

Design of Crude Fractionation Units with Preflashing or Prefractionation: Energy Targeting

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This paper uses a previously developed systematic procedure to obtain design targets for heat integration in conventional crude fractionation units that use prefractionation columns or preflash drums. It is shown that, under the same high product yield conditions, prefractionation or preflashing followed by a conventional unit are not as energy efficient as straight conventional crude units. This is in great part due to the loss of the carrier effect that light components have in separating heavy gas-oil fractions in the flash zone. Such a carrier effect is discussed. In addition, if one accepts the yield of atmospheric gas oil to be smaller, then these prefractionation/preflash options consume less energy.

Introduction

In previous work,¹ a systematic procedure for the design of conventional crude fractionation units was presented. This procedure is based on a step-by-step combination of rigorous simulation and heat integration. The procedure starts with a column without pump-around circuits, and as heat is transferred from the condenser to pump-around circuits with higher temperature, a tradeoff between steam usage and furnace savings is established. This transfer of heat is possible because of the well-known operating and design flexibility that crude fractionation installations exhibit, knowledge that was formalized in detail by Bagajewicz.² The procedure presented by Bagajewicz and Ji¹ makes use of rigorous simulations and heat supply demand diagrams similar to those introduced by Andrejovich and Westerberg³ and Terranova and Westerberg.⁴ Based on these targets, a heat-exchanger network design procedure to handle crudes of different density at maximum efficiency was developed.^{1,5} These design procedures did not use any preflashing or prefractionation schemes.

In this paper, a rigorous procedure to obtain targets of energy consumption and column reflux arrangements for prefractionation/preflashing units is presented. Figures 1 and 2 depict both schemes.

In the prefractionation scheme (Figure 1), light products obtained in the prefractionation column are not sent to the main column. In the preflash scheme (Figure 2), the intended effect is to avoid unnecessary heating of light components in the furnace, short-circuiting them to be injected at an appropriate tray in the column for further fractionation.

There are several unresolved questions regarding these designs. Among others, there are the following:

- In the case of preflashing, what is the optimum temperature of the flash drum?
- In the case of prefractionation, what is the optimum temperature of the prefractionation column feed?
- In the case of preflashing, what is the feed tray that one needs to use to feed the vapor from the drum into the column?

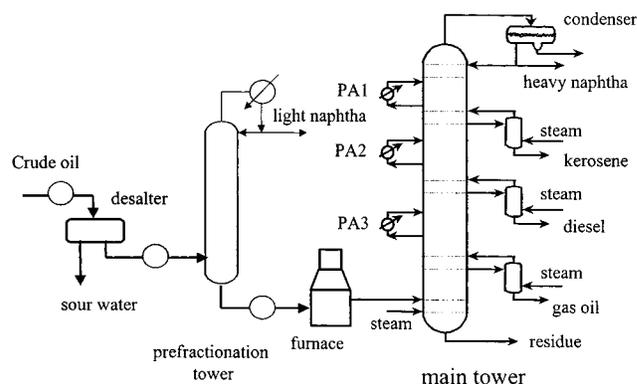


Figure 1. Basic prefractionation design (the heat-exchanger network is omitted).

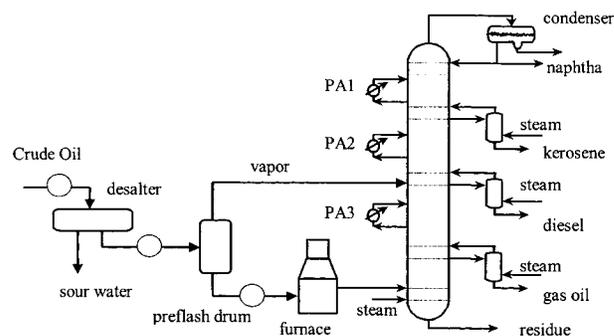


Figure 2. Basic preflash design (the heat-exchanger network is omitted).

(d) What are the loads of the pump-around circuits and what are the steam flow rates for stripping that produce the most energy efficient combination?

In addition, both schemes avoid introduction of lights in the feed tray (flash zone) with the aforementioned expectation that one would avoid heating of these lights unnecessarily. However, the so-called carrier effect of lights, which is known to increase the separation of atmospheric gas oil from the residue,⁶ calls for keeping them in the feed stream. Because steam can act as a carrier too, the question is whether it can actually substitute for light hydrocarbons. Finally, if yield is to be sacrificed, what is the extent of yield reduction compared to energy savings? In other words, would it be worth reducing the yield to achieve energy efficiency?

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Table 1. Comparison between Conventional and Preflash Design (Light Crude)

	conventional	preflashing
naphtha, m ³ /h	249.6	249.3
kerosene, m ³ /h	144.1	143.9
diesel, m ³ /h	70.5	69.7
gas oil, m ³ /h	118.6	101.9
residue, m ³ /h	212.5	230.6
product gaps, °C		
naphtha–kerosene	25.1	24.5
kerosene–diesel	5.0	3.2
diesel–gas oil	0.6	–0.9
heating utility, MW	103.6	97.3
energy consumption, MW	113.0	106.8

All of these questions have to be answered by taking into account the fact that a specific quality of products (given by TBP or ASTM D86 points) needs to be guaranteed equally for all options that are compared.

Effect of Preflashing on Products

In this section the step-by-step procedure proposed by Bagajewicz and Ji¹ is applied to both prefractionation and preflash schemes. In the case of prefractionation, the method can be applied directly. In the case of preflashing, the introduction of the vapor feed from the flash drum changes the heat load distribution in the column affecting the heat duty distribution of the pump-around circuits. Thus, to better study the effect of the drum temperature and the feed tray, a column without pump-around circuits is first used. Pump-around circuits are systematically added later.

