CHAPTER ONE

Refinery Distillation

Crude oil as produced in the oil field is a complex mixture of hydrocarbons ranging from methane to asphalt, with varying proportions of paraffins, naphthenes, and aromatics. The objective of crude distillation is to fractionate crude oil into light-end hydrocarbons (C_1–C_4), naphtha/gasoline, kerosene, diesel, and atmospheric resid. Some of these broad cuts can be marketed directly, while others require further processing in refinery downstream units to make them saleable.

The first processing step in the refinery, after desalting the crude, is separation of crude into a number of fractions by distillation. The distillation is carried out at a pressure slightly above atmospheric. This is necessary for the following considerations:

1. To raise the boiling point of the light-end carbons so that refinery cooling water can be used to condense some of the C_3 and C_4 in the overhead condenser.
2. To place the uncondensed gas under sufficient pressure to allow it to flow to the next piece of processing equipment.
3. To allow for pressure drop in the column.

Crude oil is preheated in exchangers and finally vaporized in a fired furnace until approximately the required overhead and sidestream products are vaporized. The furnace effluent is flashed into the crude column flash zone, where the vapor and liquid separate. The liquid leaving the flash zone still contains some distillate components, which are recovered by steam stripping. After steam stripping, the bottom product, also known as *reduced crude*, is discharged from the tower. The bottom temperature is limited to 700–750°F to prevent cracking.

The atmospheric resid is fed to a furnace, heated to 730–770°F and next to a vacuum tower operated at a minimum practical vacuum (80–110 mm Hg). The operating conditions are dictated by cracking and product quality required. The objectives of vacuum distillation is generally to separate vacuum gas oil (VGO) from reduced crude. The VGO may become feedstock for FCCU or hydrocracker units or used to make lube base stocks. Depending on the end use, there may be one or more sidestreams. The bottom stream from the vacuum distillation unit may be used to produce bitumen or used for fuel oil production after mixing it with small amounts of cutter stocks (in the diesel/kerosene range).

If the crude contains very high percentages of light-ends, a flash drum or a prefractionator with an overhead condensing system is added ahead of atmospheric tower. The prefractionator is designed to recover most of the light-ends and a part of the light naphtha. The bottom stream from prefractionator becomes feed to atmospheric tower.

PROCESS VARIABLES

The following variables are important in the design of crude columns:

1. The nature of the crude—water content, metal content, and heat stability. The heat stability of the crude limits the temperature to which crude can be heated in the furnace without incipient cracking.
2. Flash zone operating conditions—flash zone temperature is limited by advent of cracking; flash zone pressure is set by fixing the reflux drum pressure and adding to it to the line and tower pressure drop.
3. Overflash is the vaporization of crude over and above the crude overhead and sidestream products. Overflash is generally kept in the range of 3–6 LV% (LV = Liquid Volume). Overflash prevents coking of wash section plates and carryover of coke to the bottom sidestream and ensures a better fractionation between the bottom sidestream and the tower bottom by providing reflux to plates between the lowest sidestream and the flash zone. A larger overflash also consumes larger utilities; therefore, overflash is kept to a minimum value consistent with the quality requirement of the bottom sidestream.
4. In steam stripping, the bottom stripping steam is used to recover the light components from the bottom liquid. In the flash zone of an atmospheric distillation column, approximately 50–60% of crude is vaporized. The unvaporized crude travels down the stripping section of the column containing four to six plates and is stripped of any low boiling-point distillates still contained in the reduced

crude by superheated steam. The steam rate used is approximately 5–10 lbs/bbl of stripped product.[1] The flash point of the stripped stream can be adjusted by varying the stripping steam rate.
5. Fractionation is the difference between the 5% ASTM curve of a heavy cut and the 95% point on the ASTM curve of a lighter cut of two adjacent side products. A positive difference is called a *gap*,[2] and a negative difference is called an *overlap*.

The design procedures used for atmospheric and vacuum distillation are mostly empirical, as crude oil is made of a very large number of hydrocarbons, from methane to asphaltic pitch. The basic data required, refinery crude distillation column, and a brief overview of the design procedures follow.

TRUE BOILING POINT CURVE

The composition of any crude oil sample is approximated by a true boiling point (TBP) curve. The method used is basically a batch distillation operation, using a large number of stages, usually greater than 60, and high reflux to distillate ratio (greater than 5). The temperature at any point on the temperature-volumetric yield curve represents the true boiling point of the hydrocarbon material present at the given volume percent point distilled. TBP distillation curves are generally run only on the crude and not on petroleum products. Typical TBP curves of crude and products are shown in Figures 1-1 and 1-2.

ASTM DISTILLATION

For petroleum products, a more rapid distillation procedure is used. This is procedure, developed by the American Society for Testing and Materials (ASTM), employs a batch distillation procedure with no trays or reflux between the still pot and the condenser.[3] The only reflux available is that generated by heat losses from the condenser.

EQUILIBRIUM FLASH VAPORIZATION

In this procedure,[4] the feed material is heated as it flows continuously through a heating coil. Vapor formed travels along in the tube with the remaining liquid until separation is permitted in a vapor separator or

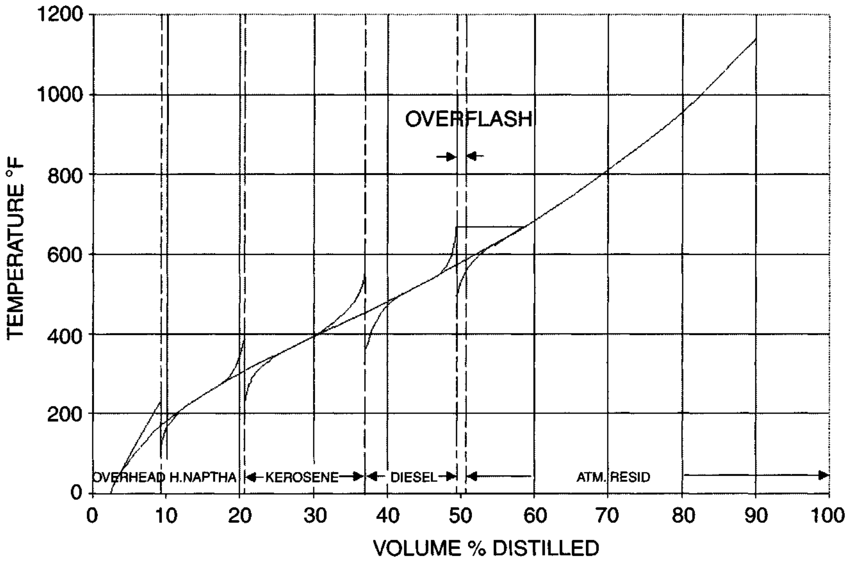

Figure 1-1. TBP curves of feed and products atmosphere distillation tower.

