Simultaneous Optimization Approach for Heat Exchanger Network Retrofit with Process Changes

J. Zhang† and X. X. Zhu*

Department of Process Integration, UMIST, P.O. Box 88, Manchester M60 1QD, England

This paper addresses the problem of a heat exchanger network (HEN) retrofit considering process changes. Traditionally it is a common practice to fix process conditions before a HEN is retrofitted. However, it has been realized that there exists strong interactions between HENs and processes, and it should simultaneously consider the changes to process conditions (e.g. flow rates, temperatures) and the HEN retrofit.In this paper, new insights for process changes and their impacts on HEN retrofit have been developed and the $T-H$ diagram is used to represent the interactions. On the basis of the newly developed understanding, a systematic method has been developed to deal with process changes and HEN topological changes in a simultaneous way. To demonstrate this, a short-cut crude distillation model has been built up that relates stream temperatures and flow rates to process conditions. This model has been combined with the model for HEN topological changes proposed by Zhu and Asante,[1] which forms a complete model for optimizing the tradeoff between HEN retrofit and process changes. The case study shows significant cost savings by considering process changes and HEN modifications simultaneously.

1. Introduction

A heat exchanger network (HEN) is commonly used in the process industries. As an example, Figure 1 is used to represent a typical combination of a process and a HEN. The feed is required to be heated to a certain temperature before being charged to main processing units. Products serve as heat sources to transfer heat to the feed, which is usually in the form of liquid. In some cases, intermediate streams within the process also need to be vaporized or condensed (e.g. through reboilers or condensers).

Traditionally, HEN retrofit is considered separately from process modifications. In other words, before modifying a HEN, supply temperatures, target temperatures, and flow rates of streams are fixed. On the basis of these fixed conditions, more surface areas are added or the necessary structural changes are made to meet the retrofit target (Tjoe and Linnhoff;[2] Floudas et al.). However, process changes sometimes may help to ease the task of HEN retrofit. For example, consider the composite curves for a process shown in Figure 2a, in which a reboiler is located at a pinch point. If the reboiler pressure could be reduced, it could relax the pinch point (Figure 2b). Then the cold curve could be shifted closer to the hot composite curve until two curves are pinched again (Figure 2c). This results in the reduction in both hot and cold utility consumption.

In general, no matter what changes are made to a process, there are mainly two effects on the HEN. The first effect is the changes to both stream supply temperatures and target temperatures. These temperatures normally change with process operating conditions and configurations. The second effect is the changes to stream flow rates. Different feeds may result in different product yields and thus product flow rates. Different product specifications can also cause changes to stream flow rates.

* Corresponding author. E-mail: F.Zhu@umist.ac.uk.
† Current Address: AspenTech Ltd., Birchwood, Warrington, WA3 7RB, United Kingdom.

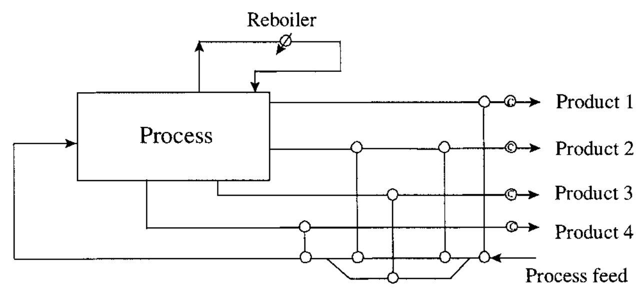

Figure 1. Typical structure of process and HEN.

Apart from these two effects on streams in the HEN, other influences also exist. For example, both stream heat capacities and film heat transfer coefficients may vary with stream temperatures. However, these impacts on heat recovery are relatively small and they can be ignored compared with changes in temperatures and flow rates.

The aim of this work is to develop a fundamental understanding of process changes to the HEN, and exploit interactions between process changes and HEN modifications. The improved understanding is then used as a basis to develop a new systematic method to deal with these two aspects simultaneously.

2. Review of Previous Work

General Guideline for Process Changes. There are few publications addressing the HEN retrofit problem by changing process conditions. One of them is by Linnhoff and Parker,[4] who proposed a method, the so-called plus/minus principle, to address the effect of process changes on utility consumption, which can be explained in Figure 3. The plus/minus principle dictates that increasing the heat load of a hot stream and decreasing the heat load of a cold stream above the process pinch reduce the hot utility consumption. In contrast, decreasing the heat load of a hot stream and increasing the heat load of a cold stream below the process pinch reduce the cold utility consumption.

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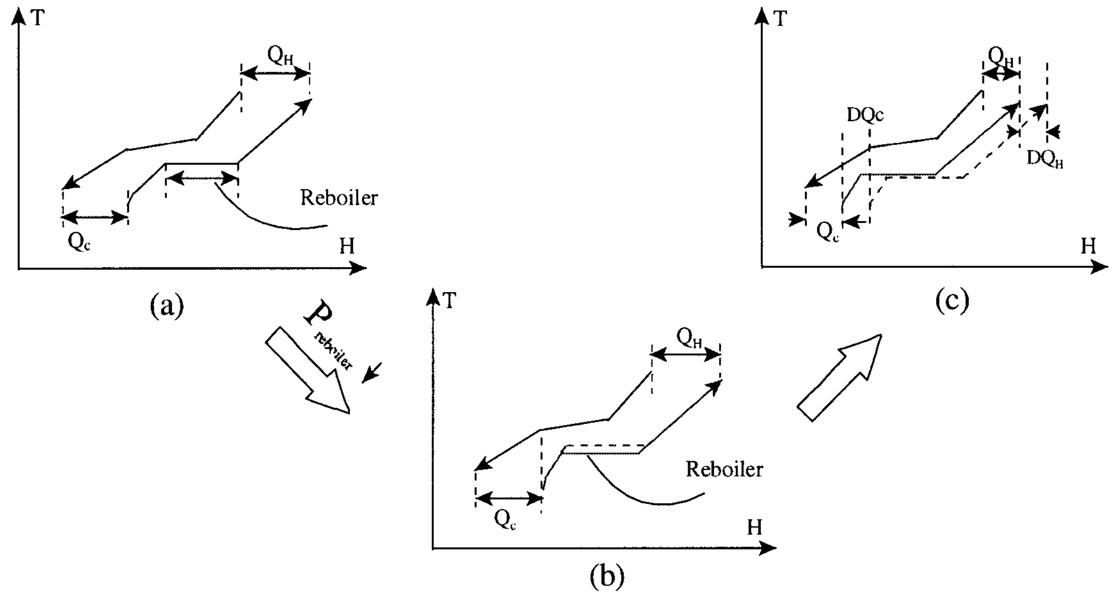

Figure 2. Benefits to the HEN from process changes.

