HEAT EXCHANGER NETWORKS FOR PETROLEUM FRACTIONATION UNITS HANDLING CRUDES OF DIFFERENT DENSITY

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Abstract

The problem of designing crude fractionation units is not only a distillation design. They have the added complexity that they should be able to process different types of crude at different times. These crudes may vary in density from heavy to light. This paper presents a summary of a recently presented rigorous methodology to design such multipurpose plant. First, a targeting procedure implemented using a commercial simulator is presented. The design of a heat exchanger network that will take advantage of the flexibility in the targets follows. Extensions to well-known heat exchanger network synthesis models (e.g. transshipment) to address this target flexibility are presented.

1. Introduction

Crude distillation is energy intensive. It consumes an equivalent of 2% of the crude processed. The conventional design used these days, consisting of a column with side strippers and pump-around circuits was first suggested 70 years ago (Miller, 1938), and still dominates in the refining industry. The current conventional design with pump-around circuits and side-strippers was formalized by Watkins (1979), who also discussed a few variants such as pump-back reflux and stripping using reboilers and provided an empirically based design procedure. There also exist some designs where a prefractionation column is used. In this paper we concentrate on the design of Figure 1.

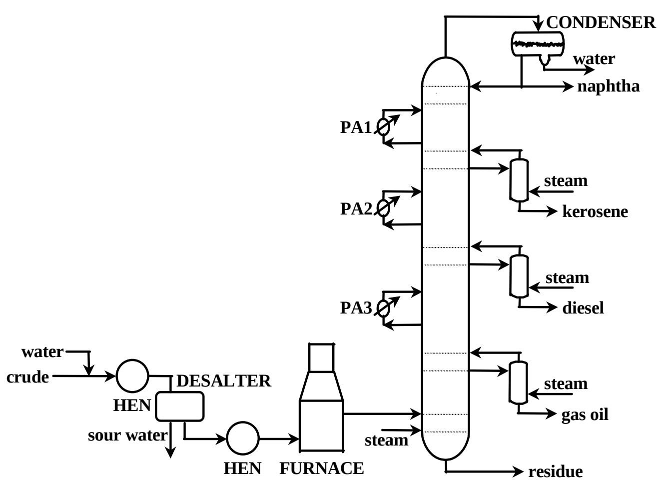

Figure 1: ***Conventional Crude Distillation***

Compared to common distillation of discrete components, crude distillation units have several unique features. They process large quantities of raw material, they are energy intensive, no reboilers are normally used, and the products are characterized by certain ASTM assays. However, the most distinctive one is the fact that these units process a variety of crudes depending on the situation of the market. The methodologies that are very effective in solving discrete component mixture problems seem inapplicable and not helpful. Nevertheless, the introduction of pseudo components made crude distillation no longer a black box and many advances were made in the simulation of these systems. However, very little has been proposed in the field of their design since Watkins (1979).

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Recently, Liebmann and his coworkers (1995, 1998) proposed an integrated design procedure in which the column, the heat exchanger network (HEN) and their interactions are considered simultaneously. They decompose a complex column into an equivalent sequence of simple columns and propose to design each simple, non thermal-coupled columns. Lastly, they merge the sequence of columns into a complex column. In contrast, addressing the column as a whole, as it is done in this work, helps determining the effect of each variable and their interaction. In addition, *crude units are multipurpose by nature: they process different crudes. Therefore it is desired that the optimal design should be such that maximum energy recovery is obtained for each case.*

Such design procedure rendering a structure suitable for the processing of different crudes each at maximum energy efficiency was recently presented in a two part series paper (Bagajewicz and Soto, 2001, Bagajewicz and Ji, 2001). This paper is a summarized version of those results.

2. Method

2.1 Heat Demand-supply Diagram

Heat supply-demand diagrams are a natural extension based on the concept of energy density-temperature diagrams, which are abundant in the literature (Hohmann, 1971; Huang and Elshout, 1976; Andrecovich and Westerberg, 1985; Terranova and Westerberg, 1989). In an energy density-temperature diagram, a stream is represented by a curve. This curve represents the product of mass flowrate and specific heat capacity (true or apparent in the case of phase changing streams). Areas below the heat demand curve but not covered by the heat supply curve represent the minimum heating utility of the unit. Such a heat demand-supply diagram is shown in figure 2.

2.2 Pump-around Circuits and Heat Recovery

The original purpose of adding pump-around circuits was to reduce vapor and liquid traffic at the top section of the column (Watkins, 1979). Without pump-around circuits, all condensation heat has to be taken out from the condenser, which results in a large vapor flowrate at the top trays. We now explore the limits of how much heat can be transferred from the condenser to the pump around circuits.

