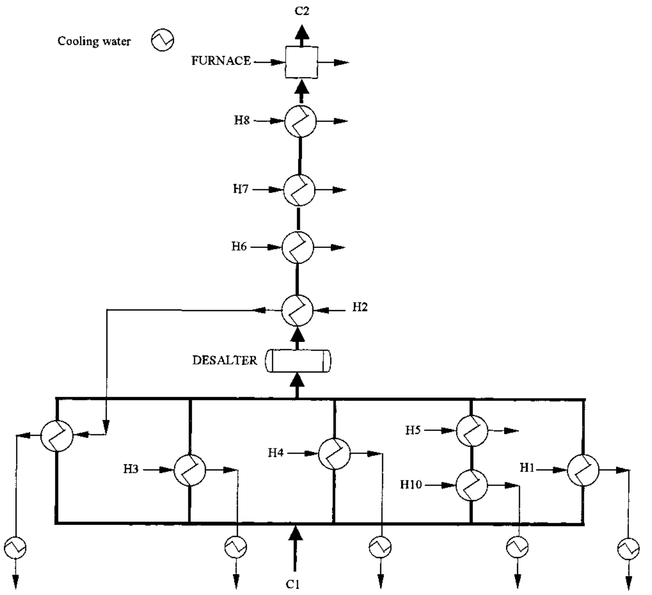

Figure 4. RTM applied to heavy crude.

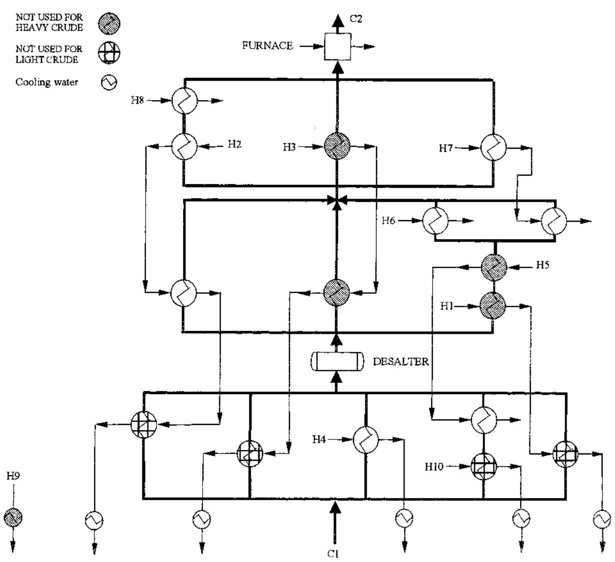

Figure 5. Multiple HEN obtained by merging Figures 3 and 4.

heat transfer of the network of the heavy crude above the desalter but cannot handle it efficiently below the desalter, and vice versa. If one merges both networks, the resulting structure is very complicated and features 22 exchangers (Figure 5). Note that, even though the network for the heavy crude above the desalter does not contain splits, the light crude structure can still be used. Next, we use model **P1** to reduce the number of exchangers and simplify the network structure.

The multiperiod model **P1** was solved assuming that the number of heat exchangers is smaller than a prespecified value N^*. By reducing N^* by one exchanger at a time, one can find the minimum number of heat exchangers that satisfies the energy targets for the light and heavy crude cases simultaneously. The model was solved using GAMS.[8] Figure 6 shows the heat exchanger network obtained using an EMAT of 5.6 °C (10 °F). This network has 20 units, and neither loops nor bypasses exist. However, it is necessary to split the crude stream into four and five branches above and below the desalter, respectively. Areas are presented in Table 4, and costs are compared with the network obtained using the RTM model in Table 5. There is one heat exchanger that requires a large area (H6−C2) because it is the one that transfers heat in the middle region of the light crude composite curves where they are almost parallel (Figure 1).

