On the Energy Efficiency of Stripping-Type Crude Distillation

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In this article, the energy efficiency of crude fractionation using stripping-type columns is analyzed to test previous reports indicating that this option is advantageous over the conventional design. Rigorous simulations demonstrate that, for the processing of a light crude, the stripping-type column cannot achieve the same small residue yield of the conventional case, unless the column is heated above the maximum allowable temperature that prevents coking. In addition, it is shown that operation of the stripping-type design at similar residue yields also provides no advantage over the conventional design.

Introduction

In conventional crude atmospheric distillation, the crude is fed at the bottom part of the column, and the main part of the column functions as a rectifying section. The conventional design is shown in Figure 1. Its connection to an indirect sequence of distillation columns is shown in Figure 2. Watkins[1] discussed the possibility of using direct and indirect rectifying sequences. In recent years, a design featuring the direct sequence (Figure 3) was proposed.[2]

In previous work,[3,4] 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 analysis. The procedure starts with a column without pump-around circuits. When heat is transferred from the condenser to the pump-around circuits, a tradeoff between the steam usage and the fuel gas savings is established. This transfer of heat is possible because of the operating and design flexibility that crude fractionation installations exhibit. Such design flexibility was studied in detail by Bagajewicz.[5] The procedure makes use of heat supply−demand diagrams similar to those introduced by Andrecovich and Westerberg[6] and Terranova and Westerberg.[7] On the basis of these targets, a universal heat exchanger network that allows several crudes to be processed with the minimal energy consumption can be found.[4]

In the stripping-type design, the crude is heated to a relative low temperature (about 150 °C) and fed at the top of the column. Because the crude temperature is low, the vapor-to-liquid ratio of the feed is small. The crude goes down the column and is heated consecutively in three heaters (upper, middle, and lower heaters). Side products are withdrawn from the vapor phases and rectified in side rectifiers.

Liebmann and Dohle[8] reported that this design featured a 5% lower utility cost than the optimized conventional design. However, the comparison was not on the basis of the same allowable temperature. The maximum allowable temperature is limited by the thermal stability of the crude being processed and is found by laboratory testing. For the Venezuela crude oil used in Liebmann and Dohle's paper,[8] the allowable temperature was 343 °C.[1] However, in their stripping-type design, the crude was heated to 370 °C in both the middle and lower heaters. It can be expected that, at this temperature, severe thermal cracking takes place. Because of the above limitation of their study, a reevaluation of the stripping-type design is necessary. This paper performs the evaluation using the method proposed by Bagajewicz and Ji,[3] taking into account the temperature limit of thermal cracking.

Crude Vaporization Patterns and Heat Demand

A major difference between the conventional design and the stripping-type design is the heating pattern of the crude oil. Because crude oil is heated step-by-step and the vapor is separated immediately after it is generated in the stripping-type design, the amount of crude to be heated in the stripping section is smaller than that in the conventional design. On the contrary, the amount of crude being heated in the conventional design is constant, that is, no vapor is separated until the heating is completed at the outlet of the furnace.

To better understand the difference between the two heating patterns, one should analyze the distillation curves. It is well-known that distillation curves can be used to approximate the distillation behavior in terms of product distribution and vaporization ratio. The two relevant distillation curves, ASTM D86 and EFV, are useful tools.

ASTM D86 distillation is carried out in a glass flask equipped with an electric heater and a water cooler. The oil sample is heated in the flask, and the resultant vapor is condensed and collected in a receiver at laboratory pressure. The temperature versus amount collected is recorded. Virtually no fractionation occurs in this distillation.

Equilibrium flash vaporization (EFV) is a test used to determine the vaporization ratios of crude oils as a function of temperature. In this test, the crude oil is continuously heated without separation of the vapor from the remaining liquid.[9] The temperature versus liquid amount is recorded. To obtain the EFV curve, a series of runs at different temperatures is carried out, and each run constitutes one point (of temperature and percentage vaporized) on the flash curve.