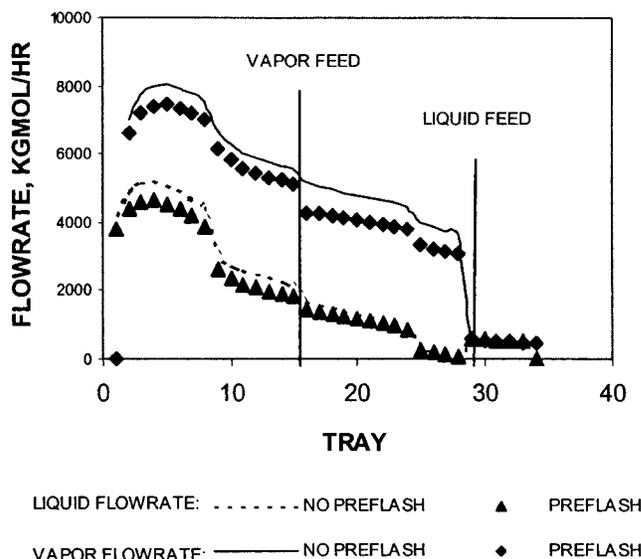
The main column has 34 trays, and side products are withdrawn at trays 9, 16, and 25, respectively. The outlet temperature of the furnace is set at 360 °C, and the condenser temperature is fixed at 32 °C. Two crude oils, one light and one heavy, are used in the study. Details of the assay data and product 95% distillation temperatures were given by Bagajewicz and Ji.¹ The over flash rate chosen is 3%, the flow rates of side withdrawals and the condenser duty are adjusted to achieve the specifications of the over flash rate, and product 95% distillation temperatures as well as the flow rates of stripping steam streams are maintained constant. Finally, the minimum approach temperature for energy targeting (HRAT) is 22.2 °C (40 °C), as suggested by Bagajewicz and Soto.⁵ Utility is minimized not only because pinch analysis is performed but also because the pump-around circuits are manipulated to decrease it.

Consider first a preflash drum operated at 163 °C and the feed tray to be tray 15. Tray 15 is chosen because the vapor concentration best matches the composition of the vapor from the drum. The effect of changing these two values will be studied later. The simulation results corresponding to a light crude with no pump-around circuits are shown in Table 1 and Figure 3. As shown by Bagajewicz,² pump-around circuits have an effect on energy recovery but a very slight effect on product yield. Thus, as we shall see later, the energy consumption shown in Table 1 should further decrease as pump-around circuits are added.

The major differences are as follows:

(1) The vapor and liquid traffic decrease as a result of decreased heat input in the feed. That is, a lower reflux is needed.

(2) The yield of gas oil decreases and the yield of residue increases as a result of less stripping ends existing at the flash zone of the column.

**Figure 3.** Comparison of vapor and liquid distribution.**Table 2. Comparison between Conventional and Preflashing Design (Heavy Crude)**

	conventional	preflashing
naphtha, m ³ /h	54.5	54.2
kerosene, m ³ /h	47.9	48.4
diesel, m ³ /h	71.1	70.8
gas oil, m ³ /h	28.8	26.4
residue, m ³ /h	593.0	595.4
product gaps, °C		
naphtha–kerosene	19.8	18.6
kerosene–diesel	1.3	1.0
diesel–gas oil	–5.6	–5.1
heating utility, MW	101.3	100.5
energy consumption, MW	106.5	105.8

(3) The gaps between the different products decrease.

As we shall see later, the difference in yield for the gas oil and the residues cannot be reconciled. Essentially, the presence of lights in the flash zone for the conventional case, which provide the so-called carrier effect, cannot be easily substituted by other stripping means in the case of the preflash or prefractionation designs. The comparison with a prefractionation design is omitted because the same effect is expected for the reasons just outlined.

The same comparison was performed for the case of a heavy crude (Table 2). In this case, no pump-around circuits were used either.

Specifically, the yield of gas oil decreases. Because the heavy crude contains less light distillates, vaporization in the heat-exchanger network train is not severe and can be suppressed under moderate pressure. On the other hand, the heavy crude flashing at a low temperature (e.g., 163 °C) does not produce much vapor. From the viewpoint of energy savings, a small amount of vapor bypassing the furnace cannot significantly reduce the duty of the furnace. We can see from Table 2 that the energy consumption for the preflash design is only slightly lower. Therefore, the preflash design is not justified for heavy crude. The same can be said for the prefractionation design.

We now proceed to determine the optimal pump-around circuit loads. The heat is shifted from the condenser to the pump-around circuits step by step, as proposed by Bagajewicz and Ji.¹ Briefly, the method consists of transferring as much heat as possible from the condenser to the pump-around circuits until the

Table 3. Comparison between Conventional and Preflash Designs with Pump-around Circuits (Light Crude).

	conventional	preflashing
naphtha, m ³ /h	246.6	245.4
kerosene, m ³ /h	143.1	142.6
diesel, m ³ /h	73.9	74.4
gas oil, m ³ /h	119.4	102.4
residue, m ³ /h	212.5	230.6
product gaps, °C		
naphtha–kerosene	20.5	18.8
kerosene–diesel	–1.9	–6.6
diesel–gas oil	–0.3	–1.7
heating utility, MW	84.2	78.4
energy consumption, MW	93.7	87.9

overall minimum utility consumption obtained using pinch analysis reaches a minimum. The procedure is summarized next.

Step 1. Start with a column configuration suggested by Watkins with one exception: no pump-around circuits.

Step 2. Perform a simulation.

Step 3. Construct the supply demand diagram.

Step 4. Transfer the optimum amount of heat to a pump-around circuit located in the region between the top tray and the first draw.

Step 5. If needed, increase the steam in the first side stripper until the desired gap is restored.

Step 6. If there is heat surplus from the pump-around circuit just added, transfer the heat to the next region between draws in the same way as in step 4; if not, stop.

The fact that preflashing reduces the gas-oil yield reveals that the light components in the crude help the vaporization of heavy components (Table 3). However, the yield of gas oil can also be increased by using more stripping steam. In other words, both light components in the crude oil and stripping steam have a stripping effect. We now discuss both cases:

(i) The stripping agent is steam.

(ii) The stripping agent is a hydrocarbon, which exists in the mixture to be stripped.