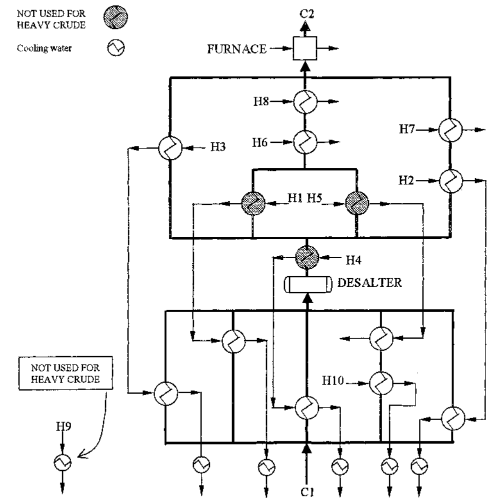

Figure 6. Heat exchanger network for HRAT/EMAT = 11.1/5.6 °C.

Table 4. Areas for HRAT/EMAT = 11.1/5.6 °C

unit	area (m^2)	% of total
H1−C1	691	1.3%
H2−C1	1021	1.9%
H3−C1	481	0.9%
H4−C1	3303	6.2%
H5−C1	1.778	3.4%
H10−C1	1623	3.1%
H1−C2	1831	3.4%
H2−C2	673	1.3%
H3−C2	6936	13.1%
H4−C2	647	1.2%
H5−C2	4953	9.3%
H6−C2	21264	40.1%
H7−C2	1294	2.4%
H8−C2	2688	5.1%
H1−CW	866	1.6%
H2−CW	472	0.9%
H3−CW	869	1.6%
H4−CW	987	1.9%
H9−CW	466	0.9%
H10−CW	227	0.4%

Table 5. Area and Costs[a] for HRAT = 11.1 °C

	combined RTM	multiperiod model
total area, m^2	45499	52959
no. of shells	57	64
operating costs, 10^6 \$/year	4.18	4.18
fixed costs, 10^6 \$/year	3.13	3.63
total costs, 10^6 \$/year	7.31	7.81

[a] Cost data: fuel gas = \$6.83/(MW h), cooling water = \$1.2287/(MW h), steam cost = \$1.76/ton. Installed cost per shell = $1168.5A^{0.65}$ (A in m^2), interest = 10%, plant life =15 years.

In comparing this solution with the one obtained using the RTM model, one finds that the required area is increased by about 16% (Table 5) and that the number of units is reduced (from 22 obtained for the RTM to 20) but that the number of shells is larger. The operational costs are the same because there is no difference in the energy consumption of the two networks. The difference between the total annualized costs is smaller (7%) than the difference in area.

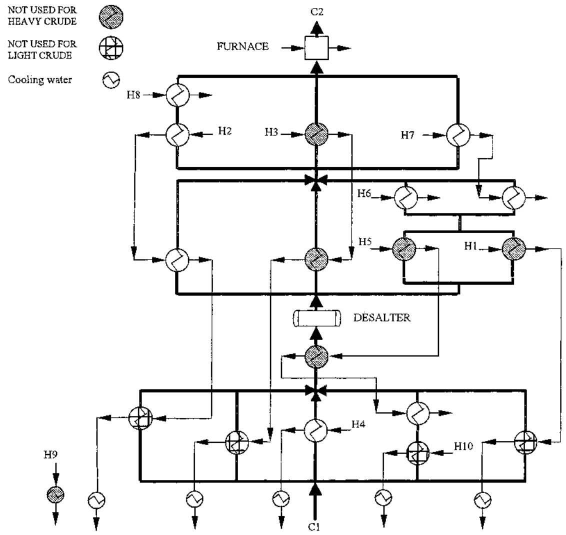

Figure 7. Network obtained using RTM and a high desalter temperature.

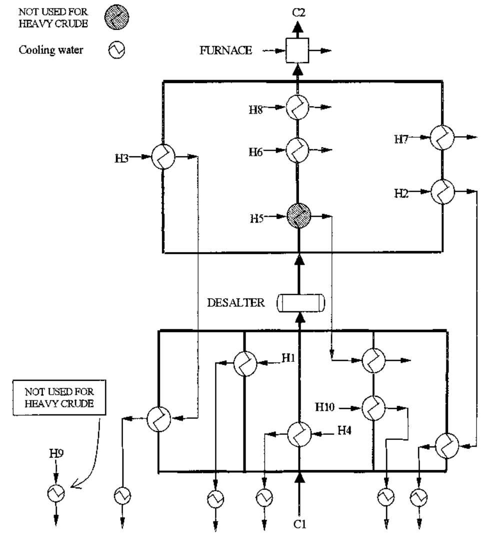

Figure 8. Solution for HRAT/EMAT = 11.1/5.6 °C and a high desalter temperature.