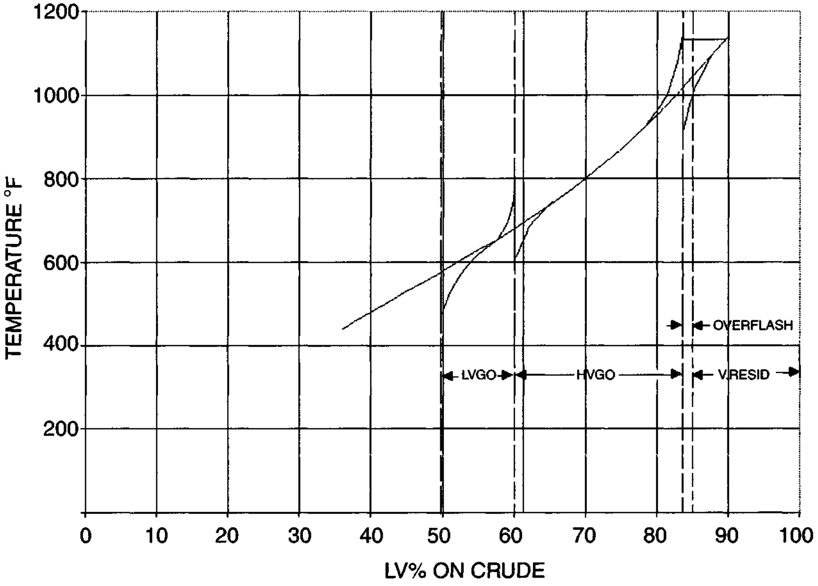

Figure 1-2. TBP curve of feed and products for vacuum tower.

vaporizer. By conducting the operation at various outlet temperatures, a curve of percent vaporized vs. temperature may be plotted. Also, this distillation can be run at a pressure above atmospheric as well as under vacuum. Equilibrium flash vaporization (EFV) curves are run chiefly on crude oil or reduced crude samples being evaluated for vacuum column feed.

CRUDE ASSAY

The complete and definitive analysis of a crude oil is called *crude assay*. This is more detailed than a crude TBP curve. A complete crude assay contains some of the following data:

1. Whole crude salt, gravity, viscosity, sulfur, light-end carbons, and the pour point.
2. A TBP curve and a mid-volume plot of gravity, viscosity, sulfur, and the like.
3. Light-end carbons analysis up to C_8 or C_9.
4. Properties of fractions (naphthas, kerosenes, diesels, heavy diesels, vacuum gas oils, and resids). The properties required include yield as volume percent, gravity, sulfur, viscosity, octane number, diesel index, flash point, fire point, freeze point, smoke point, and pour point.
5. Properties of the lube distillates if the crude is suitable for manufacture of lubes.
6. Detailed studies of fractions for various properties and suitability for various end uses.

PROCESS DESIGN OF A CRUDE DISTILLATION TOWER

A very brief overview of the design steps involved follows:

1. Prepare TBP distillation and equilibrium flash vaporization curves of the crude to be processed. Several methods are available for converting TBP data to EFV curves.
2. Using crude assay data, construct TBP curves for all products except gas and reduced crude. These are then converted to ASTM and EFV curves by Edmister,[5] 'Maxwell,'[6] or computer methods.
3. Prepare material balance of the crude distillation column, on both volume and weight bases, showing crude input and product output.

Also plot the physical properties, such as cut range on TBP and LV%, mid vol% vs. SG, molecular weight, mean average boiling point, and enthalpy curves for crude and various products.

4. Fractionation requirements are considered next. Ideal fractionation is the difference between the 5% and 95% points on ASTM distillation curves obtained from ideal TBP curves of adjacent heavier and lighter cuts. Having fixed the gaps as the design parameter, the ideal gap is converted into an actual gap. The difference between the ideal gap and actual gap required is deviation. Deviation is directly correlated with (number of plates × reflux).
5. The deviation or gap can be correlated with an *F* factor,[7] which is the product of number of plates between two adjacent side draws offstream and internal reflux ratio. *Internal reflux* is defined as volume of liquid (at 60°F) of the hot reflux below the draw offplate of the lighter product divided by the volume of liquid products (at 60°F) except gas, lighter than the adjacent heavier products. This implies that the reflux ratio and the number of plates are interchangeable for a given fractionation, which holds quite accurately for the degree of fractionation generally desired and the number of plates (5–10) and reflux ratios (1–5) normally used. The procedure is made clear by Example 1-1.

NUMBER OF TRAYS

Most atmospheric towers have 25–35 trays between the flash zone and tower top. The number of trays in various sections of the tower depends on the properties of cuts desired from the crude column, as shown in Table 1-1.

The allowable pressure drop for trays is approximately 0.1–0.2 psi, per tray. Generally, a pressure drop of 5 psi is allowed between

Table 1-1
Number of Trays between Side Draws in Crude Distillation Unit

SEPARATION	NUMBER OF TRAYS
NAPHTHA–KEROSENE	8–9
KEROSENE–LIGHT DIESEL	9–11
LIGHT DIESEL–ATM RESID	8–11
FLASH ZONE TO FIRST DRAW TRAY	4–5
STEAM STRIPPER SECTION	4–6

Table 1-2
Typical Separation Obtainable in Atmospheric and Vacuum Towers

SEPARATION	(5–95) GAP°F
NAPHTHA–KEROSENE	12°F GAP
KEROSENE–LIGHT DIESEL	62°F OVERLAP
LIGHT DIESEL–HEAVY DIESEL	169°F OVERLAP
HEAVY DIESEL–VGO	70°F OVERLAP
VGO–VACUUM BOTTOMS	70°F OVERLAP

OVERLAP IS A GAP WITH A NEGATIVE SIGN.

the flash zone and the tower top. Flash zone pressure is set as the sum of reflux drum pressure and combined pressure drop across condenser and trays above the flash zone. A pressure drop of 5 psi between the flash zone and furnace outlet is generally allowed.

FLASH ZONE CONDITIONS

The reflux drum pressure is estimated first. This is the bubble point pressure of the top product at the maximum cooling water temperature. The flash zone pressure is then equal to reflux drum pressure plus pressure drop in the condenser overhead lines plus the pressure drop in the trays.

Before fixing the flash zone temperature, the bottom stripping steam quantity and overflash are fixed. The volume percentage of strip-out on crude is calculated using available correlations.[8] If D is the sum of all distillate streams, V is percent of vaporization in the flash zone, OF is overflash, and ST is strip out, then

$$V = D + OF - ST$$

From the flash curve of the crude, the temperature at which this vaporization is achieved at flash zone pressure is determined. This temperature should not exceed the maximum permissible temperature. If it does, the quantity of overflash and stripping steam are changed until a permissible temperature is obtained.

