Overview of Crude Units

Crude units are the first units that process petroleum in any refinery. There objective is to separate the mixture into several fractions like naphtha, kerosene, diesel and gas oil. A schematic diagram of an atmospheric crude fractionation unit is shown in Figure 1-1.

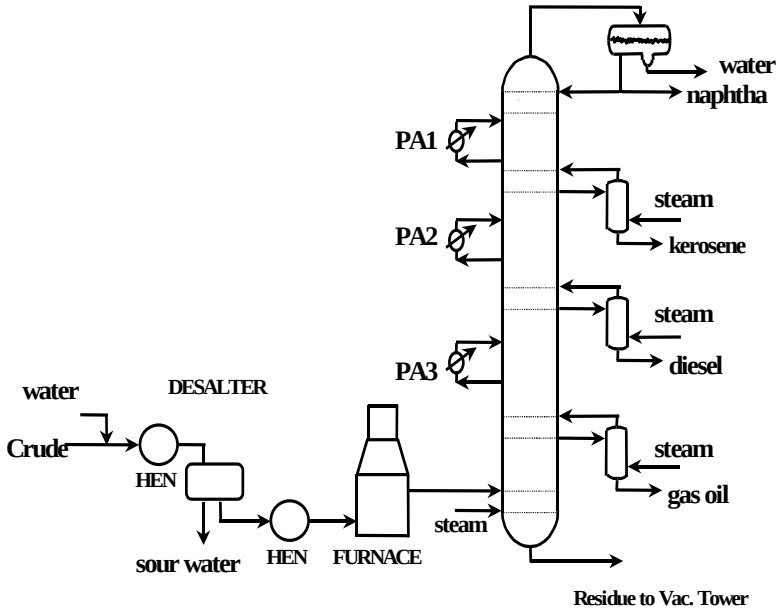

Figure 1-1: Atmospheric Crude Unit.

Crude oil contains salts which can be harmful to downstream equipment and must be removed. To remove the salts, water is mixed with the crude oil and typically heated to temperatures between about 215 ^{o}F to about 280 ^{o}F and allowed to separate in the desalter. The desalted crude enters another heat exchanger network. Both heat exchanger networks make use of the vapors of the main column condenser, the pump-around circuit streams (PA1, PA2 and PA3), and the products that need to be cooled. Then, the preheated crude enters the furnace, where it is heated to about 340-372 ^{o}C (644-700 ^{o}F). The partially vaporized crude is fed into the feed region (called flash zone) of the atmospheric column, where the vapor and liquid separate. The vapor includes all the components that comprise the products, while the liquid is the residue with a small amount of components in the range of gas oil. These components are removed from the residue by steam stripping at the bottom of the column. Products are withdrawn from the side of the column and side strippers are used to help controlling the composition of light components. In addition, to more effectively remove heat, liquid is extracted at various points of the column and cooled down to be reinjected at a different position on the column. Cooling water and sometimes air coolers are used in the heat exchangers PA1, PA2 and PA3, but it is always more advantageous to have these streams release their heat to the raw crude oil in the heat exchanger networks (HEN), usually called pre-heating

trains. Several different designs and configurations for the heat exchanger network in the conventional crude oil distillation unit are possible. Figure 1-2 shows one particular instance of a preheating train, not necessarily the best or most recommended one (efficient ones are discussed later). In addition, in some oil distillation units, gas oil is not produced and instead becomes part of the residue. Such units contain one less side-stripper and one less pump-around than those shown in Figure 1-1 and Figure 1-2. Further, in units in which gas oil is not produced, the diesel may be further separated into heavy and light diesel.

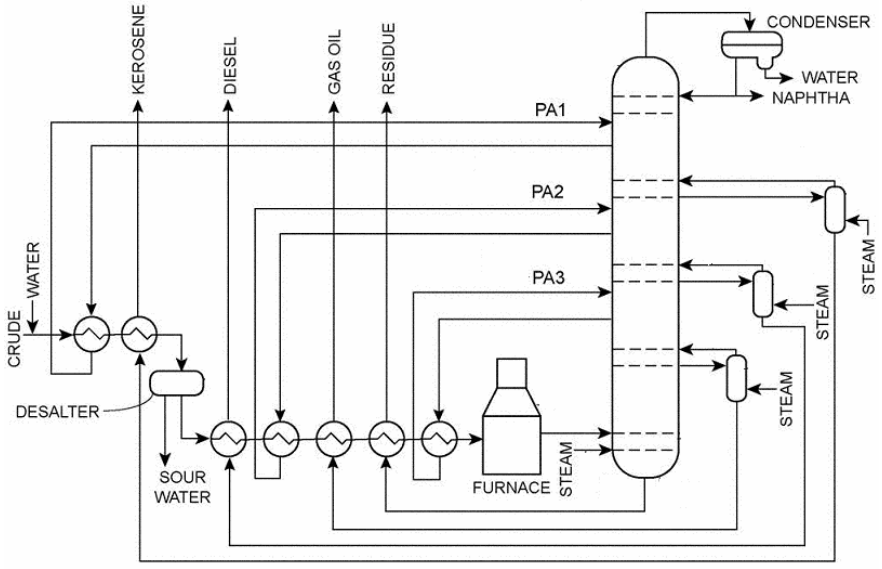

Figure 1-2: Atmospheric Crude Unit-with preheating train.

The topped crude leaving the atmospheric tower still contains significant amount of valuable oils. These oils cannot be distillated at atmospheric pressure because the temperature required would be so high that severe thermal cracking takes place. Figure 1-3 depicts such a unit. Note first that this unit does not have a condenser and does not feature side strippers either, simply because products do not have specifications on their light end. However, side strippers can be used in specific cases, such as lube base oil production.

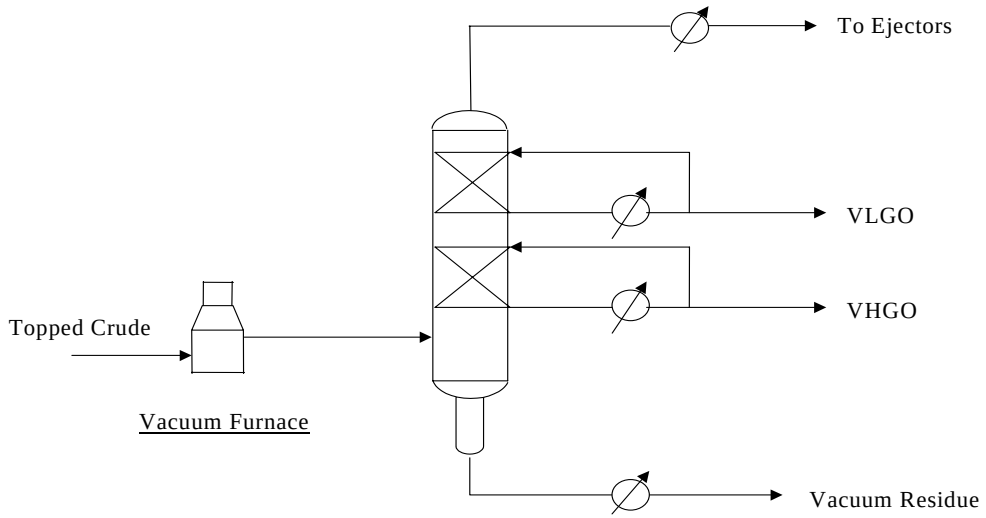

Figure 1-3: Atmospheric Crude Unit-with Preheating train.

Figure 1-3 shows that light vacuum gas oil (LVGO) and heavy vacuum gas oil (HVGO), are produced. Sometimes, depending on its properties, LVGO is blended with other products like atmospheric diesel. Both are typically used as feed to fluid catalytic cracking units. The vacuum distillation consists of the vacuum furnace, vacuum tower and the vacuum producing system. The topped crude is heated up in the vacuum furnace to about 400 °C. The temperature is controlled to be just below the temperature of thermal decomposition. Although a single cut of vacuum gas oil (VGO) is allowed in some cases, drawing LVGO and HVGO separately is more beneficial from the point of view of energy savings, because the resultant HVGO draw temperature is 90-120 °C higher than the corresponding draw temperature of a single VGO cut. Lighter components are removed from the residue by steam stripping. In addition coke formation is reduced by circulating partially cooled bottoms to quench the liquid to a lower temperature. Because the heavy crude fraction contains metal complexes (asphaltenes and porphyrines), which are catalyst poisons for downstream processes, sometimes a recirculation of “wash oil” in the bottom part (not shown in the figure) is included to prevent these compounds to reach the HVGO.

Vacuum distillation is widely used to produce catalytic cracking plant feed stocks of low carbon content. It is also used to produce lube oil fractions. In practice, the uses of vacuum distillates and residue depend on the type of crude oil feed, the type of refinery and its downstream processing capacities. The major specifications for the most common products are (Watkins, 1979):

- *Catalytic cracking feedstock*: Gas oils for catalytic cracking feedstock require a strict separation between distillate and residue. The amount of carbon content in these oils should be minimal while not sacrificing gas oil recovery. This is necessary to reduce

coke forming on the cracking catalyst. The metals content, particularly vanadium and nickel, should be controlled because they are severe catalyst poisons.

- *Hydrotreaters or Hydrocrackers Feedstock*: These can tolerate a slightly higher metal content because the hydrogenation catalysts contain these metals. However, the amount of carbon and asphalt materials should be minimal to prevent coke formation.
- *Distillate fuel oils*: Distillate fuel oils to be used directly are specified by API gravity, viscosity, metal content and flash point. Atmospheric boiling ranges can be used to define the separation.
- *Vacuum residue*: The properties of vacuum residue can be defined in several ways. When distillate production is to be maximized, the amount of gas oil allowed remaining in the bottoms stream must be minimized. The residue is blended into residual fuels. In this operation, one can normally set the volume percent of either the whole crude or the topped crude that is to be yielded as vacuum residuum.

In the preheating train, the crude is under pressure to suppress vaporization. In the case of a light crude, the pressure required to suppress vaporization is too high. The solution is to separate some light components before heating the crude further in the preheat train (Figure 1-4). The light components separated in the pre-flash drum are sent to the column directly.

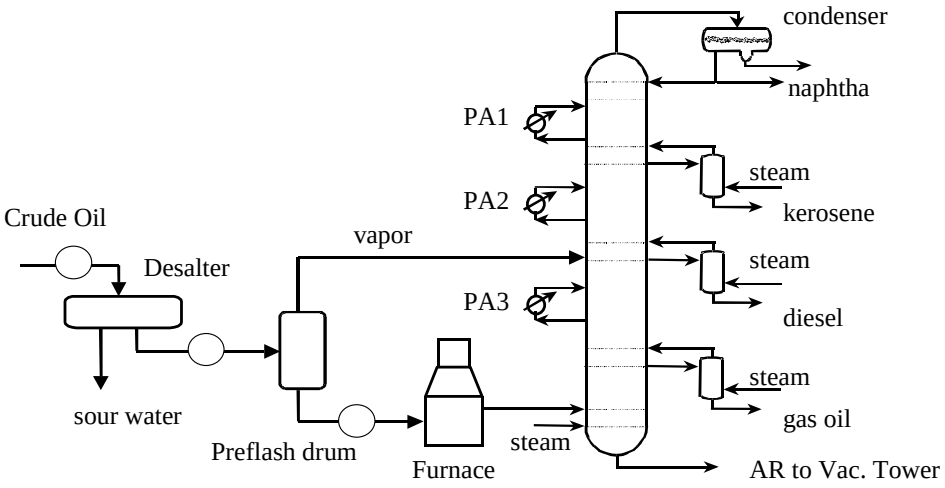

Figure 1-4 : Pre-flash design

In the pre-fractionation design (Figure 1-5), the light components are separated in a pre-fractionation column. Thus, in the pre-flash design, components in the range of naphtha are condensed in the condenser of the atmospheric tower, while in the pre-fractionation design, these components are split into two fractions: light naphtha condensed in the condenser of the pre-fractionation condenser and heavy naphtha condensed in the condenser of the atmospheric tower.