Equilibrium flash vaporization (EFV) distillation provides a relation between system temperature and the percentage of vapor generated. EFV distillation is a batch process during which no vapor is separated. In principle, the heating process in EFV is similar to that in conventional distillation. In practice, the EFV curve

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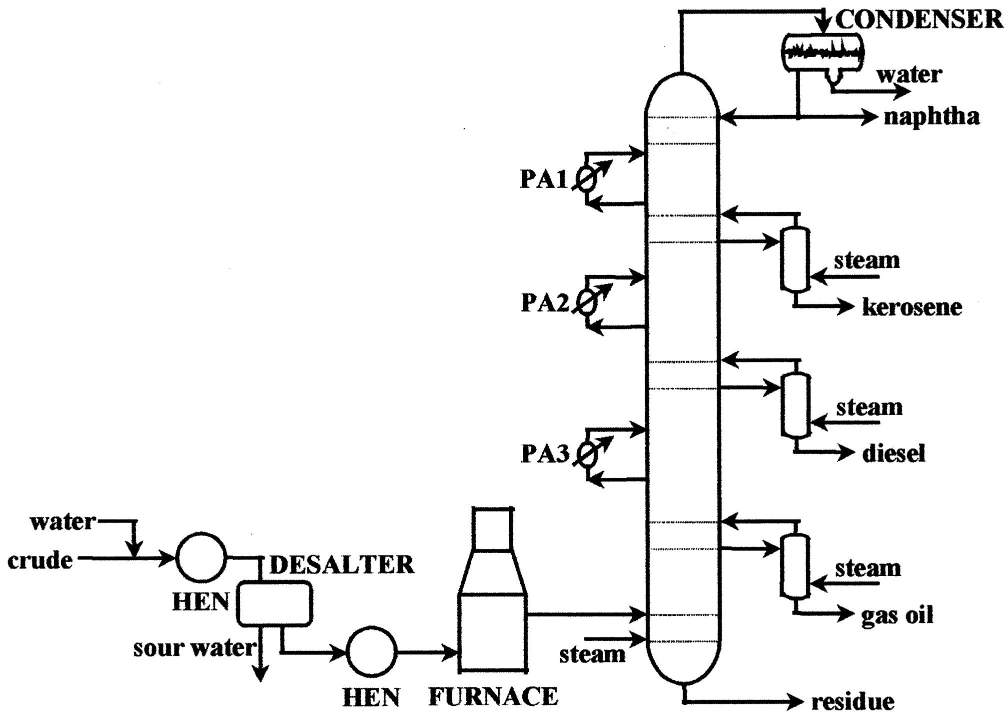

Figure 1. Conventional crude distillation.

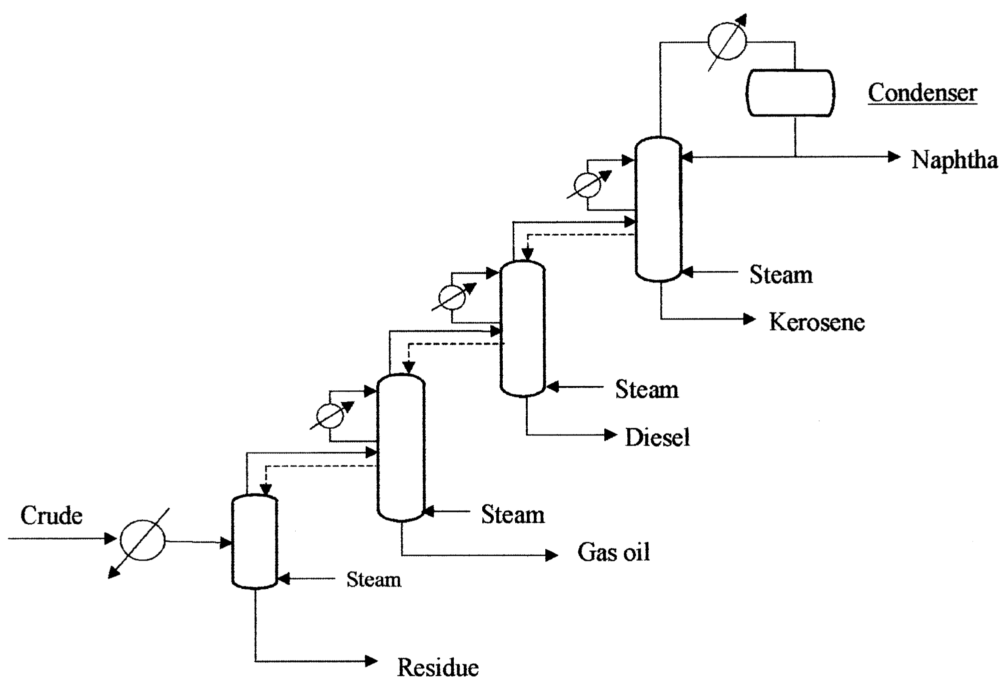

Figure 2. Indirect sequence.[1]

is used to estimate the vaporization ratio for the conventional design.[9]

Similarly, a close relation between ASTM D86 distillation and stripping-type distillation is apparent. For a stripping-type tower with an infinite number of heaters, a vapor-withdraw line between adjacent heaters, and zero pressure drop in the tower, one would obtain exactly the same temperature vs vaporization curve as for ASTM D86 distillation.