In the rest of the paper, we will first explore the carrier effect and its intricacies in the case of crude fractionation and then present a comparison between the preflash design and the conventional design for processing the light crude oil on the basis that both designs produce the same amount of residue.

Carrier Effect

For the purpose of our analysis, we define the stripping effect of a component to be the differential change in the amount of residual liquid over the differential change of the amount of this component in the feed, provided that the system temperature, pressure, and quantities of other components are constant. Thus, when the addition of a component to the feed decreases the liquid rate, the assumption is that it happens at the expense of vaporizing more of the heavy or intermediate components. Translated into the crude fractionation field, the presence of light ends decreases the residue yield and, therefore, increases the gas-oil yield at its expense. We now analyze both stripping agents separately.

Case 1: Steam Stripping. Stichlmair and Fair⁷ present a few charts obtained numerically, showing that the liquid yield in a flash is lowered when light components are added to a mixture. We attempt to

explain this theoretically. We start using the well-known flash equations

$$x_i = \frac{z_i}{\frac{L}{F} + \left(1 - \frac{L}{F}\right)K_i}$$

$$\sum_i x_i = 1$$

$$K_i = P_i^s/P \quad (1)$$

where x_i and z_i are the molar fractions of component i in the liquid and in the feed, P_i^s is the saturated pressure of component i , P is the total pressure of the system, and L and F are the liquid flow rate and the feed, respectively. The set of equations (1) is referred to as the no-steam equation. The vapor–liquid-phase behavior is assumed to follow Raoult's law, which is appropriate for hydrocarbons.

Let z_i^* and y_i^* be the dry compositions and F^* and V^* the dry flow rates. The temperature is assumed to be high enough so that any water in the liquid is neglected. A simple balance tells that

$$y_i = y_i^*(1 - P_{st}/P^*) \quad (2)$$

where P_{st} is the partial pressure of steam and P^* is the system pressure. For atmospheric distillation, P^* is somewhere between 0.24 and 0.37 MPa. Because $P^* - y_i = P_i^s x_i$, we obtain

$$\frac{y_i^*}{x_i^*} = \frac{P_i^s}{P^* - P_{st}} \quad (3)$$

For convenience, we define $K_i^* = y_i^*/x_i^*$. K_i^* is called the modified vapor–liquid equilibrium constant of component i . Finally, combining a dry component balance ($F_i^* z_i^* = Lx_i^* + Vy_i^*$)

$$x_i^* = \frac{z_i^*}{\frac{L}{F^*} + \left(1 - \frac{L}{F^*}\right)K_i^*}$$

$$\sum_i x_i^* = 1$$

$$K_i^* = \frac{P_i^s}{P^* - P_{st}} \quad (4)$$

We now compare eqs 1 and 4. For a given hydrocarbon mixture flashing at a fixed temperature, we have $z_i^* = z_i$ and $F^* = F$. Suppose the flash without steam and the flash with steam take place at the same temperature; then the saturated pressure of component i does not change. When $P = P^* - P_{st}$, we have $K_i^* = K_i$ and $x_i^* = x_i$. This means that injecting steam is equivalent to reducing the system pressure. Note that eq 3 still holds if steam is replaced by a gas that is insoluble in the hydrocarbon mixture. The use of steam, however, has some operational limits that will be explored later.

Case 2: Hydrocarbon Stripping. We now consider the stripping effect of light components that distribute

Table 4. Mixture Composition and K Values

	molar composition	K value		molar composition	K value
water	0	267	n -C ₂₀	0.089	2.56
ethane	0	138	n -C ₂₂	0.098	1.75
propane	0	115	n -C ₂₄	0.108	1.14
n -butane	0	85.6	n -C ₂₆	0.117	0.79
n -hexane	0	45.2	n -C ₂₈	0.121	0.54
n -octane	0	27.4	n -C ₃₀	0.107	0.38
n -decane	0	17.8	n -C ₃₂	0.081	0.26
n -dodecane	0.04	12.2	n -C ₃₆	0.045	0.13
n -tetradecane	0.053	8.68	n -C ₂₀	0.089	2.56
n -hexadecane	0.064	5.84			
n -octadecane	0.077	3.79			

in both phases. Consider the following flash equations:

$$l_i = \frac{f_i}{1 + \frac{F-L}{L}K_i} \quad (5)$$

$$L = \sum_j l_j \quad (6)$$

where l_i is the flow rate of component i in the liquid phase. Substituting eq 5 into eq 6, differentiating, and considering that $\partial F/\partial f_i = 1$, one obtains

$$\left(\frac{\partial L}{\partial f_i}\right)_{T,P,f_{j \neq i}} = \frac{l_i/f_i - \sum_j y_j l_j/f_j}{1 - (F/L)\sum_j y_j l_j/f_j} \quad (7)$$

Because the liquid-phase partition ratio l_i/f_i is less than 1, the summation in eq 7 is less than 1. Accordingly, the denominator is positive. Therefore, whether the derivative is negative or not depends on the numerator. The numerator consists of two terms. The first term represents the liquid-phase partition ratio of component i in the liquid phase, and the second term is the average partition ratio of the mixture. If the partition ratio of component i is smaller than the average partition ratio, the derivative is negative and the component has a stripping effect. Thus, eq 7 can be used to determine if, at a given pressure and temperature, a component is a stripping agent; that is, if increasing its composition in the feed leads to a decrease of the liquid flow rate in the flash.

The stripping effect of each component in a mixture of aliphatic hydrocarbons is studied next. The composition of a mixture and the K values of each component are shown in Table 4. The mixture is allowed to flash at 0.1 MPa and 393 °C. Figure 4 shows the residue/feed ratio as a function of K values of the added components. The molar ratios of the pure component added to the feed are 1/100, 5/100, 10/100, 20/100, and 30/100, respectively.

