

## CRUDE DISTILLATION

### 4.1. INTRODUCTION

Crude distillation unit (CDU) is at the front-end of the refinery, also known as topping unit, or atmospheric distillation unit. It receives high flow rates hence its size and operating cost are the largest in the refinery. Many crude distillation units are designed to handle a variety of crude oil types. The design of the unit is based on a light crude scenario and a heavy crude scenario. The unit should run satisfactorily at about 60% of the design feed rate. Seasonal temperature variation should be incorporated in the design because changes in the cut point of gasoline can vary by 20 °C (36 °F) between summer and winter.

The capacity of the CDU ranges from 10,000 barrels per stream day (BPSD) or 1400 metric tons per day (tpd) to 400,000 BPSD (56,000 metric tpd). The economics of refining favours larger units. A good size CDU can process about 200,000 BPSD. The unit produces raw products which have to be processed in downstream unit to produce products of certain specifications. This involves the removal of undesirable components like sulphur, nitrogen and metal compounds, and limiting the aromatic contents.

Typical products from the unit are:

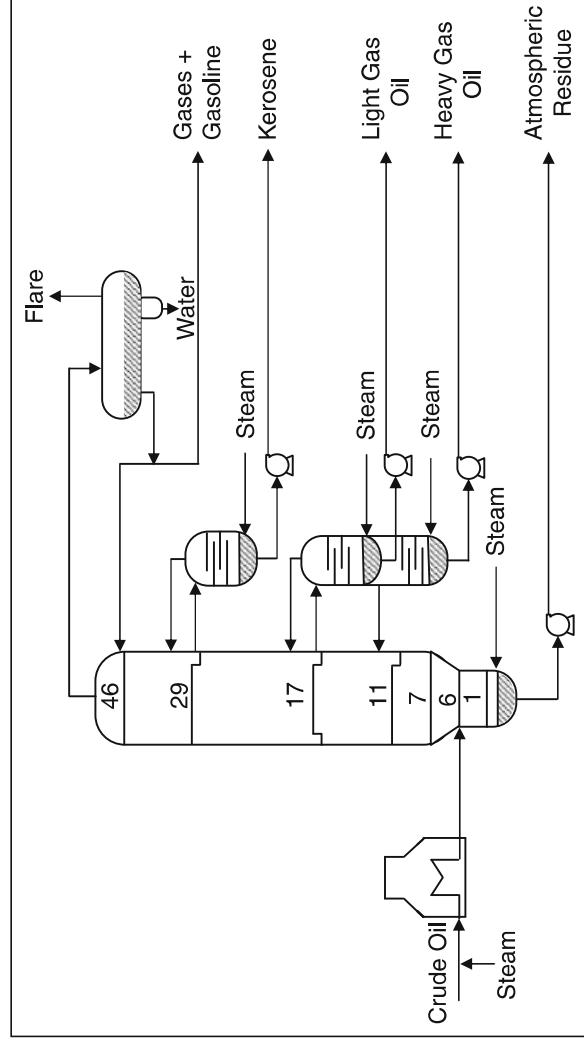
- Gases
- Light straight run naphtha (also called light gasoline or light naphtha)
- Heavy gasoline (also called military jet fuel)
- Kerosene (also called light distillate or jet fuel)
- Middle distillates called diesel or light gas oil (LGO)
- Heavy distillates called atmospheric gas oil (AGO) or heavy gas oil (HGO)
- Crude column bottoms called atmospheric residue or topped crude.

## 4.2. PROCESS DESCRIPTION

The process flow diagram of a typical crude distillation unit is shown in Figure 4.1. Crude oil is pumped from storage tanks where it is freed from sediments and free water by gravity. It goes through a series of heat exchangers where it is heated with hot products coming out from the distillation column and by the exchange with heat from the pumparound liquid streams. The temperature of the crude feed can reach 120–150 °C (248–302 °F).

The crude oil contains salt in the form of dissolved salt in the tiny droplet of water which forms a water-in oil emulsion. This water cannot be separated by gravity or through mechanical means. It is separated through electrostatic water separation. This process is called desalting. In the electrostatic desalter, the salty water droplets are caused to coalesce and migrate to the aqueous phase by gravity. It involves mixing the crude with dilution water (5–6 vol%) through a mixing valve.

The crude is further heated in product heat exchangers. The preheating of the crude using the hot products cools down the products to the desired temperature for pumping to the storage tanks. This is essential for the economics of the unit in terms of energy conservation and utilization. Of course, preheating is not enough, as the crude has to be partially vaporized to the extent that all products, except for the atmospheric residue have to be in the vapour phase when the crude enters the atmospheric distillation column. Thus a furnace is required to boost the temperature to



**Figure 4.1** Process flow diagram of an atmospheric distillation unit

between 330 and 385 °C (626 and 725 °F) depending on the crude composition.

The partially vaporized crude is transferred to the flash zone of the column located at a point lower down the column and above what is called the stripping section. The main column is typically 50 m (164 ft) high and is equipped with about 30–50 valve trays. The vapour goes up in tremendous amounts and at a high flow rate, necessitating a large diameter column above the flash zone. At the bottom of the stripping section, steam is injected into the column to strip the atmospheric residue of any light hydrocarbon and to lower the partial pressure of the hydrocarbon vapours in the flash zone. This has the effect of lowering the boiling point of the hydrocarbons and causing more hydrocarbons to boil and go up the column to be eventually condensed and withdrawn as side streams. As the hot vapours from the flash zone rise through the trays up the column, they are contacted by the colder reflux down the column. In the overhead condenser, the vapours are condensed and part of the light naphtha is returned to the column as reflux. Further reflux is provided by several pumparound streams along the column.

In the distillation tower, heat required for separation is provided by the enthalpy of the feed. For effective separation heat has to be removed from the tower, in this case, by the overhead condenser and several pumparound streams along the tower length. The pumparound stream is a liquid withdrawn at a point below a side stream tray that is cooled by the cold crude feed as part of the preheat exchangers train. It is then returned to the column a few trays above the draw tray. This pumparound cooling accomplishes a number of tasks. First, the cold liquid condenses more of the rising vapours thus providing more reflux to compensate for the withdrawal of products from the column. Second, heat is removed from the column at higher temperatures. This is in addition to the heat removal from the condenser which takes place at relatively lower temperatures, thus the thermal efficiency of the column is improved and the required furnace duty is reduced. Third, pumparound streams reduce the vapour flow rate throughout the column. Therefore, the required column is smaller than what would otherwise be required if pumparound streams were not there. The drawback to using more pumparound streams is that they tend to reduce the fractionation because a more fractionated liquid is mixed after cooling with a less fractionated liquid a few trays above.

The side draw products are usually stripped to control their initial boiling point. The strippers contain several trays and the stripping is done using steam at the bottom of the stripper or reboiler type side stream strippers. The end boiling point of the side stream is controlled by the flow rate of the side stream product.

The overhead vapour is condensed at the top of the tower by heat exchange with the cool crude coming into the unit and by air and cooling water. The liquid product is called light straight run naphtha. Part of this product is returned to the column as an external reflux. Down the column,

**Table 4.1** Tray distribution in a crude distillation unit

Zone	Number of trays
Overhead product to kerosene	10
Kerosene to light gas oil	8
Light gas oil to heavy gas oil	6
Heavy gas oil to flash zone	6
Flash zone to atmospheric residue	6
Pumparounds	3–4

other products are withdrawn, such as heavy straight run naphtha, kerosene or jet fuel, LGO and HGO. All of these products are withdrawn above the feed tray. The atmospheric residue is withdrawn from the bottom of the column.