The selected value of the HRAT resulted in a large amount of area, which is impractical. More important, the number of shells needed is unrealistic. Consequently, the HRAT was changed to 22.22 °C (40 °F) and 44.44 °C (80 °F). At the same time, the EMAT was changed to 16.66 °C (30 °F) and 33.33 °C (60 °F), respectively. Before the impact of the HRAT/EMAT changes is shown, the role of the desalter temperature and the pump-around flexibility is discussed.

Desalter Temperature

As already discussed, while the light crude controls the network structure above the desalter, the heavy one does the same below it. Consequently, if the desalter temperature is increased for the light crude, one should expect a decrease in the network area above the desalter, as the light crude can now use the excess area below the desalter, which is there to serve the heavy crude heat recovery. In turn, increasing the desalter temperature for the heavy crude will require more area and matches because the region of temperatures below the desalter is limiting for this crude.

The previous RTM and multiperiod designs (Figures 5 and 6) used a desalter temperature of 104.4 °C (220 °F) for both crudes. This temperature was increased to 137.8 °C (280 °F) only for the light crude set. Applying these changes, the RTM design gives the network of Figure 7. In turn, the solution of problem **P1** gives the network structure shown in Figure 8, which has two exchangers, less than the structure in Figure 6. Furthermore, the required area is lower and compares well with the value obtained using the RTM, as shown in Table 6. On the other hand, the RTM not only gives a very complicated network structure above the desalter but also indicates the same below the desalter. As we can see in Figure 7, the RTM renders a network that is even more complicated than the one shown in Figure 5. The multiperiod model not only renders a smaller cost but also renders fewer shells. *In conclusion, the higher possible temperature in the desalter should be used for the light crude, while the lower one should be selected for the heavy crude, so that the minimum network area is achieved*.

Pump-Around Flexibility

As discussed in Part I,[1] some degrees of flexibility in handling the load of the different pump-around circuits exist. For example, in the case of the light crude, one can take some surplus heat from pump-around 1 (the one above the demand curve in Figure 7 of Part I) and return it to the condenser. (See Figure 9.) In doing so, one would be just shifting heat from one cooler to another hoping that one cooler could be eliminated. However, the network obtained (Figures 6 or 8) does not use any cooling water for PA1 (stream H5), and therefore no exchanger can be eliminated by the proposed shift in the energy load distribution. Only the steam consumption could be decreased by this change. Such a shift, although beneficial, represents a small variation and is not explored any further.

Economic Comparison for Different HRAT/EMAT Values

Because the number of shells is excessive for an HRAT of 11.1 °C, the effect of changing the HRAT to 22.2 °C (40 °F) and 44.4 °C (80 °F) was studied. EMAT values of 16.7 °C (30 °F) and 33.3 °C (60 °F), respectively, were used. As pointed out above, smaller values of the EMAT can be used (around 10 °C), especially for large values of the HRAT. Each time an HRAT value is selected, the energy targets should be determined again because the heat load distribution throughout the column pump-around circuits changes. Not only does the minimum utility (which is not even a linear function of the HRAT as in straight pinch analysis because the heat capacities of the streams change) change, but the pump-around duties are also modified, and consequently different flow rates are obtained. In addition, a high temperature in the desalter was used for the light crude 137.8 °C (280 °F). The solutions are shown in Figures 10 and 11, while the total areas, numbers of shells, and costs are shown in Table 7. Tables 8 and 9 show the area distribution. As we can see, the solution in Figure 10 has 18 exchangers (40 shells). Although it has the same number of exchangers as the solution in Figure 8, the total area and costs are much lower. When the HRAT/EMAT is increased (Figure 11), the number of exchangers is reduced by 1 and the number of shells by

Table 6. Area and Costs for HRAT = 11.1 °C and Different Desalter Temperatures

desalter temperature	104.4 °C (220 °F)		137.8 °C (280 °F)	
	combined RTM	multiperiod model	combined RTM	multiperiod model
total area, m^2	45499	52959	47882	48218
no. of shells	57	64	62	60
operating costs, 10^6 \$/year	4.18	4.18	4.18	4.18
fixed costs, 10^6 \$/year	3.13	3.63	3.34	3.32
total costs, 10^6 \$/year	7.31	7.81	7.52	7.50