The temperature at which a crude oil begin to undergo thermal decomposition varies from crude to crude, depending on its composition (naphthenic, paraffinic, or aromatic base) and the trace metals present in the crude. Decomposition temperature can be determined only by actual test runs. For most paraffinic and naphthenic crudes, it is in the range of 650–670°F.

COLUMN OVERHEAD TEMPERATURE

The column top temperature is equal to the dew point of the overhead vapor. This corresponds to the 100% point on the EFV curve of the top product at its partial pressure calculated on the top tray.

A trial and error procedure is used to determine the temperature:

1. The temperature of reflux drum is fixed, keeping in view the maximum temperature of the cooling medium (water or air).
2. Estimate a tower overhead temperature, assuming steam does not condense at that temperature.
3. Run a heat balance around top of tower to determine the heat to be removed by pumpback reflux. Calculate the quantity of pumpback reflux.
4. Calculate the partial pressure of the distillate and reflux in the overhead vapor. Adjust the 100% point temperature on the distillate atmospheric flash vaporization curve to the partial pressure.
5. Repeat these steps until the calculated temperature is equal to the one estimated.
6. Calculate the partial pressure of steam in the overhead vapor. If the vapor pressure of steam at the overhead temperature is greater than the partial pressure of steam, then the assumption that steam does not condense is correct. If not, it is necessary to assume a quantity of steam condensing and repeat all steps until the partial pressure of steam in the overhead vapor is equal to the vapor pressure of water at overhead temperature. Also, in this case, it is necessary to provide sidestream water draw-off facilities.
7. To calculate overhead gas and distillate quantities, make a component analysis of total tower overhead stream consisting of overhead gas, overhead distillate, pumpback reflux, and steam. Next make a flash calculation on total overhead vapor at the distillate drum pressure and temperature.
8. The overhead condenser duty is determined by making an enthalpy balance around the top of the tower.

BOTTOM STRIPPING

To determine the amount of liquid to be vaporized by the stripping steam in the bottom of the tower, it is necessary to construct the flash curve of this liquid (called the *initial bottoms*). The flash curve of the reduced crude can be constructed from the flash curve of the whole crude.[9] It is assumed that the initial bottom is flashed in the presence of stripping steam at the pressure existing on top of the stripping plate and at the exit temperature of liquid from this plate.

Approximately 50–60% of the crude is vaporized in the flash zone of the atmospheric tower. The unvaporized crude travels down the stripping section of the tower, containing four to six plates, and is stripped of any remaining low-boiling distillates by superheated steam at 600°F. The steam rate used is approximately 5–10 lb/bbl of stripped product. The flash point of the stripped product can be adjusted by varying stripping steam rate.

SIDESTREAM STRIPPER

Distillate products (kerosene and diesel) are withdrawn from the column as sidestream and usually contain material from adjacent cuts. Thus, the kerosene cut may contain some naphtha and the light diesel cut may contain some kerosene-range boiling material. These side cuts are steam stripped using superheated steam, in small sidestream stripper columns, containing four to six plates, where lower-boiling hydrocarbons are stripped out and the flash point of the product adjusted to the requirements.

REFLUX

In normal distillation columns, heat is added to the column from a reboiler and removed in an overhead condenser. A part of the distillate condensed in overhead condenser is returned to the column as reflux to aid fractionation. This approach is not feasible in crude distillation because the overhead temperature is too low for recovery of heat. Also the vapor and liquid flows in column increase markedly from bottom to top, requiring a very large-diameter tower. To recover the maximum heat and have uniform vapor and liquid loads in the column, intermediate refluxes are withdrawn, they exchange heat with incoming crude oil before entering the furnace and are returned to the plate above in the column (Figure 1-3).

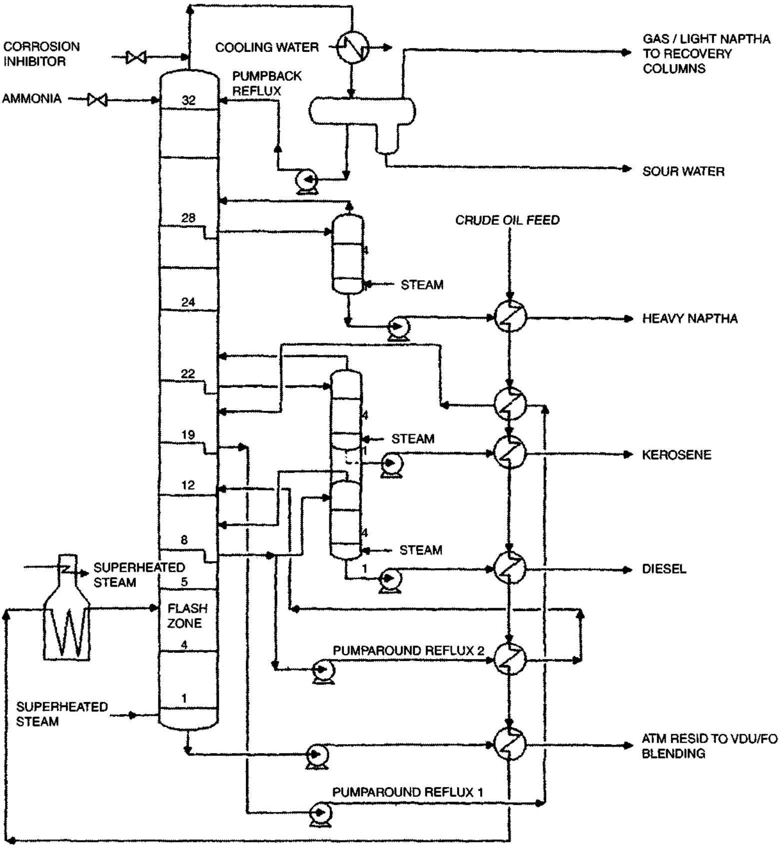

Figure 1-3. Atmospheric crude column with pumpback and pumparound reflux.