Now, the ASTM D86 and EFV curves can be used to answer two questions: (i) For the same temperature limit, which distillation option produces more vapor, that is is, more distillates yield? (ii) To achieve the same vaporization ratio, which distillation option demands more heat?

The first question can also be rephrased as follows: To achieve the same vaporization ratio, which distillation requires a higher temperature? This question is important because we want to achieve the maximal vaporization at the maximum allowable temperature.

Figure 4 shows a comparison of the two curves. At high vaporization, the ASTM curve is located above the EFV curve. This means that, at the same temperature, the ASTM distillation produces less vapor than EFV distillation. In this sense, the stripping-type design will likely produce a lower amount of distillates and more

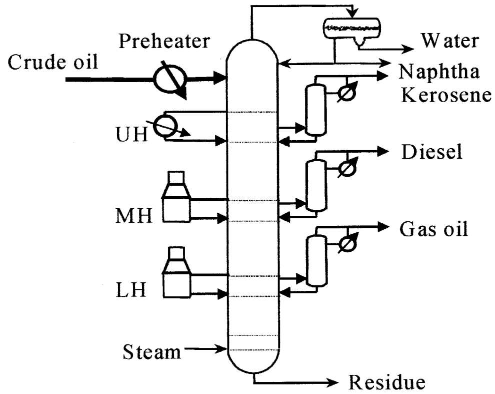

Figure 3. Stripping-type crude distillation column.

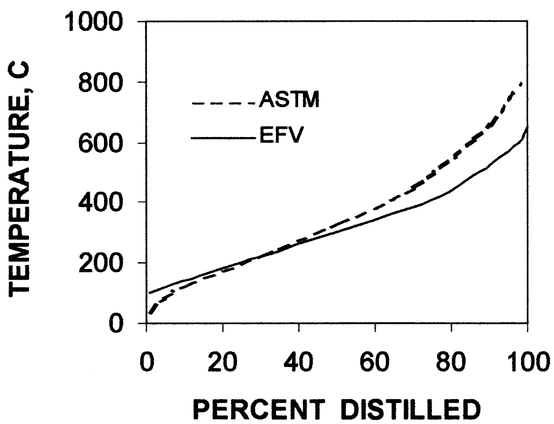

Figure 4. ASTM curve and EFV curve of the Venezuela crude. (Properties given in Appendix A.)

residue at the same temperature. Whether other operating variables, such as steam, could correct this problem remains to be answered.

The second question asks which heating pattern is more energy-efficient. From the viewpoint of energy efficiency, the stripping-type design reduces the heat demand from the light components (naphtha and kerosene), which are only heated to low temperatures. Conversely, in the conventional design, light components are heated to the temperature limit. The heating pattern for the stripping-type design (ASTM) is, however, not advantageous for the heavy components (gas oil). As mentioned before, the distillation temperature required for the ASTM distillation to achieve the same vaporization ratio is higher. This means that the heavy components vaporize at a higher temperature than they would in the EFV distillation. Therefore, the heat demand for heavy components is higher.

Whether the ASTM process requires less energy to achieve the same vaporization ratio, ignoring the temperature limit of thermal cracking, will depend on the tradeoff between the heat savings for light components and the larger energy consumption for heavy components. To obtain the ASTM heating curve, nine heaters and nine separators were used to simulate the ASTM distillation. Figure 5 shows part of the flow sheet used for this simulation. The crude was heated at 760 mmHg step-by-step, and vapors were withdrawn at each step. The resulting curves for heat demand are shown in Figure 6.

Figure 6 reveals that, for small vaporization ratios (less than 0.25), the two curves coincide. For larger vaporization ratios, however, the heat demand for ASTM distillation is always higher than that for EFV distillation. To explain this finding, a comparison of temperatures and heat demands needed to achieve 0.473 vaporization is listed in Table 1.

Figure 7 shows the ASTM and EFV curves for a heavy crude. It is expected that heavy crude should exhibit behavior similar to that of light crude. For heavy crude, the desire to increase the yield of distillates is stronger. In this sense, the stripping-type design is even worse for a heavy crude than for a light crude.

We therefore conclude that one should expect less vaporization of heavy fractions in the stripping-type design than in the conventional design when both are operated at the same maximum temperature. In addition, the duty is expected to be higher in the stripping-type design for the same level of vaporization, i.e., for the same distillates yield.