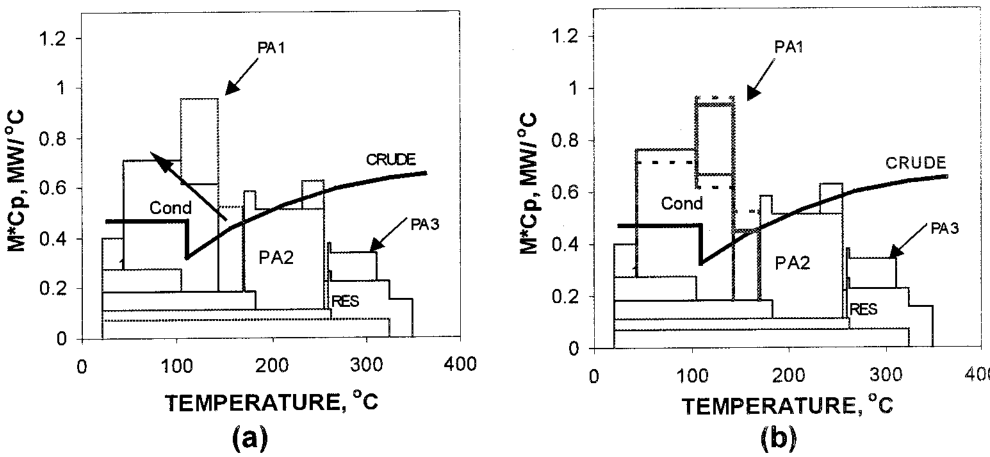

Figure 9. Moving heat back to the condenser.

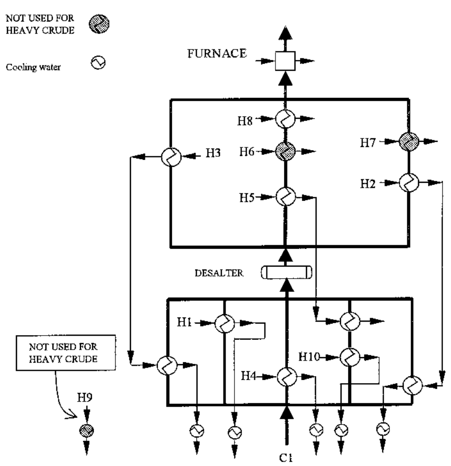

Figure 10. Heat exchanger network for HRAT/EMAT = 22.2/16.7 °C.

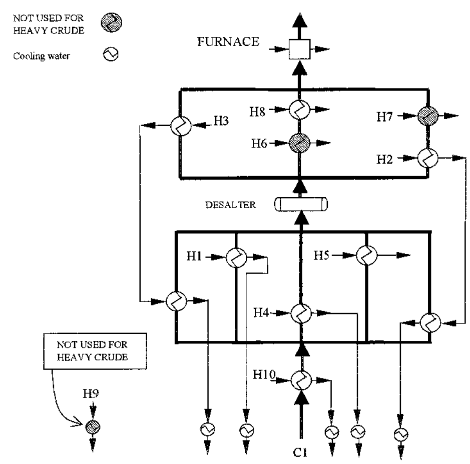

Figure 11. Heat exchanger network for HRAT/EMAT = 44.4/33.3 °C.

11, while the total cost remains lower than those of the previous cases. To obtain an even lower capital cost, one can attempt to use a higher HRAT value. The exercise is the same, and we do not repeat it here.

Universal Heat Exchanger Network

The designs shown above address an energy-efficient scheme for only two crudes. The conjecture is that it can also handle any crude of intermediate density. To test this conjecture, simulations of the entire system using an intermediate crude were performed. Details of this crude are shown in Tables 10 and 11. Minimum heating and cooling utilities for HRAT = 22.2 °C are 61.1 and 16.9 MW, respectively. Those values were obtained during the analysis made in Part I of this paper.