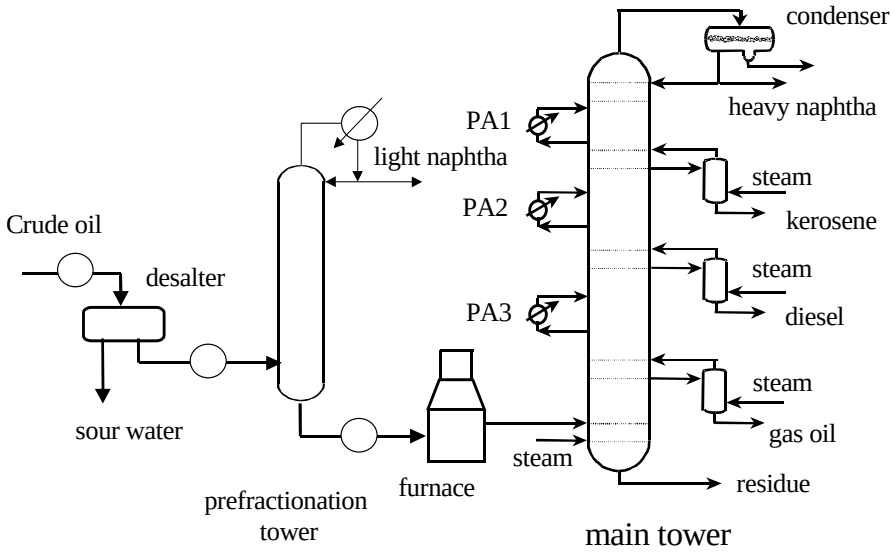

Figure 1-5: Pre-fractionation design

The pre-fractionation design, however, is considered to be similar to the pre-flash design when energy consumption is considered (Bagajewicz and Ji, 2002). Because the condensation heat for naphtha (or light naphtha plus heavy naphtha) is constant, the only difference is that the pre-fractionation design provides this heat in two condensers with different temperatures. When there is significant heat surplus in the temperature range of the condensers (intermediate and light crudes), the difference does not affect energy consumption.

Finally, in recent years, an alternative stripping-type process was proposed (Figure 1-6), the crude is heated to a relative low temperature (about 150°C) and fed close to the top of the column. Because the crude temperature is low, the vapor ratio of the feed is small. The crude goes down the column and is heated consecutively in three heaters (the upper heater, middle heater and lower heater). Side products are withdrawn from the vapor phases, and rectified in side rectifiers. Although some preliminary studies indicated that these units could be advantageous from the point of view of energy consumptions, it was later proven that they are actually less energy efficient.

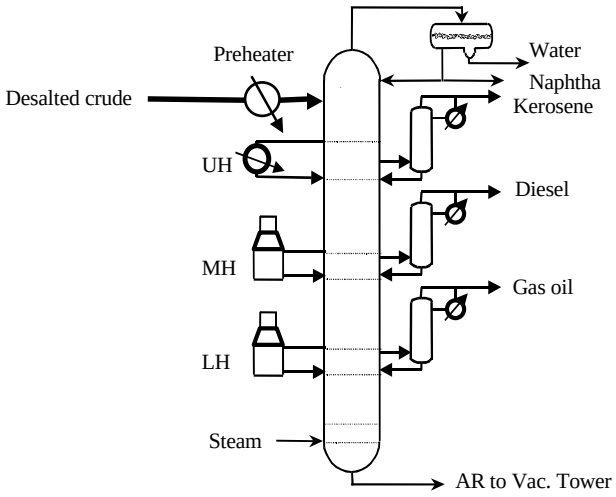

Figure 1-6: Stripping-Type design.

In the framework of sharp split distillation sequences, the conventional design is an indirect sequence and the stripping-type design is a direct sequence (Liebmann, 1996). In other words, these two designs are located at the two ends (complete direct and indirect sequences) of the spectrum of distillation sequences. The pre-fractionation design is a sequence somewhere between these two extremes. Indeed, the lightest component is separated first and an indirect sequence follows. The pre-flash design does not fall exactly into regular sharp split distillation theory, but can be considered somehow similar to the pre-fractionation design.

In the rest of this unit, we will overview some crude characterization procedures and measures, will discuss how crude fractionation products are specified and we will be discussing the flexibility of conventional atmospheric crude fractionation units. These concepts will be used in Unit 2 to discuss various ways of simulating these units and in the rest of the course, how can this be used to properly design and retrofit these plants.

Crude Oil Characterization

Petroleum is a mixture of hundreds of hydrocarbons of all type (Alyphatic, Aromatic, Cycloalyphatic, etc) with water, salts, sulfur and nitrogen containing compounds and some metal complexes. Thus, instead of straight composition, alternative measures are used to characterize it: density, distillation curves, etc. We now discuss some of those that are relevant to the use of modern procedures of simulation and design (density, K factor as well as TBP and ASTM distillation curves). Notwithstanding their importance, we omit discussing, salt, sulfur, nitrogen as well as metal content, mainly because most of these factors are more relevant for processes downstream the crude fractionation unit. A complete account of all these methods can be found in Wauquier (1995).

Density

The main classification of crude is by referring to their density. Light crudes have a specific gravity (d) smaller than 0.825, medium crudes between 0.825 and 0.875, heavy crudes between 0.875 and 1.000 and extra heavy crudes larger than 1.000.

While specific gravity may suffice, the petroleum industry uses another scale (API):

$$API = \frac{141.5}{d} - 131.5 \tag{1-1}$$

Thus, light oils have a large API density and heavy ones a low one. Typical low values are close to 15 (some Venezuelan oils) to about 40 (some middle east oils).

Crudes are also characterized as paraffinic, asphaltic, napthenic and aromatic (plus combinations of those). This primary classification intends to reflect the relative abundance of different types of compounds, as their name indicates. These can also be more characterized in more detail by the density of their heavy gasoline cut (boiling point

between 100 and 200 ^{o}C), d_1, and the residue (boiling point larger than 350 ^{o}C), d_2. For example a paraffinic crude has d_1<0.76 and d_2<0.93, an aromatic crude has d_1>0.80 and d_2 between 0.93 and 0.975, an asphaltic crude has d_1>0.78 and d_2>0.975, and so on. This is obviously related to the relative abundance of hydrogen and carbon. Nevertheless, there is a need for a unique number, rather than a range, to characterize the type of crude and this is somewhat reflected by the Watson factor K_W.

Watson Factor K_W

This factor, which was introduced by researchers at Universal Oil products Company (UOP) and for this reason also known as K_{UOP}, is based on the observation that the H/C ratio is related to the boiling points of mixtures, as it is to the individual components. Thus, for pure components, we have

$$K_W = \frac{(1.8\,T)^{1/3}}{d} \tag{1-2}$$

where T is the boiling point (Kelvin) and d the specific gravity at 15.6 ^{o}C (60 ^{o}F). This value is typically 13 for paraffins, 12 for hydrocarbons whose chain and ring weights are equivalent, 11 for pure naphthenes and 10 for pure aromatics.

Because mixtures do not have a unique boiling point, to use this factor in mixtures one needs to resort to the use of some average temperature. Thus, one needs to record the temperatures for which 10, 20, 50, 80 and 90% of the volume has distilled from the sample. Thus if TBP distillation (true boiling point), which is discussed below, is used, we use

$$T = \frac{T_{20} + T_{50} + T_{80}}{3} \tag{1-3}$$

For ASTM distillations (see below), the following is used

$$T = \frac{T_{10} + 2T_{50} + T_{90}}{4} \tag{1-4}$$

While use of this factor has been extremely useful to characterize crudes, the latest tendency, *albeit more expensive and time consuming*, is to rely on better characterization methods based on modern physical tests, like Refractive index and Molecular weight (ndM method), mass spectrometry, ultraviolet (UV) spectrometry, carbon atom distribution by infrared spectrometry and nuclear magnetic resonance, or by chromatographic (gas and liquid) techniques.

True Boiling Point (TBP) Method for Crude Characterization

The true boiling point curve is basically a plot of the boiling point of each component of the mixture as a function of the cumulative volumetric fraction distilled. It implies that a batch distillation column with a large number of trays and a large reflux ratio is used, which will ensure that each component is separated at a time. In practice, the ASTM

D2892 test method (ASTM is an acronym for American Society for testing and Materials) is used. It consists of using a 15 to 18 tray distillation column operating at a 5:1 reflux ratio. The distilled volume is measured and the temperature in the reboiler is recorded. An example of such curve is shown in Figure 1-7. In simulators, one enters a few points and a spline interpolation is performed to get the entire curve.

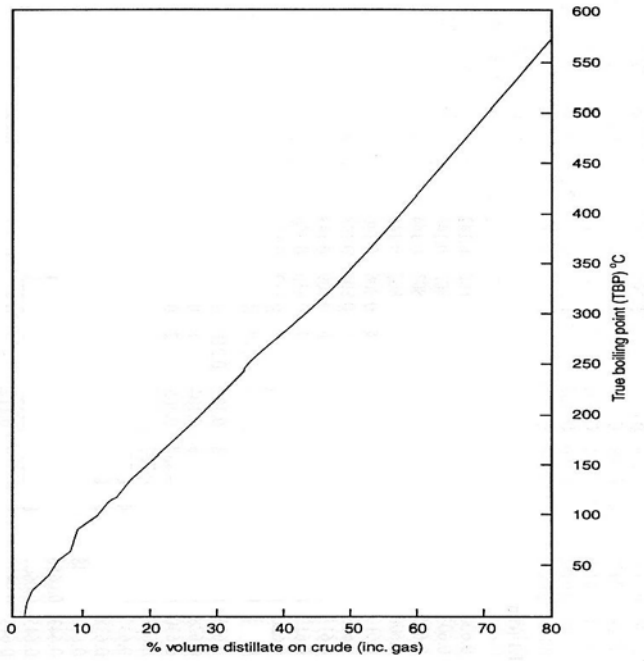

Figure 1-7: TBP curve for Kuwait Crude Sp. Gr. 08685 (Taken from "Elements of Petroleum Processing", Jones D., 1995)

ASTM Methods for Crude Characterization

In practice, obtaining the TBP curves are time consuming and expensive. To ameliorate these inconveniences other quicker methods have been developed. One such method, used for almost all petroleum fractions, except for heavy fractions, is the ASTM D86 method. In this test, the sample is boiled and the vapors are collected and condensed as they are produced. Vapor temperatures are then recorded as a function of cumulative liquid volume collected. Because no reflux is used and there is only one equilibrium stage, the initial boiling point of this assay is larger and the final point is lower than the one in the TBP curve.

Another distillation method used for heavy fractions is the ASTM D1160. There is need to use this other method because hydrocarbons start to crack at temperatures higher than about 340 ^{o}C. The ASTM D1160 is conducted at vacuum conditions (1 to 50 mmHg). There are several other ASTM methods, but we concentrate on these two because these are the only ones used to specify products. We briefly mention here the simulated distillation method ASTM D2887, which consists of a chromatographic separation of the crude into different fractions.

Finally, the Equilibrium Flash vaporization (EFV) method consists of heating up a sample keeping vapor and liquid together in the same vessel, that is, not separating it from the liquid phase while it is produced, as in the other ASTM D86 and D1160 methods. The temperature is recorded as a function of the percentage vaporized. Separation is poorer than in ASTM methods, and therefore, the initial point is higher and the final point lower. We illustrate the differences between these curves in Figure 1-8.