Features of a Stripping-Type Design

As discussed in the previous section, the stripping-type design has to heat the crude to a higher temperature to achieve the same amount of vaporization as in the conventional design. In addition, light components help vaporize heavy components at lower temperatures. This is called the carrier effect and it was discussed in detail by Ji and Bagajewicz.[10] Such light components are not present in the stripping-type design.

The second feature is the different pattern of product composition. In the conventional design, all light components come from the bottom section of the tower and go through the trays where side products are withdrawn. As a result, light components appear in each side product withdraw line and have to be stripped off in the side strippers. In industry, this is called controlling the flash point. In contrast, the light components in the stripping-type design come from the top section and are withdrawn as soon as they are vaporized, so they do not reach the trays where diesel and gas oil are withdrawn. Thus, the problem of the presence of light components does not exist. However, heavy components have a large chance of being present in these side products and raise their end points.

The last feature is that the products are withdrawn in the vapor phase rather than in the liquid phase. For the same composition of mixture, the dew point is higher than the boiling point. Thus, by withdrawing vapor, one can obtain the heat at a higher temperature level. This is advantageous from the point of view of energy efficiency. However, vapor-phase withdrawal also brings the problem of corrosion in the side condensers if water is present.

Energy Targeting

The main column in Figure 1 contains 34 trays. A crude of API 36 is used in the simulation. The flow rate used is 5000 BBL/h. The crude data were taken from a previous article.[3] The properties of the crude are listed in Appendix B. Table 2 shows the product specifications and limiting values of gaps.

In simulating the stripping-type distillation, the feed temperature; the duties of the upper, middle, and lower heaters; and the duties of the condensers are allowed to change to minimize the residue flow, that is, to maximize the distillates yield while meeting the specifications and not overheating any tray above the maximum temperature.

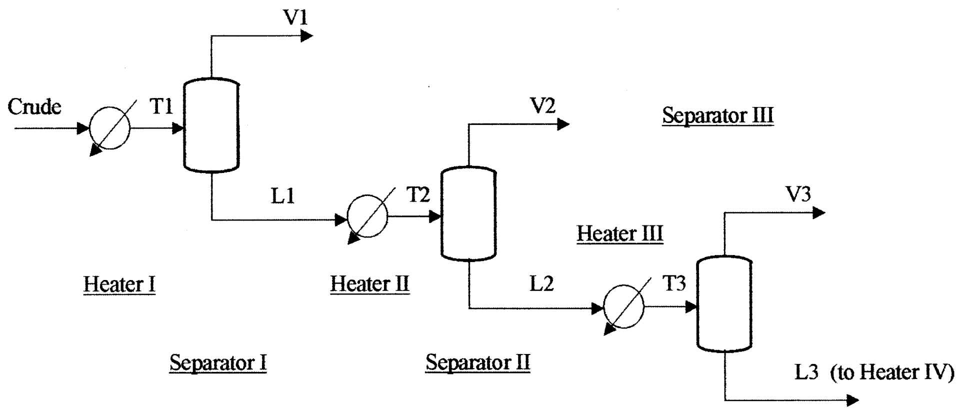

Figure 5. Simulation of ASTM D86 distillation. (Heaters 4−9 and separators 4−9 are not shown in this figure.)

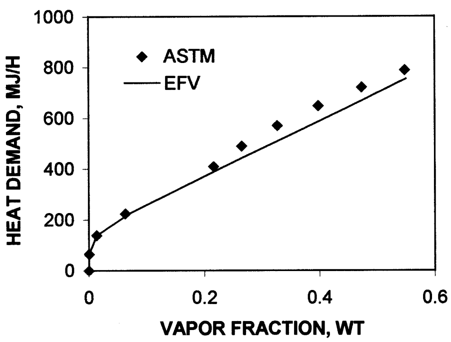

Figure 6. Heat requirements for ASTM distillation and EFV distillation at 760 mmHg (1 M^3/h,Venezuela crude; see Appendix A).

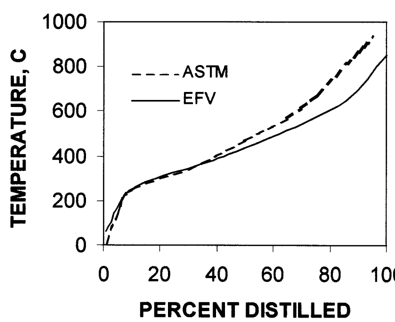

Figure 7. ASTM and EFV curves for a heavy crude (see Appendix B).