Originally, the residue/feed ratio is 0.43. The residue/feed ratio decreases with an increasing K value of the added component. Note that all curves intersect at $K = 1.35$, a value at which $(\partial L/\partial f_i)_{T,P,f_{j \neq i}}$ equals zero. When the K value of the added component is greater than 1.35, the residue/feed ratio is lower than 0.43. This means these components act as stripping components. With the increase of the K value, the residue/feed ratio decreases. When the K value is greater than 30, the residue/feed ratio is almost constant. In this sense, water, ethane, propane, and butane have the same stripping effect on

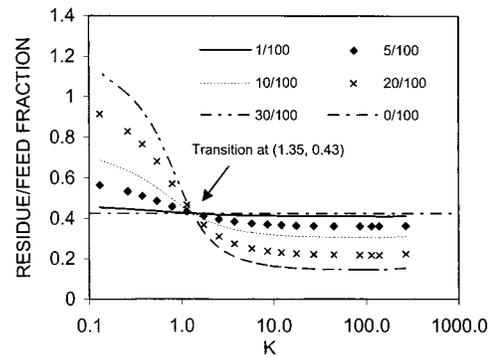


Figure 4. Stripping effect as a function of K values (C10–C36 mixture, 0.1 MPa, 393 °C).

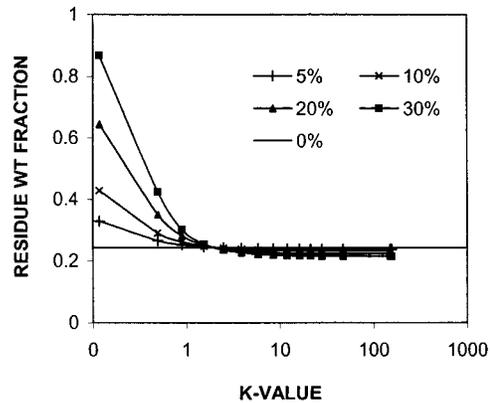


Figure 5. Stripping effect for a light crude.

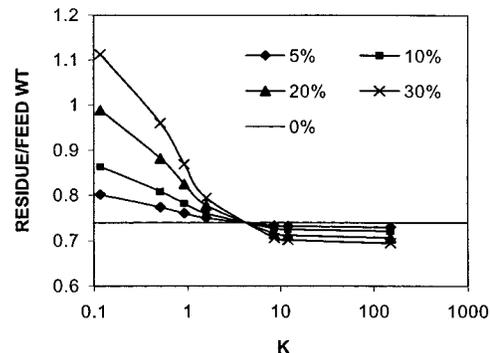


Figure 6. Stripping effect for a heavy crude.

a molar basis. At the same time, it can be seen that, for a given component with a K value greater than 1.35, the residue/feed ratio decreases with an increasing amount of the added component. When the K value is less than 1.35, the residue/feed ratio is higher than that without a stripping component. This means that the component added is not a stripping agent. Instead, it prevents vaporization.

The phenomenon can be explained as follows. An added component helps the vaporization by increasing its partial pressure, but it also hurts the vaporization of other components by reducing their liquid molar compositions. Furthermore, the part of the component distributed in the liquid phase will contribute to a larger residue/feed ratio. When the K value of a component is large, it generates a large partial pressure, and most of it stays in the vapor phase, so the net effect is stripping. On the other hand, when the K value is small or, in other words, the component is heavy, a large portion of the component stays in the liquid phase, resulting in a larger residue/feed ratio.

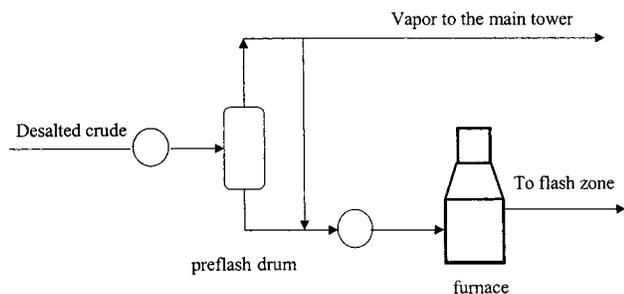


Figure 7. Schematic diagram for preflash vapor splitting.

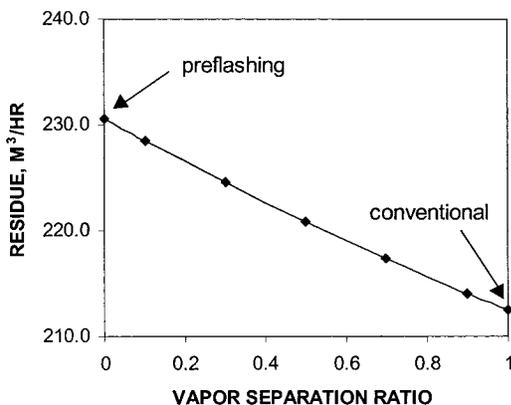


Figure 8. Residue yield as a function of the vapor separation ratio in preflash design.

To determine if this effect manifests quantitatively in the same approximate way for crudes, the light crude was mixed with stripping agents and flashed at 360 °C. The heavy crude was mixed with stripping agents and flashed at 343 °C. The results are shown in Figures 5 and 6, which depict trends similar to that seen in Figure 4. The residue/feed ratio in light crude stripping, however, does not go down as much because of the existence of a large amount of light components in the mixture. The effect is shown parametric as percentages, which are the amount of stripping agent divided by the amount of the crude on a molar basis.

In practice, the crude experiences flashing at the flash zone of the main tower, and the resultant liquid is further stripped with steam at the bottom part of the main tower. To show how light components affect the flow rate of the residue leaving the main tower, we split

the vapor from the preflash drum (163 °C) into two parts as shown in Figure 7.

One part goes to the main tower and the other mixes with the remaining crude again; the resultant mixture enters the furnace. Now, concentrate on the flash zone and examine the effect of the light components entering into it. The outlet temperature of the furnace is fixed at 360 °C. By changing the remixing ratio, one can control the amount of light components entering the furnace. The result is shown in Figure 8.

The amount of residue increases constantly with an increasing vapor separation ratio. The vapor remixing ratio of one corresponds to the conventional design, and the ratio of zero corresponds to the common preflash design. It is seen that the yield of the residue decreases constantly with the increasing vapor remixing ratio. The trend is consistent with what we saw in Figures 4–6.