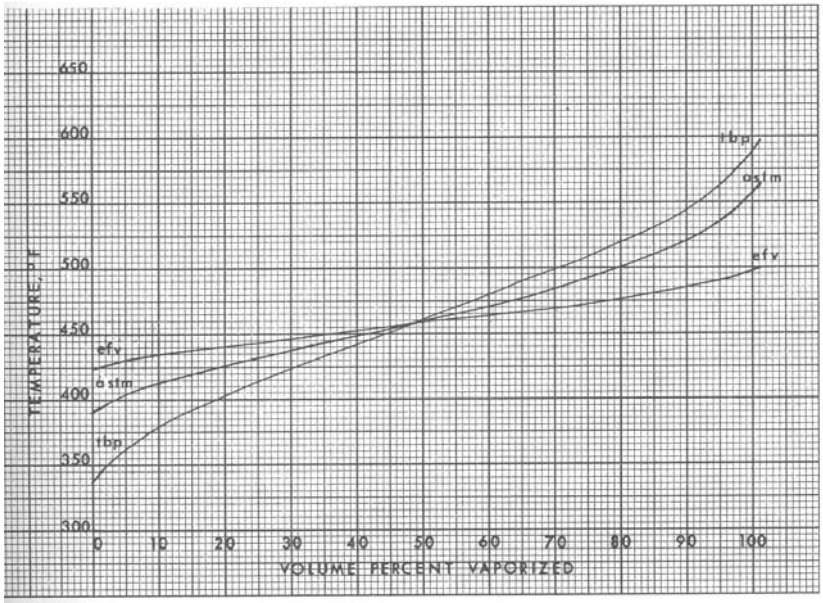

Figure 1-8: TBP, EFV and ASTM curves for a 39.7 degree API light distillate (Copied from "Petroleum refinery Distillation", Watkins, R.N., 1979).

Formulas that allow the transformation between these assays exist and therefore, one can perform all the analysis, using one of them. Typically the TBP is used as the central reference curve, but the products are often specified using ASTM D86 or D1160 points.

We now illustrate the conversion formula used to relate TBP values to ASTM D86 values. API recommends the Riazi (1979) formula:

$$T_{ASTM} = aT_{TBP}^{b} \qquad (1\text{-}5)$$

where a and b depend on the fraction distilled, as shown in Table 1-1 and degrees Kelvin should be used. Similar formulas exist for other conversions.

Table 1-1: Coefficients to convert TBP to ASTM (Equation 1-5) (Riazi Method)

% volume distilled	Coefficient *a*	Coefficient *b*
0	1.08947	0.99810
10	1.71243	0.91743
30	1.29838	0.95923
50	1.10755	0.98270
70	1.13047	0.97790
90	1.04643	0.98912
95	1.21455	0.96572

Product Properties

The TBP curve, together with the density curve can be used to somewhat characterize the products that will result from fractionation. This is important, especially if one wants to generate some initial values to aid simulations to converge. To do this, one needs to define cut points. These are the temperatures that represent the limits of the fraction of the TBP curve that will be produced. This is illustrated in Figure 1-9.

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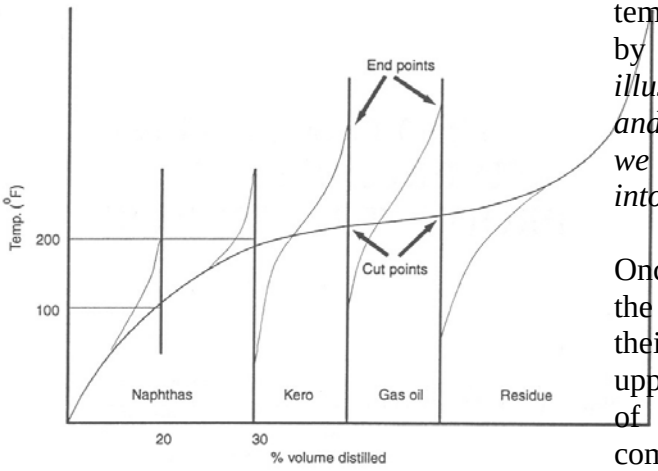

Figure 1-9: Cut points and end points.

A boiling range will then be the range of temperatures for each product established by two cuts. *In this section we will illustrate how one can obtain the ASTM and density of one such cut. In addition, we will show how such cut is separated into different pseudo components.*

Once the cut points are established, when the fractions are analyzed independently, their end points will be higher than the upper cut point temperature. This is due, of course, to the absence of the heavy components. The same can be said for the initial points, which will be lower than that of the mixture. This is also illustrated in Figure 1-9

Once the cut points are established, the procedure follows the following steps:

1. Determine the corresponding volumetric range, which gives the volumetric flowrate of the fraction directly.
2. Determine the end point and one other point of the ASTM distillation curve of the fraction.
3. Determine the complete TBP distillation curve for the fraction
4. Determine the specific gravity of the fraction and the mass flowrate of the product is then determined, if needed.

Step 1) The volumetric range is already in percentage fraction, so this is the fraction of the total flowrate.

Step 2) To calculate the initial and end points (as well as all those in between), one needs to resort to empirical correlations. There are two procedures that are used. Both are based on obtaining two ASTM points of the curve. In one case, end points and 90% are obtained using the graph of Figure 1-10, which provides the difference between the corresponding TBP and the ASTM values for selected cases differentiated by temperature. Alternatively, UOP recommends the linear relation shown in Figure 1-11 for

the end point and a simple rule for the initial point. The simple rule consists of subtracting 15 ^{o}F from the lower cut point. This applies to all fractions except the lightest and all vacuum gas oil cuts, for which 50 ^{o}F are subtracted.

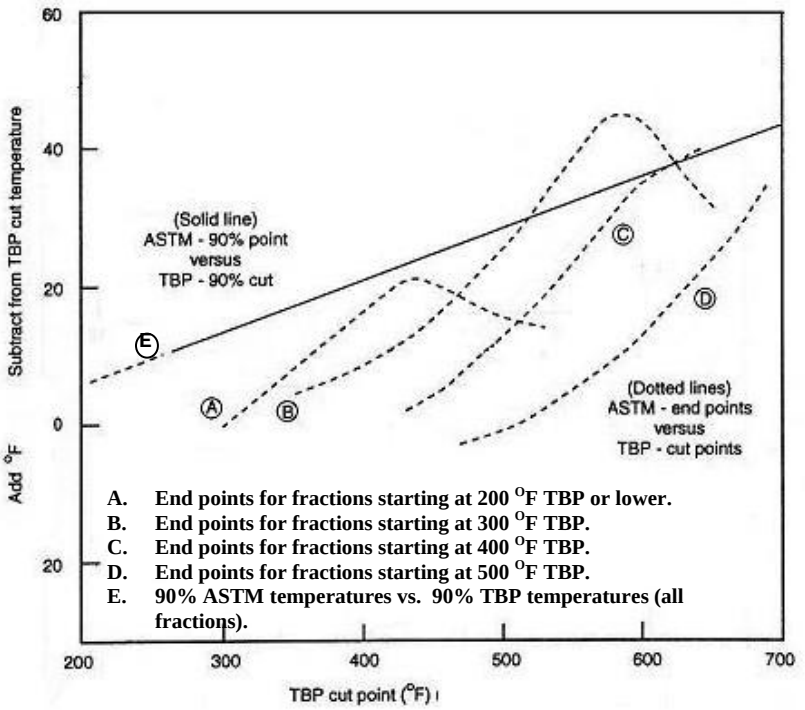

Figure 1-10: Cut points and end points (Adapted from "Elements of Petroleum Processing", Jones D., 1995).

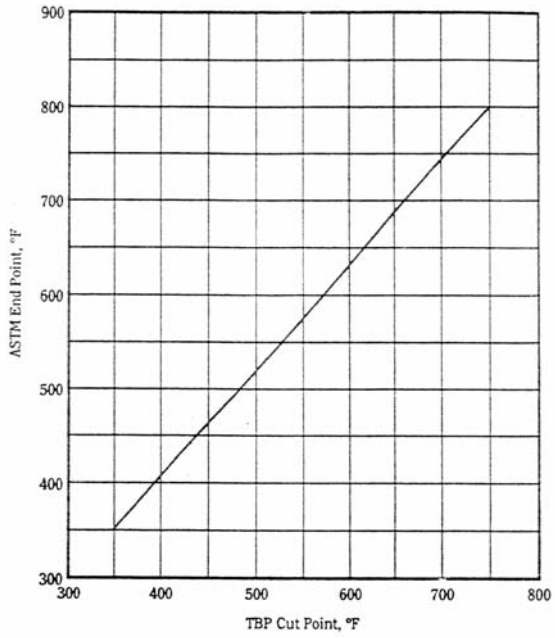

Figure 1-11: Cut points and end points (Source UOP).

Step 3) To obtain the whole ASTM curve, the nomogram of Figure 1-12 is used. The two points obtained are located and a straight line is drawn through them. Next, one needs to make the corresponding correction between ASTM and TBP values (equation 1-5).

Step 4) To obtain the density of the fraction one can calculate the mid boiling point of each fraction (Figure 1-13). This is the temperature corresponding to the middle of the corresponding % volume interval. The specific gravity is now easily obtained if the density vs. volume fraction, or density vs. TBP curve is available. However, sometimes, the density curve is not available and only the specific gravity of the whole crude is given. In such case, one proceeds as follows: The factor K_W is calculated using equation (1-2) with the temperature obtained using (1-3). Next, equation (1-2) is inverted to obtain the specific gravity of the fraction, using the value of K_W just calculated and the midpoint temperature.

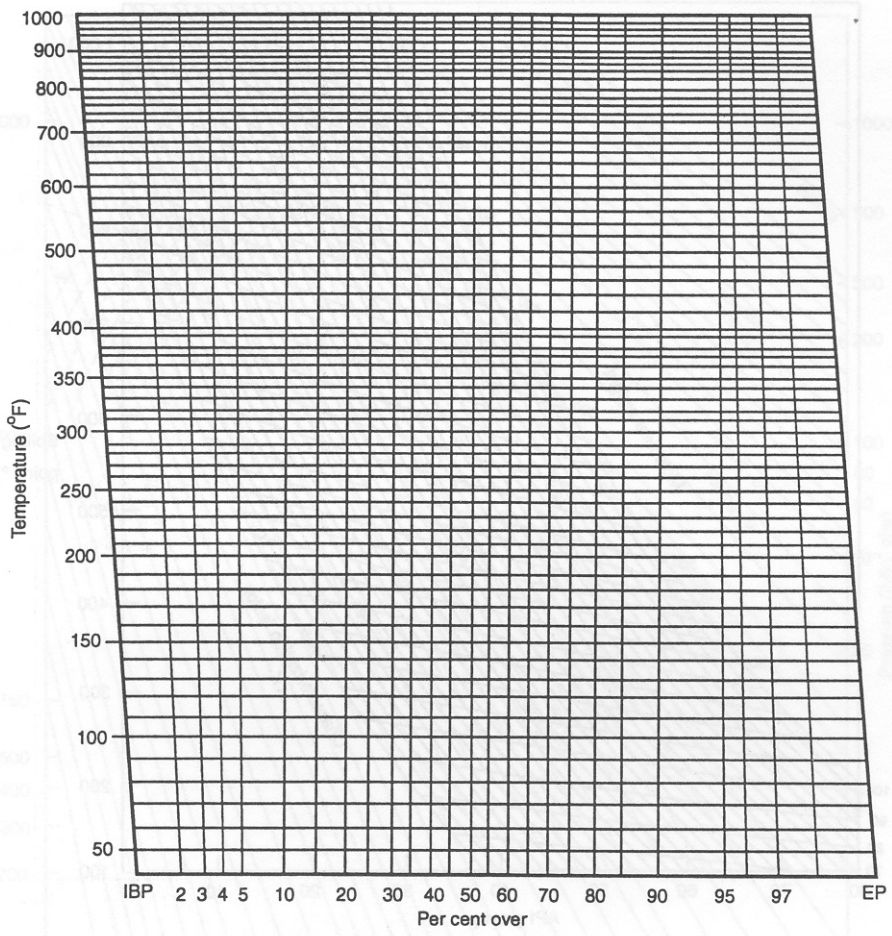

Figure 1-12: Nomogram used to obtain full ASTM curves (Probabilistic plot). (Copied from "Elements of Petroleum Processing", Jones D., 1995).