The main column is equipped with between 30 and 50 valve trays. Typical designs have the trays distribution between products as shown in Table 4.1.

### 4.3. OPERATION OF CRUDE DISTILLATION UNITS

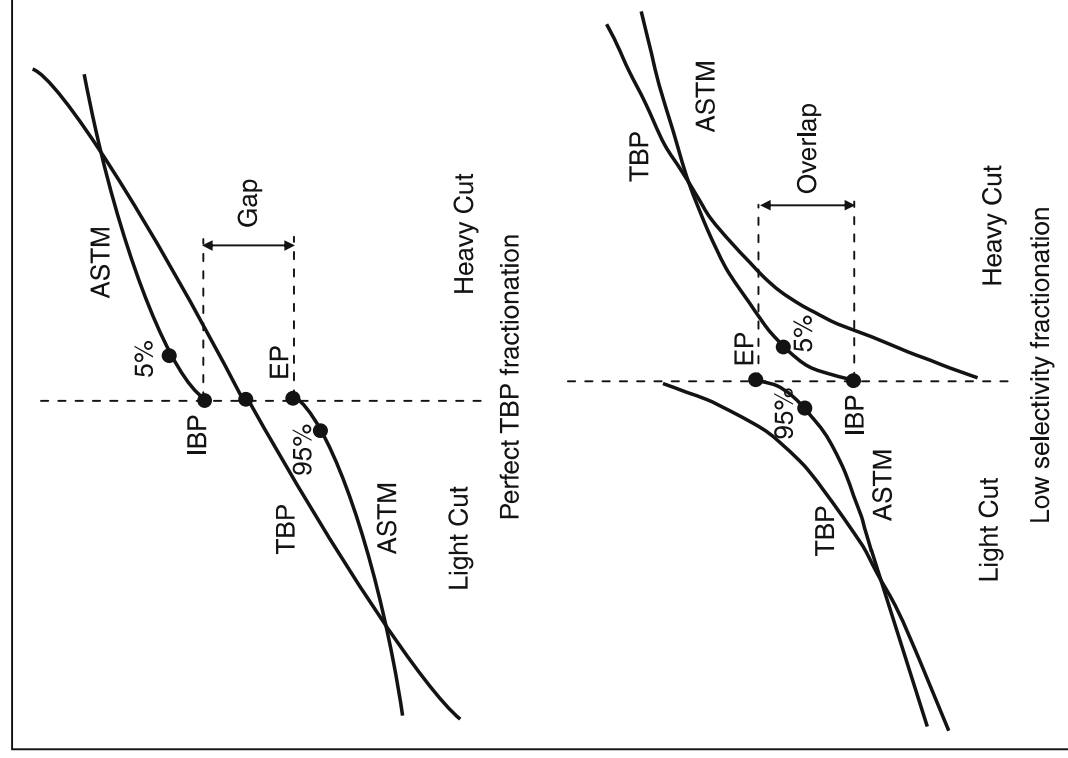
The CDU can be looked at from the point of view of a process engineering as a multicomponent distillation column. Indeed, the commercial process simulation program models CDU as a case of multicomponent distillation with undefined pseudo-components instead of the normally encountered defined components. However, because we are dealing with a mixture of thousands of compounds and due to the limitation of any distillation column in terms of its capacity to fractionate these components, there are specific operational aspects which characterize the CDU operation. In addition, there are some practical aspects in meeting the required specifications and boiling range of the required transportation fuels. In this section, the factors which affect the design and operation of the unit are explored.

#### 4.3.1. Fractionation

The degree of fractionation in a crude unit is determined by the gap or overlap between two adjacent side stream products. Hence we can talk about the gap or overlap in the boiling point range between kerosene and LGO for example. In the ideal case there would be no overlap between these products and the end boiling point of kerosene would be the initial

boiling point of the LGO. However, if we compare the ASTM distillation boiling points, and since ASTM distillation does not give perfect fractionation, the ASTM end point of kerosene is higher than the initial ASTM boiling point of LGO. This is called fractionation overlap.

Since determining the initial and end point on the laboratory test is not always possible or accurate, the fractionation gap is defined as the difference between the ASTM 5% boiling point of the product and the 95% point of the lighter product. When this difference is positive, we have a gap indicating good fractionation. A negative difference is called an overlap indicating that some of the light product is still in the heavier product and vice versa. Figure 4.2 shows the gap and overlap concept. By controlling the cut point of any two consecutive products we can affect the degree of fractionation.



**Figure 4.2** Gap and overlap (Ptak *et al.*, 2000)

#### 4-3.1.1. Cut Points

The cut points in the CDU are controlled by the overhead vapour temperature which determines how much vapour goes to the condensers to produce light naphtha and by the flow rate of the various products straight from the column or the side stream strippers. The atmospheric residue level control inside the column determines its flow rate and thus its initial cut point.

The amount of light naphtha is determined by the dew point of the naphtha at its partial pressure, which is close to the overhead temperature. Changing the drawoff rate of any product affects the cut points of the heavier product below it. For example, lowering the kerosene flow rate will lower its end point (make it lighter), but will also modify the initial cut points of the LGO and HGO and the initial cut point of the atmospheric residue. The residue flow rate, the internal reflux rate, the drawoff temperatures and the pumparounds are also affected.

Therefore, if the cut point of one stream is changed through a change in its withdrawal rate, the flow rate of the heavier product next to it should be changed in the reverse and by the same amount in order to make the changes in the desired stream only. For example, if the end point of kerosene is lowered by decreasing the kerosene flow rate by a certain amount, the flow rate of LGO has to be increased by the same amount. If this action is taken, only the cut point of kerosene is affected and the cut points of the other products remain unchanged.

The side stream rate also affects the temperature at the withdrawal tray and lowers the internal reflux coming out of that tray. The internal reflux rate affects the degree of fractionation. It can be increased by increasing the heater outlet temperature, and by lowering the pumparound duty in the lower section of the column. When less heat is removed by the lower pumparound, more vapours will be available up the column and more internal reflux is produced as the vapours are condensed.

#### 4-3.1.2. Degree of Fractionation

The fractionation quality between two consecutive streams is affected by several factors such as the vapour and liquid flow rates in the column zone between these two streams, the number of trays, and the heat extracted by the pumparound. Fractionation quality is formulated in terms of gap or overlap of the products. For perfect fractionation, zero gap and overlap are required. This means that the EBP of the light cut would be the IBP of the heavier cut and so on.

#### 4-3.2. Overflash

In order to fractionate the crude oil into the various products, it has to be heated to a temperature between 330 and 385 °C (626 and 725 °F), depending on the crude composition. The partially vaporized crude is

transferred to the flash zone of the column located at a point lower down the column. The furnace outlet temperature should be enough to vaporize all products withdrawn above the flash zone plus about 3–5 vol% of the bottom product. This overflash has the function of providing liquid wash to the vapours going up the column from the flash zone, and improving the fractionation on the trays above the flash zone, thereby improving the quality of the HGO and reducing the overlap with the bottom products below the flash zone. This necessitates that there must be few trays in the region between the flash zone and the HGO drawoff. The overflash provides heat input to the column in excess to that needed to distill the overhead products. It also prevents coke deposition on the trays in the wash zone.