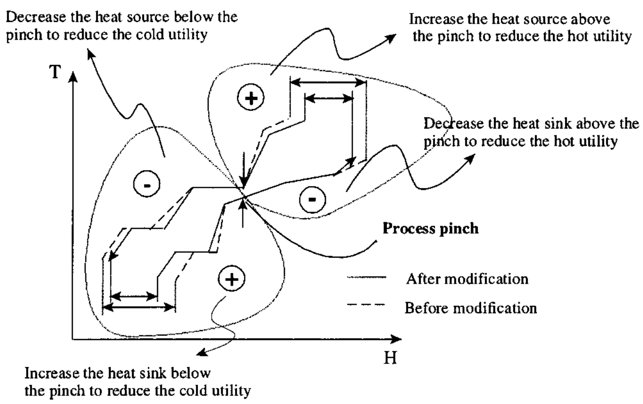

Figure 3. Plus/minus principle.

Following this rule, temperatures and flow rates of process streams can be adjusted to minimize the total utility consumption. However, the plus/minus principle uses the process pinch as the reference point, which is more applicable in grass-roots design. As a result, the plus/minus principle cannot address constraints in an existing HENs. Therefore, although the plus/minus principle provides insights to process changes, it is not suitable for use in retrofit scenarios.

Duran and Grossmann[5] also proposed an optimization model for considering stream temperature and flow rate changes in process synthesis. In this approach, however, the existing network constraints have not been considered and therefore it is only suitable for the grass-roots design. Although Lang et al.[6] and Grossmann et al.[7] proposed a modified model, they are still in the scope of flow sheet synthesis rather than process retrofit.

Heat Exchanger Network Retrofit. Methods proposed for the HEN retrofit design may be divided into two groups. The first group of methods are those based on “pinch techniques” (e.g. Toje and Linnhoff;[3] Polley et al.;[8] Shookoya and Kotjabasakis;[9] Carlsson et al.[10]). Although this class of approaches places the designer in control throughout the design process, the manual nature of the pinch methods means that the design process can be time-consuming and tedious and the designer must be conversant with design methods. The other group of methods is based on the use of mathematical programming techniques (e.g. Jones et al.;[11] Saboo et al.;[12] Ciric and Floudas;[13] Yee and Grossmann[14]). The common feature of these approaches is that the retrofit problem is formulated into an optimization problem, in which a general superstructure is constructed including all possible changes to the existing network. However, the general superstructure could contain too many changes for a large retrofit problem, many of which may be infeasible and impractical. This could create too large an optimization problem to be solved with too many combinations of modification options.

To overcome the drawbacks of these two groups of methods while combining their advantages, Asante and Zhu[15] recently introduced the concept of network pinch. The network pinch reveals the actual bottlenecks in an existing HEN and indicates the heat recovery limit, which cannot be relaxed by adding new areas to the existing HEN. This concept is used as the basis for developing a decomposition approach which consists of several MILP models. This approach significantly reduces the number of possible structural changes to be considered and results in a much smaller optimization problem.

In all the above methods, however, HEN retrofit is considered after process conditions are fixed. In other words, stream temperatures and flow rates are treated as constants in HEN retrofit. Basically, no methods have been developed so far to exploit synergetic interactions between process changes and HEN modifications.

3. Relaxing the Network Pinch by Process Changes

This work is built on the concept of the network pinch and uses process changes to relax the network pinch, the actual bottleneck in an existing HEN. In general, a process can be changed to give different temperatures, flow rates, and duties that are three major parameters in a HEN. In these three parameters, only two of them can be changed independently. There are several combinations of these parameter changes that have different impacts on relaxing the network pinch.

Case 1: Flow Rate and Duty Changes. When the flow rate of a stream in a process increases while both

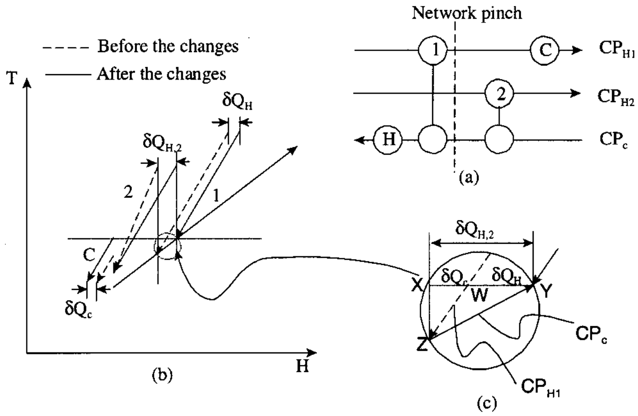

Figure 4. Network pinch shifting due to changes in flow rate and duty.

stream supply and target temperatures are kept constant, this change increases the stream heat duty. This corresponds to the scenario when a process throughput is increased but the process is still operated at the same conditions (e.g. same reaction severity, same fractionation).

A simple example is used to illustrate the effect of changes to flow rate and duty on the HEN. The grid diagram (Figure 4a) shows that the network consists of two process exchangers and one heater and one cooler. The network pinch occurs at the cold side of exchanger 1, which is represented on the $T-H$ diagram (Figure 4b). Matches 1 and 2 are located above and below the network pinch, respectively. As the flow rate of hot stream 2 goes up, the slope of match 2 becomes flatter and its duty increases by δQ_{H2}, shown as the solid line in Figure 4b. This represents the case where heat is shifted from below to above the network pinch. To make the network feasible, match 1 has to move upward to avoid overlap between the two matches. As a result, the hot utility can be reduced by δQ_H. However, δQ_H, the reduction in hot utility, is less than δQ_{H2}. The difference ($\delta Q_{H2} - \delta Q_H$) goes to the cooling consumption (δQ_C), which is increased to bridge the gap between the new outlet temperature of match 1 and the inlet temperature of the cooler.

To derive equations to calculate both δQ_{H2} and δQ_H, let us magnify triangle XYZ in the $T-H$ diagram (Figure 4c). In this triangle, since the existing position and new position of match 1 are parallel, WY represents the reduction in hot utility (δQ_H) while XY is the total duty change in stream H2 (δQ_{H2}). From this triangle, we can derive eqs 1 and 2.

$$\delta Q_H = \left(1 - \frac{CP_{H1}}{CP_C}\right)\delta Q_{H2} \qquad (1)$$

$$\delta Q_C = \delta Q_{H2} - \delta Q_H \qquad (2)$$

where CP_{H1} and CP_C are heat capacity flows of hot stream 1 and the cold stream, respectively, defined as the product of the flow rate and the specific heat capacity. From eq 1, the more the heat load (δQ_{H2}) of match 2 is shifted from below to above the network pinch, the more reduction in the hot utility consumption. At the same time, the penalty on the cold utility can be calculated by eq 2. The case of a match above the network pinch can be described similarly.

Case 2: Temperature Changes Only. For example, exchanger 2 is pinched at the hot inlet (Figure 5).

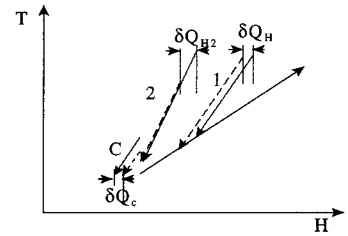

Figure 5. Temperature and duty changes.