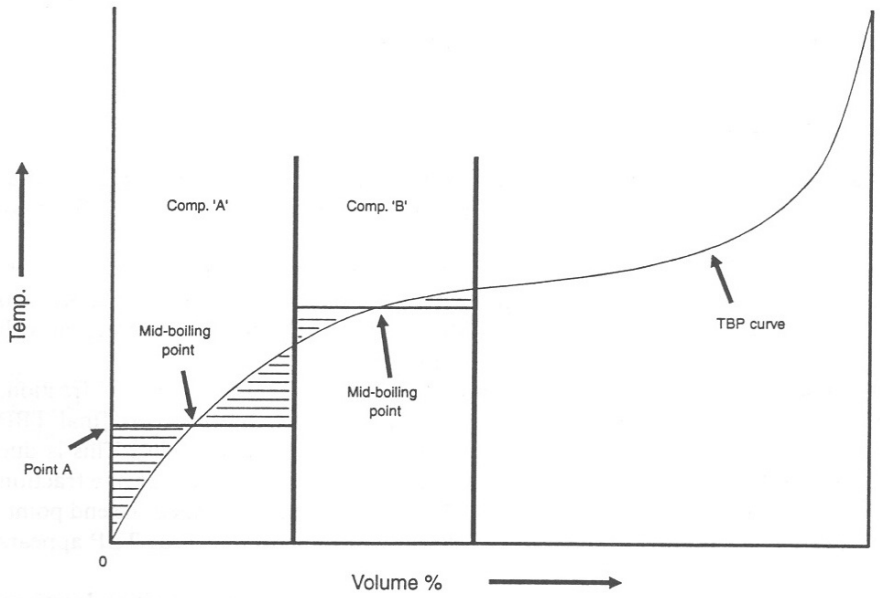

Figure 1-13: Cut points and end points

Example 1-1:

Consider the 197-222 ^{o}C (387-432^{o}F) boiling range for the TBP curve given in Figure 1-7.

***Step* 1)** *Volumetric range*: According to the figure, the volumetric range corresponding to these cuts is 27.3-31.2% yield of the fraction is 3.9% vol. This gives us directly the volumetric flowrate of this fraction, that is, a 3.9% of the total flowrate.

***Step* 2)** *ASTM end and 90% points*: According to Figure 1-10, curve B, the difference corresponding to the upper cut is 13 ^{o}F. Thus, the ASTM end point will be 419 ^{o}F (432 ^{o}F -13 ^{o}F). By using equation (1-5), one obtains the corresponding TBP value (to obtain the coefficients *a* and *b*, we use here the fraction corresponding to 95% and above, that is the last row of Table 1-1 and we use ^{o}K), which is 435.5 ^{o}F. According to this figure to obtain the 90% ASTM value, we need the 90% TBP point. We assume that this is equal to the 90% cut point, which is 219.5 ^{o}C. Curve E from Figure 1-10 indicates that the difference is 22 ^{o}F. Thus, the 90% ASTM point is 405 ^{o}F (427 ^{o}F -22 ^{o}F). Now, if one uses Figure 1-11 one obtains 439 ^{o}F as the end point. To get the corresponding initial point we subtract 15 ^{o}F from 387 ^{o}F, obtaining 372^{o}F.

***Step* 3)** Two straight lines connecting the two sets of points obtained in step 2 are drawn. The first set of points is: (100%, 435.50 ^{o}F) and (90%, 405^{o}F). The second set of points is: (100%, 440 ^{o}F) and (0%, 372^{o}F). The complete curves are then built using the ASTM values provided by these two lines.

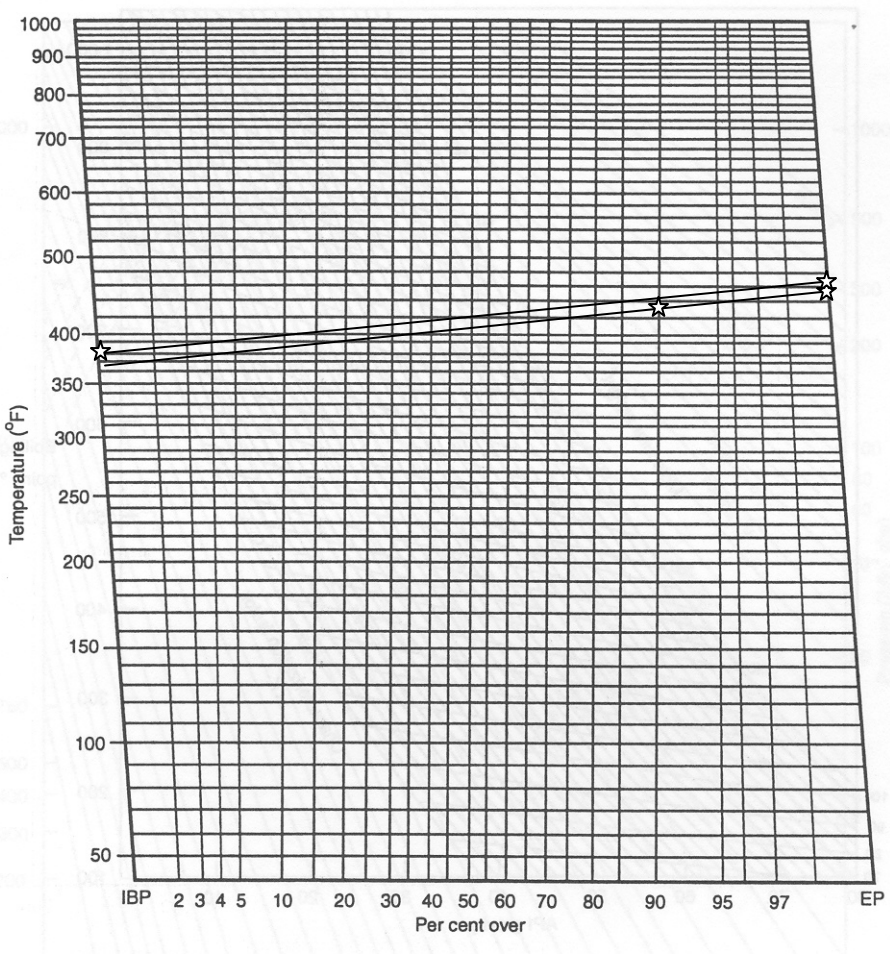

Figure 1-14: Step 3) of Example 1-1.

***Step* 4)** The middle point of this fraction is 410 ^{o}F. Now, from Figure 1-15, one obtains the specific gravity: d~0.798. If the curve is not available, one proceeds to calculate the K_W factor using (1-2) and the temperature equation (1-3). To do this, one obtains first the three needed temperatures: T_{20}=148 ^{o}C, T_{50}=258 ^{o}C, T_{80}=570 ^{o}C, which together with the specific gravity of 0.8685, renders K_W =11.80. Now, the specific gravity of the cut is obtained using T=410 ^{o}F and K_W =11.80 in formula (1-2). The value is: 0.808, remarkably similar!

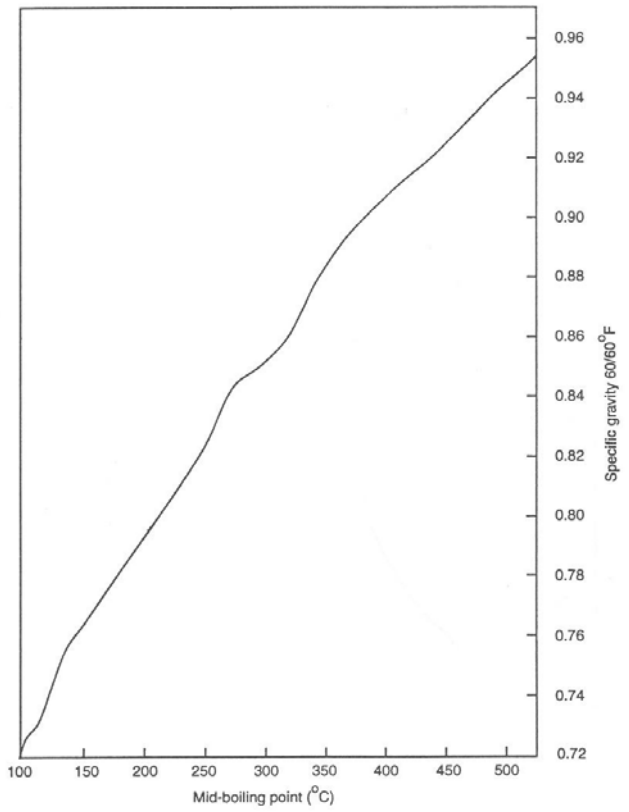

Figure 1-15: Specific gravity vs. TBP curve for Kuwait Crude Sp. Gr. 08685. Copied from "Elements of Petroleum Processing", Jones D., 1995).

Exercise 1-1:

Consider the Kuwait Crude Sp. Gr. 08685, whose data are given in Figure 1-7 and Figure 1-15. Consider a diesel cut with boiling range 432-595 ^{o}F.

1) Obtain the TBP curve of this cut using the method described above.
2) Consider the following volume intervals 13%, 16.5%, 21%, 18%, 18.5% and 13% from the IBP to the EP, respectively. Calculate the middle point temperature, and the specific gravity of each cut.

Crude Breakup

Now that the procedure to establish the properties of a given fraction is known, we proceed to discuss the procedure needed to determine the complete crude breakup. Examples are given in Table 1-2 and Table **1-3**.

Table 1-2: Example TBP cut for crude fractionation. (Jones pp. 23)

Product	TBP Cut Points
LPG and gasoline	IBP-300 ^{o}F
Naphtha	300-385 ^{o}F
Kerosene	380-480 ^{o}F
Diesel	480-610 ^{o}F
Atmospheric Gas Oil (AGO)	610-690 ^{o}F
Vacuum Gas Oil (VGO)	690-1000 ^{o}F

Table 1-3: Product Specifications (According to Watkins)

Product	Specification
Naphtha	D86 (95% point) =182 °C
Kerosene	D86 (95% point) =271 °C
Diesel	D86 (95% point) =327 °C
Gas Oil	D86 (95% point) =377-410 °C

In general, one can start with the values shown in Table 1-3. However, there a detailed analysis on what is the downstream use of these products sets the quality with more accuracy. Consider for example a refinery oriented to produce auto grade diesel, LPGs, kerosene (for cutter stock) and gasolines. The specs for the products in this plant are given in Table 1-4.

Table 1-4: Example ASTM ranges for Atmospheric crude fractionation

Product	Max Naphtha
Naphtha feed to Naphtha splitter	347 ^{o}F (ASTM EP)
Kerosene	335 ^{o}F (ASTM 5%)-370 ^{o}F (ASTM 95%)
Light Diesel	385 ^{o}F -595 ^{o}F (TBP cut points)
Heavy Diesel	570 ^{o}F (5% ASTM)-700 ^{o}F (TBP cut point)

To generate these some specific properties of these products, other properties than ASTM temperatures are also used. We summarize some, because they are often used in simulation too.

We first introduce the concept of *Gap*, which is used to quantify the overlap between adjacent fractions. Consider two adjacent fractions. The (*5-95*) *Gap* is the difference between the 5% ASTM temperature of a heavy fraction (H) minus the 95% ASTM temperature of the lighter fraction (L), that is:

$$Gap_{H,L} = T_5^H - T_{95}^L \tag{1-6}$$

The *Gaps* recommended by Watkins are given in Table 1-5.