The furnace outlet temperature is controlled to keep coking inside the furnace tubes and in the column flash zone to a minimum. However, the composition of the crude plays a part in determining the maximum temperature allowed. Paraffinic crude oils cracks more readily than an aromatic or asphalt-base crude. Therefore, the furnace outlet temperature for paraffinic crude oils is lower than that for other crude types.

### 4.3.3. Column Pressure

The pressure inside the CDU column is controlled by the back pressure of the overhead reflux drum at about 0.2–0.34 bar gauge (3–5 psig). The top tray pressure is 0.4–0.7 bar gauge (6–10 psig) higher than the reflux drum. The flash zone pressure is usually 0.34–0.54 bar (5–8 psi) higher than the top tray.

### 4.3.4. Overhead Temperature

The overhead temperature must be controlled to be 14–17 °C (25–31 °F) higher than the dew point temperature for the water at the column overhead pressure so that no liquid water is condensed in the column. This is to prevent corrosion due to the hydrogen chloride dissolved in liquid water (hydrochloric acid).

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#### Example E4.1

If the overhead stream contains 8.5 mol% water at a pressure of 34.7 psia (2.36 bars), calculate the overhead temperature for safe operation.

#### Solution:

The saturation temperature of water at the partial pressure of water in the overhead vapour.

Water partial pressure =  $0.085 \times 2.36 = 0.2$  bars

From the steam tables:

Saturated steam temperature at 0.2 bars = 61 °C

Safe overhead operating temperature =  $61 + 17 = 78$  °C

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### 4.3.5. Pre-flash Columns and Crude Column Capacity

The crude flow rate to the CDU determines the capacity of the whole refinery. A crude column is typically designed for 80% loading, which means that the unit can be operated at 20% throughput more than the design value. The capacity of the column is limited by the vapour flow rate with a velocity between 2.5 and 3.5 ft/s (0.76 and 1.07 m/s). The vapour flow rate increases as the vapours rise from the flash zone to the overhead. To keep the vapour velocity within the limits mentioned above, the pumparounds, which are installed at several points along the column, extract heat from the column. This results in condensing the rising vapours and reducing the vapour velocity.

To expand crude capacity, the most used technique is to introduce a pre-flash column before the crude heater. The crude oil after preheating in the hot products and pumparound heat exchangers is flashed into a column where the lightest products are removed. The bottoms from the pre-flash column are introduced into the crude heater and then to the crude column. The amounts of the light ends in the crude are now less, and this reduces the vapour loading up the column. Although the unit throughput is increased, the furnace duty is not increased, since the crude rate going to the furnace is not affected due to the removal of the light ends. Pre-flash columns are also introduced in the original design of the CDU when the crude oil is light, and when it contains a lot of light ends in the naphtha range.

## 4.4. CRUDE OIL DESALTING

When the crude oil enters the unit, it carries with it some brine in the form of very fine water droplets emulsified in the crude oil. The salt content of the crude measured in pounds per thousand barrels (PTB) can be as high as 2000. Desalting of crude oil is an essential part of the refinery operation. The salt content should be lowered to between 5.7 and 14.3 kg/1000 m<sup>3</sup> (2 and 5 PTB). Poor desalting has the following effects:

- Salts deposit inside the tubes of furnaces and on the tube bundles of heat exchangers creating fouling, thus reducing the heat transfer efficiency;
- Corrosion of overhead equipment; and,
- The salts carried with the products act as catalyst poisons in catalytic cracking units.

### 4.4.1. Types of Salts in Crude Oil

Salts in the crude oil are mostly in the form of dissolved salts in fine water droplets emulsified in the crude oil. This is called a water-in-oil emulsion, where the continuous phase is the oil and the dispersed phase is the water.



The water droplets are so small that they cannot settle by gravity. Furthermore, these fine droplets have on their surfaces the big asphaltene molecules with the fine solid particles coming from sediments, sands or corrosion products. The presence of these molecules on the surface of the droplets acts as a shield that prevents the droplets from uniting with each other in what is called coalescence. The salts can also be present in the form of salts crystals suspended in the crude oil. Salt removal requires that these salts be ionized in the water. Hence, wash water is added to the crude to facilitate the desalting process as will be explained later.

Going back to the subject of salt types, these are mostly magnesium, calcium and sodium chlorides with sodium chloride being the abundant type. These chlorides, except for NaCl, hydrolyze at high temperatures to hydrogen chloride:



On the other hand, NaCl does not hydrolyze. Hydrogen chloride dissolves in the overhead system water, producing hydrochloric acid, an extremely corrosive acid.

#### 4.4.2. Desalting Process

To remove the salts from the crude oil, the water-in oil emulsion has to be broken, thus producing a continuous water phase that can be readily separated as a simple decanting process. The process is accomplished through the following steps (Abdel-Aal *et al.*, 2003):

- *Water washing:* Water is mixed with the incoming crude oil through a mixing valve. The water dissolves salt crystals and the mixing distributes the salts into the water, uniformly producing very tiny droplets. Demulsifying agents are added at this stage to aide in breaking the emulsion by removing the asphaltenes from the surface of the droplets.
- *Heating:* The crude oil temperature should be in the range of 48.9–54.4 °C (120–130 °F) since the water–oil separation is affected by the viscosity and density of the oil.
- *Coalescence:* The water droplets are so fine in diameter in the range of 1–10 μm that they do not settle by gravity. Coalescence produces larger drops that can be settled by gravity. This is accomplished through an electrostatic electric field between two electrodes. The electric field ionizes the water droplets and orients them so that they are attracted to each other. Agitation is also produced and aides in coalescence. The force of attraction between the water droplets is given by:

$$F = KE^2d^2\left(\frac{d}{s}\right)^4 \quad (4.3)$$

where  $E$  is the electric field,  $d$  is the drop diameter and  $s$  is the distance between drops centres and  $K$  is a constant.

- *Settling*: According to Stock's law the settling rate of the water droplets after coalescence is given by

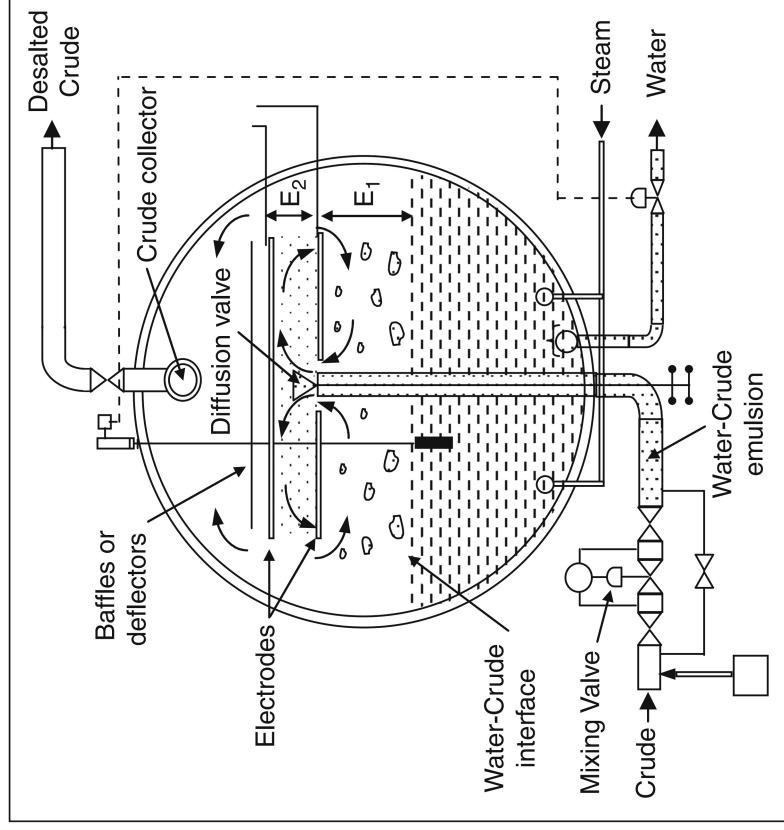
$$\text{Settling rate} = \frac{k(\rho_{\text{H}_2\text{O}} - \rho_{\text{oil}})d^2}{\mu_{\text{oil}}} \quad (4.4)$$

where  $\rho$  is the density  $\mu$  is the viscosity,  $d$  is the droplet diameter and  $k$  is a constant.