SIDESTREAM TEMPERATURE

The flash curve of the product stream is determined first. This product is completely vaporized below the sidestream draw-off plate. Therefore, the 100% point of the flash curve is used. To determine the partial pressure of the product plus reflux vapor, both of which are of same composition, the lighter vapors are considered inert.

$$\begin{array}{l}\text{Partial pressure}\\ \text{of side stream}\end{array} = \frac{(\text{moles of sidestream} + \text{moles of reflux})}{(\text{total moles of vapor below plates})} \times \text{total pressure}$$

EXAMPLE 1-1

The 95% point of heavy naphtha is 315°F and the 5% ASTM distillation point of kerosene is 370°F. The flash point of kerosene is 127.2°F. Calculate the deviation from actual fractionation between heavy naphtha and kerosene for the steam-stripped kerosene fraction and the number of plates and reflux required for separation.

$$\text{Ideal gap} = 370 - 315, \text{ or } 55°\text{F}$$

The actual 5% ASTM distillation point of a fraction can be correlated from its flash point (known), by following relation:

$$\text{Flash point (°F)} = 0.77 \times (\text{ASTM 5\% point, °F}) - 150$$

The actual 5% point on the ASTM distillation curve of kerosene, by this correlation, equals 360°F, which is 10° less than ideal. Since kerosene is to be steam stripped, 95% of heavy naphtha will be 325°F. Therefore,

$$\begin{aligned} \text{Actual gap} &= (360 - 325), \text{ or } 35°\text{F} \\ \text{Deviation from ideal fractionation} &= (55 - 35), \text{ or } 20°\text{F} \end{aligned}$$

From the Packie's correlation, an F factor of 11.5 is required.

CHARACTERIZATION OF UNIT FRACTIONATION

In commercial atmospheric and vacuum units, the distillation is not perfect. For example, a kerosene fraction with a TBP cut of 300–400°F will have material (referred to as *tails*) that boils below 300°F and other material that boils above 400°F. Because of these tails, the yield of the required product must be reduced to stay within the desired product quality limits.

The size and shape of the tails of each product depends on the characteristics of the unit from which it was produced. The factors affecting the fractionation are the number of trays between the product draw trays, tray efficiency, reflux ratio, operating pressure, and boiling ranges of the products.

Several approaches are possible to characterize fractionation in an operating unit. One approach is to characterize the light tail at the front end of a stream in terms of two factors:

V_l is the volume boiling below the cut point, expressed as LV% of crude.

T_f is the temperature difference between the cut point and the TBP initial boiling point (1 LV% distilled) of the stream.

Consider the TBP distillation of products from an atmospheric distillation column (Figure 1-1). The front-end tail of kerosene (TBP cut 300–400) contains 1.5% material on crude boiling below 300°F (see Table 1-3); therefore, $V_f = 1.5$.

The initial boiling point of kerosene cut (1 LV% distilled) is 240°F and the temperature difference between the cut point (300°F) and IBP is 60°F; therefore, $T_F = 60$.

The shape of the front tail can be developed using these two parameters on a probability plot. Having established these parameters, the same values are used for the front end tails of kerosenes on this unit for different cut-point temperatures (e.g., for different flash-point kerosenes).

A similar approach is used for back end tail; in the preceding example, the lighter heavy straight-run (HSR) naphtha cut is before kerosene. The

Table 1-3
Front and Back Tail Characterization of a Typical Atmospheric Crude Unit

	FRONT END TAIL		BACK END TAIL	
STREAM	**V_F LV%**	**$T_F \Delta T$**	**V_B LV%**	**$T_B \Delta T$**
C_4	—	—	0.0	0.0
LSR	—	—	1.0	35.0
HSR	1.0	40.0	1.5	50.0
KEROSENE	1.5	60.0	2.0	65.0
LIGHT DIESEL	2.0	70.0	3.5	120.0
RESID	3.5	160.0	—	—

NOTE:
KEROSENE V_F = HSR V_B.
LIGHT DIESEL V_F = KEROSENE V_B.
RESID V_F = LIGHT DIESEL V_B.

volume of HSR material boiling above the kerosene cut point of 300°F must be 1.5 LV% (on crude), equal to the front end tail volume on kerosene. Let us call it V_B; therefore,

$$V_F = V_B = 1.5\% \text{ (LV on crude)}$$

The HSR end point (99% LV distilled) is 250°F and the cut point is 300°F. Therefore,

$$T_B \text{ for HSR} = 300 - 250 = 50°\text{F}$$

The shape of the back end tail can be estimated using a probability paper. Similarly the shape of front and back end tails for all cuts on vacuum units can also be determined (Table 1-4).

Having established these parameters, the same values are used, for example, for all kerosene cuts on this unit at different front end cut temperatures. This is an excellent approximation, provided the changes in cut point and boiling range are not too large.

Having established the appropriate unit fractionation parameters, the individual product distillations can be established based on selected TBP cut temperatures. These are defined by the points where the produced yield cuts the crude TBP curve. For example, referring to Figure 1-1, the yield of a product lighter than kerosene is 20.4 LV%; hence, the kerosene

Table 1-4
Front and Back Tail Characterization of a Typical Vacuum Unit

	FRONT END TAIL		**BACK END TAIL**	
STREAM	**V_F LV%**	**$T_F \Delta T$**	**V_B LV%**	**$T_B \Delta T$**
WET GAS OIL	—	—	—	—
DRY GAS OIL	—	—	1.0	32.0
HEAVY DIESEL	1.0	60.0	2.2	108.0
VACUUM RESID	2.2	100.0	—	—

NOTE:
HEAVY DIESEL V_F = DRY GAS OIL V_B.
VACUUM RESID V_F = HEAVY DIESEL V_B.
RESID V_F = LIGHT DIESEL V_B.

initial cut point is 300°F where the crude volume percent distilled is 20.4. The kerosene back end TBP cut point is 448°F where the crude volume percent distilled is 36.8, giving the required kerosene yield of 16.4 LV% on the crude.

The product volume and product qualities can be determined by breaking the distillation into narrow cuts, called *pseudocomponents*, and blending the qualities of these using the properties of the narrow cuts from the crude assay data.

GENERAL PROPERTIES OF PETROLEUM FRACTIONS

Most petroleum distillates, especially those from the atmospheric distillation, are usually defined in term of their ASTM boiling ranges. The following general class of distillates is obtained from petroleum: liquefied petroleum gas, naphtha, kerosene, diesel, vacuum gas oil, and residual fuel oil.

DISTILLATES

Liquefied Petroleum Gas

The gases obtained from crude oil distillation are ethane, propane, and *n*-butane isobutene. These products cannot be produced directly from the crude distillation and require high-pressure distillation of overhead gases from the crude column. C_3 and C_4 particularly are recovered and sold as liquefied petroleum gas (LPG), while C_1 and C_2 are generally used as refinery fuel.

Naphtha

C_5-400°F ASTM cut is generally termed *naphtha*. There are many grades and boiling ranges of naphtha. Many refineries produce 400°F end-point naphtha as an overhead distillate from the crude column, then fractionate it as required in separate facilities. Naphtha is used as feedstock for petrochemicals either by thermal cracking to olefins or by reforming and extraction of aromatics. Also some naphtha is used in the manufacture of gasoline by a catalytic reforming process.