Table 1-5: Gap Specifications (According to Watkins)

Products	Specification
Kerosene –Naphtha	(5-95) Gap ≥ 16.7 °C
Diesel- Kerosene	(5-95) Gap ≥ 0 °C
Gas Oil- Diesel	(5-95) Gap -5.6 °C to –11 °C

Another property needed to specify a particular product is its flash point, which is defined as the temperature at which the vapor above the oil will momentarily flash or explode. Usually, this is determined in the laboratory. It is also related to the ASTM 5% point by means of the following correlation:

$$Flash\ Point\ (^{O}\mathrm{F}) = 0.77\ (ASTM\ 5\%\ ^{O}\mathrm{F} - 150\ ^{O}\mathrm{F}) \tag{1-7}$$

Other properties are also of importance. They are:

- Pour point, defined as the temperature at which the oil ceases to flow. The pour point is obviously related to pumping problems, which are obviously avoided by heating up the mixture.
- Cloud point, which is temperature at which the oil becomes cloudy or hazy.
- Octane number, related to knocking properties and used for gasolines. The octane number is important downstream from the crude fractionation unit and is adjusted by a procedure called blending in which different products are mixed to reach the specification.
- Cetane number, or Diesel Index, is related to ignition quality and defined as the time period between the start of injection and start of combustion (ignition) of the fuel. The fluids of reference are: n-hexadecane (Cetane Number = 100) and alpha methyl naphthalene (Cetane Number = 0). Higher cetane fuels will have shorter ignition delay periods than lower cetane fuels. The calculated cetane number is often used: this is given by the API gravity times the aniline point (in ^{o}F) divided by 100. In turn, the aniline point is the lowest temperature (°F or °C) at which equal volumes of aniline ($C_6H_5NH_2$) and the oil form a single phase. It correlates roughly with the

amount and type of aromatic hydrocarbons in an oil sample. A low AP is indicative of higher aromatics, while a high AP is indicative of lower aromatics content.

- Sulfur content in its various forms. Although there are ways to predict the sulfur content of products, given those of the crude, we omit these calculations here.

Cuts corresponding to different products are typically obtained using a subset of the following specs (although others can be also added to the list):

- 95% ASTM point (usually the D86 point is used)
- Allowable *Gaps.*
- Minimum Flash point
- Desired pour point
- Desired cetane number.
- Sulfur content

Consider for example the specs of Table 1-4. They come from the following considerations:

- ***Naphtha:*** It will be fed to a catalytic reformer typically requires a charge with ASTM IBP=194 ^{o}F and EBP=347 ^{o}F.
- ***Light Diesel and Heavy Diesel:*** These will be later blended to produce the required diesel. Such diesel has a pour point of 10 ^{o}F. This limits the end point of the heavy diesel cut to be 700 ^{o}F. Sulfur is of concern here, so the light gas oil is to have less that 0.8% sulfur (Heavy diesel will be hydrotreated). This can be directly manipulated through the vol% in the TBP curve (details omitted). If a yield of 18% is chosen, then the cut range obtained is 387-595 ^{o}F (cut points). The light diesel flash point is 175 ^{o}F.
- ***Kerosene:*** A specification of 95% ASTM Temperature=335 ^{o}F and a minimum Gap of 20 ^{o}F with the naphtha and a 10 ^{o}F Gap with the light diesel.

Exercise 1-2: Obtain the %vol of crude for the specifications given in Table 1-4, using the crude of Figure 1-7. For a total of 59689 BPD processed, obtain the corresponding product rates.

Other Properties

To successfully simulate a crude fractionation unit, one also needs to be able to predict properties like enthalpy, vapor pressure and density, viscosity, surface tension, etc. The way this is performed is by generating pseudo-components from portions of the TBP curve, that is, picking a certain interval of volume fraction for example and using the mid temperature of that interval to establish its boiling point. If a density curve is known, then one can also obtain the density of each pseudo-component. If not, one can use the aid of

the Watson K_W factor to obtain it (the simplifying assumption here is that K_W is constant for all the different fractions).

Since the principal characteristics needed to define a pure component are its boiling point, standard density and molecular weight, the first two are then given for each pseudo-component. The molecular weight can be related to these two by a special formula. In turn critical constants and especial coefficients (e.g. acentric factor, coefficients for equations of state, enthalpy, entropy expressions, etc) are needed to calculate properties of mixtures. For these, there are also expressions that help calculate them as a function of normal boiling temperature and standard density. Since this is done automatically inside simulator, we omit the details here. See Wauquier (1995) for details.

Crude Units Design and Operations

The design of crude units is reviewed by Watkins (1979), where different design alternatives are presented and a method to design atmospheric and vacuum columns is suggested. Watkins' book is still the main reference for crude unit design and no other comprehensive attempt has been presented since. Crude fractionation columns are classified by Watkins into top reflux (type U), pump-around reflux (type A) and pump-back reflux (type R). A top reflux corresponds to the conventional top condenser of distillation columns. In a pump-around reflux, liquid is drawn from a plate and after being cooled down it is returned to a few plates above. Finally in a pump-back reflux, liquid in equilibrium is sub-cooled and sent to the tray immediately below. Figure 1-16 depicts an atmospheric crude fractionation unit containing all the features that characterize types U, A and R. In this figure the heat exchanger network is omitted to highlight design parameters unrelated to heat integration.

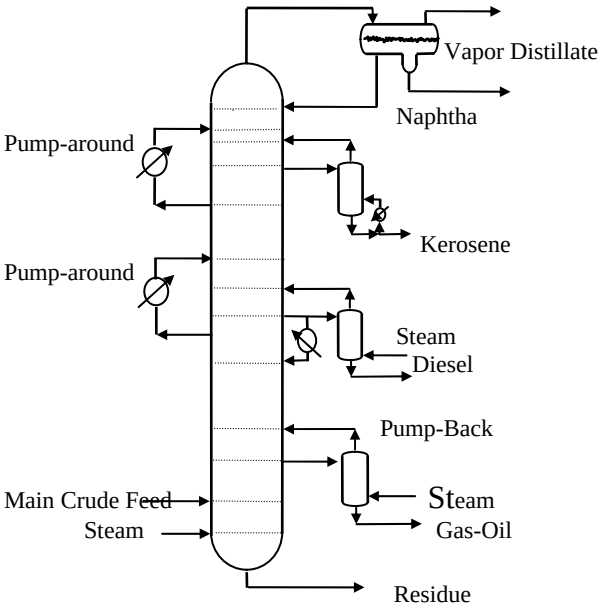

Figure 1-16: Atmospheric Crude Unit. Sketch of three reflux methods.

Crude is preheated to a certain temperature, typically 260 ^{o}F, for desalting purposes. The column has typically three (sometimes four) side draws to produce kerosene, diesel oil and atmospheric gas oil(s). To produce a sharper separation, specifically, to strip some of the lighter fraction of the draw, a stripper is set for each case. Steam is used to produce this additional separation, but reboilers have also been considered in the past. Figure 1-16 contains one such reboiler in the gas-oil stripper. Type U column, corresponds to a column where no pump-around circuit or pump-back is present. Thus, all the cooling is performed at the top condenser.

In type A columns, two to three pump-around circuits are typically used to cool down some of the liquid and they may still share the cooling duty with a top condenser. Finally Type R columns with pump-back reflux can also contain a top reflux. The reason for this intermediate cooling was explained first by Watkins as the desire to reduce flow variations between the top and the bottom trays. Watkins shows that these flowrate variations are typically reduced by 60%, avoiding therefore an over-sizing of the column diameter. One additional advantage is that heat is available at higher temperature level, so energy integration can be enhanced, something that was not a concern before the 1973 energy crisis (Watkins published is book shortly thereafter). The disadvantage of the pump-around circuits is however identified by Watkins as a loss of theoretical trays. On the other hand also according to Watkins, although sub-cooling takes place in pump-back refluxes there is no loss of separation efficiency.

Overflash is a term coined for the ratio of the volumetric flowrates of liquid flowing from the tray above the feed and the feed. This value is set to 2-3%. The purpose is, according to Watkins, to guarantee that the plates between the flash zone and the first draw above it do not dry up.

Steam injection has been used for years as an alternative to reboilers. Nelson (1949) first pointed out that reboilers, usually run by hot oil as is still the practice in several rerun and stabilizing columns, "do increase the efficiency of fractionation, but a satisfactory degree of separation can usually be obtained more cheaply by the use of a (steam) stripping section". This type of analysis has been repeated in one way or another throughout the years. For example, Hsie and McAvoy (1990) state that "steam also reduces the partial pressure of the hydrocarbons and thus lowers the required vaporization partial pressure of the hydrocarbons and thus lowers the required vaporization temperature". The obvious implication of all these statements is that steam injection is better, because it lowers the temperature, and eventually by implication, lowers the energy requirements. Finally, side-strippers have been used to improve the *Gap* and steam stripping has been used over the years to provide the necessary vaporization.

Over the years, these designs have been applied in a variety of combinations. They have all evolved to configurations containing type A (pump-around) reflux, with or without top reflux. When top reflux is not used, it is usual to have a pump-around discharging sub-cooled liquid on the top tray. Pump-back reflux (type R) seems to have been abandoned in practice. Similarly steam injection as opposed to the use of reboilers is the current practice. Industry has only in appearance settled with these designs. There is an active pursuit for de-bottlenecking to increase throughput and energy efficiency retrofits, especially in the last 15-20 years. One such recent study is offered by Lin et. al. (1996).

Fractionation of crude is energy intensive, and the main driving force for better designs stems from the sizable savings that can be achieved with proper heat integration. Pinch Technology (Linnhoff, 1993) has contributed to a great extent to profound changes in the heat exchanger network configuration of crude units, especially with the preheating network. When applied to the analysis of crude units, the first discovery of pinch

technology was that a lot of heat was transferred across the pinch, due in great part to the fact that the preheating crude was not split but rather heated up in a series of heat exchangers one after the other, whereas pinch designs usually call for stream splitting. Linnhoff et. al. (1982) proposed the concept of appropriate placement, in which process decisions are made in light of the impact on heat recovery through the pinch diagram. Linnhoff and Parker (1984) extended the idea to an iterative scheme by proposing the Plus/Minus approach, where pinch diagrams suggest process changes that would in turn improve the heat recovery. Unfortunately, this method was described in the context of case studies, but never generalized nor automated. Bagajewicz (1998) has proposed such automation and has studied energy and air emission saving horizons for entire crude units. In this study, several parameters, such as pump-around rates and return temperatures, as well as steam injection rates have been varied to achieve maximum savings. The study takes advantage of the findings about column flexibility that this paper presents.

Structural alternatives to current designs have been proposed as early as in 1929. Piromoov and Beiswenger (1929) proposed the Carrier Distillation process, which was implemented by Lummus and apparently later abandoned. The purpose is to take part of the topping column overhead and mix it with the feed crude, with the resulting effect of lowering the boiling point of the mixture. Radical departures from the current designs have been also proposed. For example Brugma (1941) proposed an arrangement that has startling resemblance to energy integrated (Petlyuk type) configurations. According to Nelson (1949), Brugma configurations have been abandoned because the entire length of the three towers is equal to the conventional configuration. These were the years where energy consumption was not of much concern, so revisiting these old ideas might produce some surprises. Also, crude pre-flash columns have been used to stabilize the crude (Nelson, 1949). After the fertile period of the forties, industry evolved implementing the three designs that Watkins (1979) explores in his book, to later abandon pump-back reflux and reboilers. Currently there are efforts to perform modifications of columns for de-bottlenecking studies. Mainly, these modifications have been related to plate efficiency enhancement and in the case of vacuum columns lower pressure drop (Snell and Juno, 1996). In a recent paper, Liebmann and Dhole (1995) have shown that radically different designs can realize substantially higher savings.

Consider a type A column (column with pump-around circuits), as described in Table 1-6 and Figure 3 depicts the TBP curve corresponding to the crude used. Its average API is 28.6. Light-ends composition is described in Table 1-7. To perform these studies the PRO/II Simulator (Simsci) was used. Table 1-8 shows the set of independent design variables of the base case.

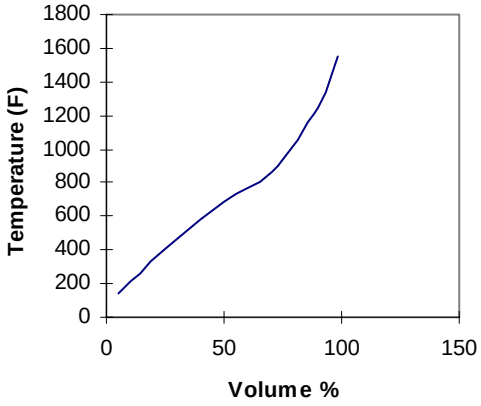

Figure 1-17: Crude TBP curve.