Table 1. Comparison of Temperature and Heat Demand

	temperature (°C)	wt fraction of crude vaporized	weight of liquid remaining	heat demand (MJ/h)
EFV	100	0.014	0.986	138.8
	307	0.473	0.527	670
ASTM	307	0.345	0.655	582
	380	0.473	0.527	720.3

Figure 8 is a heat demand−supply diagram for the stripping-type design. The solid bold line represents the crude heat demand curve, which is composed of the preheating section and the upper, middle, and lower side heaters. The heat supply includes heat from the main condenser, side condensers, products, and residue. The demand curve is not continuous because of the temperature gaps between adjacent heat sinks.

The discontinuity in the heat demand curve is an important feature. It affects the match between the heat demand and the heat supply. Consider the region between the preheating curve and the upper heater. In this region, heat supply is available from the side condensers and the residue. However, there is no heat demand in this region. The only way to make use of this heat supply is to cascade the heat to regions at lower temperatures. In the preheating region, a heat deficit is present (shaded region), but it is much smaller than the aforementioned heat supply. Therefore, a large part of the heat supply cannot be utilized. Finally a small match between supply and demand exists at the level of the upper heater.

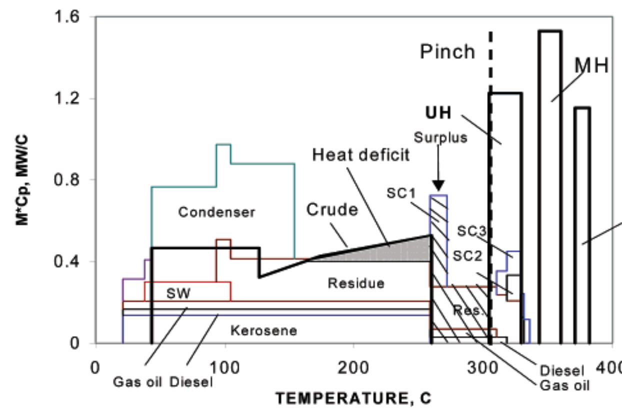

Figure 8. Heat demand−supply diagram. (SW is saline water; SCi represents side condensers; UH, MH, and LH are the upper, middle, and lower heaters, respectively).

Table 2. Stripping-type Column Specification

	specification
naphtha D86 95% point (°C)	182.2
kerosene D86 95% point (°C)	271.1
diesel D86 95% point (°C)	326.7
tray 29 temperature (°F)	360
(5−95) gaps (°C)	
kerosene−naphtha	16.7
diesel−kerosene	0
gas oil−diesel	−13.9

The location of the pinch point can be easily obtained from this diagram. It is the lowest temperature at which the demand is larger than the supply after shifting and area matching has been performed (Figure 8). The heating utility is given by the unmatched demand on the right, and the cooling utility is given by the extra supply on the left.

Effect of Feed Temperature. The heating utility can be reduced by raising the temperature of the feed, that is, by shifting part of the upper heater duty into the crude preheater. The solid curve in Figure 9 represents the new heat demand curve for a higher feed temperature, and the dashed line corresponds to the

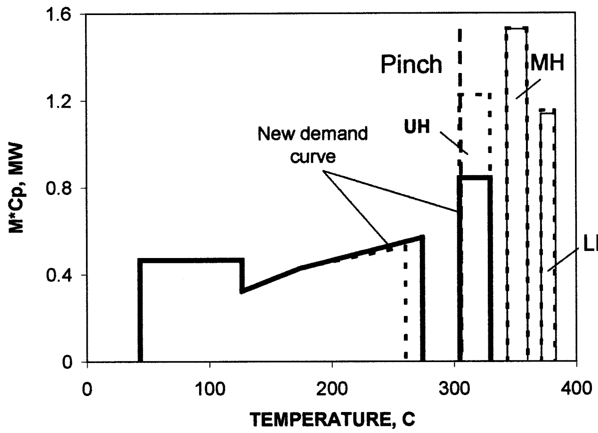

Figure 9. Effect of increasing the feed temperature on heat demand (UH, MH, and LH are the upper, middle, and lower heaters, respectively.)