Consider k pump around circuits. It can be shown using simple balances that the maximum heat that the pump around circuits can remove is given by:

$$\sum_k Q_k = (\sum_i P_i + L_0)\Delta H_v \tag{1}$$

where P_i, is the mass flowrate of a draw above the section of trays of the pump around circuit, L_0, is the mass flowrate of the liquid leaving the pump around circuit section and ΔH_v, is the average heat of vaporization in the section.

Thus, several configurations or levels of heat consumption can be obtained by shifting heat from the condenser to the pump around circuits, or from one pump around circuit to the lower one. It is well known that this shifting of heat causes some deterioration of the quality of the separation. First, the presence of the pump-around circuit decreases the number of effective ideal trays (Bagajewicz, 1998). The effect can be even more detrimental to separation, if the flowrate of a pump-around circuit is increased. As we shall see later, these effects can be compensated to a certain extent by increasing the steam rate in the side strippers. Another alternative is to increase the number of trays. However, with the total number of trays kept constant, the relation between heat recovery and steam consumption can be incorporated into the design procedure.

2.3 Column Design Procedure

We now summarize the technique for designing a multipurpose energy efficient atmospheric column. First, the total number of trays is picked and the Watkins design method is used to get an initial scheme. Then the heat supply-demand diagram is created and analyzed to determine the direction of heat shifting needed for maximum energy efficiency. This procedure is repeated for at least the lightest crude and the heaviest crude that will be processed. Thus, the design procedure is divided in two parts, the targeting procedure and the multipurpose heat exchanger network design. Typically, since the light crudes are the ones that need larger reflux, they will exhibit larger amount of pump-around circuits. After these targets are determined, there is still some flexibility to move heat from one pump-around to another. This feature is useful in the final design.

2.4 Multipurpose Heat Exchanger Network Design Procedure

The targets that were obtained in the column design can be now used to develop a multipurpose heat exchanger network (HEN), which is capable of operating at the optimum whenever it processes a light or a heavy crude. The design procedure is mainly based on the well-known transshipment model proposed by Papoulias and Grossmann (1983). However, some modifications to achieve the multipurpose HEN were introduced. These modifications are the following:

1) The objective function is defined as the summation of the total heating utility plus the steam consumption. The number of matches is included as a constraint.
2) The crude stream is divided in sub-streams, and heat can be "cascaded" upwards similarly to hot streams.
3) The solution must satisfy the targets obtained for the heavy and light crude cases simultaneously. To achieve this goal, the matches between the streams must be at the same temperature intervals.

The proposed model can be written in the following way:

$$\min E = \alpha_L[Q_F^L + 0.7 * H_S^L] + \alpha_H[Q_F^H + 0.7 * H_S^H]$$

s.t. (Model 1)

Heat balances in hot streams

$$\left.\begin{aligned} & R_{i,T}^k - R_{i,T-1}^k + \sum_{j\in C} V_{i,j,T}^k = WH_{i,T}^k \\ & R_{F,T}^k - R_{F,T-1}^k + \sum_{j\in C} V_{F,j,T}^k = \theta_{F,T}^k \\ & \sum_{T\in N} \theta_{F,T}^k = Q_F^k \end{aligned}\right\} \quad \begin{aligned} & \mathbf{i \in H, T \in N} \\ & \mathbf{k \in P} \end{aligned}$$

Heat balances in cold streams

$$\left.\begin{aligned} & D_{j,T}^k - D_{j,T+1}^k + \sum_{i\in H} V_{i,j,T}^k = WC_{j,T}^k \\ & D_{C3,TCs_{C3}+1}^k = D_{C2,TCe_j}^k \end{aligned}\right\} \quad \begin{aligned} & \mathbf{j \in C, T \in N} \\ & \mathbf{k \in P} \end{aligned}$$

Heat is allowed to be transferred from the sub-stream C2 to sub-stream C3.

Cold streams:

$$FC_j^k Cpc_{j,T}^k \Delta T_T = WC_{j,T}^k \qquad \mathbf{j \in \{C1,C2\},\ T \in N,\ k \in P}$$

Steam consumption:

$$H_S^k = a * Q_{COND}^k + b * Q_{PA1}^k + c * Q_{PA2}^k + d * Q_{PA3}^k + e$$

$\mathbf{k \in P}$

Matches in intervals:

$$\left.\begin{aligned} & V_{i,j,T}^k - U * Y_{i,j,T}^k \le 0 \\ & Y_{i,j,T}^L = Y_{i,j,T}^H \\ & \sum_{j\in C} Y_{i,j,T}^L \le N_T \end{aligned}\right\} \mathbf{i \in H, j \in C, T \in N, k \in P}$$

The last equation limits the amount of matches in interval T.