Table 1-7: Light-ends Composition

COMPOUND	Vol %
Ethane	0.1
Propane	0.2
Isobutane	0.3
n-Butane	0.7
Isopentane	0.5
n-Pentane	1.2

Table 1-6: Column Data

Number of Plates	30
Number of trays (Side Strippers)	2
Pump-around 1 (PA1) Draw	4
Pump-around 1 (PA1) return	Tray 2
Pump-around 2 (PA2) Draw	Tray 15
Pump-around 2 (PA2) return	Tray 13
Pump-around 3 (PA3) Draw	Tray 21
Pump-around 3 (PA3) return	Tray 19
Kerosene Draw	Tray 12
Kerosene Side-Stripper return	Tray 11
Diesel Draw Tray	Tray 18
Diesel Side-Stripper return	Tray 17
Gas-Oil Draw	Tray 24
Gas-Oil Side-Stripper return	Tray 23
Crude Feed	Tray 27

Table 1-8: Design Variables

VARIABLE	VALUE
Overflash	3%
Main Steam (60 psig, 600 ^{o}F)	10000 lb/hr
Kerosene Stripper Steam (60 psig, 600 ^{o}F)	2000 lb/hr
Diesel Stripper Steam (60 psig, 600 ^{o}F)	4500 lb/hr
AGO Stripper Steam (60 psig, 600 ^{o}F)	4000 lb/hr
Condenser Temperature	110 ^{o}F
Kerosene D86 95% Temperature	520 ^{o}F

Diesel D86 95% Temperature	665 ^{o}F
AGO TBP 95% Temperature	885 ^{o}F
Pump-around 1 (PA 1) Return Temperature	175 ^{o}F
Pump-around 2 (PA 2) Return Temperature	310 ^{o}F
Pump-around 3 (PA 3) Return Temperature	450 ^{o}F
Pump-around 1 (PA 1) Heat Rate	50 10^{6} Btu/hr
Pump-around 2 (PA 2) Heat Rate	50 10^{6} Btu/hr
Pump-around 2 (PA 2) Heat Rate	40 10^{6} Btu/hr
Tray efficiency main column	60%
Tray efficiency side strippers	50%

To characterize the feeds and products pseudo-component flowrate distributions are used. Figure 1-18 shows the curves for the base case as obtained from a simulation. These plots consist of the molar flowrate of each pseudo-component as a function of their boiling point. The solid curve in this figure corresponds to the crude feed. The curves corresponding to naphtha, kerosene, gas-oil and residue are also included. These curves add up to the crude feed.

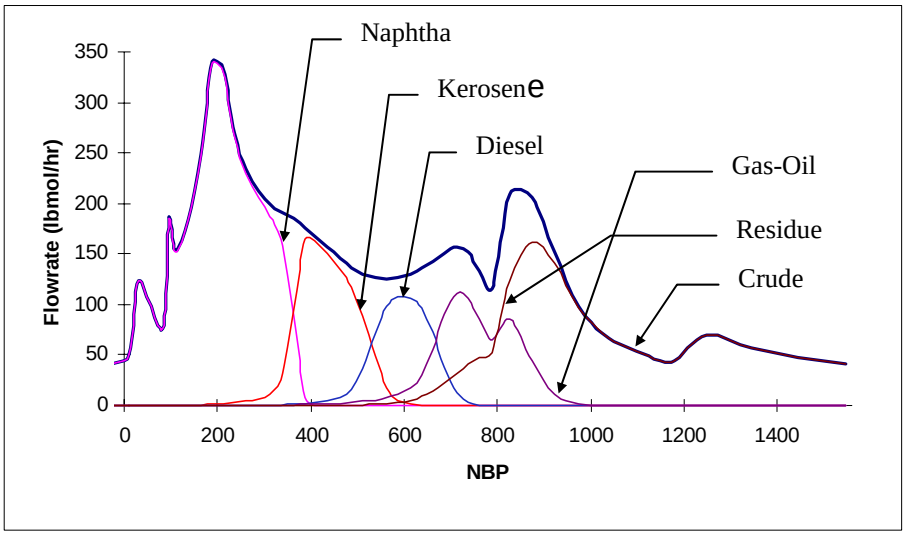

Figure 1-18: Base Case Pseudo-Component Flowrate Distribution

PUMP-AROUND DUTY

Consider eliminating pump-around (PA1), i.e. changing the specification form from 50 10^{6} Btu/hr to zero. The flowrate of the product streams are summarized in Table 1-9 indicating that a variation of less than 0.9 % is observed.

Table 1-9: Effect of Eliminating Pump-Around 1 (PA1)
(Flowrates on Dry Basis)

PRODUCT	FLOWRATE Base Case (lbmol/hr)	FLOWRATE No Pump-Around PA1 (lbmol/hr)

Overhead	2895	2900
Kerosene	1038	1033
Diesel	682	682
Gas-Oil	743	743

The pseudo-component rate distribution is not shown, as differences between the base case and the design without a pump-around are unnoticeable. The same effect was observed by varying the heat duty of all the other pump-around circuits. Additionally, the exact amount of heat eliminated from the pump-around is introduced in the condenser.

Thus, pump-around duties have essentially no effect on the production rate, nor they change the product composition in any sizable amount.

Figure 1-19 shows the effect of eliminating pump-around PA1 on the vapor and liquid tray rates of the column.

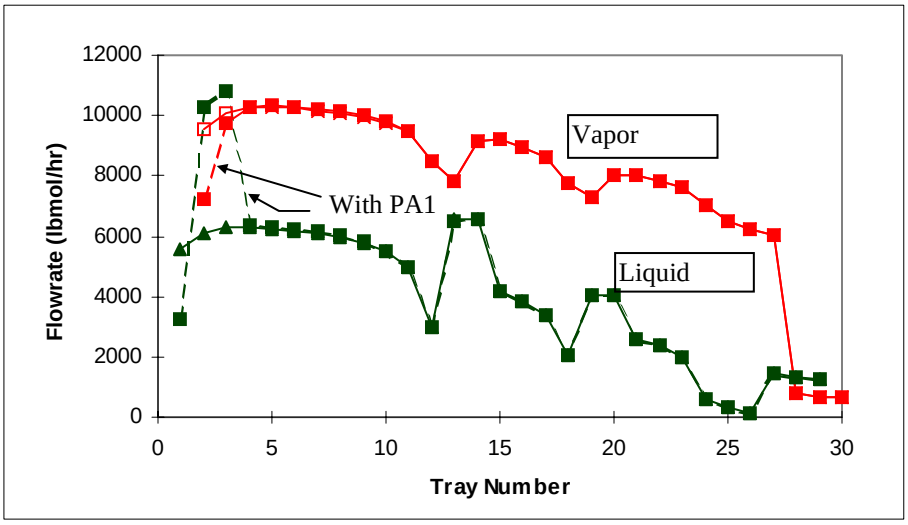

Figure 1-19: Base Case and Column without PA1 flowrate profile.

In the presence of the pump-around, liquid flowrates increase, as expected, to later decrease to a value in the first tray that is lower than the one observed without the pump-around. Vapor flowrates, decrease as expected. This contradicts column profiles presented by Watkins (1979) where both liquid and vapor flowrates are reduced. The same is shown in Perry (page 13-86).

Figure 1-20 shows the temperature profile for the base case. As the first pump-around is eliminated, the effect on the temperature profile is only seen in the first 4-5 trays. The differences are shown in Figure 1-21. To further compare the effect of pump-around circuits, all of them have been eliminated. Flowrate profiles are compared in Figure 1-22.

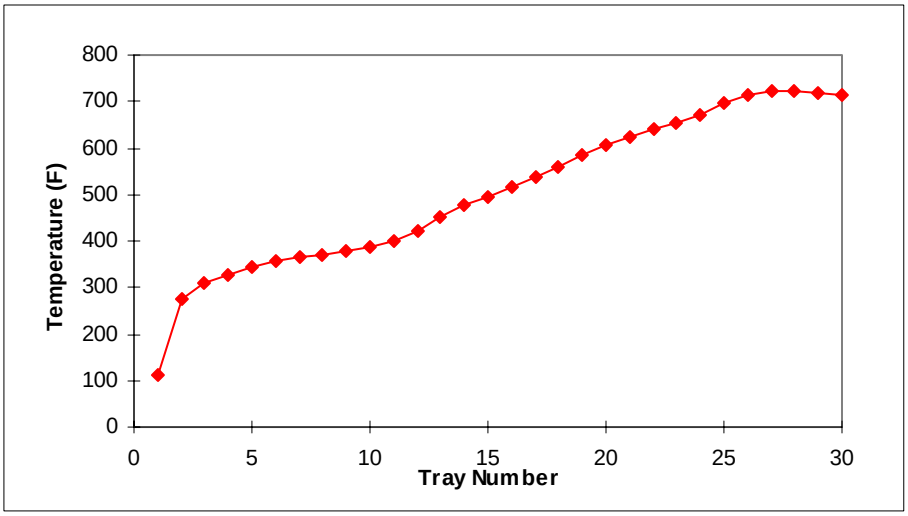

Figure 1-20: Base Case and Column without PA1 Temperature profiles.

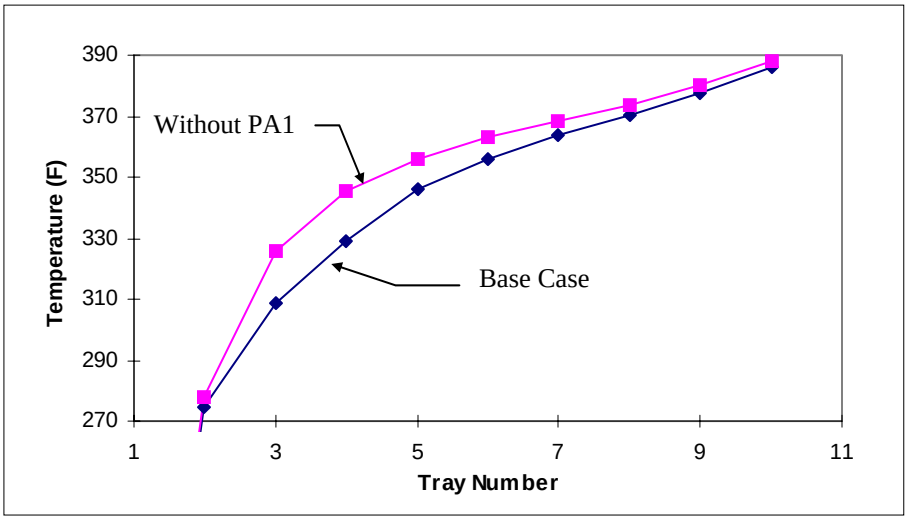

Figure 1-21: Temperature Profiles.

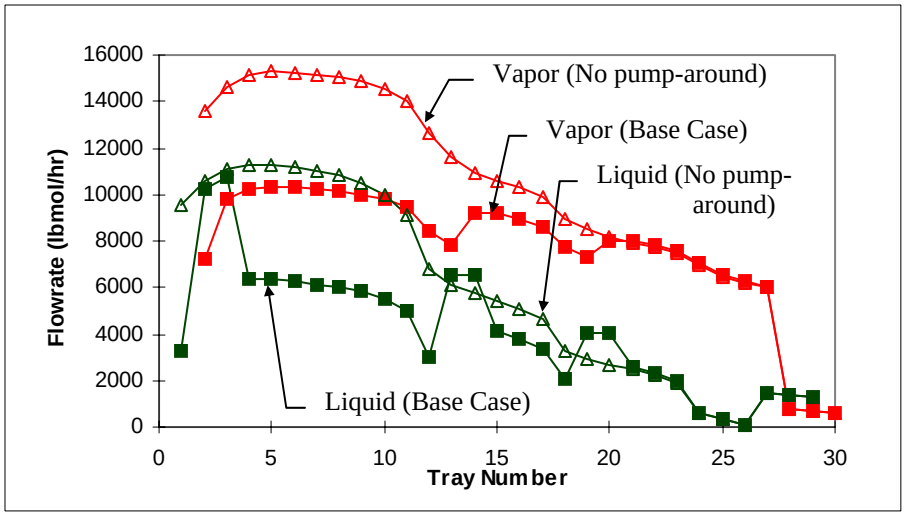

Figure 1-22: Base Case and Column without Pump-Around flowrate profiles.