Following, we first discuss the effect of steam temperature and the location of its injection. We then investigate the energy requirement associated with the use of stripping agents, and last we study the effect of process pressure.

Steam Stripping: Effect of the Flow Rate and Temperature. Typically, there is a temperature drop of about 17 °C between the flash zone and the bottom tray of the atmospheric column.⁸ The temperature drop is related to the vaporization of relatively light components being directly stripped by the steam. The question is whether a higher steam temperature reduces the residue yield. Because the atmospheric column does not allow a large flow rate of steam, which would result in the formation of free water on top trays, we use the flowchart of Figure 9 to perform this study.

The stripping power of steam as a function of the flow rate at different temperatures is depicted in Figure 10. To show the limit of the stripping power, the value of the steam/feed ratio is allowed to reach high values. Some of these values are unrealistic, but they help to illustrate the point. When the steam/feed ratio is small, the temperature of the steam does not affect the yield of residue because the heat taken from the oil is relatively small. When the steam/feed ratio is raised to about 40, the amount of heat taken by the steam becomes significant and the temperature of the steam starts to make a difference. Steam at 177 °C is worse than steam at 260 °C. On the 177 °C steam curve, a minimum point is observed at a steam/feed ratio of 175.

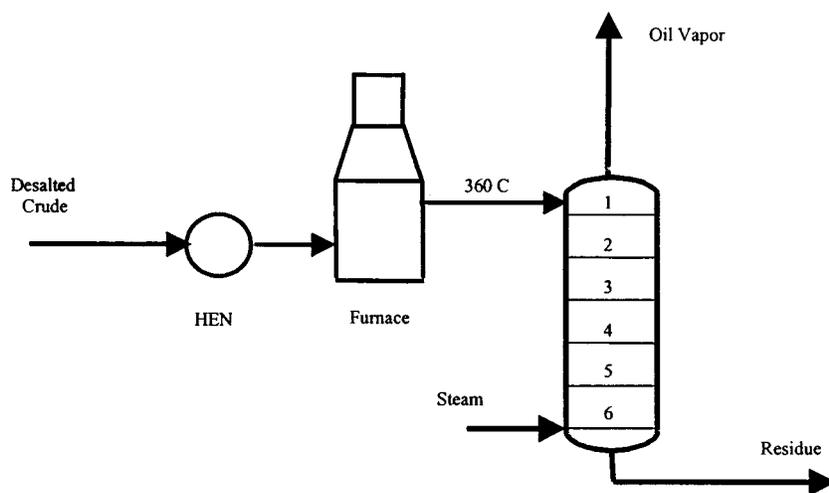


Figure 9. Flowsheet for studying the effect of light components.

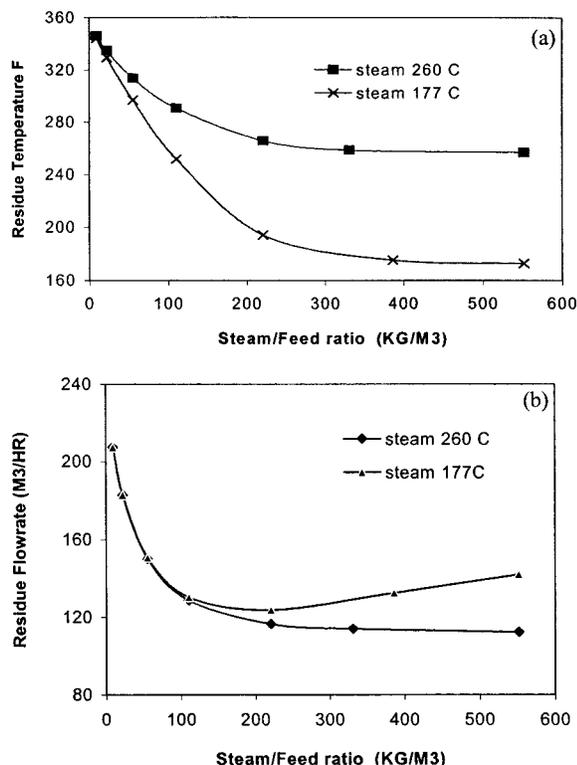


Figure 10. Residue yield and temperature as functions of the flow rate of stripping steam: (a) residue flow rate; (b) residue temperature.

This means that, above this ratio, the cooling effect of the steam dominates and the yield of residue goes up with an increasing amount of stripping steam.

Steam Injection. In this section, we discuss two cases: injecting steam into the crude before and after the furnace. Figure 11 is used for the simulation. The flash zone pressure is assumed to be the same as that of the outlet of the furnace. The results are shown in Table 5.

In both cases, the yield of residue decreases with the increased steam/feed ratio. For the same steam/feed ratio, however, injecting before the furnace produces less residue. This is because, in this scheme, the steam–crude mixture leaving the furnace has achieved a

Table 5. Effect of Steam Injection Location

steam/feed, kg/m ³	residue for SBF, ^a m ³ /h	residue for SAF, ^a m ³ /h	flash zone temp for SAF, ^a °C
0.0	207.3	207.3	360.0
3.3	202.7	204.1	359.4
5.5	199.8	202.0	359.1
11.0	193.0	197.2	358.3
22.1	181.6	188.6	357.0

^a SBF = steam injected at the inlet of the furnace. SAF = steam injected at the outlet of the furnace.

vapor–liquid equilibrium and no temperature change takes place between the outlet of the furnace and the flash zone. In the other case, the hot crude, which is a vapor–liquid mixture at 360 °C, meets with the injected steam (also 360 °C). Because of the stripping effect, more hydrocarbons vaporize. Because the vaporization process is adiabatic, the temperature of the system goes down. The last column in Table 5 shows the temperatures in the flash zone. Therefore, injecting steam before the furnace is more advantageous.