Increasing the flow rate and the duty of the hot stream cannot relax the network pinch. In this case, one should increase the inlet temperature of the hot stream while keeping its duty and flow rate constant, shown as the solid line. Then the gap is created between match 2 and the cold composite curve, which allows the cold composite curve to be shifted horizontally toward the left-hand and thus reduces the hot utility consumption by δQ_H.

Case 3: Temperature and Duty Changes. This situation leads to similar effects as those for case 1, that is, flow rate and duty changes, and eqs 1 and 2 can be used to predict changes in hot and cold utility consumption.

Case 4: Temperature, Flow Rate, and Duty Changes. Finally, let us consider the case of changing stream temperature, flow rate, and duty simultaneously. This case can be explained by the crude oil distillation in a refinery, where the cut points of side streams can be adjusted within a certain range. The change in cut points will result in changes to temperatures, flow rates, and stream duty of side-draw products. Taking Figure 6a as an example, the duty of match 1 increases by δQ_{H1} due to the increase of its flow rate and the hot inlet temperature. As a result, the hot utility consumption is reduced by δQ_H. The overall effect of these changes can be analyzed in two steps. In the first step (Figure 6b), only the temperature and the duty are changed in match 1. This leads to the match duty increase by δQ^T_{H1} and the hot utility reduction by δQ^T_H, which can be calculated by using eq 1. In the second step (Figure 6c), only the flow rate and the duty are changed and the hot utility saving δQ^F_H can be predicted by eq 1 as well. By adding up these two separate effects, the overall heating duty can be saved by ($\delta Q_H = \delta Q^T_H + \delta Q^F_H$) while the cold utility is increased by ($\delta Q_C = \delta Q^T_C + \delta Q^F_C$).

An Example. The example of Figure 7 is used to illustrate the application of the plus/minus principle based on the network pinch. Figure 8 shows that the process pinch occurs at 124 °C while the network pinch occurs at the region 142−164 °C defined by exchangers 5 and 6 (EMAT = 10 °C). For convenient visualization, the two pinching matches are shown touching the cold composite curve.

From the process point of view, it is possible to make a small change to the supply temperatures by 7 and 5 °C, respectively, for streams H6 (on which exchanger 4 is located) and H7 (on which exchangers 6 and 8 are located). When we use the process pinch as the reference point, according to the plus/minus principle, we could reduce the hot utility consumption by increasing the supply temperatures for these two streams and hence the duties for these three exchangers. We could estimate the increase in heat duties for both H6 and H7, respectively, as $0.132 \times 7 = 0.92$ and $0.154 \times 5 = 0.77$. Therefore, the plus/minus principle would estimate that the total reduction in the hot utility is 1.69 MW.

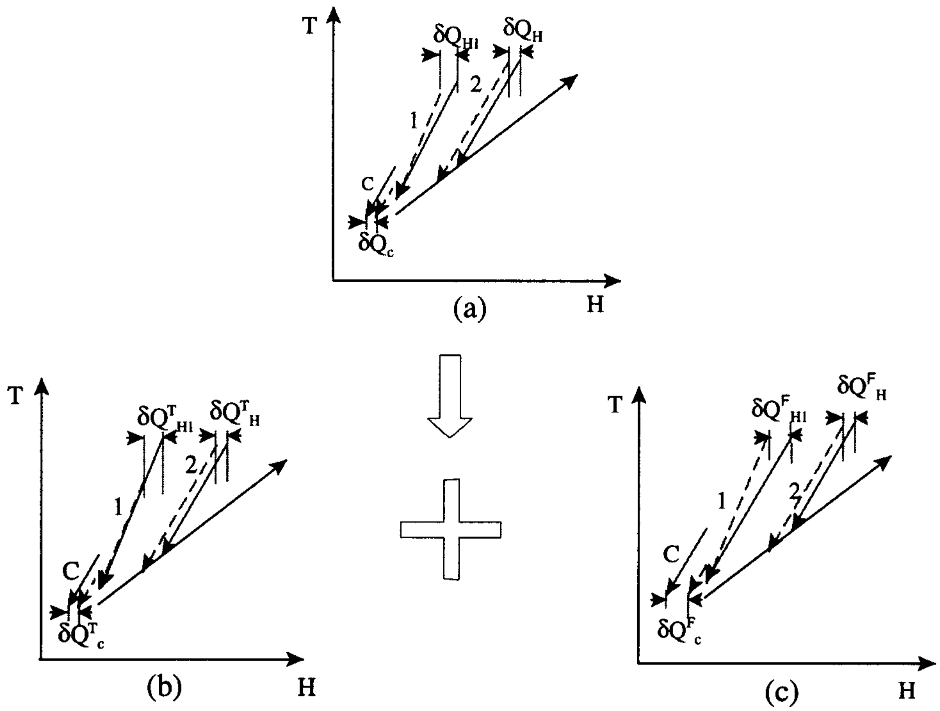

Figure 6. Relaxing the network pinch by temperature, flow, and duty changes simultaneously.

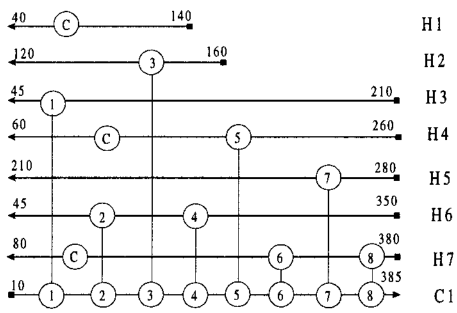

Figure 7. HEN structure in the example.

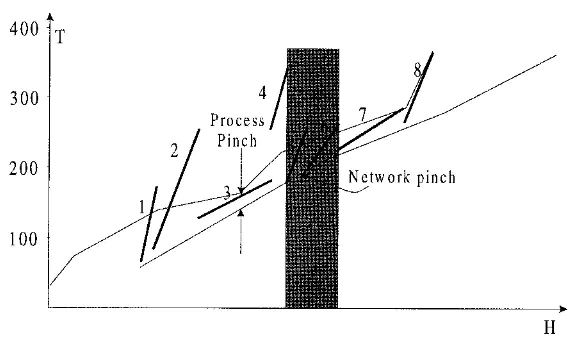

Figure 8. Individual match of the HEN and its composite curves.

However, when the network pinch is used as the reference point, exchanger 4 locates below the network pinch while exchangers 6 and 8 locate above the network pinch. Thus, when increasing the supply temperature for H6, the heat shifts from below to above the network pinch. According to the discussion in case 2, only the part of heat shifted contributes to the reduction in the hot utility by 0.55 MW (instead of 0.92 MW) at the cost of the cold utility of 0.37 MW, both of which can be calculated by using eqs 1 and 2. As a result, by increasing the supply temperatures for both H6 and H7, the total reduction in the hot utility is 1.32 MW (0.77 + 0.55) with an increase in the cold utility by 0.37 MW. This result is confirmed by simulation, which is very different from the above estimate using the process pinch as the reference point.