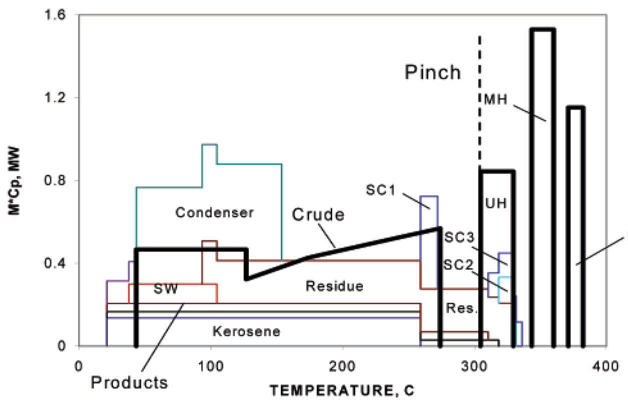

Figure 10. Complete heat demand−supply diagram for the higher feed temperature (SW is saline water; SCi represents side condensers; UH, MH, and LH are the upper, middle, and lower heaters, respectively).

base case. Simply put, there is a transfer of heat demand from the upper heater to the crude preheater. This favors heat integration, as can easily be seen in Figure 10, which shows the complete diagram for the higher feed temperature.

Note that the location of the pinch does not change at all, however the heat deficit in the upper heater region decreases significantly. The demands of the middle and lower heaters remain the same.

A comparison of the strategy used above with that for the conventional design reveals a few important issues. With a fixed heat demand curve (the crude curve) in the conventional design, the strategy is to move the extra heat supply to higher-temperature regions, which is achieved by redistributing the pump-around duties.[3] The heat demand curve in the stripping-type design can be modified by redistributing heat among the preheater and the side heaters. Here, the strategy is to move some heat demand from the high-temperature region (the upper heater region) to the low-temperature region (the preheater region).

However, there is a limit on the heat shift from the upper heater to the preheater. Table 3 shows that the naphtha−kerosene gap decreases with increasing feed temperature. This is because the vapor arising from the upper heater decreases in amount as a result of a lower upper heater duty. In the section of the column between the feed tray and the upper heater, the vapor strips off light components from the descending crude.

Table 3. Effect of Feed Temperature in Stripping-type Design

	237.8 °C	260 °C
preheater duty (MW)	60.0	72.4
upper heater duty (MW)	42.3	31.9
naphtha yield (M^3/h)	244.8	235.2
kerosene yield (M^3/h)	141.0	155.1
naphtha−kerosene gap (°C)	18.8	5.3

Table 4. Comparison of Product Gaps, Yields, and Energy Consumption

	conventional	stripping-type
naphtha−kerosene gap (°C)	16.7	16.7
kerosene−diesel gap (°C)	0.0	13.6
diesel−gas oil gap (°C)	−13.2	−13.9
naphtha yield (M^3/h)	244.7	242.1
kerosene yield (M^3/h)	137.8	157.4
diesel yield (M^3/h)	52.8	35.9
gas oil yield (M^3/h)	43.9	35.0
residue yield (M^3/h)	316.4	324.5
heating utility (furnaces, MW)	45.6	49.5
steam consumption (MW)	19.1	16.4
energy consumption (MW)	59.0	61.0

When the vapor flow rate is too low, the vapor cannot efficiently remove the light components from the crude, and the crude entering the upper heater will carry a significant amount of light components. In the upper heater, these components vaporize along with kerosene and enter the kerosene rectifier. As the rectifier has no way to remove light components, all of the light components are present in the kerosene product, contributing to a lower 5% D86 boiling temperature.

Comparison with the Conventional Design

A comparison can be made with a conventional column operating at maximum energy efficiency with the same product specifications. Such a conventional column is also run to minimize the residue yield. This is achieved by simply running the flash zone at the maximum allowable temperature and maintaining a small overflash ratio (3%). The comparison is therefore fair because two optimized designs are used, the quality of the products is the same, and a temperature limit is enforced in both cases.

Such a comparison reveals that the residue yield of the stripping-type design is about 70% larger than that of the conventional design for the same temperature limit of 360 °C. This yield cannot be further reduced either by increasing steam injection or by adjusting the feeding temperature and/or the duties of the heaters. Therefore, it is impossible to make a meaningful comparison of energy efficiency when the same level of separation is not achievable. Therefore, to determine whether the stripping column could be competitive at a similar residue yield, the conventional design was run at this large yield. As a result, the maximum temperature in the conventional column decreased to 324 °C. Although the yields of naphtha are almost the same, the yield of kerosene is significantly higher, and the yield of diesel is lower in the stripping-type design (Table 4). This is because, in the conventional design, some of the light components, which constitute the kerosene, are carried away by the diesel and the gas oil streams.