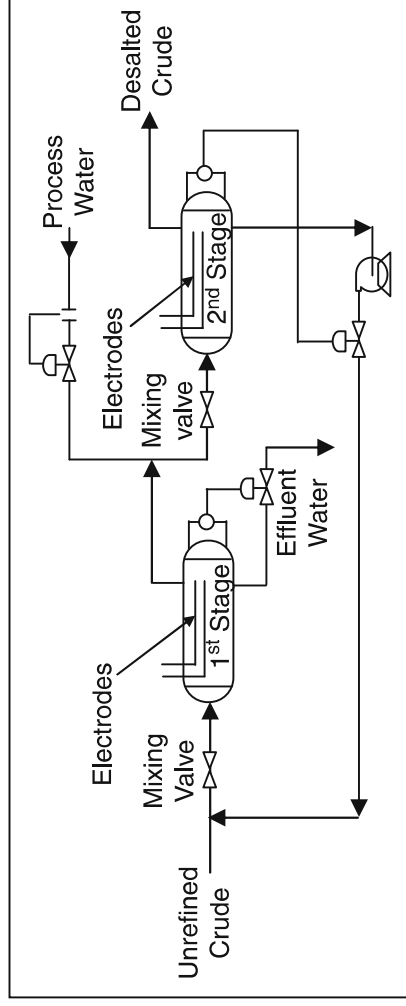
#### 4.4.3. Description of Desalter

A typical desalter contains two metal electrodes as shown in Figure 4.3. A high voltage is applied between these two electrodes. For effective desalting the electric fields are applied as follows:

- A high voltage field called the “secondary field” of about 1000 V/cm between the two electrodes is applied. The ionization of the water droplets and coalescence takes place here



**Figure 4-3** Simplified flow diagram of an electrostatic desalter (Prak *et al.*, 2000)



**Figure 4.4** Two-stage desalting

- A primary field of about 600 V/cm between the water–crude interface and the lower electrode is applied. This field helps the water droplets settle faster.

The desalter of this design achieves 90% salt removal. However 99% salt removal is possible with two-stage desalters as shown in Figure 4.4. A second stage is also essential since desalter maintenance requires a lengthy amount of time to remove the dirt and sediment which settle at the bottom. Therefore, the crude unit can be operated with a one stage desalter while the other is cleaned.

#### 4.4.4. Desalter Operating Variables

For an efficient desalter operation, the following variables are controlled:

- *Desalting temperature:* The settling rate depends on the density and viscosity of the crude. Since increasing the temperature lowers the density and viscosity, the settling rate is increased with temperature based on the crude gravity, typical desalting temperature can vary between 50 and 150 °C (122 and 302 °F).
- *Washing water ratio:* Adding water to the crude oil helps in salt removal. Hence, increasing the wash water rate increases the coalescence rate. Depending on the desalting temperature, a minimum value should be used. For example, Kuwait crude (31.2 API) requires 7–8 vol% water addition relative to the crude rate.
- *Water level:* Raising the water level reduces the settling time for the water droplets in the crude oil, thus improving the desalting efficiency. However, if the water level gets too high and reaches the lower electrode, it shorts out the desalter. Since the primary electric field depends on the distance between the lower electrode and the water–crude interface, it is always better to keep the level constant for stable operation.
- *Washing water injection point:* Usually the washing water is injected at the mixing valve. However, if it is feared that salt deposition may occur in

the preheat exchangers, part or all of the washing water is injected right after the crude feed pump.

- *Demulsifier injection rate:* Demulsifiers are basic copolymers with one end being hydrophilic (loves water and attaches to the surface of the water droplet), and the other end being hydrophobic (loves the oil and is directed to the oil side). When these compounds are adsorbed on the droplet surface, they stabilize the droplet. The demulsifier is added to the crude after the feed pump or before the mixing valve at levels between 3 and 10 ppm of the crude.
- *Type of washing water:* Process water in addition of fresh water is used for desalting. The water should be relatively soft in order to prevent scaling. It should be slightly acidic with a pH in the range of 6. It should be free from hydrogen sulphide and ammonia so as to not create more corrosion problems. Therefore, distillation overhead condensates and process water from other units can be used after stripping.
- *Pressure drop in the mixing valve:* Mixing the washing water with crude oil is necessary in order to distribute the water and dissolve any suspended salts crystals. The pressure drop across the mixing valve determines the mixing efficiency. On the other hand, the mixing process produces finer (smaller diameter) droplets which tend to stabilize the emulsion and make water separation more difficult. Therefore, there is a compromise in selection of the appropriate pressure drop across the mixing valve. A pressure drop between 0.5 and 1.5 bar (7.4 and 22 psi) is used.

One variable which is not mentioned above is the desalter pressure. The operation of the desalter requires that the crude be in the liquid phase during desalting. A typical pressure of 12 bar (176 psia) is necessary to achieve this purpose. When the process control variables are properly adjusted, a 90% salt rejection (2–5 PTB of salts in the desalted crude relative to the raw crude) can be achieved. With a two stage operation the salt rejection can reach 99%. Any remaining salts are neutralized by the injection of sodium hydroxide which reacts with the calcium and magnesium chloride to produce sodium chloride.



NaCl does not hydrolyze to the corrosive hydrogen chloride.

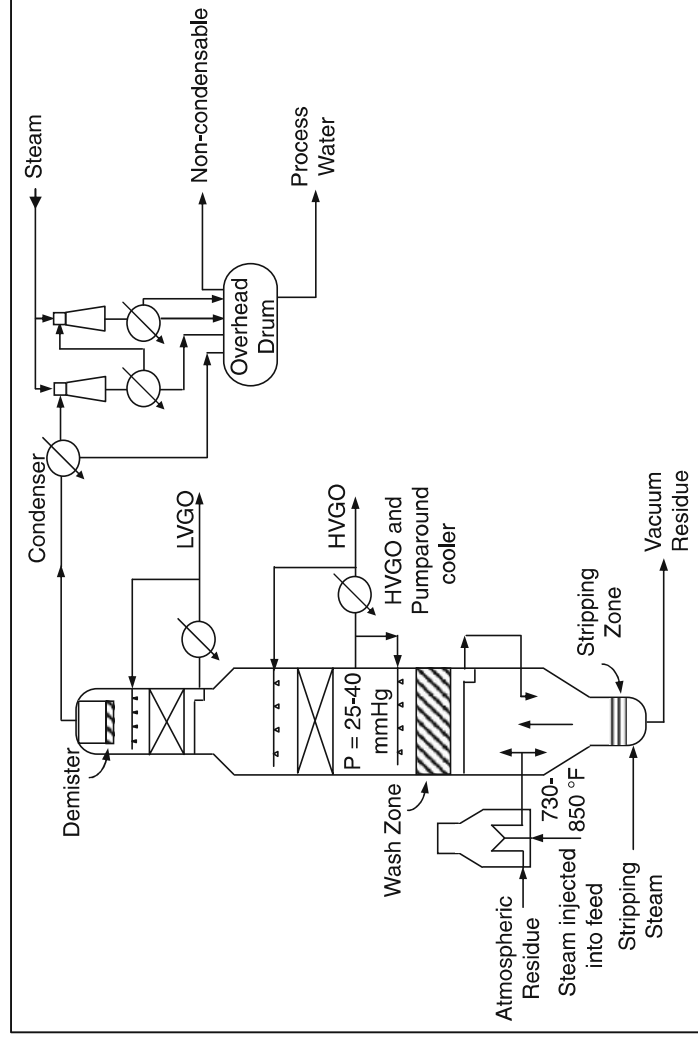
## 4.5. VACUUM DISTILLATION

To extract more distillates from the atmospheric residue, the bottom from the atmospheric CDU is sent to the vacuum distillation unit. The vacuum unit distillates are classified as light vacuum gas oil (LVGO),