As a summary, it is found that in addition to HEN topological modifications, process changes (temperature, flow rate, and duty) can also relax the network pinch. When changing process conditions to relax the network pinch, the general plus/minus principle still applies but under the condition of using the network pinch as the reference point. The reason is that the network pinch reflects the actual bottleneck in an existing HEN while the process pinch reveals the limit in a grass-roots design.

4. Simultaneous Approach for HEN Retrofit

The above discussions reveal the impacts of process changes on stream temperatures and flow rates, hence in a HEN retrofit problem. To consider these impacts together with topological changes to a HEN network, a modeling method has been developed here.

Process Model Development. With a process model, we want to investigate the variation of flow rates and temperatures of process streams and predict the impact of these changes on a HEN. The model consists of mass balance, energy balance, phase equilibrium, and so forth. Since the model development heavily depends on the process to be considered, we use a crude distillation column in the refinery for the purpose of illustration.

A typical crude column configuration is shown in Figure 9. Crude oil is charged into the main column and separated into several fractions such as naphtha, distillates, and residue. After being stripped by steam, products with relatively high temperatures transfer heat with cold crude oil to reduce the furnace duty. To further reduce the furnace duty and the condenser duty of the column, intermediate side streams (pumparounds) are drawn out from the main column to exchange the heat with crude oil and are then sent back to the column. In this process, column products and pumparounds serve as the heat sources and crude oil

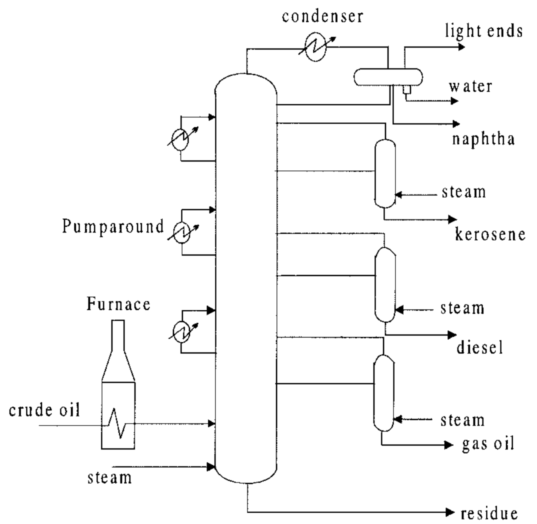

Figure 9. Typical configuration of the crude distillation column.

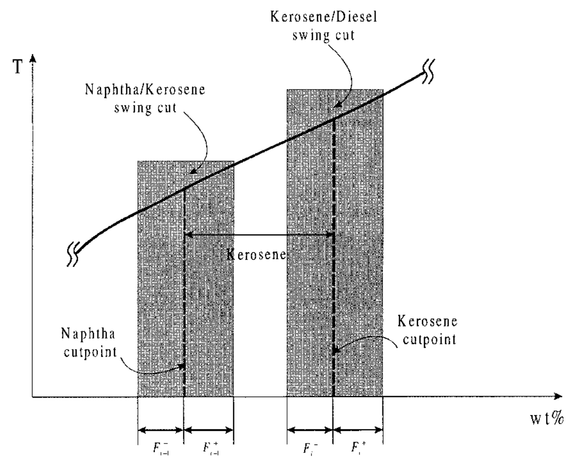

Figure 10. Cut point and the swing cut of the distillation fraction.

is the heat sink. The objective of developing this process model is to predict the product temperatures, the flow rates, and the temperatures and flow rates of pumparounds, which are the major parameters in a preheat train. For the crude oil, its flow rate depends on the column capacity and the column inlet temperature (target temperature) depends on the crude oil composition. For simplicity, we assume only one crude oil is used and then its target temperature is fixed.

Mass Balance on the Column. This work adopts the swing-cut concept to express the overall mass balance on the column (eqs 3 and 4). In eq 3, it calculates the flow rate (F_i) of each product on the basis of weight percentage (wt_i) of the total crude oil flow (F). The flow rate is allowed to vary within the range of swing cuts (F_{i-1}^{+} and F_i^{-}) shown in Figure 10. In eq 4, the total swing-cut flow rate, which accounts for the percentage of the crude oil flow (ws_iF), is the sum of the amount going to the product above (F_i^{-}) and the amount going to the product below (F_i^{+}). The summation of the percentage of all products including swing cuts must be unity (eq 5).

$$F_i = wt_iF + F_{i-1}^{+} + F_i^{-} \tag{3}$$

$$ws_iF = F_i^{+} + F_i^{-} \tag{4}$$

$$\sum_i (wt_i + ws_i) = 1 \tag{5}$$

Energy Balance on the Column. The overall energy balance on the column can be represented by eq 6. This equation covers both cases of fixed and variable throughput. In the latter case, crude throughput (F) and enthalpy (h) are treated as variables. In this equation, the enthalpy of the stripping steam (H_s) is known as well due to the specified steam temperatures and pressures. On the top of the column, the enthalpies of water (h^{w}) and naphtha are fixed at a specified condensing condition. It is assumed that the stripping steam is condensed to water and completely removed from the products. However, since the product enthalpy (h_i) varies with the temperature and flow rate (F_i) of the corresponding product, as a result, the energy balance becomes nonlinear. The column pumparound duty (Q_p) is calculated as the product of specific flow ($\overline{cp}_pF_p$) and temperature difference (ΔT_p) (eq 7), which also forms bilinear terms in the equation. These nonlinear terms are converted to linear ones, which will be discussed in the section on linearization.

$$Fh + \sum_s F_sH_s = \sum_i F_ih_i + \sum_s F_sh^{w} + \sum_p Q_p + QC \tag{6}$$

$$Q_p = \overline{cp}_pF_p\Delta T_p \tag{7}$$

Temperature Model Development. In this section, we aim at developing a linear model to predict changes in the column product temperatures (stream supply temperatures in the HEN). In the rigorous column model, the column temperature profile is predicted by the equation of state (EOS). However, it is quickly found that it is virtually impossible to linearize each a nonlinear term due to highly complex formulations in the EOS; thus, alternative approaches have to be used.

Although the crude oil distillation is a complicated system, it is realized that the change in stream temperature is approximately linear with the change in fraction flow rate. This can be explained by using the equilibrium flash vaporization (EFV) curve of the crude oil (Figure 11). In the crude distillation, the EFV curve can be used to predict column side-stream temperatures at atmospheric pressure. In this figure, the shadow area indicates three swing cuts between the naphtha, kerosene, diesel, and gas oil. As the fraction flow rate varies with the cut point, the EFV temperature changes within a narrow range. For instance, as the diesel cut point changes from wt_2 to wt_1, and the diesel side-draw temperature decreases from T_2 to T_1. In this region, the temperature change is roughly linear with the diesel cut point change. This observation applies to other fractions. Furthermore, the stripping ratio of each fraction normally stays constant; thus, the change in the column product temperature can be expressed as a linear function of the cut point change and consequently the product flow rate change.