Total number of matches:

$$\left.\begin{aligned} & \sum_{T\in N} Y_{i,j,T}^L - U * Z_{i,j} \le 0 \\ & \sum_{i,j} Z_{i,j} \le N^* \end{aligned}\right\} \quad \mathbf{i \in H, j \in C, T \in N}$$

$V_{i,j,T}^{k} \geq 0$, $Y_{i,j,T}^{k} = \{0,1\}$, $Z_{i,j} = \{0,1\}$, $R_{i,T}^{k} \geq 0$,
$D_{i,T}^{k} \geq 0$

$i \in H$, $j \in C$, $T \in N$, $k \in P$

2.5 Illustration

We now illustrate the column design procedure for the light crude. There are 34 trays in the main column and 4 trays in each stripper. The flowrates of stripping steams are estimated and adjusted to 10 LB per barrel product, as suggested by Watkins. Total energy consumption is calculated according to:

$$E = U + 0.7 * \sum H_i^s \qquad (2)$$

where U is the minimum heating utility of this system, determined using straight pinch analysis, and $\sum H_i^s$ is the summation of enthalpies of all steam consumed. As low-pressure steam is cheaper than fuel gas on the same amount of heat base, the steam term is multiplied by a weight factor of 0.7. Following, all the heat possible is transferred from the condenser to the pump around circuit one. Then heat is transferred from a higher pump around to a lower one. In this process the value of the energy consumption keeps lowering until, at a certain point in the transferring of heat to the third pump around circuit, the steam consumption takes over. Table 1 shows the effect of the transferring of heat in the energy consumption. This procedure is repeated for the heavy crude. The final schemes for both crudes are shown in figure 2 and figure 3. Note that in the heavy crude design, the heat transfer stops in the first pump around circuit.

*Table 1. Effect of Pump-around Duties on Energy Consumption**

PA1 Duty	PA2 Duty	PA3 Duty	Energy Consumption (MMBTU/hr)
0	0	0	282.53
71	0	0	278.03
27	50.5	0	278.07
27	25.5	25	276.94

units: MMBTU/hr

The distribution of cooling duties among pumparounds has great impact on the design of heat exchange network. This means theoretically that each individual crude requires a specific heat exchange network to get the optimal savings. In a multipurpose design, it is necessary to sacrifice individual efficiency to achieve overall efficiencies. This means designs for light or heavy crudes may be shifted a bit from the optimum.

To illustrate the design procedure for the heat exchanger network, the original transshipment model (Papoulias and Grossmann, 1983) was used to get the minimum number of units for the light and heavy crude targets separately. The results of this model indicate that a HEN with 18 heat exchangers can handle the light crude and 16 heat exchangers will be necessary for the heavy one. In both cases, the minimum heating utility was used.

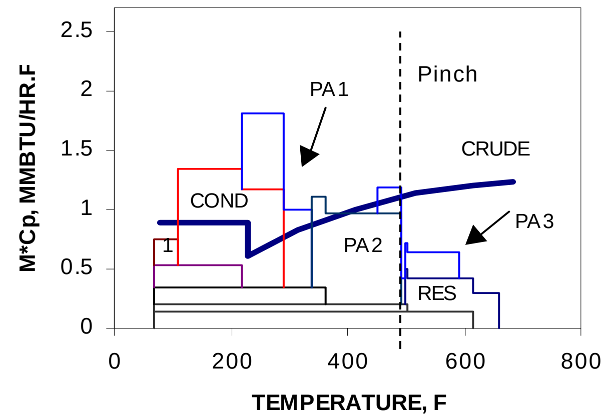

Figure 2: *Heat Demand-Supply Diagram for the Light Crude*

1: Naphtha and condensed water. COND: condenser. PAi: pump-around i. RES: residue. Products: from top to bottom are kerosene, diesel and gas oil.

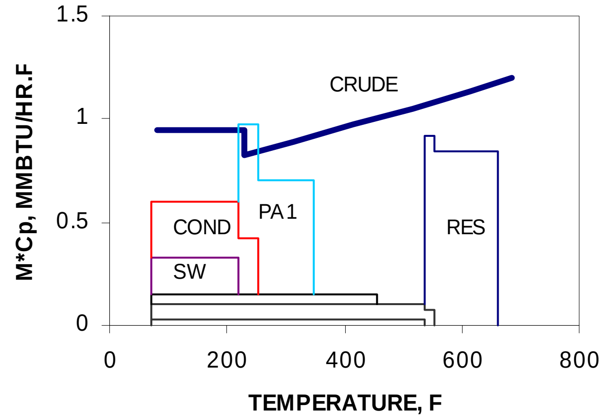

Figure 3: *Heat Demand-Supply Diagram for the Heavy Crude*

Once the minimum values for Q_F are known, one can use the proposed model to find the matches between the hot and cold streams that satisfy both sets of conditions simultaneously. The result is shown in figure 4. The HEN that was obtained is capable of processing the light or the heavy crude at the optimum conditions. The dotted exchangers above the desalter represent those exchangers that do not transfer any heat when the heavy crude is being processed. In a similar way, the stripped exchangers below the desalter represent those exchangers that do not operate when the light crude is processed. There is an alternative solution in which, almost all the heat exchangers can be in operation at the optimum condition too.