Kerosene

The most important use of kerosene is as aviation turbine fuel. This product has the most stringent specifications, which must be met to ensure the safety standards of the various categories of aircraft. The most important specifications are the flash and freeze points of this fuel. The initial boiling point (IBP) is adjusted to meet the minimum flash requirements of approximately 100°F. The final boiling point (FBP) is adjusted to meet the maximum freeze point requirement of the jet fuel grade, approximately −52°F. A full-range kerosene may have an ASTM boiling range between 310 and 550°F. Basic civil jet fuels are

1. Jet A, a kerosene-type fuel having a maximum freeze point of −40°C. Jet A-type fuel is used by mainly domestic airlines of various countries, where a higher freeze point imposes no operating limitations.
2. Jet A-1, a kerosene-type fuel identical with Jet A but with a maximum freeze point of −47°C. This type of fuel is used by most international airlines. Jet A and Jet A-1 generally have a flash point of 38°C.
3. Jet B is a wide-cut gasoline-type fuel with a maximum freeze point of −50 to −58°C. The fuel is of a wider cut, comprising heavy naphtha and kerosene, and is meant mainly for military aircraft.

A limited number of additives are permitted in aviation turbine fuels. The type and concentration of all additives are closely controlled by appropriate fuel specifications. The following aviation turbine fuel additives are in current use:

- *Antioxidants.* Its use is mandatory in fuels produced by a hydrotreating process, to prevent formation of hydrogen peroxide, which can cause rapid deterioration of nitrile rubber fuel system components.
- *Static dissipators*, also known as *antistatic additives* or *electrical conductivity improvers*. Its use is mandatory to increase the electrical conductivity of the fuel, which in turn promotes a rapid relaxation of any static charge build-up during the movement of fuel.
- *Fuel system icing inhibitor* (FSII). The main purpose of FSII is to prevent fuel system blockage by ice formation from water precipitated from fuels in flight. Because of the biocidal nature of this additive, it is very effective in reducing microbiological contamination problems in aircraft tanks and ground fuel handling facilities.

As most commercial aircrafts are provided with fuel filter heaters, they have no requirement for the anti-icing properties of this additive. FSII is therefore not usually permitted in civil specifications, its use is confined mainly to military fuels.

- *Corrosion inhibitor/lubricity improver.* Its use is optional to protect storage tanks and pipelines from corrosion and improve the lubricating properties of the fuel.

Diesel

Diesel grades have an ASTM end point of 650–700°F. Diesel fuel is a blend of light and heavy distillates and has an ASTM boiling range of approximately 350–675°F. Marine diesels are a little heavier, having an ASTM boiling end point approximately 775°F. The most important specifications of diesel fuels are cetane number, sulfur, and pour or cloud point. Cetane number is related to the burning quality of the fuel in an engine. The permissible sulfur content of diesel is being lowered worldwide due to the environmental pollution concerns resulting from combustion of this fuel. Pour point or cloud point of diesel is related to the storage and handling properties of diesel and depends on the climatic conditions in which the fuel is being used.

Vacuum Gas Oil

Vacuum gas oil is the distillate boiling between 700 and 1000°F. This is not a saleable product and is used as feed to secondary processing units, such as fluid catalytic cracking units, and hydrocrackers, for conversion to light and middle distillates.

Residual Fuel Oil

Hydrocarbon material boiling above 1000°F is not distillable and consists mostly of resins and asphaltenes. This is blended with cutter stock, usually kerosene and diesel, to meet the viscosity and sulfur specifications of various fuel oil grades.

VACUUM DISTILLATION PRODUCTS

In an atmospheric distillation tower, the maximum flash zone temperature without cracking is 700–800°F. The atmospheric residuum, commonly known as *reduced crude*, contains a large volume of distillable oils that can be recovered by vacuum distillation at the maximum permissible flash zone temperature. The TBP cut point between vacuum gas oil and vacuum resid is approximately 1075–1125°F. The cut point is generally optimized, depending on the objectives of the vacuum distillation, into asphalt operation and pitch operation.

Asphalt Operation

Given the specification (penetration) of the asphalt to be produced, the corresponding residuum yield can be determined from the crude assay data. The total distillate yield is determined by subtracting asphalt yield from the total vacuum column feed. In case a number of lubricating oil distillates is to be produced, the distillation range of each has to be specified, and the corresponding yields can be determined from the crude assay data. Lube cuts are produced as sidestreams from the vacuum column.

In asphalt operation, some gas oil must remain in the pitch to provide the proper degree of plasticity. The gravity of an asphalt stream is usually between 5 and 8° API. Not all crudes can be used to make asphalt. Experimental data for asphalt operation are necessary to relate asphalt penetration to residual volume. The penetration range between 85 and 10, are possible and the units are generally designed to produce more than one grade of asphalt.

The principal criteria for producing lube oil fractions are viscosity, color, and rejection to residuum the heavy impurities and metals. These oils are further refined by solvent extraction, dewaxing, and other types of finishing treatment, such as hydrotreating. Vacuum towers for the manufacture of lubricating oils are designed to provide same relative degree of fractionation between streams as in the atmospheric tower. Sidestreams are stripped in the external towers to control front-end properties. The number of trays between the draw trays is set arbitrarily. Generally, three to five trays are used between draws. Sieve trays are more popular for vacuum column service.

Pitch Operation

The objective in this case is to produce maximum distillate and minimum pitch, which is used for fuel oil blending. In this case, the TBP cut point between the distillate and pitch has to be set by unit design, generally around 1100°F. From the crude assay data, the total distillate yield from the crude up to the cut point is known; deducting the total distillates yield in the atmospheric column, the total yield of vacuum distillate can be estimated. The light vacuum distillate yield is set at approximately 30% of the total vacuum gas oil, to facilitate heat recovery at two levels of heat.

The unit design has to specify the amount of overflash, depending on the purity of the heavy vacuum gas oil (HVGO) required. If the color requirements or level of metal contaminants is not severe, 1–2 vol% (volume %) of vacuum feed is taken as overflash.

Vacuum column design calculation is similar to atmospheric column design with some differences in technique as follows:

- A material balance is made for vacuum feed vs. the products—vacuum bottoms, sidestream products (vacuum gas oils), and overhead condensable hydrocarbons. The assumed quantity of noncondensables is not carried in the material balance nor considered in the flash zone calculations but must be estimated for vacuum ejector calculations.
- The construction of flash vaporization curve (AFVC, atmospheric flush vaporization curve) of the reduced crude, feed to vacuum distillation unit is done in the same manner as for the whole crude.