medium vacuum gas oil (MVGGO), and heavy vacuum gas oil (HVGO). In addition a vacuum residue is produced. If the distillates are feed to down stream conversion process, their the sulphur, metal and asphaltene content should be reduced by hydrotreating or hydroprocessing. In some refineries the whole atmospheric residue is hydroprocessed before vacuum distillation. The vacuum unit can also be used to produce lubrication oil grade feed stocks. This depends on the quality of the crude oil feed to the refinery as only special types of crude can produce lube grade feed stocks.

#### 4.5.1. Process Description

Figure 4.5 shows the flow diagram of the vacuum distillation unit. The atmospheric residue can be sent directly to the vacuum unit after heat extraction in the crude preheat exchangers train. If it is sent to storage, the temperature should not be below 150 °C (300 °F) to control the viscosity necessary for proper flow. It is then heated in several exchangers by the hot products and pumparounds of the vacuum unit. Final heating to 380–415 °C (716–779 °F) is done in a fired heater. To minimize thermal cracking and coking, steam is injected in the heater tube passes. The feed enters the vacuum tower at the lower part of the column. As in the case of atmospheric distillation, a 3–5 vol% overflash is maintained (i.e., 3–5 vol% vapours are produced more than the total products withdrawn above the flash zone). This is to provide some fractionation between the HVGO



**Figure 4-5** Process flow diagram of the vacuum distillation unit

drawoff tray and the flash zone, thereby controlling its end point. The distillate is withdrawn as LVGO and two other cuts, MVGO and HVGO. The two cuts of MVGO and HVGO are necessary to extract heat from the tower at a more advantageous level from the HVGO pumparound.

Vacuum distillation columns are equipped with packing for fractionation and heat exchange zones. This is in order to reduce the pressure drop in the column which is necessary for creating a low vacuum in the lower section of the column. The bottom zone is equipped with valve trays. The vapours from the flash zone go through a wash and fractionation zone where the heavy ends are condensed with HVGO reflux. Further up, the column sections (consisting of a heat exchange and fractionation zone) are separated by sprays of liquid from the pumparound or the internal reflux.

Vacuum distillation units have a system to create the vacuum that uses either ejectors or a combination of ejectors and liquid ring pumps. Ejectors recompress the gases through a nozzle where vapours from the column are sucked into the venturi section of the nozzle by a stream of medium or low pressure steam. The vapour phase at the ejector exit is partially condensed in an exchanger with cooling water. The liquid phase is then sent to the overhead drum. The vapour phase goes from the condenser to another ejector–condenser stage.

Liquid ring pumps are similar to rotor gas compressors. One pump can replace two or three stages of ejectors in dry or wet type vacuum distillation. They do not use steam and can significantly reduce hydrocarbon-rich aqueous condensates in a system using ejectors. Systems with ejectors are much more flexible and rapid to put into operation. The higher investments required by liquid ring pumps are offset by reduced steam consumption and lower installation costs.

## 4.6. CRUDE DISTILLATION MATERIAL BALANCE

In this section, we will consider the product slate from the crude distillation unit (atmospheric and vacuum distillation). For a given crude oil feed rate, the flow rates and the properties of the various products are calculated. Furthermore, the properties of the various cuts are estimated. For this purpose, crude assay data have to be provided. In addition, the desired products from the atmospheric and vacuum distillation towers are assigned along with their respective boiling point ranges.

### 4.6.1. Crude Assay Data

Laboratory crude assay reports provide TBP for the whole crude as explained in Section 3.2.3. The maximum temperature that can be measured with this test is in the range of 496–526 °C (925–975 °F), depending on the crude oil. The actual end point of the crude oil can be

as high as 790 °C (1454 °F). Extrapolation of the measured portion of the TBP curve to a volume percent approaching the end point is discussed in Sections 3.3.1 and 3.3.2.

The products from the crude distillation unit (atmospheric and vacuum distillation) are then chosen. Typical products with their end points are shown in example E4.2. The volume percent of each cut is determined by calculating the cumulative volume percent for each fraction at its end boiling point from any of the fit procedures explained earlier in Section 3.3 and then calculating the difference. The volume fraction of the vacuum residue is obtained as 100 minus the cumulative volume percent at the end boiling point of the vacuum distillate.

#### Example E4.2

Consider the Kuwait export crude with the following TBP–vol% data:

Volume %	TBP (°C)
5	40
10	85
30	215
50	340
70	495

Determine the volumetric yield, average boiling point, molecular weight and specific gravity for the products shown in Table E4.2.1.

**Table E4.2.1** Typical CDU products and their end boiling points

Cut #	Product	End point (°C)
1	Off gas	10
2	Light straight run naphtha	70
3	Naphtha	180
4	Kerosene	240.0
5	Light diesel	290.0
6	Heavy diesel	340.0
7	Atm. gas oil	370.0
8	Vacuum gas oil	390.0
9	Vac. distillate	550.0
9	Vac. residue	—

#### Solution:

From the fifth order polynomial fit of TBP data versus vol% (Section 3.3.1) and given the end point of each cut, we can calculate the cumulative volume

percent. For example, the cumulative volume percent for the off gas and the light straight run naphtha is 1.33 and 8.6%, respectively. Therefore, the volume percent of the light straight run naphtha is 8.6–1.33 or 7.27%. The average normal boiling point for each cut is calculated at the mid percent of each cut. This means that we can use polynomial fit again but this time, at a cumulative volume percent of the cut (in this case the light straight run naphtha)  $1.33 + 7.27/2$  or 4.965%. The average boiling point is then calculated to be 43.6 °C. From equation (3.29) with this average boiling point, the molecular weight is 71.2. The specific gravity is calculated, from the molecular weight and average boiling point using equation (3.11), to be 0.680. The calculations for the other cuts are shown below. These calculations can be done using the Excel spreadsheet Crude Unit. Table E4.2.2 shows the results.

**Table E4.2.2** Volumetric yield of cuts and their properties

Cut #	Product	Vol% of cut	End point °C	Average NBP, °C	Molecular weight	Specific gravity
1	Off gas	1.33	10	2.5	55.6	0.635
2	Lt. St. Run Naph	7.27	70	43.6	71.2	0.680
3	Naphtha	16.56	180	131.5	112.4	0.754
4	Kerosene	10.05	240.0	209.9	159.8	0.803
5	Light Diesel	7.83	290.0	264.4	200.1	0.831
6	Heavy Diesel	6.99	340.0	314.3	243.6	0.854
7	Atm. Gas Oil	3.84	370.0	354.8	284.7	0.870
8	Vacuum Gas Oil	2.43	390.0	379.9	313.2	0.880
9	Vac. Distillate	18.34	550.0	466.6	434.7	0.910
9	Vac. Residue	26.70	–	688.6	1150.3	0.969

The TBP curve for each product can also be obtained from the volume percent of each cut as shown in the example E4.3.