Some of these light components are carried out with the diesel and the gas oil. In the stripping-type design, the crude oil enters the column at the top. Therefore, the carrying of kerosene in the heavy products is ruled

Table 5. Relative Costs for the Two Designs

	conventional	stripping-type	reason for higher cost
main tower	similar	similar[a]	
furnace	lower	higher	heating more cold streams
heat exchangers	similar	similar	
side columns	same	same	
desalter	same	same	
main condenser	same	same	
PA exchangers/ side condensers	lower	higher	corrosion
total cost	lower	higher	

[a] 10% smaller, but fouling problem arises.

out. As expected, in the stripping-type design, the gap between kerosene and diesel is much higher than that in the conventional design. Finally, the energy consumption is calculated by adding the minimum heating utility of the heat exchanger network and 70% of the steam consumption to account for cost differences. The minimum utility is calculated using pinch analysis.

Simulations were also performed for the stripping-type design using a higher temperature limit (399 °C) to obtain the same yield of residue as in the conventional column (360 °C). It is at such higher temperatures, which are not recommended in practice, that the stripping-type design achieves around 5% lower total energy consumption, as reported earlier.[8] These conditions were used in a series of simulations using the light crude of this paper, and the same trend was observed.

The above comparison shows that the stripping-type design cannot achieve the same distillates yields as the conventional design for the same allowable high heating temperature. At low maximum temperatures, when similar residue yields are achievable, the energy consumptions are similar, and one might speculate that the investment cost might make a difference because, in the stripping-type design, the crude is heated to a lower temperature, and therefore, a less complex preheating train might be needed. Thus, the tradeoff between operating and investment costs needs to be analyzed.

The relative costs of the two designs are shown in Table 5. Preliminary size estimation shows that the main tower for the stripping-type design is slightly smaller (about 10%) than that of the conventional design. However, in the stripping-type design, the dirty crude goes through most of the trays. The dirty material reduces the tray efficiency and causes fouling problems. Therefore, additional trays are needed to offset the lower efficiency, and special designs are required to reduce fouling. It is expected that the saving in tower size is canceled by the additional cost for the trays.

The side condensers used in the stripping-type design are considered more expensive than pump-around heat exchangers in the conventional design because the side condensers have to deal with corrosion problem caused by steam condensation. The corrosion problem becomes more severe when processing high-sulfur crudes.

The investment cost for the furnace in the stripping-type design is considered higher because it handles as many as four cold streams. The internal structure will be more complex than that of the conventional design where only one cold stream exists. The energy consumption is higher as well.

The relative cost for all of the other heat exchangers is not easy to estimate accurately without designing the heat exchanger network. However, a comparison can be made on the basis of the total number of exchangers and estimates of the area.

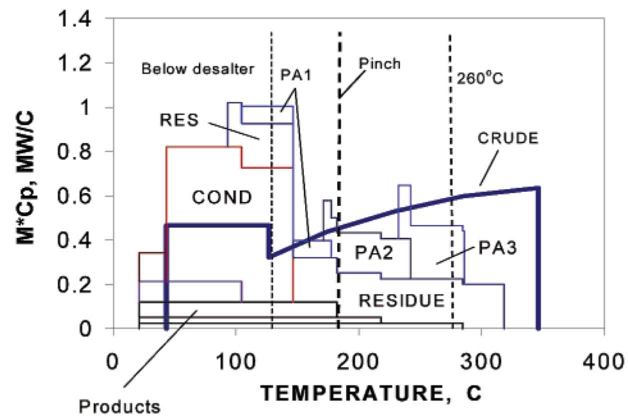

Figure 11. Heat demand−supply diagram for conventional design. (Products from the bottom up are gas oil, diesel, kerosene, and naphtha; PAi is the pump-around heat duty; RES is the residue).

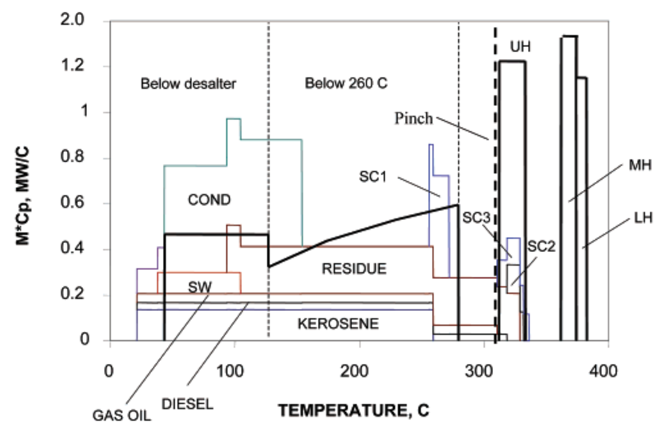

Figure 12. Heat demand−supply diagram for stripping-type crude distillation. (SW is saline water; SCi represents side condensers; UH, MH, and LH are the upper, middle, and lower heaters, respectively).