To test the ability of the network of Figure 10 to process this crude at maximum heat efficiency, model **P1** was run with the integer variables set corresponding to the matches to one. The desalter temperature was varied until a solution with the minimum energy consumption (61.1 and 16.9 MW, respectively) was obtained. However, the required area for some heat exchangers was higher than the one calculated before, especially below the desalter. This is shown in Table 12. Therefore, the conjecture that the design obtained can eventually handle an intermediate density crude at maximum energy efficiency (minimum utility) is confirmed only in relation to the network structure. Adjustments in the area of some heat exchangers are still needed. These adjustments amount to seven additional shells, corresponding to an additional 6210 m^2 and 420000 \$/year of additional fixed cost. Quite clearly, the

Table 7. Area and Costs for Different HRAT/EMATs

HRAT/EMAT (°C)	22.2/16.7	44.4/33.3
total area, m^2	28470	18485
no. of shells	40	29
operating costs, 10^6 \$/year	4.51	5.13
fixed costs, 10^6 \$/year	2.04	1.37
total costs, 10^6 \$/year	6.55	6.50

Table 8. Areas for HRAT/EMAT = 22.2/16.7 °C

unit	area (m^2)	% of total
H1−C1	980	3.5%
H2−C1	891	3.2%
H3−C1	631	2.2%
H4−C1	2729	9.7%
H5−C1	2772	9.9%
H10−C1	662	2.4%
H2−C2	1044	3.7%
H3−C2	3304	11.8%
H5−C2	1717	6.1%
H6−C2	6891	24.5%
H7−C2	1881	6.7%
H8−C2	1690	6.0%
H1−CW	608	2.2%
H2−CW	328	1.2%
H3−CW	604	2.1%
H4−CW	926	3.3%
H9−CW	286	1.0%
H10−CW	156	0.6%

Table 9. Areas for HRAT/EMAT = 44.4/33.3 °C

unit	area (m^2)	% of total
H1−C1	1290	7.1%
H2−C1	448	2.5%
H3−C1	577	3.2%
H4−C1	1109	6.1%
H5−C1	2504	13.7%
H10−C1	265	1.5%
H2−C2	626	3.4%
H3−C2	1499	8.2%
H6−C2	3769	20.7%
H7−C2	1497	8.2%
H8−C2	1528	8.4%
H1−CW	525	2.9%
H2−CW	346	1.9%
H3−CW	644	3.5%
H4−CW	1164	6.4%
H9−CW	285	1.6%
H10−CW	166	0.9%

Table 10. Stream Data Set for Medium Crude

stream	CP (MW/°C)	T_{in} (°C)	T_{out} (°C)
H1 (kerosene)	0.0534	182.2	21.1
H2 (diesel)	0.0360	272.2	21.1
H3 (AGO)	0.0279	302.8	21.1
H4 (condenser)	0.3130	144.4	43.3
H5 (PA1)	0.2300	170.6	104.4
H6 (PA2)	0.3032	260.0	173.9
H7 (PA3)	0	−	−
H8 (residue)	0.3191	353.9	260.0
H9 (naphtha)	0.0580	43.3	21.1
H10 (sour w.)	0.0924	137.8	21.1
C1	0.4915	21.1	137.8
C2	*a*	137.8	360.0

[a] See Table 11.

Table 11. $C_P = f(\Delta T)$ for Medium Crude

C2	
T (°C)	C_P (MW/°C)
137.8	0
148.9	0.4405
176.7	0.4622
204.4	0.4963
232.2	0.5275
260.0	0.5628
287.8	0.5973
315.6	0.6252
360.0	0.7171

medium crude has a higher residue stream than the light crude and sufficient duty in the pump-around circuits, so that all of the structure prepared for the light crude above the desalter cannot handle the heat loads.

Table 12. Area Differences for Medium Crude, HRAT = 22.2 °C

unit	area Figure 10 (m^2)	A required (m^2)
H1−C1	993	2714
H2−C1	903	1830
H3−C1	640	1400
H4−C1	2765	4722
H8−C2	1712	2545
H10−CW	158	168

A similar situation takes place below the desalter. A new conjecture emerges: If the multiperiod model **P1** is used for the three crudes, the structure obtained would be able to accommodate the processing of crudes with densities in between. The possible outcomes are (a) marginal increments of furnace heat duty if additional area is not added or (b) the same efficiency if additional area is added.