### Example E4.3

Plot the true boiling point curve for the kerosene product from Table E4.2.2.

#### Solution:

Volume percent of kerosene = 9.51%

The cumulative vol% at the IBP of kerosene =  $1.33 + 7.27 + 16.56 = 25.16\%$

The cumulative vol% at the EBP of kerosene =  $25.16 + 10.05 = 35.21\%$



The cumulative vol% at the 10% of kerosene cut =  $0.1(10.05) + 25.16 = 26.17\%$ . At vol% of 26.17% the estimated TBP is  $186.73\text{ }^\circ\text{C}$  using the same polynomial fit that was used in E4.2. The procedure is repeated at 20% of kerosene volume which yield 27.17% and TBP of  $189.26\text{ }^\circ\text{C}$ . Figure E4.3.1 shows the TBP curve for kerosene which starts at IBP of  $180\text{ }^\circ\text{C}$  and ends at EBP of  $240\text{ }^\circ\text{C}$ .

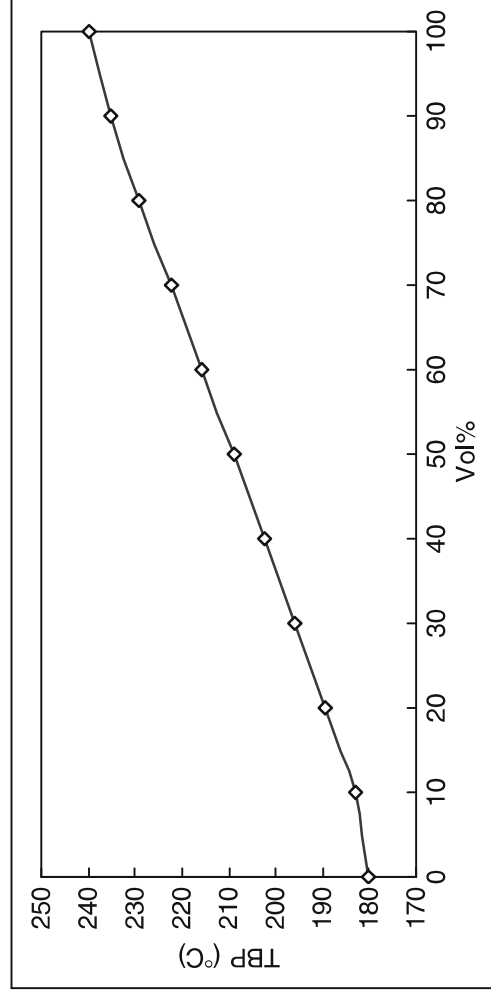


Figure E4.3.1 TBP curve for kerosene cut

#### 4.6.2. Material Balance

For a given feed flow rate in barrels per calendar day (BPCD), and given the specific gravity of the feed, the mass flow rate of the feed can be calculated. Then the volumetric throughput of each product cut in BPCD and in kg/h (lb/h) is calculated. The mass flow rate of the vacuum residue is calculated as the difference between the feed mass flow rate and the total mass flow rate of all products lighter than the vacuum residue. Similarly from the mass flow rate of the residue and its gravity, its volumetric flow rate can be calculated. The sum of the volumetric flow rates of the products may become higher than the feed. This is expected, since the volume change of mixing is negative for petroleum fractions. When the fractions are mixed, the total volume of the mixture is less than the sum of the volumes of the fractions. Example E4.4 illustrates the material balance calculations.

**Example E4.4**

For Kuwait export crude (API = 31.5) given in the example E4.2, perform a material balance CDU calculations for a feed rate of 100,000 BPCD.

**Solution:**

Mass flow rate of feed is 573,844 kg/h. The volumetric product yield is converted to mass flow rates. The mass flow rate of the residue is the difference between the feed rate and the sum of product lighter than the vacuum residue. The results are shown in Table E4.4.1

**Table E4.4.1** Crude unit material balance

	EBP, °C	Vol%	BPCD	SG	kg/h
<b>Feed</b>					
Crude oil		100	100,000.0	0.8685	573,844
<b>Products</b>					
Off gas	10	1.33	1330	0.635	5,580
Lt. St. Run	70	7.27	7270	0.680	32,664
Naphtha	180	16.56	16560	0.754	82,500
Kerosene	240.0	10.05	10050	0.803	53,322
Light diesel	290.0	7.83	7830	0.831	42,992
Heavy Diesel	340.0	6.99	6990	0.854	39,442
Atm. Gasoil	370.0	3.84	3840	0.870	22,074
Vac. Gasoil	390.0	2.43	2430	0.880	14,129
Vac. Distil	550.0	18.34	18340	0.910	110,272
Vac. Residue	–	26.70	26700	0.969	170,869
		101.33	101330		573,844

**4.6.3. Sulphur Material Balance**

For crude oil containing significant amounts of sulphur, it is necessary to make a sulphur balance around the crude unit. The sulphur content of the crude oil should be given. The sulphur content of the products should also be known. If these data are not available the estimation of the sulphur content of petroleum fractions can be calculated using the following equations in the two ranges of molecular weight  $M$  (Riazi *et al.*, 1999):

For fractions with  $M < 200$

$$\text{wt}\%S = 177.448 - 170.946R_i + 0.2258m + 4.054SG \quad (4.6)$$

And for fractions with  $M \geq 200$

$$\text{wt}\%S = -58.02 + 38.463R_i - 0.023m + 22.4SG \quad (4.7)$$

where SG is the specific gravity,  $R_i$  is the refractivity intercept defined as

$$R_i = n - \frac{d}{2} \quad (4.8)$$

where  $n$  and  $d$  are the refractive index and density of liquid hydrocarbon at 68 °F (20 °C) in g/cm<sup>3</sup>. Refer to Chapter 3, Sections 3.4.3 and 3.4.4 on how to calculate these parameters. The parameter  $m$  is defined as

$$m = M(n - 1.475) \quad (4.9)$$

After calculating the wt% S in each product except the vacuum residue, the amount of sulphur is calculated by multiplying this percentage by the mass rate of each product. The sulphur in the vacuum residue is calculated from the difference of the total sulphur in the crude feed and the total sulphur in the products.

### Example E4.5

Perform a sulphur balance on the crude unit of example E4.4, knowing that the sulphur content of Kuwait export is 2.52 wt%.

#### Solution:

The wt% S in each product except the vacuum residue is calculated using equations (4.6) and (4.7) in Table E4.5.1.