ATMOSPHERIC DISTILLATION UNIT

In Figure 1-4, the crude oil received from off-site storage tanks through booster pumps is pumped by charge pump P-101 and preheated in parallel trains of preheat exchangers with hot intermediate streams and products. A small quantity of water and demulsifier chemicals are added before preheating. The hot crude is mixed with washwater and fed to electric desalters V-106 A and B to reduce the salt content by an electric desalting process. The water phase, containing most of the dissolved salts contained in the crude, separates out. The desalted crude is dosed with an NaOH solution to a fixed chloride content. The desalted crude is further heated through two parallel trains of heat exchangers and fed to preflash tower V-117. The preflash tower overhead vapor is cooled by exchanging heat with crude oil and condensed in an overhead drum V-118. Part of this liquid naphtha is used as reflux in the column; the rest of the liquid and the vapor from the drum are sent to the naphtha processing unit.

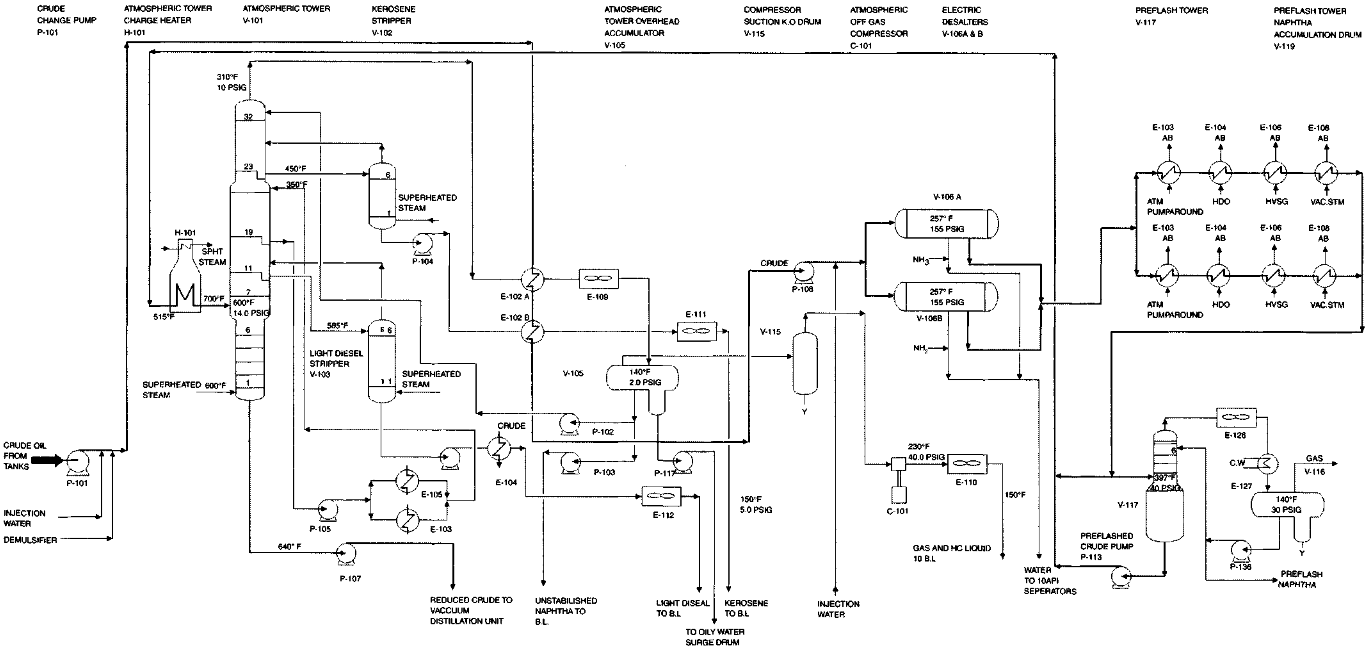

Figure 1-4. Atmospheric distillation. K.O. = knockout; C.W. = cooling water; B.L. = battery limits.

The crude from the bottom of the preflash tower is pumped through the heat exchangers, recovering heat from vacuum tower bottoms and sidestream HVGO, and sent to fired atmospheric heater H-101. The crude is partially vaporized in the fired heater before entering the flash zone of the atmospheric tower V-101. Superheated stripping steam is introduced through the bottom of the column.

The tower overhead vapor is cooled by exchanging heat with crude oil, condensed in air cooler E-109, and routed to overhead product accumulator V-105. The overhead gases from this accumulator are compressed in compressor C-101 to about 40 psig pressure and sent to the refinery gas recovery system. The condensed naphtha in the accumulator is separated from water. A part of this naphtha is sent back to the column by reflux pump P-102 and the rest is withdrawn as an intermediate product for processing in naphtha fractionation unit.

Kerosene and light diesel cuts are withdrawn as sidestreams from the atmospheric distillation tower. These are steam stripped in steam strippers V-102 and V-103, respectively. The kerosene and light diesel product streams exchange heat with crude oil feed in the crude preheat train and finally cooled in air fin coolers E-106 and E-111 and sent to storage.

The hot atmospheric bottoms or reduced crude, at approximately 660°F, is transferred by P-107 to vacuum tower heater H-102.

VACUUM DISTILLATION UNIT

The reduced crude from vacuum heater H-102 enters the flash zone of vacuum tower V-104. The column operates under vacuum by means of an ejector/condenser system to achieve the required separation between the heavy components at lower temperature. Some gaseous hydrocarbons are produced due to cracking of the feed in vacuum heater H-102. This sour gas is burnt in atmospheric tower heater H-101 while the condensate water is routed to desalter feed water surge drum V-106.

The tower is provided with a cold recycle (quench) to lower the bottom temperature and avoid coking. A superheated stripping stream is introduced at the bottom of the tower. The heavy diesel product is drawn as a sidestream and exchanges heat with crude oil in the preheat train. It is partly used as top and intermediate reflux to the column, and the balance is sent to storage after cooling in E-121 and E-113.