However, the EFV curve can only be used to analyze temperature variations qualitatively. The product temperature model with a high accuracy cannot be developed on the basis of the EFV curve. This is because (1) an accurate EFV temperature curve is not always available and (2) converting from other distillation curves such as the true boiling point (TBP) curve is much less accurate and thus large errors could be expected. In addition, the influence of other column

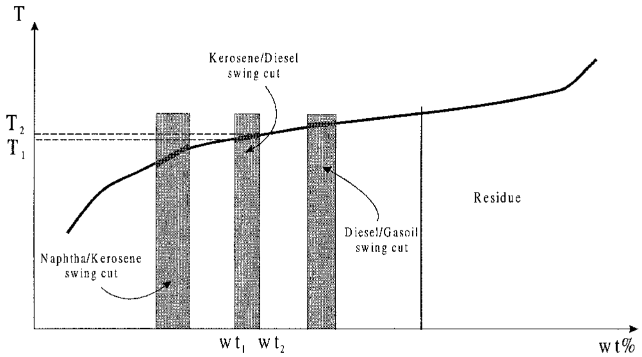

Figure 11. Temperature changes with the cut point variations in the distillation column.

operating parameters such as pumparound duties on the product temperature cannot be reflected on the EFV curve. Instead, process simulation provides an accurate and convenient way to conduct this work.

To develop the stream temperature model from the simulation, we need to select the most influential parameters on temperature changes. With this, we find that changes on both cut points and pumparound duties have significant impacts on stream temperatures. Since we are more interested in predicting incremental changes in stream temperatures, we derive the following equation with both cut points and pumparound duties in the differential forms (ΔD_i and ΔQ_p) on the basis of current operating conditions.

$$TS_i = TS_i^0 + \sum_i c_i \Delta D_i + \sum_p d_p \Delta Q_p \tag{8}$$

Once the independent variables and the model format are determined as above, the next step is to determine the coefficients c_i and d_p. This task can be fulfilled with simulation and regression tools. In the simulation, it is vital to choose correct ranges for independent variables (e.g. pumparound duty and cut point) so that they can cover operation scenarios as wide as possible. After the simulation, regression is conducted to analyze the entire set of simulation data. The model development procedure can be found in ref16. Comparing results with those of the rigorous simulation model shows that this linear temperature model can achieve a high accuracy (Table 1).

Overall Simultaneous Model. For the heat exchanger network, it is required that the model should be able to consider both changes in stream parameters (e.g. temperatures and flow rates) and topology in the HEN (i.e. exchanger resequencing, repiping, and addition). Equations 3−8 can be used to deal with the parameter changes while the MILP model proposed by Zhu and Asante[1] can be used to identify topology changes. By combining both parametric and structural optimization, we then form a complete model as follows.

According to Zhu and Asante,[1] the node representation for a HEN is used (Figure 12). A hot node k and a cold node l are assigned on each hot stream i and cold stream j. The location of an exchanger ($ijkl$) is thus defined by the node k on hot stream i and the node l on cold stream j at the exchanger inlet. Similarly, process heaters (ujl) and coolers (uik) are defined by means of the nodes at which the heater or cooler is located and the utility (u) used in the exchanger.

Objective Function. The objective function in the simultaneous optimization model is defined as eq 9 to minimize the overall hot and cold utility costs while identifying the network pinch.

$$\text{minimize } c_1\Big(\sum_{ujl\in H} CH_u Q_{ujl} + \sum_{uik\in C} CC_u Q_{uik}\Big) + c_2 \Big(\sum_{ijkl\in ME\cup NE\cup RE} p^1_{ijkl} + \sum_{ijkl\in ME\cup NE\cup RE} p^2_{ijkl}\Big) \tag{9}$$

where $c_1 \gg c_2$.

Mass Balance on the Column

$$F_i = wt_i F + F^+_{i-1} + F^+_i \tag{10}$$

where a and b are coefficients.

$$ws_i F = F^+_i + F^-_i \tag{11}$$

$$\sum_i (wt_i + ws_i) = 1 \tag{12}$$

Energy Balance on the Column

$$Fh + \sum_s F_s H_s = \sum_i F_i h_i + \sum_s F_s h^w + \sum_p Q_p + QC \tag{13}$$

$$h_i = aTS_i + b \tag{14}$$

Temperature Equations. In addition to eq 15, the product cut point is calculated from the flow rate (F_i) (eq 16) and the pumparound duty is calculated from the pumparound flow (F_p) and the temperature drop (ΔT_p) (eq 17). All stream temperatures are bounded within a certain range.

$$TS_i = TS_i^0 + \sum_i c_i \Delta D_i + \sum_p d_p \Delta Q_p \tag{15}$$

$$\Delta D_i = \frac{F_i}{F} - D_i^0 \tag{16}$$

$$\Delta Q_p = \overline{cp_p} F_p \Delta T_p - Q_p^0 \tag{17}$$

$$TS_i^{\min} \le TS_i \le TS_i^{\max} \tag{18}$$

Heat Balances on Exchangers. The match heat balances, both on the hot side (eq 19) and on the cold side (eq 20), consist of terms to account for possible topological change options, which include adding a new exchanger ($Q_{ijkl\in NE}$) and relocating a existing one ($Q_{ijkl\in RE}$). The original HEN model was developed on the basis of fixed stream flow rates and supply temperatures. However, if process changes are considered together with HEN topological changes, this requires both stream temperatures and flow rates to vary during optimization, which results in bilinear formulations.

$$T_{i,k+1} = T_{i,k} - \frac{1}{CP_i}\Big[\sum_{j,l} Q_{ijkl\in ME} + \sum_{jl} Q_{ijkl\in NE} + \sum_{j,l} Q_{ijkl\in RE} + \sum_u Q_{uik\in C} + \sum_u Q_{uik\in NC}\Big] \tag{19}$$

Topological Change Control. (a) New Process−

Table 1. Comparisons between Simulation Results and Correlation Results

column conditions	cut points, wt %	NAP/KER	KER/DSL	DSL/GSO	
		25.86	33.27	49.43	
	PA duty, mmcal/h	kerosene	diesel	gas oil	
		2.0	2.5	3.2	
	steam, kg/h	kerosene	diesel	gas oil	residue
		600	700	800	3000
temp, °C	naphtha (TOP)	kerosene	diesel	gas oil	residue
simulations	134.1	160.3	219.1	272.2	331.2
correlations	134.9	158.5	218.2	272.9	330.2

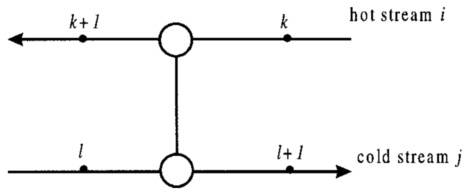

Figure 12. Definition of match *ijkl*.

$$T_{j,l+1} = T_{j,l} - \frac{1}{CP_j}[\sum_{i,k} Q_{ijkl\in ME} + \sum_{i,k} Q_{ijkl\in NE} + \sum_{i,k} Q_{ijkl\in RE} + \sum_{u} Q_{ujk\in H} + \sum_{u} Q_{ujk\in NH}] \quad (20)$$