**Table E4.5.1** Calculation of sulphur content of the products

	$T_b, K$	SG	$d, 20^\circ C$	$I$	$n, 20^\circ C$	$R_i$	MW	$m$	S, wt%
Off gas	275.7	0.635	0.6298	0.2188	1.3565	1.0416	55.6	-6.5890	0.479
Lt. St. Run	316.8	0.680	0.6754	0.2324	1.3814	1.0437	71.2	-6.6710	0.287
Naphtha	404.7	0.754	0.7496	0.2540	1.4219	1.0471	112.4	-5.9731	0.161
Kerosene	483.1	0.803	0.7989	0.2681	1.4488	1.0493	159.8	-4.1919	0.379
Light diesel	537.5	0.831	0.8271	0.2760	1.4642	1.0506	200.1	-2.1700	1.054
Heavy Diesel	587.4	0.854	0.8499	0.2823	1.4766	1.0516	243.6	0.3786	1.541
Atm. Gasoil	627.9	0.870	0.8666	0.2869	1.4857	1.0523	284.7	3.0372	1.880
Vac. Gasoil	653.1	0.880	0.8764	0.2896	1.4910	1.0528	313.2	5.0039	2.069
Vac. Distil	739.7	0.910	0.9070	0.2979	1.5077	1.0541	434.7	14.1990	2.591

The Sulphur material balance is shown in Table E4.5.2.

**Table E4.5.2** Sulphur material balance

	Mass (kg/hr)	wt% S	kg/hr S
<b>Feed</b>	573,844	2.52	14,461
<b>Products</b>			
Off gas	5,580	0.479	26.7
Lt. St. Naphtha	32,664	0.287	93.8
Naphtha	82,500	0.161	132.8
Kerosene	53,322	0.379	202.1
Light diesel	42,992	1.054	453.1
Heavy diesel	39,442	1.541	607.8
Atm. Gas Oil	22,074	1.880	415.0
Vac. Gas Oil	14,129	2.069	292.3
Vac. Distillate	110,272	2.591	2857.1
Vac. Residue	170,869	5.490	9380.3
Total	573,844	2.52	14,461

## 4.7. DESIGN OF CRUDE DISTILLATION UNITS USING PROCESS SIMULATORS

Although the CDU contains many pieces of equipment with different unit operations, the most important one is the distillation column. The design and simulation of the atmospheric and vacuum distillation columns is the most important part of the whole unit design. It determines the quality of the product and since it is energy intensive, it can have a substantial effect on the economics of the unit. Thus, improving the design or operation of the distillation columns can increase the profitability of the refinery. The simulation or design of the distillation columns involves dividing the crude oil into pseudo-components as outlined in Chapter 3. Then a thermodynamic model is chosen for vapour liquid equilibrium and thermodynamic properties calculations. A good model is the cubic equations of state, and the Peng–Robinson equation is one of the most widely used models for hydrocarbon and petroleum mixtures. Next, the unit operations stage-wise or “tray to tray” distillation calculations are performed. The mass, energy balance and vapour liquid equilibrium relations for each tray are written and solved together, subject to certain specification for the products. Computer simulation programs such as UNISIM are used for quick simulation of CDU units.

**Example E4.6**

Perform a material balance for a CDU using UNISIM for 100,000 BPCD of 29 API crude with the following assay.

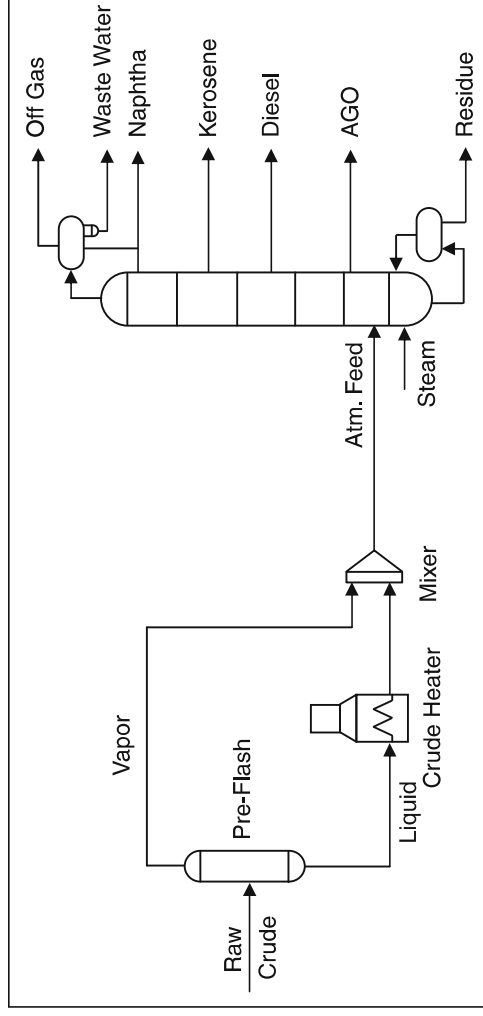
vol%	TBP (°C)	vol%	TBP (°C)
0.0	-9.44	40.0	273.33
4.5	32.22	50.0	326.67
9.0	73.89	60.0	393.33
14.5	115.56	70.0	473.89
20.0	154.44	76.0	520.56
30.0	223.89	80.0	546.11

The crude is fed to a pre-flash separator operating at 450 °F and 75 psia. The vapour from this separator bypasses the crude furnace and is remixed with the hot (650 °F) liquid leaving the furnace. The combined stream is then fed to the distillation column (Figure E4.6.1). The column operates with a total condenser, three side strippers and three pumparounds (Figure E4.6.2). (UNISIM 2007).

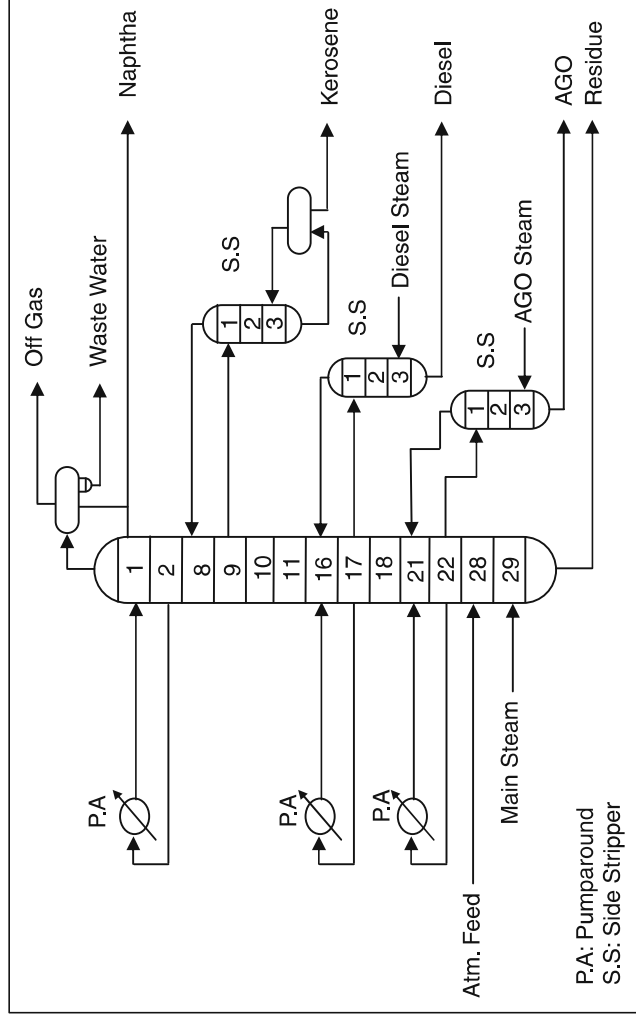
**Solution:**

In the oil environment and oil manager data entry of the UNISIM software, the crude assay is entered as vol% and TBP. The yield distribution of the products is shown in Figure E4.6.3.