Energy Requirements for Stripping Agents. We showed that substances with large K values have the same stripping effect on a molar basis. However, the energy requirement for heating a gas to an elevated temperature varies. From the viewpoint of energy efficiency, gases requiring the lowest energy are the best. Figure 12 shows heating curves for water, N₂, H₂, and light hydrocarbons. Water has the largest heating requirement because of the heat of vaporization. Nitrogen and hydrogen have the smallest energy demands, and light hydrocarbons lie in between, with the energy demand increasing with their molecular weight. Therefore, nitrogen and hydrogen are the best stripping agents. The replacement of steam by N₂ or H₂ has the advantage that corrosion related to water formation in the distillation system is reduced or even eliminated. The drawback is that they need to be separated from the overhead product and recycled back. The flowsheet containing N₂ or H₂ stripping is therefore slightly more complex, and an economic analysis is needed to determine whether they are economically more beneficial than the conventional steam stripping.

Effect of the Pressure. The atmospheric tower operates at a pressure in the range of 0.2–0.3 MPa. Figure 13 shows the heating curves for *n*-butane and

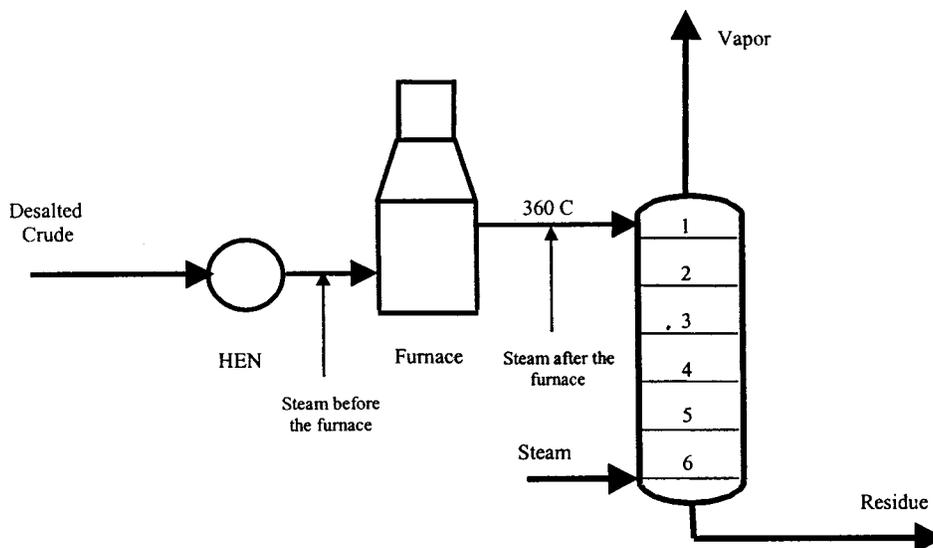


Figure 11. Flowsheet for studying the effect of steam injection.

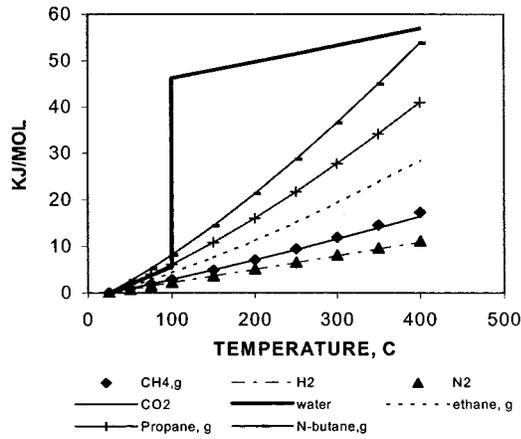


Figure 12. Enthalpy increase as a function of the final temperatures (base: 25 °C, 1 atm).

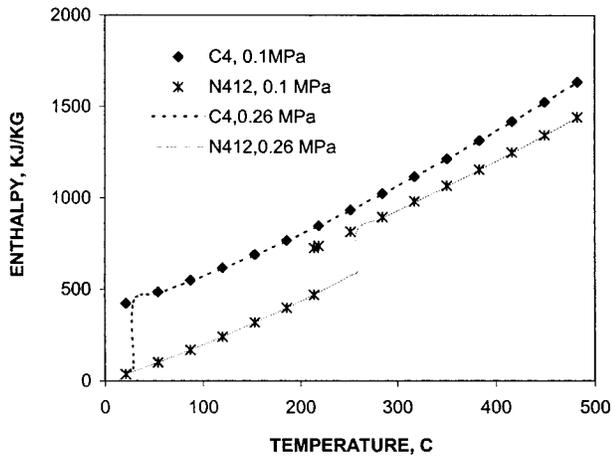


Figure 13. Effect of pressure on heating curves.

N412, which is a pseudocomponent with a molecular weight of 167.1, at 0.1 and 0.26 MPa, respectively. The figure shows that pressure only shifts the temperature at which vaporization takes place. It is also seen that, on a weight basis, the heat requirements for components with a large difference in molecular weight are very close.

From the above study, we reach the following conclusions:

(i) Steam stripping enhances vaporization by lowering the partial pressure of the hydrocarbon vapor. In the comparison made, it is possible to substitute the carrier effect of lights by injecting steam. For the light crude this is achieved by almost doubling (from 15 000 to 27 500 lb/h) the steam injection. The steam consumption is about 3 MW more, and the minimum utility increases by about 2.5 MW. However, the increased steam will either increase the flooding factor, in the case of an existing column, or cause the column to have a larger diameter, in the case of a grassroots design.

(ii) The stripping effect of a hydrocarbon not only depends on its property but also depends on the property of the mixture to be stripped.

(iii) The temperature of the bottom stripping steam does not affect the yield of the residue when the steam/feed ratio is less than 40.

(iv) Injecting steam before the furnace produces less residue than injecting steam after the furnace.

(v) N_2 , H_2 , and light hydrocarbons have the same stripping ability as water but require less heat to be

Table 6. Comparison between Conventional, Preflash, and Prefractionation Designs (Light Crude)^a

	conventional	preflashing	prefractionation
naphtha, m ³ /h	246.1	245.5	222.5
kerosene, m ³ /h	144.5	144.8	148.8
diesel, m ³ /h	70.7	71.6	71.5
gas oil, m ³ /h	103.6	102.9	105.7
residue, m ³ /h	230.5	230.6	227.5
product gaps, °C			
naphtha–kerosene	19.9	19.1	18.2
kerosene–diesel	0.4	0.0	−0.4
diesel–gas oil	−2.5	−1.9	−1.7
heating utility, MW	82.2	76.0	80.4
energy consumption, MW	92.5	87.7	90.4

^a Preflash temperature: 163 °C. Vapor feed at tray 15. $\Delta T = 22.2$ °C.