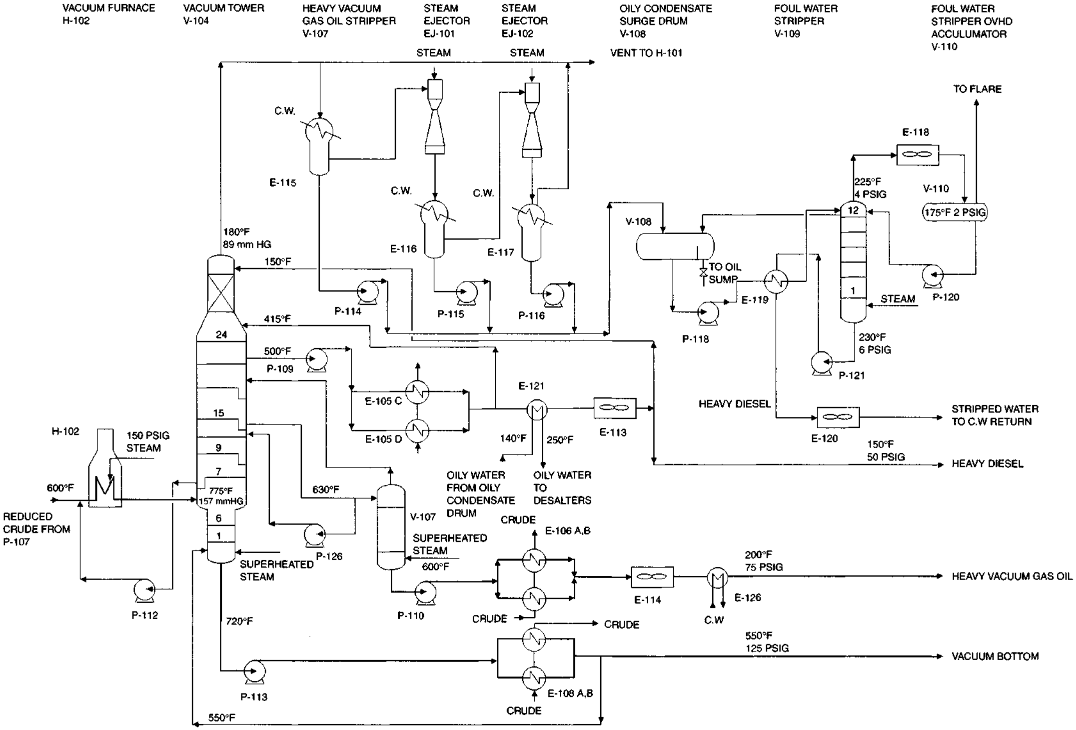

Figure 1-5. Vacuum distillation unit. C.W. = cooling water; OVDH = overhead.

Vacuum gas oil drawn as bottom sidestream is stripped in steam stripper V-107 and cooled by exchanging heat with crude in the preheat train and finally in air cooler E 114 and trim cooler E-126 before being sent to off-site storage tanks.

The bottom product, the vacuum residue, exchanges heat with crude coming from the preflash tower bottoms and in the preheat train before being sent to off-site storage.

To control corrosion, a 3% ammonia solution and inhibitor is injected into the top of preflash, atmospheric, and vacuum towers.

Foul water is generated in the overhead accumulator drum of atmospheric distillation column and in the ejector/condenser system of the vacuum distillation column. All foul water streams are combined in oily condensate surge drum V-108. From V-108, the oily water is transferred by P-118 to foul water stripper V-109. Superheated steam is admitted at the bottom of this 12-plate tower for stripping H_2S and NH_3 from the foul water.

The overhead gases are cooled and condensed in air fin cooler E-118. Noncondensable gases are routed to the flare header. Condensed and concentrated H_2S/NH_3 liquid is returned to the column as total reflux. The hot stripped water from the column bottom is partly recycled to desalters and the rest is to a water treatment plant.

The typical operating conditions for an atmospheric and vacuum distillation towers are shown in Tables 1-5 through 1-7.

CRUDE DESALTING

Crude desalting is the first processing step in a refinery (see Figure 1-6). The objectives of crude desalting are the removal of salts and solids and the formation water from unrefined crude oil before the crude is introduced in the crude distillation unit of the refinery.

Salt in the crude oil is in the form of dissolved or suspended salt crystals in water emulsified with the crude oil. The basic process of desalting is to wash the salt from crude oil with water. Problems occur in efficient and economical water/oil mixing, water wetting of suspended solids, and separation of oil from wash water. The separation of oil and washwater is affected by the gravity, viscosity, and pH of the crude oil and the ratio of water/crude used for washing.

An important function of the desalting process is the removal of suspended solids from the crude oil. These are usually very fine sand and soil particles, iron oxides, and iron sulfide particles from pipelines,

Table 1-5
Atmospheric Tower Operating Conditions

OPERATING PARAMETER	UNITS	
TEMPERATURES	°F	
TRANSFER LINE		660
FLASH ZONE		657
TOWER TOP		359
KEROSENE DRAW-OFF		469
PUMPAROUND DRAW OFF		548
PUMPAROUND RETURN		345
LIGHT DIESEL DRAW OFF		603
TOWER BOTTOM		648
PRESSURE	psig	
REFLUX DRUM		2.0
TOWER TOP		10.3
FLASH ZONE		14.7
REFLUX RATIO, REFLUX/LIQUID DIST.		0.6
STRIPPING STEAM		
TO ATMOSPHERIC TOWER	lbs/bbl RESID	5.5
TO KEROSENE STRIPPER	lbs/bbl RESID	5.9
TO DIESEL STRIPPER	lbs/bbl RESID	2.1
ATMOSPHERIC HEATER		
PROCESS FLUID CONDITIONS		
TEMPERATURE IN	°F	453
TEMPERATURE OUT	°F	660
PRESSURE DROP	psi	138
TUBE SKIN TEMPERATURE (AVG)	°F	735
STACK GAS TEMPERATURE	°F	725
FRACTIONATION EFFICIENCY		
95%–5% ASTM DISTRIBUTION GAP		
ATMOSPHERIC NAPHTHA-KEROSENE		GAP + 10
KEROSENE-LIGHT DIESEL		GAP – 36

NOTE: BASIS 154000 BPSD KUWAIT CRUDE RUN.

tanks or tankers, and other contaminants picked up in transit or from processing.

Until recently, the criteria for desalting crude oil was 10 lb salt/1000 bbl (expressed as NaCl), but due to more stringent requirements of some downstream processes, desalting is now done at the much lower level of 1.0 lb/1000 bbl or lower. Reduced equipment fouling and corrosion

Table 1-6
Vacuum Tower Operating Conditions

OPERATING PARAMETER	UNITS	
TEMPERATURES	°F	
TRANSFER LINE		740
FLASH ZONE		711
TOWER TOP		307
HEAVY DIESEL DRAW-OFF		447
TOP REFLUX TEMPERATURE		121
HVGO DRAW-OFF		613
TOWER BOTTOM		670
PRESSURE	mmHg	
TOWER TOP		64
FLASH ZONE		125
TOP REFLUX RATIO; REFLUX/FEED		0.15
HOT REFLUX RATIO; REFLUX/FEED		0.97
WASH OIL RATIO; WASH OIL/FEED		0.14
BOTTOM QUENCH OIL RATIO; QUENCH/FEED		0.24
STRIPPING STEAM		
TO VACUUM TOWER	lbs/bbl RESID	8.0
TO HVGO STRIPPER	lbs/bbl RESID	4.6
VACUUM HEATER		
PROCESS FLUID CONDITIONS		
TEMPERATURE IN	°F	645
TEMPERATURE OUT	°F	736
PRESSURE DROP	psi	73
TUBE SKIN TEMPERATURE (AVG)	°F	850
STACK GAS TEMPERATURE	°F	845
FRACTIONATION EFFICIENCY		
95%–5% ASTM DISTRIBUTION GAP		
LIGHT DIESEL–HEAVY DIESEL		GAP – 145
HEAVY DIESEL-HVGO		GAP + 25

NOTE: BASIS 154000 BPSD KUWAIT CRUDE RUN.