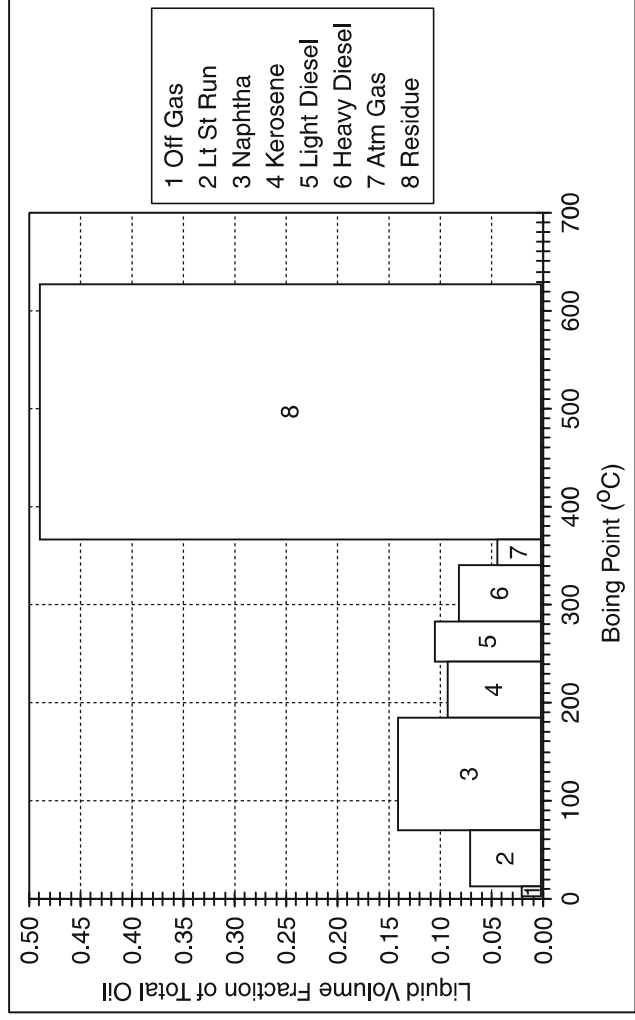
The distillation column has three inlet steam streams, with pressures and flow rates listed in Table E4.6.1. The main distillation column contains 29 stages (see Figure E4.6.2). The overhead condenser operates at 19.7 psia and the bottoms at 32.7 psia. The side stripper connections are also shown in Figure E4.6.2.



**Figure E4.6.1** Crude distillation flowsheet



**Figure E4.6.2** Distillation column



**Figure E4.6.3** Crude distribution curve

The distillation column specifications are listed in Table E4.6.1.

**Table E4.6.1** Column specifications

Item	Specification
Kero-SS product flow	61.61 m <sup>3</sup> /h
Diesel-SS product flow	127.5 m <sup>3</sup> /h
AGO-SS product flow	29.81 m <sup>3</sup> /h
Pumparound 1 rate	331.2 m <sup>3</sup> /h
Pumparound 1 duty	-5.8 × 10 <sup>7</sup> kJ/h
Pumparound 2 rate	198.7 m <sup>3</sup> /h
Pumparound 2 duty	-3.7 × 10 <sup>7</sup> kJ/h
Pumparound 3 rate	198.7 m <sup>3</sup> /h
Pumparound 3 duty	-3.7 × 10 <sup>7</sup> kJ/h
Naphtha product rate	152.4 m <sup>3</sup> /h
Liquid flow	23.19 m <sup>3</sup> /h
Kero reboiler duty	7.913 × 10 <sup>6</sup> kJ/h
Vapour product flow	0.0
Reflux ratio	1

The solved material and energy balance is shown in Table E4.6.2.

**Table E4.6.2** Material and energy balance

Name	Vapour fraction	<i>T</i> (°C)	<i>P</i> (kPa)	Molar flow (kg mol/h)	Mass flow (kg/h)	Molar enthalpy (kJ/kg mol)
Hot crude	0.6447	343.3	448.2	2843	5.829 × 10 <sup>5</sup>	-2.673 × 10 <sup>5</sup>
Main steam	1.0	190.6	1034	188.8	3402	-2.359 × 10 <sup>5</sup>
Diesel steam	1.0	148.9	344.7	75.54	1361	-2.370 × 10 <sup>5</sup>
AGO steam	1.0	148.9	344.7	62.95	1134	-2.370 × 10 <sup>5</sup>
Off gas	1.0	41.92	135.8	0	0	-
Naphtha	0.0	41.92	135.8	1283	1.124 × 10 <sup>5</sup>	-1.921 × 10 <sup>5</sup>
Waste water	0.0	41.92	135.8	316.8	5707	-2.841 × 10 <sup>5</sup>
Kerosene	0.0	236.5	205.8	320.9	5.079 × 10 <sup>4</sup>	-2.678 × 10 <sup>5</sup>
Diesel	0.0	253.0	213.6	509.8	1.106 × 10 <sup>5</sup>	-3.583 × 10 <sup>5</sup>
AGO	0.0	300.4	218.6	91.7	2.708 × 10 <sup>4</sup>	-4.493 × 10 <sup>5</sup>
Residue	0.0	355.1	225.5	647.8	2.822 × 10 <sup>5</sup>	-5.953 × 10 <sup>5</sup>

## QUESTIONS AND PROBLEMS

4.1. With 100,000 BPD of the following crude (API = 36), estimate the products of the atmospheric distillation column. If the atmospheric residue of the crude is taken at 650+ °F. It enters in a vacuum distillation tower to give three products: light vacuum gas oil (650–850 °F), heavy vacuum gas oil (850–1050 °F) and vacuum residue (1050+ °F). Calculate the mass flow rate of these products. Then calculate the sulphur content (lb/hr) for each product.

ASTM D86 (°F)	vol%	Cum vol%	SG
86	0.0	0.0	
122	0.5	0.5	0.6700
167	1.2	1.7	0.6750
212	1.6	3.3	0.7220
257	2.7	6.0	0.7480
302	3.1	9.1	0.7650
347	3.9	13.0	0.7780
392	4.7	17.7	0.7890
437	5.7	23.4	0.8010
482	8.0	31.4	0.8140
527	10.7	42.1	0.8250
584	5.0	47.1	0.8450
636	10.0	57.1	0.8540
689	7.8	64.9	0.8630
742	7.0	71.9	0.8640
794	6.5	78.4	0.8890
–	20.8	99.2	0.9310

- 4.2. What is the concentration of salt in the washed solution for a desalted oil containing 1 lb of NaCl/1000 bbl of oil, if the inlet oil contains 20 lb of NaCl/1000 bbl of oil? Assume the amount of water in oil before and after desalting is the same and the wash water used is 10 vol % of oil.
- 4.3. Sketch a combined atmospheric and vacuum distillation for a crude oil feed.
- 4.4. Sketch a side view for the desalter.



## REFERENCES

- Abdel-Aal, H. K., Aggor, M., and Fahim, M. A. (2003). "Petroleum and Gas Field Processing," Marcel Dekker, New York.
- Ptak, C., Bonfils, P., and Marty, C. (2000). "Distillation Absorption and Stripping in the Petroleum Industry," Chapter 5, In "Separation Processes," Petroleum Refining, Vol. 2, Wauquier, P., ed., TECHNIP, France.
- Riazi, M. R., Nasimi, N., and Roomi, Y. (1999). Estimating sulfur content of petroleum products and crude oils. *IEC Res.* **38**, 11.
- UNISIM Design Suite R370, Honeywell Process Solutions, Calgary, Alberta, Canada.