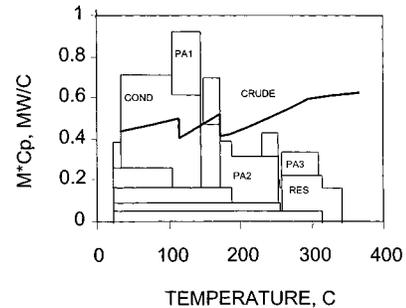


Figure 14. Final heat supply demand diagram for a preflash design ($\Delta T = 22.2$ °C). C: condenser. PA1, PA2, and PA3: pump-around circuits. RES: residue.

heated. From the standpoint of energy savings, they are better than water.

(vi) Pressure does not affect the heating curve except postponing the vaporization.

Finally, we can naturally envision two ways to increase the vaporization ratio at a fixed temperature, without resorting to increasing steam injection disproportionately. One is to put more light components into the crude, which can be achieved by adding stripping gases or recycling light distillates such as naphtha or kerosene. The other way is to remove heavy components from the crude. We leave the investigation of these points for the future.

Comparison for Lower AGO Yield

We now compare the preflash and prefractionation options with the conventional option at *lower AGO yields*. Table 6 summarizes the results. In this comparison, the stripping steams were adjusted so that both designs have the same amount of residue and the same product gaps.

The heat demand–supply diagrams corresponding to the optimal solution are shown in Figure 14. Not all of the heat possible to transfer from the second pump-around (PA2) to the third (PA3) was transferred because of the tradeoff between this load and the increased steam consumption created.

It is shown that the energy consumption of the preflash design is 4.8 MW lower than that of the conventional design. Thus, the energy consumption of the prefractionation design lies between the two designs. To understand the results, two aspects have to be taken into consideration: one is the temperature level of the heat to be recovered and another is the amount of components bypassing the furnace. Because the pre-

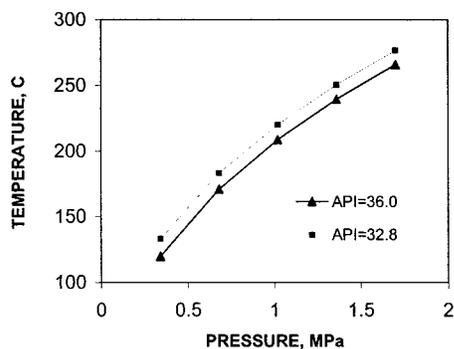


Figure 15. Crude oil bubble temperature as a function of pressure.

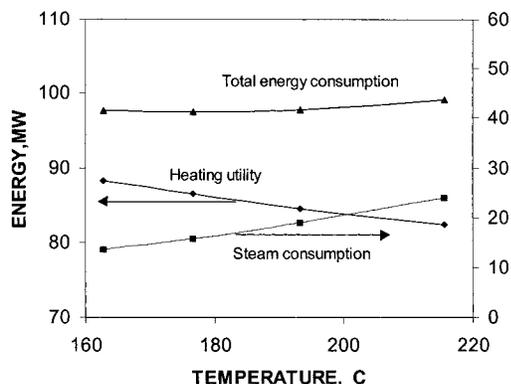


Figure 16. Effect of the pre-flashing temperature on total energy consumption.

fractionation column operates at a low temperature where the heat supply is in surplus, distributing heat between the prefractionation condenser and the main column condenser does not affect the heating utility requirement. Therefore, the only aspect affecting the energy consumption would be the amount of components bypassing the furnace. When the temperature of the feed entering the prefractionation column is the same as that entering the preflash drum, the amount of components bypassing the furnace in the former would be lower because a part of the vapor entering the column is condensed as the reflux and then mixes with the crude at the bottom of the prefractionation column. In terms of energy consumption, such a prefractionation design would be equivalent to a preflash design operating at a lower flashing temperature.

Note that product distributions vary considerably from one design to another, although the product gaps are exactly the same. For example, the preflash design gives more diesel and less gas oil compared to the conventional design. This might be preferable in some cases.

Effect of the Preflash Temperature. The necessity of a preflash drum depends on the property of the crude, the temperature of the crude entering the furnace, and the operating pressure. The bubble temperatures of the lighter crude are higher than that of the heavier crude (Figure 15). The higher the operating pressure, the higher is the bubble temperature. If a crude considerably exceeds the bubble temperature before entering the furnace, a preflash drum would be necessary.

The effect of the preflash drum temperature on the low AGO yield arrangement is shown in Figure 16. With the increase of the preflash temperature, the heating utility decreases continuously because of more light

Table 7. Effect of Vapor Feeding Locations

	tray 8	tray 15	tray 18	tray 27	tray 29
naphtha, m ³ /h	249.5	249.3	249.3	249.3	249.5
kerosene, m ³ /h	143.8	143.9	143.5	143.4	143.9
diesel, m ³ /h	69.6	69.7	70.3	68.6	68.7
gas oil, m ³ /h	101.9	101.9	101.8	103.5	102.3
residue, m ³ /h	230.6	230.6	230.6	230.6	231.0
product gaps, °C					
naphtha-kerosene	24.9	24.5	24.5	24.5	24.9
kerosene-diesel	3.8	3.2	2.1	2.5	4.0
diesel-gas oil	-0.9	-0.9	-0.8	-4.7	-0.9

components bypassing the furnace. On the other hand, the steam consumption increases rapidly as more steams are needed for both keeping the gas oil yield and fixing the product gaps. Although the total energy consumption reaches a minimum at 177 °C, the differences in the curve are relatively small. However, if steam is in surplus in the refinery, heating utility only counts and lower temperatures should be used.