Process Match Placement Control. A new exchanger must be controlled by additional equations (eqs 21 and 22). These equations make sure that the exchanger duty is forced to be zero when the match is not placed ($z_{ijkl\in NE} = 0$) and that the total number of new exchangers to be considered is specified by constant MODS1.

$$Q_{ijkl\in NE} \le \Delta H_i z_{ijkl\in NE} \quad (21)$$

$$\sum_{ijkl\in NE} z_{ijkl} \le \text{MODS1} \quad (22)$$

(b) Exchanger Relocation Control. Similarly, a binary variable ($z_{ijkl\in RE}$) is introduced to make sure that an existing exchanger is relocated to a new position while simultaneously it is removed from the original position. The constant MODS2 is used to control the number of the relocated exchanger at a time.

$$Q_{ijkl\in ME} \le \Delta H_i(1 - \sum_{j,l} z_{ijkl\in RE}) \quad (23)$$

$$Q_{ijkl\in ME} \le \Delta H_i(1 - \sum_{i,k} z_{ijkl\in RE}) \quad (24)$$

$$\sum_{ijkl\in RE} z_{ijkl\in RE} \le \text{MODS2} \quad (25)$$

(c) Overall Control of the Number of Modifications. The total number of new exchangers and relocated exchangers can be limited by the constant MODS3.

$$\sum_{ijkl\in NE} z_{ijkl} + \sum_{ijkl\in RE} z_{ijkl} \le \text{MODS3} \quad (26)$$

Match Temperature Approach Constraints. The temperature approach of a new exchanger and a relocated exchanger is controlled above the EMAT boundary.

$$DT^1_{ijkl\in NE\cup RE} = T_{i,k+1} - T_{j,l} + c_3(1 - z_{ijkl\in NE\cup RE}) \quad (27)$$

$$DT^2_{ijkl\in NE\cup RE} = T_{i,k} - T_{j,l+1} + c_3(1 - z_{ijkl\in NE\cup RE}) \quad (28)$$

$$DT^1_{ijkl\in NE\cup RE} \ge \text{EMAT}$$

$$DT^2_{ijkl\in NE\cup RE} \ge \text{EMAT} \quad (29)$$

Identification of Pinching Matches. Two continuous variables p^1_{ijkl} and p^2_{ijkl} are defined for identification of the pinching matches. Since the variables p_{ijkl} are minimized in the objective function, they always tend to move toward their lower bounds. This means that, for the exchangers that are pinched, the value of the corresponding p_{ijkl} will be one; hence, DT_{ijkl} = EMAT. Otherwise, the value of the corresponding p_{ijkl} will be zero. For the exchangers whose DT is less than one degree larger than the EMAT, however, the value of the corresponding p_{ijkl} will be equal to the difference between the DT and EMAT.

$$(1 - p^1_{ijkl}) \le DT^1_{ijkl} - \text{EMAT} \qquad ijkl \in NE \cup RE \quad (30)$$

$$(1 - p^2_{ijkl}) \le DT^2_{ijkl} - \text{EMAT} \qquad ijkl \in NE \cup RE \quad (31)$$

where $0 \le p^1_{ijkl} \le 1$ and $0 \le p^2_{ijkl} \le 1$.

Connection of Process Model and HEN Model. Stream temperatures and flow rates become interfaced, linking the HEN model and the column model. In the column model, the stream specific heat capacity (cp_i) is specified at the average temperature ($\bar{T}_i$). Therefore, the stream flow rate can be linearly linked with the heat capacity flow (CP_i) used in the HEN model (eq 32). In addition, the first-node temperature of the hot stream must be the same as the stream supply temperature (TS_i) (eq 33). Apart from the link in the temperature and flow rate, column pumparound duties and condenser duty are also connected with the corresponding stream duties (eqs 34 and 35).

$$CP_i = \overline{cp}_i F_i$$

where

$$\overline{cp}_i = a\bar{T}_i^{\,2} + b\bar{T}_i + c$$

and

$$\bar{T}_i = \frac{TS_i^{\min} + TS_i^{\max} + 2TT_i}{4} \quad (32)$$

$$TS_i = T_{i,k} \quad \text{for } k = 1 \quad (33)$$

$$Q_p = \sum_{j,l,k} Q_{pjkl\in ME} + \sum_{j,l,k} Q_{pjkl\in NE} + \sum_{j,l,k} Q_{pjkl\in RE} + \sum_{u,k} Q_{upk\in C} \quad (34)$$

$$QC = \sum_{j,l,k} Q_{ijkl\in ME} + \sum_{j,l,k} Q_{ijkl\in NE} + \sum_{j,l,k} Q_{ijkl\in RE} + \sum_{u,k} Q_{uik\in C} \quad \text{for } i = \text{condensing stream} \quad (35)$$

Linearization. Two linearization methods, that is, the logarithmic linearization and piecewise linearization, are used in this paper to convert original nonlinear terms to linear. The first one can be applied to the bilinear terms in eq 13. For instance, for the bilinear term $X \times Y$, a new variable Z is introduced as $Z = X \times Y$. By using the logarithm on both side of the equation, the bilinear term becomes separable (eq 36). Each logarithmic term is linearized and then linked using eq 37.

$$\ln Z = \ln X + \ln Y \quad (36)$$

$$aZ - bX - cY + d = 0 \quad (37)$$

This logarithmic linearization method is, however, not suitable to the linearization of eqs 17, 19, and 20 because of two reasons. First, in a HEN optimization, it is likely that some of the exchanger duty may evolve to zero, which would cause infeasibility with a logarithm function. Second, temperature changes could vary within a wide range, and use of the above linearization method to fit the whole range would produce large errors.

Instead, to maintain the necessary precision, the second method is adopted. Two new artificial variables X and Y are defined first (eq 38) to separate bilinear terms in the form of $CP\Delta T$ and the energy balance ($Q = CP\Delta T$). Then a piecewise linearization formulation can be applied on both X^2 and Y^2, respectively. Note that the accuracy of the solution can be improved with the increase of linear segments but at the cost of introducing more binary variables.

$$X = \frac{1}{2}(CP + \Delta T); \quad Y = \frac{1}{2}(CP - \Delta T) \quad (38)$$

Remarks. The overall simultaneous model consists

$$Q = X^2 - Y^2 \quad (39)$$

of eqs 9−39, which are used to relax bottlenecks (the network pinch) in a heat recovery system by HEN topological changes and possible process modifications.

The overall HEN retrofitting procedure involves two stages, screening and optimization, which are summarized as Figure 13. Bottlenecks in the existing HEN, which are defined as pinching matches, are identified first, and the maximum heat recovery potential of the existing HEN (R_{max}) is calculated. If R_{max} is less than the required value, then HEN topological changes and process modifications are inevitable. In the screening stage, the HEN is evolved to increase its heat recovery by structural changes and process changes simultaneously until the required heat recovery is satisfied. Then the HEN with the new structure and new stream temperatures and flow rates is submitted to the nonlinear optimization. In this cost optimization stage, the tradeoff between energy and capital costs is optimized to minimize the total retrofit cost.