Watkins (1979) states that the flowrate profiles of both liquid and vapor are reduced by the same percentage (40% in his Figure 2.5) when pump-around circuits are introduced. Reduction takes place in the case of vapor, and this reduction is, naturally, dependent on the duty of the pump-around circuits. The claimed reduction in liquid rates takes place, although at the top of the column flowrates are similar in both cases. In the presence of pump-around circuits the temperature throughout the column reduces. Figure 1-23 shows the effect. Heat duty differences are summarized in Table 1-10. The difference in furnace duty corresponding to the change in crude feed temperature is 1.7 10^6 Btu/hr. Assuming a furnace inlet of 450 ^oF, this difference is about 0.3% of the total furnace duty. Table 1-11 compares the product flowrates.

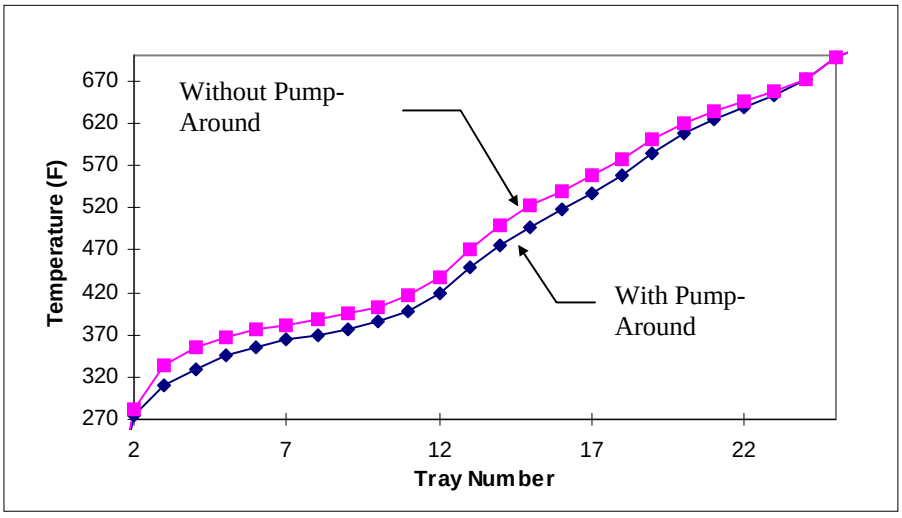

Figure 1-23: Base Case and Column without Pump-around Temperature profile.

Table 1-10: Heat Duty comparison between Type A and Type U columns.

	TYPE A (Base Case)	TYPE U (No Pump-around circuits)
Condenser (Btu/hr)	$-150\ 10^6$	$-287.1\ 10^6$
PA1 (Btu/hr)	$-50\ 10^6$	0
PA2 (Btu/hr)	$-50\ 10^6$	0
PA3 (Btu/hr)	$-40\ 10^6$	0
Feed Temperature (^oF)	724.5	723.1

Table 1-11: Comparison of Performance
(Flowrates are on a dry basis)

PRODUCT	FLOWRATE TYPE A (lbmol/hr)	FLOWRATE Type U (lbmol/hr)
Overhead	2895	2921
Kerosene	1038	1047
Diesel	682	671
Gas-Oil	743	709
Residue	1188	1198

Type A columns (with pump-around circuits) require a slightly higher furnace duty (higher crude inlet temperature) and as a result a slightly higher cooling duty. The presence of pump-around circuits slightly decreases the amount of overhead and kerosene, increasing the amount of heavy products. Thus, although the effect is mild, TYPE U columns have better separation efficiency. This deteriorates the Gaps.

PUMP-AROUND RETURN TEMPERATURE

The effect of return temperature was investigated, while maintaining the same duty. The return temperature of the higher pump-around (PA1) was reduced by 15 ^oF. Since the duty is maintained, this is equivalent to reducing the pump-around flowrate. The effect on the liquid flowrate profile is negligible.

REPLACEMENT OF PUMP-AROUND CIRCUITS BY COOLERS

Direct cooling in a tray is equivalent to a pump-around circuit with the same duty discharging on that tray. Figure 1-24 shows the effect on the column traffic. Product rates are exactly the same as in the case were the pump-around has been eliminated.

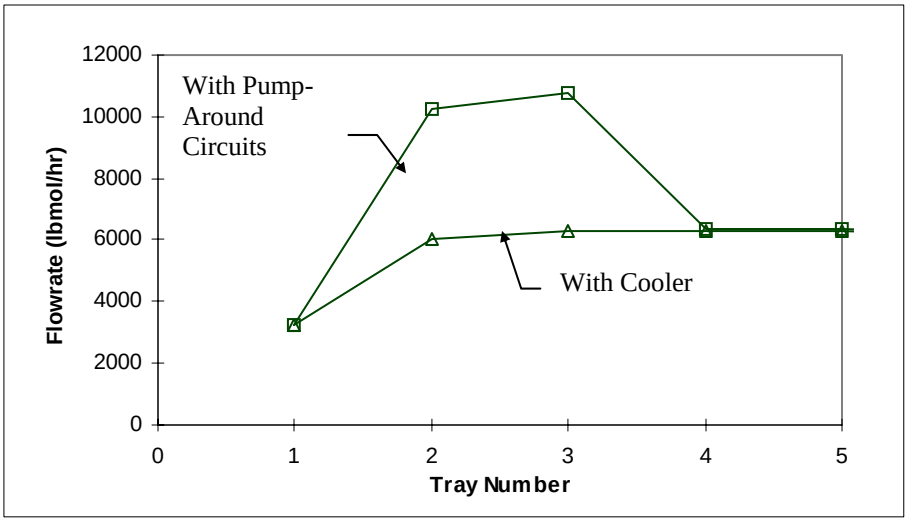

Figure 1-24: Effect of Replacing a Pump-Around with Tray Cooling on Column Traffic.

PUMP-AROUND ROLE IN HEAT INTEGRATION

In view of all the above results, one can conclude that pump-around circuits have no positive effect in crude fractionation. Not even the benefits claimed by Watkins and reproduced in Perry that the traffic in the column is balanced. One might ask, then, what is the advantage of Type A columns?

The real advantage of Type A columns versus Type U columns might rely in the fact that cooling takes place in pump-around circuits at a higher temperature. Therefore the opportunities for heat integration may increase. The pinch point of both Type A and Type U arrangements is (not surprisingly) at 329 ^{o}F when a minimum approach of 10 ^{o}F is used. Table 1-12 summarizes these opportunities for the present case study.

Table 1-12: Energy Integration Horizons for Type A and Type U columns

MINIMUM UTILITY	TYPE A	Type U
Overall (10^6 Btu/hr)	258.7	363.3
Furnace(10^6 Btu/hr)	167.7	175.1
Steam (600 ^{o}F) (10^6 Btu/hr)	91.0	188.2

The advantage of type A columns is unquestionable. Moreover, most of the reduction is obtained by reducing heat demand below 600 ^{o}F, as the temperature level of the pump-around circuit suggests. Thus

Pump-around circuits are efficient means of providing higher level temperature sources that are of great help in increasing the energy efficiency of crude units

STEAM STRIPPING

Steam stripping is another feature of crude fractionation that has been investigated. Steam injected to side-strippers has been varied. As a result, essentially the same product distribution is obtained, if the appropriate draw rates are used. Table 1-13 summarizes the results of varying the steam rates to the kerosene side-stripper. The effect on kerosene concentration of pseudo-products is almost unnoticeable, except in the lighter fractions, which are removed.

Table 1-13: Effect of Steam in Side-strippers.

Kerosene Stripping Steam Rate (lb/hr)	Overhead (lbmol/hr)	Overhead Water (lbmol/hr)	Kerosene Draw (lbmol/hr)	Kerosene Stripper Return (lbmol/hr)	Kerosene Product (Dry basis) (lbmol/hr)	Condenser Duty (Btu/hr)
4000	2898	1077	1302	477	1038	$-150\ 10^6$
2000	2887	967	1227	280	1051	$-146\ 10^6$
400	2864	881	1139	78.6	1078	$-142\ 10^6$
20	2844	862	1101	1	1099	$-140\ 10^6$

The effect on the overhead product distribution is also imperceptible. The stripping effect is really mild. In terms of energy supply, it is apparent that a high usage of steam and cooling water is used to recover only a small fraction of light components. For example, by comparing the base case and the smallest steam injection case (20 lb/hr), one finds that in the base case one spends approximately $5.3\ 10^6$ Btu/hr more in a form of steam injection and 10^7 Btu/hr more in condenser cooling duties to just recover 1.9% of the overhead. This corresponds to about 12 gallons of naphtha. A similar study was performed varying the rate of steam injection in other side-strippers observing similar results.

Steam injection in side-strippers does not affect product distribution significantly, with the adverse effect of increasing the condenser duty unnecessarily. It only removes light components and therefore controls the flash point.

Indeed, Table 1-14 shows the different ASTM D86 (760 mmHg) *Gaps.*

Table 1-14: Effect of Steam in Side-strippers.

Kerosene	Naphtha/

Stripping Steam Rate (lb/hr)	Kerosene Gap (^{o}F)
4000	39
2000	35
400	28
20	16

MAIN STEAM INJECTION

Main column steam injection has obvious effects on product component distribution. The effect, depicted in Figure 1-25, shows the effect of doubling the steam rate on the gas-oil distribution.

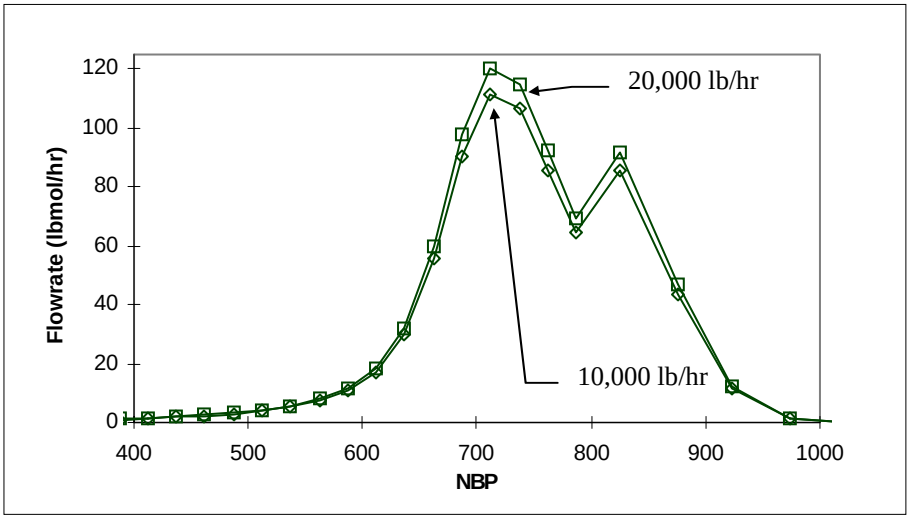

Figure 1-25: Gas-Oil Distribution as a function of Steam Rates.

The overall rate of gas-oil on a dry basis is 55 lbmol/hr larger (6.9%). The effect on diesel is to increase it in 1.3%. The effect on lighter products is insignificant. The condenser duty increases to 164 10^{6} Btu/hr. The column temperature profile is affected by a few degrees (5 ^{o}F smaller in the bottom to 2-3 ^{o}F smaller in the top portion) and obviously the column liquid and vapor rates increase. The most important effect is the one observed on the furnace duty. The column feed temperature was lowered to 720 ^{o}F resulting in a saving of 5.3 10^{6} Btu/hr. However, doubling the steam rate resulted in an increase of 13.3 10^{6} Btu/hr in the steam.