Feed Location of Preflash Drum Vapors. The vapor produced in the preflash drum at 163 °C has properties very similar to those of naphtha. It contains 53% light ends (C₂-C₅). Its 98% boiling point is 169 °C, lower than the 95% temperature of naphtha (182 °C). We compared the effect of locating this feed in different trays in the column. During the comparison, steam flow rates have been kept constant, allowing the gaps to change. The results are given in Table 7.

The following is observed:

(1) The naphtha yield and the gap between naphtha and kerosene are constant. This is because the vapor feed is basically naphtha components and the naphtha-kerosene separation section (trays 1-9) is almost not affected in the above feeding locations.

(2) When the vapor feed location is lower than that of the gas-oil withdrawal (tray 25), the yield of diesel is significantly lower and the yield of gas oil is higher. This is because the temperature of the vapor feed is much lower than that of the vapor arising from the flash zone and a portion of the hot vapor is condensed with the effect of some diesel components going to the gas-oil draw stream. In Table 7, we can see that when the vapor feed tray is 27, not only does the yield of diesel decrease but also the diesel-gas oil gap moves in the negative direction. Therefore, the vapor feed between the flash zone and gas-oil withdrawal tray is not appropriate.

(3) When the vapor feed is sent to a tray above the flash zone, the residue yield is constant. This is because the residue yield depends on four factors: the composition and temperature of the crude at the outlet of the furnace, the flow rate of stripping steam entering from the bottom of the main column, and the over flash rate. In the above situations, all of the four factors are the same.

The temperature profiles are shown in Figure 17. The tray temperatures are highest when the vapor is fed at tray 8, the highest location. When the vapor is fed at tray 18, the tray temperatures above the vapor-feeding tray are lower, which is due to the cooling effect of the low-temperature feeding vapor. The tray temperatures below the vapor feed tray are exactly the same as when the vapor is fed at tray 8. When the vapor feed location continues to move down, the same tendency appears, as expected.

Finally, when the vapor location reaches its lowest location, the flash zone (tray 29), the temperatures of trays 29-34 are much lower than all of the others with

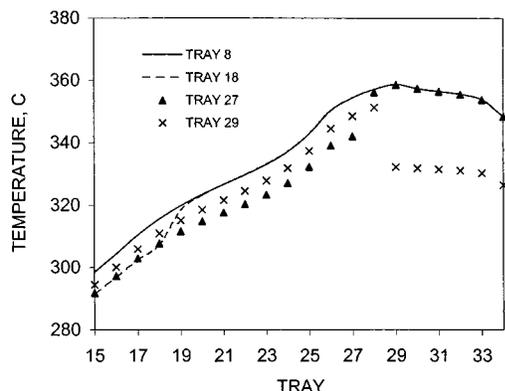


Figure 17. Temperature profiles for various vapor feeding locations.

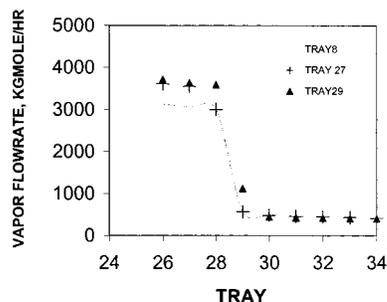


Figure 18. Vapor profiles as a function of the vapor feeding tray.

other vapor feed locations. At the flash zone, the hot crude flashes and splits into rising hot vapor and descending hot unstripped residues. The hot vapor meets with the cold preflashed vapor immediately, and condensation occurs. The condensed liquid, basically gas-oil and diesel components, mixes with the unstripped residue on tray 29. However, the condensed liquid does not stay with the unstripped residue and is stripped out when the mixture of the unstripped residue and the condensate meets the steam coming from tray 30.

Figure 18 illustrates that the vapor flow rate of tray 29 is about twice as large as that when the preflashed vapor is fed at other locations. The overall effect is equivalent to that of conventional design with a lower outlet temperature of the furnace. It can be predicted that heating the preflashed vapor to higher temperature would reduce the temperature drop at tray 29 and help increase the output of distillates.

From the above comparison, one can conclude that product yields and gaps are not sensitive to the vapor feed location. Specifically, the residue yield stays constant. Because a higher temperature profile is favorable in heat recovery through pump-around circuits, a higher vapor feed is preferred. However, as pointed out by Golden,⁶ a higher vapor feed would have the risk of contaminating the products withdrawn below the feed tray when liquid oil entrains.

Conclusions

A rigorous targeting for the design of a crude distillation unit with preflash drums and prefractionation is presented. The major conclusion of this study is that preflashing and prefractionation designs can only be

advantageous from the point of view of energy consumption if one accepts the yield of gas oil to be reduced. Other findings of the study show that, in practice, the equivalent of the carrier effect of lights cannot always be easily obtained using steam.

In addition, the introduction of a flash drum into an already complicated heat-exchanger network needs to be considered carefully. For example, it might be advantageous to pay the penalty of higher energy consumption just to see the number of units in the preheating train reduced. Finally, heat efficiency and network simplicity are desired for a variety of crudes simultaneously. This, together with measures to improve the performance of these units, will be analyzed in future work.

Nomenclature

- f_i = flow rate of component i in the feed
- F = feed flow rate, kgmol/h
- F^* = dry flow rate of the feed, kgmol/h
- K_i = vapor–liquid equilibrium constant for component i
- K_i^* = modified vapor–liquid equilibrium constant for component i
- L = liquid flow rate, kgmol/h
- l_i = flow rate of component i in the liquid phase
- P_{st} = partial pressure of steam in the vapor phase
- P = total pressure of the system without steam
- P^* = total pressure of the system with steam
- P_i^s = saturated pressure of component i at the system temperature
- V = vapor flow rate, kgmol/h
- V^* = vapor dry flow rate, kgmol/h
- x_i = molar fraction of component i in the liquid phase
- x_i^* = dry molar composition of component i in the liquid phase
- y_i = molar fraction of component i in the vapor phase
- y_i^* = dry molar composition of component i in the vapor phase
- z_i = molar fraction of component i in the feed
- z_i^* = dry molar composition of component i in the feed

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