For the case of the conventional column in the region above the desalter (Figure 11), there are one cold and seven hot streams (PA1, PA2, PA3, residue, gas oil, diesel, and kerosene). Because of the existence of a pinch, a rigorous design of the heat exchanger network, as reported by Bagajewicz and Soto[4], would be quite complicated. However, the use of the $(N - 1)$ rule[11] can still provide a good approximation of the number of units. If the aforementioned rule is used, then seven heat exchangers and a heater are needed. The condenser is not included because the heat is in surplus and is cascaded into the region below the desalter. In the region below the desalter, there is a large heat surplus. Because the heat from the condenser is large enough, only one exchanger is needed. In addition, the four distillates and the condenser have to be cooled to the final temperatures, requiring five coolers. Therefore, the total number of exchangers is 13.

For the stripping-type design above 260 °C, three exchangers are needed for the crude to extract heat from SC2, SC3, and the residue (Figure 12).

The heat supply to the left of the upper heater has to be cascaded to lower temperatures for utilization. Above the desalter and below 260 °C, there are five hot streams: residue, SC1, kerosene, diesel, and gas oil. Similarly to the conventional design, the condenser is not used. The number of exchangers is five in addition

Table A1. TBP Data for Tia Juana Light

percent distilled	temperature (°C)	percent distilled	temperature (°C)
0	−3.0	70	462.9
5	63.5	90	680.4
10	101.7	95	787.2
30	221.8	100	894.0
50	336.9		

Table A2. Light Ends for Tia Juana Light

component	percent of assay
C2	0.048 76
C3	0.3762
I−C4	0.2774
N−C4	0.8908
total	1.5932

Table B1. Feedstock Used for the Design

crude	density (kg/m^3)	throughput (m^3/h)
light	845 (36.0 API)	795
heavy	934 (20.0 API)	795

Table B2. TBP Data

	temperature (°C)	
vol %	light crude	heavy crude
5	45	133
10	82	237
30	186	344
50	281	482
70	382	640
90	552	N/A

Table B3. Light-Ends Composition[a] of Crude

compound	light crude	heavy crude
ethane	0.13	0
propane	0.78	0.04
isobutane	0.49	0.04
n-butane	1.36	0.11
isopentane	1.05	0.14
n-pentane	1.30	0.16
total	5.11	0.48

[a] Composition in volume percent.

to the heater. Below the desalter, one exchanger is counted for heat exchange between the condenser and the crude. Four coolers are required for cooling the products. The total number of the exchangers aside from the heater is 13.

Both designs use roughly the same number of heat exchangers. If the number of exchangers is the dominating factor in pricing, we expect the costs for the exchangers to be similar. However, if the area is an issue, then the demand−supply diagrams also provide a means to estimate these areas comparatively. Indeed, the area is proportional to the area below the hot stream curves (as these streams are the ones that are in surplus). A quick comparison reveals that the areas are also similar. It is then concluded that the stripping-type design does not require a lower investment cost than the conventional design.

Conclusions

In this article, the stripping-type and conventional designs for crude fractionation were first compared when they are run at minimum residue yield, with the temperature in the columns required to stay below a maximum limit and the same product quality maintained. The result of this comparison was that the stripping-type design cannot achieve the same low yield of residue as the conventional design, which makes an energy efficiency comparison pointless. Next, the same yield of residue was imposed, with the temperature required to stay below a maximum value and the same quality of products maintained. The result is that, for the crude chosen, the energy efficiencies are similar. When the maximum temperature limit was increased significantly over the coking limits for the stripping-type design, the energy efficiency matched the 5% improvement over the conventional case reported earlier. Finally, the tradeoff between capital and energy costs was also investigated to determine whether a large reduction in capital would justify using the stripping-type distillation, revealing that this is not the case. The conclusion is that this stripping-type design is not competitive.

Appendix A

The properties of the light crude are presented in Tables A1 and A2. They are taken from Watkins.[1] The crude is Tia Juana light (Venezuela) having an API gravity of 31.6.

Appendix B

The properties of the crude (API 36) used in Tables 3 and 4 are reported in Tables B1−B3. They are taken from Bagajewicz and Ji.[3]

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