An increase in steam rate decreases the furnace heat duty in an amount that is lower than the heat added by the additional steam. Lowering the steam is therefore desirable as long as large temperatures are not reached.

COLUMN OVERFLASH

Column overflash is, according to Watkins (1979), "the extra amount of vaporization above that required by material balance considerations, which is taken to ensure that adequate reflux will be available in the trays between the flash zone and the lowest side-stream product draw tray". Specifically, it is defined as the quotient of the volumetric flowrate of liquid flowing off the tray above the feed and the volumetric flowrate of the feed. Figure 1-26 compares the flowrate profiles for an overflash of 3% and an overflash of 1%. Figure 1-27 shows the temperature profiles for the lower portion of the column.

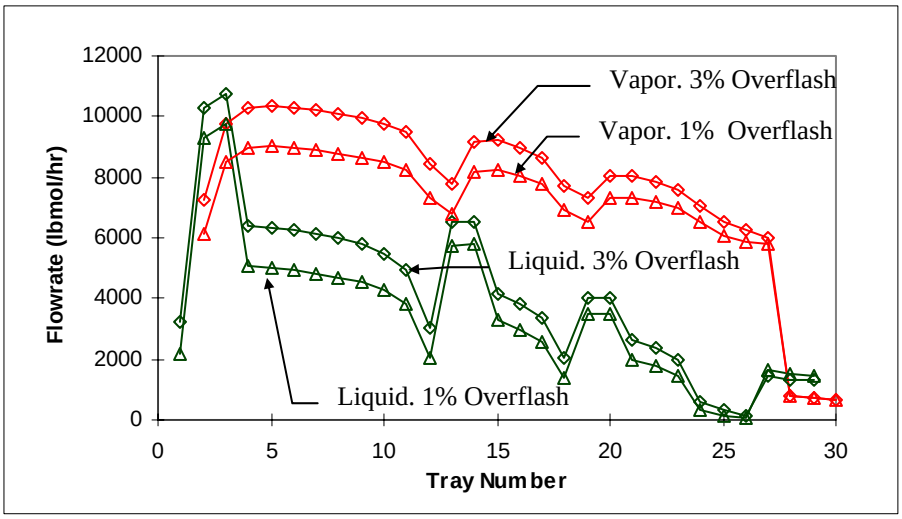

Figure 1-26: Flowrate Profiles. Base Case vs. Column with 1% Overflash.

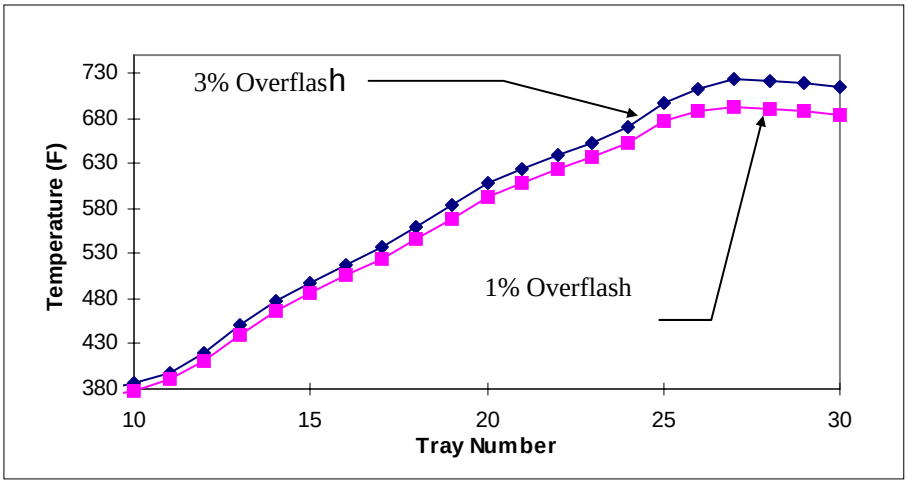

Figure 1-27: Temperature profiles. Base Case vs. Column with 1% Overflash.

Figure 1-26 indicates that for a smaller overflash the temperatures in the column are lower and Figure 1-27 shows a lower traffic upwards. In turn, even though all the specifications are met, i.e., the different products have the corresponding 95% D86 (or TBP) temperature, the flowrates of products are altered. Figure 1-27 shows the product distribution changes for Gas-Oil and the Residue. Changes for Diesel are lower than 2.5% whereas the changes for Kerosene and Naphtha are imperceptible.

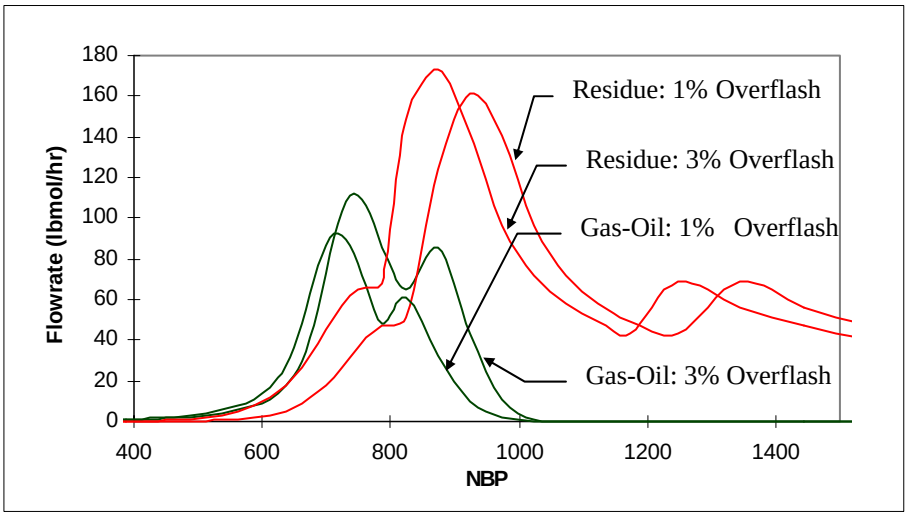

Figure 1-28: Flowrate profiles. Base Case vs. Column with 1% Overflash.

Mostly, the reduction in rate is uniform for all components, reflecting the same degree of separation. In addition the condenser duty is lowered to 126.5 10^6 Btu/hr (15.6%

reduction). Finally the feed temperature reduces to 692 ^{o}F, which represents a reduction of 39.7 10^{6} Btu/h.

An increased overflash has the effect of increasing the gas-oil rate and the residue rate, while affecting very little the rate of other products. The gas-oil composition profile remains mostly the same. A significant reduction in furnace duty is observed.

The above result has its importance in various aspects. First, there are the intricacies of how this will affect the design of the vacuum column, a subject that is not studied in this paper. Second, one may ask how a change in main steam rate can compensate for this reduction of rates. Indeed, as noticed above, an increase in steam rates increases the gas-oil production shifting the component distribution towards more lights. For this purpose the flowrate of steam was varied until the flowrate of gas-oil (on a dry basis) reached the same value as in the base case. The flowrate needed was 39,665 lb/hr. Table 1-15 shows the resulting product rates. Gas-Oil Product distributions are compared in Figure 1-29.

Table 1-15: Base case vs. Column with 1% overflash and increased main steam (Flowrates on Dry Basis)

PRODUCT	FLOWRATE Type A (lbmol/hr)	FLOWRATE Type U (lbmol/hr)
Overhead	2895	2887
Kerosene	1038	1036
Diesel	682	696
AGO	743	743
Residue	1188	1184

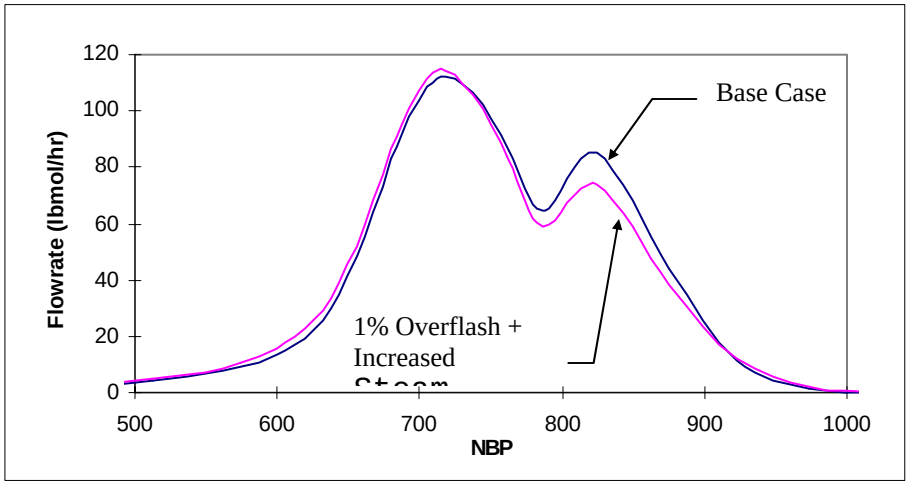

Figure 1-29: Gas-Oil Distribution for 1% overflash and increased Main Column Steam.

USE OF REBOILERS

Various claims have been made regarding the advantage of the use live steam instead of reboilers. As it was analyzed in the introductory discussion, steam injection has been considered advantageous because it lowers the temperature, and eventually by implication, lowers the energy requirements. Figure 1-30, taken from Liebmann and Dhole (1995), shows that for low degree of vaporization steam stripping is more efficient, whereas when larger degrees of vaporization are needed, reboiling should be used.

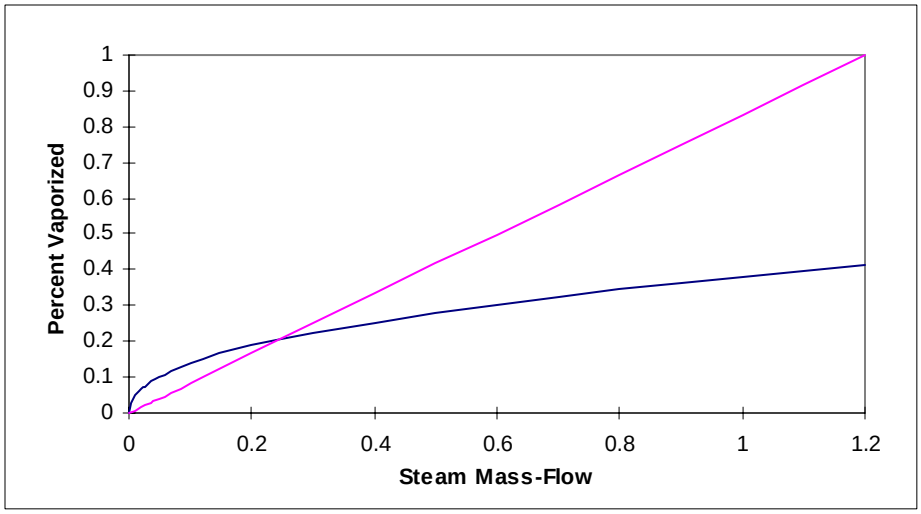

Figure 1-30: Steam vs. Reboiler efficiency.

The first effect noticed is that the use of reboilers requires, as anticipated, larger temperatures. The steam injection was substituted by a reboiler with the same duty. In comparison to the base case, the feed temperature increases to 729 ^{o}F and as a consequence the furnace duty is increased by 6.25 10^{6} Btu/h. Not only are the temperature profiles larger, especially in the reboiler where the temperature raises from 713 ^{o}F to 755 ^{o}F. Flowrate profiles are lower, and additionally, product rates are lower (11% and 9% less diesel is produced). When a reboiler was added in the side-strippers it was found that almost twice as much energy is needed to obtain the same separation pattern, with only some gain in condenser cooling duty, as in the previous case.

The usage of a reboiler in the main fractionation column and side-strippers as opposed to steam injection has been ruled out due to the high temperatures needed for a certain degree of separation. This study confirms this statement.

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