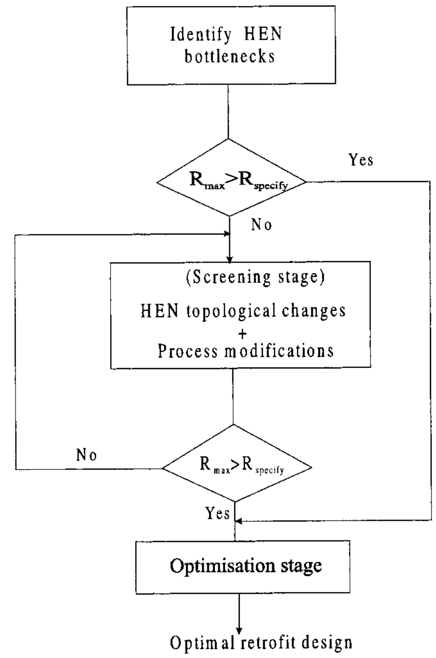

Figure 13. Overall procedure of the HEN retrofit.

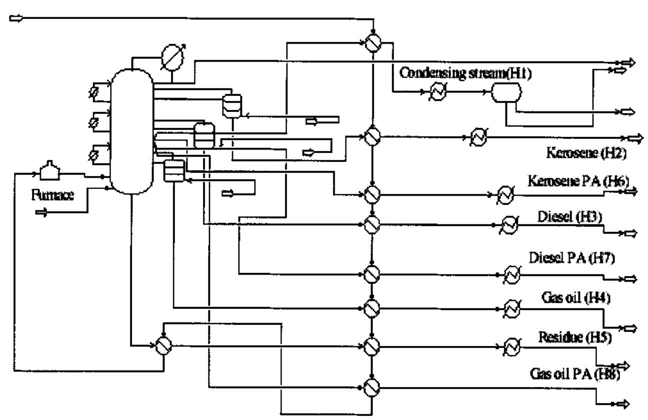

Figure 14. Crude distillation column and its preheat train in the case study.

5. Case Study

Figure 14 is a flow sheet of a crude distillation column with its preheat train, from which the network structure (Figure 15) of the preheat train can be extracted. It is required to increase the throughput by 10%, which is possible since the column has spare capacity. After the throughput increase, the column reaches its maximum capacity but the fuel consumption in the fired heater increases to 33 MMBTU/D, which exceeds its maximum capacity limit of 30 MMBTU/D. Therefore, the preheat train has to be retrofitted to increase heat recovery and thus reduce the fuel consumption. On the basis of the EMAT of 27 F permitted in the HEN operation, the maximum heat recovery limit in the existing HEN is 31.55 MMBTU/D and the network pinch location is indicated in Figure 15. Hence, it implies that topology changes are inevitable in the HEN retrofit design.

By using the network pinch method, it is found that no exchanger resequencing and repiping is beneficial to the heat recovery. In addition, stream splitting is prohibited due to the practical constraints. Therefore,

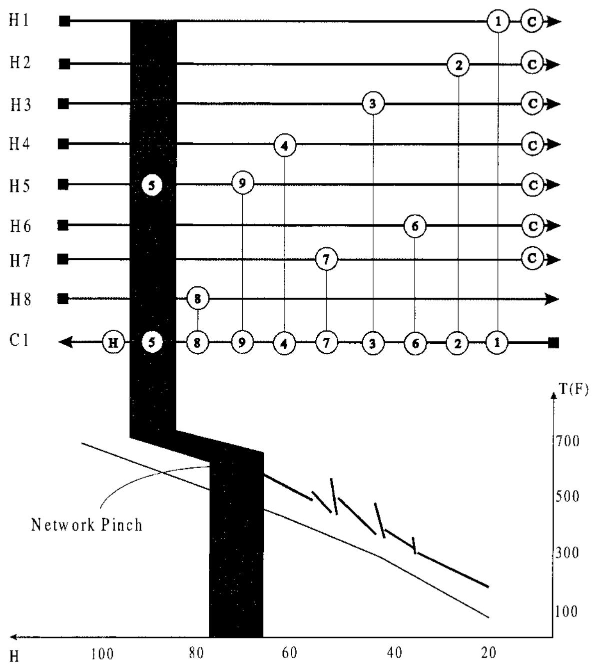

Figure 15. Limits of the existing HEN in the case study.

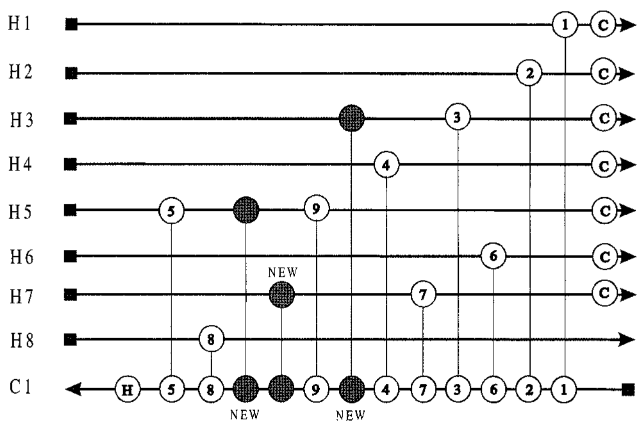

Figure 16. HEN retrofit design with HEN topological changes only.

the only option for the HEN topology modifications is the addition of new exchangers. After three new exchangers are added, the fuel consumption in the fired heater is reduced to 27.86 MMBTU/D, which is less than the fired heat maximum capacity (Figure 16). Then this new HEN topology is submitted to the nonlinear optimization to optimize the capital/energy tradeoff. After the optimization, 13 205 ft^2 of additional exchanger area is required and the hot utility consumption is 30.0 MMBTU, which is same as the capacity limit.

Instead of changing the HEN topology above, the changes to column operating conditions are considered to ease the task of HEN retrofit and hence to reduce the total costs. In this case, the range of cut point variation of each fraction and the temperature drop in each pumparound are defined in Table 2. Note that the pumparound flow rate is kept the same as existing values to avoid the hydraulic problem in the column after the throughput increase.