3. Conclusions

In this paper, a rigorous design procedure has been proposed for the design of conventional crude distillation units. This procedure is an improvement over the existing procedures for several reasons. First, this procedure aims at finding the best scheme for a multipurpose crude distillation unit that processes a variety of crudes. Second, the procedure is more straightforward and complex calculations are avoided without losing accuracy. Heat demand-supply diagrams, instead of grand composite curves are used as a guide directing the search for optimal schemes. An apparent advantage of heat demand-supply diagram is that the role of each stream, heater or cooler in the total energy consumption is clearly shown. Third, the approach is rigorous. The trade off between different operating parameters is considered and decision is based on quantitatively calculations instead of simple assumptions.

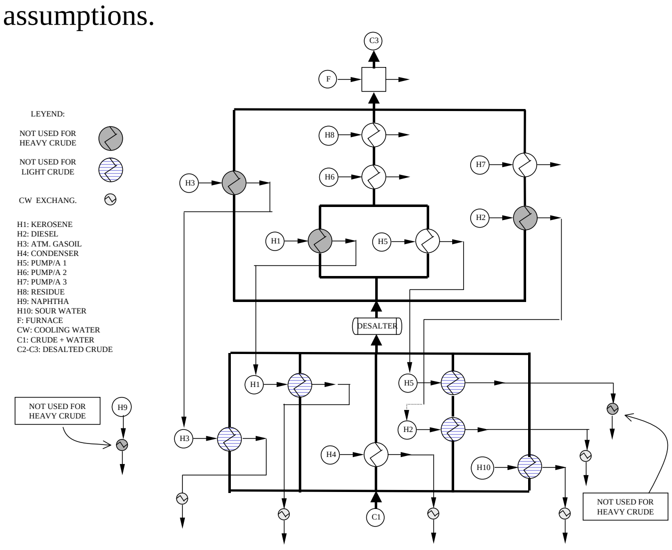

Figure 4: *Multipurpose Heat Exchanger Network*

On the other hand, a multipurpose HEN can be obtained by modeling the targets corresponding to the different types of crudes that can be processed. In this way, the common solution satisfies the optimum operation condition of the distillation column under the different scenarios.

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Nomenclature

Sets:

$C = C1 \cup C2 \cup \{C3\} \cup \{CW\}$

$C1 = \{C1\}$

$C2 = \{C2\}$

$H = H1 \cup H2 \cup \{F\}$

$H1 = \{$Kero, Dies, Ago, Res, Naph, Sour-water$\}$

$H2 = \{$Cond, PA1, PA2, PA3$\}$

$N = \{T_0, T_1, \ldots, T_N\}$

$P = \{$Light, Heavy$\}$

Parameters:

α_L, α_H : Fraction of the year each crude is processed.

$Cp\zeta^k_{j,T}$ $j \in C$: Heat capacity of cold stream *j* in interval T

$WH^k_{i,T}$ $i \in H1$: Heat load of hot stream *i* in interval T. Used only for some streams.

$WC^k_{j,T}$ $i \in CW$: Heat load of cold stream *j* in interval T. Used only for some streams.

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_T : Maximum amount of heat exchangers in parallel in interval T.

N^* : Maximum amount of heat exchangers.

U : Maximum amount of heat that can be transferred.

Variables:

Q^k_F : Furnace Load

$R^k_{i,T}$: Heat surplus from hot stream i cascaded *down* to the next interval.

$D^k_{j,T}$: Excess heat demand from cold stream j cascaded *up* to the next interval.

$V^k_{i,j,T}$: Heat transferred from hot stream *i* to cold stream j in interval T.

ΔT^k_T : Temperature change of interval T in plant k.

$Y^k_{i,j,T}$: Match between i & j in interval T in plant k.

$\theta^k_{F,T}$: Heat load of the furnace stream in interval T.

$\theta^k_{CW,T}$: Heat load of the cooling water stream in interval T.

H^k_S : Enthalpy of the stripping steam as a function of PA heats.