Discussion

Several aspects in the design of heat exchanger networks for crude distillation units have been addressed. First, by using the regular transshipment model, one can obtain the minimal energy consumption network. However, the level of complexity required to achieve this energy consumption is not practical. By using the method presented in this paper, the designer can merge exchangers above and below the pinch so that the complexity is reduced, although the required area increases.

Second, the desalter temperature can be modified to optimize the area available in the multipurpose network. Because the light crude controls the area above the desalter and the heavy crude does the same below it, one should select the highest desalter temperature for the light crude and the lowest temperature for the heavy crude. This will minimize the area in each section of the network.

Third, to decrease the network area, one can choose a higher HRAT value and allow the energy consumption to increase. This also reduces the number of exchangers, the number of shells, the network complexity, and the total cost. On the other hand, the complexity can be addressed by forcing the network to have a prefixed number of splits in the crude stream and allowing the energy consumption to increase accordingly. This approach was studied in a separate paper.[9]

Finally, several of the observations made throughout this two-part paper are of relatively high value for retrofit scenarios. In principle, one should attempt to obtain the right heat distribution of pump-around heat loads, which can be obtained using the procedure from Part I,[1] to obtain a first operational target. This has been found to be beneficial to operational costs even without changes in the preheating train, because usually heat is transferred to PA2 or PA3, thus representing greater differences of temperature in existing exchangers and therefore better heat recovery. The discussions about the temperature of the desalter for each crude, as well as the return temperatures of pump-around circuits and the flexibility of shifting loads are also valuable information for retrofitting. Once the operational targets are obtained, one can attempt to adapt some of the existing methods to retrofit heat exchanger networks to obtain a solution that will maximize the energy efficiency of both crudes, subject to a reasonably short payout.

Conclusions

The design of a crude distillation unit based on the minimum energy consumption can be addressed by following the step-by-step methodology proposed in this work. The energy targets obtained by the demand–supply diagrams can be used to develop a heat exchanger network that fits those operating conditions for different feedstocks. The model takes advantage of the flexibility identified and is able to predict the desalter temperature that must be used. The conjecture that a design for extreme crudes would be able to accommodate the processing of an intermediate crude is only partially confirmed. Quite possibly, the design needs to be performed for the three crudes simultaneously, in which case the cost will rise, but the flexibility of the network to handle crudes of densities between those used for design can be improved. This approach will be explored in future work. It is possible that, by using smaller EMAT values, the number of exchangers obtained will not be smaller, but the risk of having larger areas will increase. This is another feature left for future work. Finally, targeting for the number of shells instead of units and the more complicated problem of retrofitting will be considered in future work.

Nomenclature

Parameters

α^L, α^H = Fraction of the year each crude is processed
$Cp_{j,T}^k$ $j \in C$ = Heat capacity of cold stream j in interval T, crude k
$FC_{j,T}^k$ $j \in C$ = Flow rate of crude streams
$WH_{i,T}^k$ $i \in H1$ = Heat load of hot stream i in interval T
TCs_j $j \in C$ = Interval at which cold stream j starts
TCe_j $j \in C$ = Interval at which cold stream j ends
N^* = Maximum number of heat exchangers
$\Gamma_{i,j}^k$ = Maximum amount of heat that can be transferred from stream i to stream j, for crude k
$Qv_{i,j}^k$ = Maximum heat that can be transferred vertically between streams i and j
ΔT_i = Interval of temperature i

Variables

E = Total energy consumption
$R_{i,T}^k$ = Heat surplus from hot stream i cascaded down to the next interval
$V_{i,j,T}^k$ = Heat transferred from hot stream i to cold stream j in interval T
U = Minimum utility usage in the heat exchanger network
$Y_{i,j,T}^k$ = Match between i and j in interval T in plant k
$S_{i,j}^k$ = Vertical heat transfer from stream i to stream j, for crude k
ΔT_T^k = Temperature change of interval T in plant k
H_S^k = Enthalpy of the stripping steam as a function of PA heats
Q_F^k = Furnace load for crude k
Q_{CW}^k = Heat removed by cooling water

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