Table 1-7
Atmospheric and Vacuum Crude Distillation Utility Consumption

UTILITY	UNITS	CONSUMPTION
ELECTRICITY	kWhr	8.7
FUEL	mmBtu	0.6
STEAM	mmBtu	0.09
COOLING WATER	MIG*	0.31
DISTILLED WATER	MIG*	0.02

**THOUSAND IMPERIAL GALLONS.*

NOTE: THE UTILITY CONSUMPTIONS (PER TON FEED) ARE FOR AN INTEGRATED CRUDE AND VACUUM UNIT.

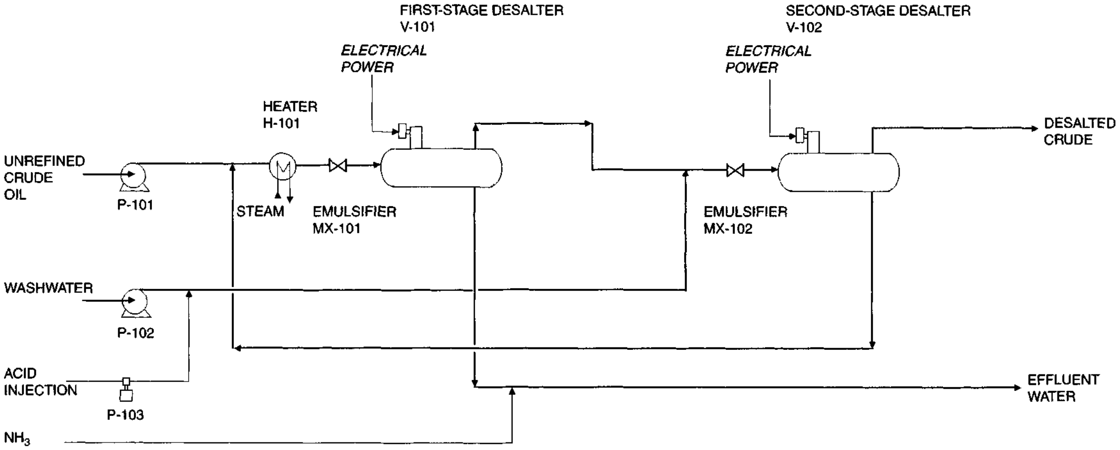

Figure 1-6. Two-stage desalter.

and longer catalyst life in downstream processing units provide justification for this additional treatment.

Desalting is carried out by emulsifying the crude oil with 3 to 10 vol% (volume %) water at a temperature of 200–300°F. Both the ratio of water to oil and the operating temperature are functions of the gravity of the crude oil. Typical operating conditions are given in Table 1-8.

The salts are dissolved in the washwater and oil and water phases are separated in a settling vessel either by adding chemicals to assist in breaking up the emulsion or by the application of an electrostatic field to coalesce the droplets of saltwater more rapidly (see Table 1-9). Either an AC or DC field may be used (see Table 1-10) and potentials of 16,000–35,000 V are used to promote coalescence. Efficiencies up to 90–95% water removal are achieved in a single stage and up to 99% in a two-stage desalting process.

Heavy naphthenic crudes form more stable emulsions than most other crudes, and desalters usually operate at lower efficiency when handling them. The crude oil densities are close to density of water, and temperatures above 280°F are required.

It is necessary to adjust the pH of the brine to obtain a value of 7 or less. If the pH of the brine exceeds 7, emulsions are formed because of the presence of sodium naphthenate and sodium sulfide. For most crude oils, it is desirable to keep the pH below 8. Better dehydration is obtained in electrical desalters when they are operated at a pH of 6. The pH is controlled by the addition of acid to the incoming or recycle water.

Makeup water is added to the second stage of a two-stage desalter. The quantity is 4–5% on crude oil volume. For very heavy crude oil (API < 15), gas oil is added as a diluent to the second stage to obtain more efficient separation. The gas oil is recovered in the crude column and recycled to the desalter. Frequently, the washwater used is from the vacuum crude unit barometric condenser or other refinery sources containing phenols. The phenols are preferentially soluble in crude oil, thus reducing the phenol content of the water sent to the refinery wastewater handling system.

Suspended solids are another major cause of water-in-oil emulsions. Wetting agents are frequently added to improve the water wetting of solids and reduce oil carry under in the desalters. Oxyalkylated phenols and sulfates are the most frequently used wetting agents.

Table 1-8
Washwater Requirements of Desalters

CRUDE API	WASHWATER, VOL%	TEMPERATURE, °F
API > 40	3–4	240–260
30 < API < 40	4–7	260–280
API < 30	7–10	280–300

Table 1-9
Operating Conditions

PARAMETER	UNITS	VALUE
CRUDE TO DESALTER*	bpsd	98000
WATER TO DESALTER	gpm	145
WATER TO CRUDE RATIO	%	5
DEMULSIFIER INJECTION	ppmw	10–15
PRESSURE		
CRUDE TO DESALTER	psig	125
DELTA P MIXING VALVE	psig	20
TEMPERATURE		
CRUDE TO DESALTER	°F	270
WATER TO DESALTER	°F	265
CRUDE FROM DESALTER	°F	260
ANALYSIS RESULTS		
CRUDE INLET SALT	lb/1000 bbl	3.94
CRUDE INLET SALT	ppmw	12.87
CRUDE OUTLET SALT	ppmw	1.2
OUTLET BS&W	% MASS	0.05
WATER		
INLET SALT CONTENT	ppm	100
OUTLET SALT CONTENT	ppm	310
INLET OIL CONTENT	ppm	7
OUTLET OIL CONTENT	ppm	10
pH INLET		6.5
OUTLET pH		6.5
OUTLET pH AFTER NH3 INJECTION		7

**30.4 API CRUDE.*

NOTE: BASIS 98000 BPSD CRUDE.

Table 1-10
Utility Consumption

UTILITY	UNITS	CONSUMPTION
ELECTRICITY	kWhr	0.014–0.070
WATER	GALLONS	10–18

NOTE: PER TON FEED.

NOTES

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4. J. B. Maxwell. *Data Book on Hydrocarbons.* Princeton, NJ: Van Nostrand, 1968. W. C. Edmister. *Applied Hydrocarbons Thermodynamics.* Houston: Gulf Publishing, 1961. W. L. Nelson. *Petroleum Refinery Engineering.* New York: McGraw-Hill, 1958.
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