After solving the simultaneous optimization model that consists of both the HEN and column models, the optimized cut points and temperature drops in pumparounds are listed in Table 3 along with the existing values. From the results, it is found that the naphtha

Table 2. Variation Range of the Column Operating Conditions

	cut points (wt %)	
	lower bound	upper bound
naphtha	23.63	25.86
kerosene	33.27	35.97
diesel	47.54	49.43
	pumparound temp drop, F	
	lower bound	upper bound
kerosene	90	180
diesel	90	180
gas oil	90	180

Table 3. Fraction Flows, Temperatures, and Pumparound Duties

	fraction flow rate, 10^3 lb/d	
	existing	optimized
naphtha	59.0	62.1
kerosene	23.1	24.3
diesel	34.2	32.3
gas oil	27.6	25.2
residue	96.1	96.1
	stream temp, F	
	existing	optimized
top of column	262.8	274.9
kerosene	324.7	325.0
diesel	443.4	450.0
gas oil	531.2	523.4
residue	638.4	638.1
	pumparound duty, MMBTU	
	existing	optimized
kerosene	11.9	12.2
diesel	7.9	7.2
gas oil	11.9	13.6

and kerosene flow rates should increase while diesel and gas oil flows should decrease. The reason is that although the naphtha fraction has a low temperature, its heat capacity flow is considerably larger than those of other fractions, which gives more heat to the crude oil. Consequently, it contributes to the reduction of fuel consumption more significantly than the others. In addition, the gas oil pumparound duty is increased by 1.7 MMBTU, but at the same time, the gas oil stream duty decreases by 0.6 MMBTU. This is due to the fact that the gas oil temperature will decrease when its pumparound duty increases. But these two changes result in the net heat duty gain (1.1 MMBTU) and are beneficial to the reduction of total fuel consumption.

With these process changes, only one new exchanger on the gas oil pumparound is required (Figure 17). As a result, to achieve the fuel consumption of 30 MMBTU/D, the overall additional area required for the new exchanger and the existing coolers is 5534 ft^2 in comparison with 13 205 ft^2 obtained without considering process changes. Clearly, the retrofit design with simultaneous consideration of HEN modification and process changes requires much cheaper capital investments (Table 4) than the design considering HEN topological changes only.

It should be pointed out that although the crude distillation model developed in this work only incorporates parameter changes in terms of cut points and pumparound temperatures, the column structure might also be modified to improve the heat recovery in terms

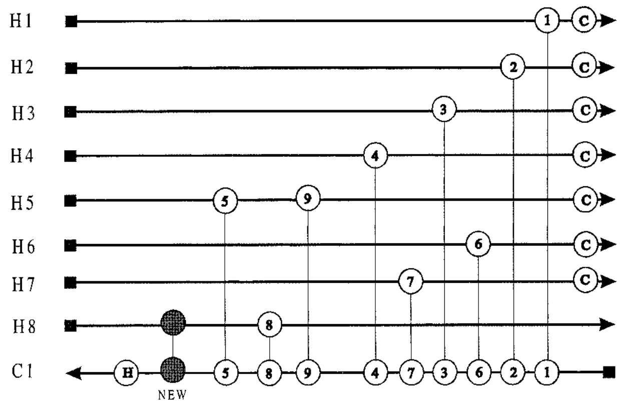

Figure 17. HEN retrofit design with simultaneous HEN and process changes.

Table 4. Comparison between Two Retrofit Designs

	HEN topology changes only	simultaneous process and HEN changes
additional area requirement (ft^2)	13205.3	5534
new matches	3	1
process changes	0	1. cut point changes 2. changes in pumparound return temperature
area costs	186.7 k$	77.5 k$

of side draw locations and pumparound locations and so on. In that case the column model becomes a MILP formulation.

6. Conclusions

In this paper, a new guideline is proposed for process modifications. The general rule for selecting process changes in a HEN retrofit dictates that increasing the heat above the *network pinch* reduces the hot utility without the penalty in the cold utility while shifting heat from below to *above the network pinch* improves the heat recovery but at the cost of the cold utility. To determine the option of minimum capital investments, a simultaneous model is proposed to exploit complex interactions between the process and the HEN. A linear HEN model and a newly developed process model with high accuracy are used to consider all possible options for the HEN retrofit both from HEN modifications and process changes. The results of the case study show that the HEN retrofit with both HEN topological changes and process changes requires much less capital than that of considering HEN topological changes only.

Although the complex distillation column is used as a process example in this paper, the proposed work has discovered a major benefit by considering both HEN retrofit and process changes simultaneously. This finding may become the motivation for considering different kinds of processes by extending the principal and basic idea of this work.

Finally, it should be emphasized that as far as process change concerns, although process changes can bring significant benefit to the HEN retrofit in terms of capital cost savings, sometimes changes to the stream flow rate and quality might not have positive effects on downstream processes. In this case, process changes should be justified in the context of overall plant optimization (Zhang et al.[17]).

Nomenclature

Sets and Indices

$C = \{c|c$ is a cooler$\}$
$I = \{i|i$ is a product stream$\}$
$J = \{j|j$ is a cold process stream$\}$
$K = \{k|k$ is a node on a hot stream$\}$
$L = \{l|l$ is a node on a hot stream$\}$
$P = \{p|p$ is a pumparound$\}$
$S = \{s|s$ is stripping steam$\}$
$ME = \{ijkl|ijkl$ is an existing process match$\}$
$NE = \{ijkl|ijkl$ is a possible new exchanger match$\}$
$RE = \{ijkl|ijkl$ is an existing match possibly relocated$\}$
$U = \{u|u$ is a utility$\}$

Parameters

$\overline{cp_i}$ = average specific heat capacity of pumparound stream i
$\overline{cp_p}$ = average specific heat capacity of pumparound stream p
D_i^0 = product cut point at the base operating conditions
EMAT = exchanger minimum approach temperature
h = enthalpy of crude oil in the column inlet
h^w = water enthalpy on the column condensing condition
H_s = enthalpy of stripping steam
MODS1 = number of new exchangers allowed
MODS2 = number of relocated exchangers allowed
MODS3 = total number of modification options
Q_i^0 = pumparound duty at the base operating conditions
TS_i^0 = stream supply temperature at the base conditions
$TS_i^{\min}$ = lower bound of the supply temperature of stream i
$TS_i^{\max}$ = upper bound of the supply temperature of stream i
$\bar{T}_i$ = average temperature of stream i
ΔH_i = maximum stream duty change
wt_i = weight percentage of product i on the crude oil
wt_s = weight percentage of swing-cut on the crude oil

Variables

CP_i = heat capacity of stream i
CP_j = heat capacity of crude oil j
DT^1 = approach temperature difference at the cold side of a match
DT^2 = approach temperature difference at the hot side of a match
F_i = flow rate of product stream i
F = crude oil flow rate
$F_{i,k}$ = flow rate of product stream i at node k
F_i^- = flow rate of swing-cut in stream $i-1$ between streams $i-1$ and i
F_i^+ = flow rate of swing-cut in stream i between streams $i-1$ and i
h_i = enthalpy of stream i
p^1 = variable to measure pinching conditions at the cold side of a match
p^2 = variable to measure pinching conditions at the hot side of a match
Q_p = heat duty of pumparound p
Q_{ijkl} = heat load of match $ijkl$
QC = column condenser duty
$T_{i,k}$ = temperature at node k in stream i
TS_i = product stream supply temperature
ΔD_i = cut point changes of stream i
ΔQ_p = duty changes in pumparound p
ΔH_i = enthalpy changes in stream i

z_{ijkl} = binary variable for match *ijkl*

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