

Design of Atmospheric Petroleum Distillation Column

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Received: November 11, 2021; Published: November 26, 2021

Abstract

This work summarizes the design procedure for Atmospheric Distillation Column for petroleum oil. The design and operation of crude distillation unit are still done almost exclusively on an empirical basis. This is because crude petroleum and its products are made of large number of hydro-carbons. The complex mixture due to large number of fractions, side stripper, and reflux made the work of improving energy efficiency into a tedious manner. This creates adversity in designing the distillation column. The varied products are distinguished by performing true boiling point experiment on the basis of boiling ranges as the exact composition in terms of hydrocarbon. The primary separation process in any refinery is that of distillation and separation is done on the basis of boiling point distribution.

The feed capacity of crude oil considered for designing of atmospheric distillation column is 675800 lb/hr of a 36.3 °API crude oil. The design calculations are carried out for five petroleum product viz. Light Naphtha (LN), Heavy Naphtha (HN), Light Distillate (LD), Heavy Distillate (HD) and Atmospheric Gas Oil (AGO).

Key words: Petroleum; Crude Distillation; Crude Assay; Atmospheric Distillation column; True Boiling Point

Introduction

The design associated operation of petroleum distillation units are still done solely on an empirical basis. This is because of crude oil and its products, to one side light end products, are created from large number of individually separate and distinct hydrocarbons—each compound being comparatively present in comparatively small amount. It's easier then to speak about boiling ranges once describing the gross properties of crude and its fractions [1].

Basic Principles

First let us consider some of the essential steps for separating petroleum into fractions while not regard for the result of side strippers

or stripping steam. The separation mentioned here can involve a vapour overhead product, five side stream products and a residual liquid bottom product. A small amount of extra vaporization called over flash are used to assure the desired vaporization happens and to provide some reflux for bottom section of the tower.

Usually, it's not possible to use a reboiler on a crude tower. So, the charge generally is heated to its maximum temperature before being fed into the column. All distillate products are volatilized as they enter the tower. So, the total heat required by the method should be contained within the feed because it leaves the charge furnace.

Various ways in which of producing distillate products from the feed vapour as shown in FIG.1. The symbols used to designate a number of these strategies were chosen for simple association with the type of separation: Type U (Top Tray Reflux) designates that the side draw points are un-refluxed even if there's reflux at the top of the last column. Type A (Pump Around Reflux) uses associate external-circulated fluid that behaves as an absorption oil regarding heat removal. Type R (Pump Back Reflux) designates the utilization of external reflux that is in equilibrium liquid with respect to the tray that re-enters the tower.

Series of Flash Drums

Before discussing the preceding three strategies of separation, take into account however fractions having the specified boiling ranges is created employing a series of flash drums. Consider a top train of figure 1 [4], within which the furnace effluent enters the first drum wherever the residual liquid W is separated from the vapour. The 1st drum effluent vapour is cooled only enough to condense the over flash (LOF) in order that it'll be separated as a liquid within the 2nd drum. The over flash leaves the system as a part of the residual liquid W.

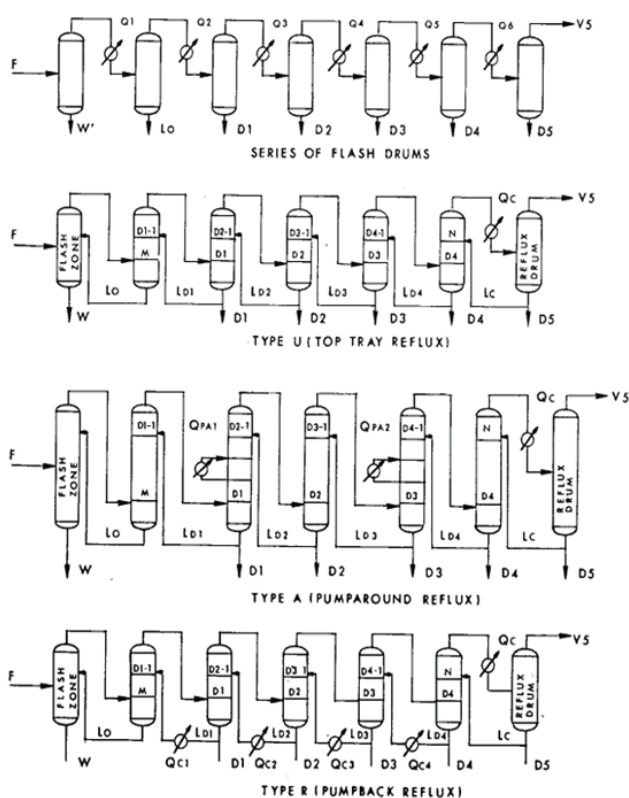


Figure 1: Processing schemes for atmospheric distillation of crude oil [4].

The vapour from the 2nd drum is cooled once more to some present temperature and enters a 3rd flash. There the heaviest distillate product D1 is removed as a liquid. The method is recurrent to supply in turn lighter fractions D2, D3, D4 and D5. The overhead from the 7th drum is a vapour product V5.

Each of those separations happens from cooling and equilibrium condensation in order that the equilibrium liquid contains little concentrations of elements lighter than the specified distillation. Conjointly the equilibrium vapour contains a number of the components that belong within the liquid in order that next fraction to be condensed will have some undesirable heavy ends. But it would be possible to use liquid reflux to wash back the heavy components from the vapour and to use stripping vapours to strip light component from the liquid.

Design Procedure [1]

In the sense that engineers perceive distillation of separate compounds, petroleum fractionation remains a black art. Within the separation of with chemically similar materials, it's usually true that the greater the distinction in boiling points of individual fractions, the better it's to separate the fractions. The sharpness of separation sometimes is outlined because the gap or overlap of the boiling ranges of adjacent fractions. For a given separation, a large number of trays can result in a low reflux demand. Likewise, a higher reflux rate would result a lower number of trays.

Product Specifications

The properties of every fraction is varied as required by sales demands, however solely at the expense of the adjacent fractions. Usually, the method designer should estimate the material balance on the premise of the specifications given for the required products. The premise for many product specifications for a crude column are derived from the method projected by American Society for Testing Materials (ASTM). This technique reports the temperatures at those certain parts of the material are vaporized.

ASTM End Points

One technique for setting product specifications is to state the most allowable end points for the fractions. The heaviest distillate product known as atmospheric gas oil, is excluded from this specification since it's withdrawn from the column to produce adequate fractionation between the other liquid distillates and also the reduced crude.

If the crude processing facilities contain a vacuum column, economics favour the production of the most possible quantity of all distillates within the atmospheric column. Then the size of the vacuum column is reduced. A side profit is that the maximization of the crude heat before getting into the furnace.

For the instance mentioned here, the identity of the petroleum fractions beginning with the lightest are: overhead vapour, light naphtha, heavy naphtha, light distillate, heavy distillate, gas oil and reduced crude. Isolation of these fractions ought to be computed on two representative crudes, one light and one heavy. The material balance are based on alternately increasing the production of gasoline (naphthas), light distillate and heavy distillate.

Key Temperatures

Intermediate temperatures generally are fixed relative to the product streams. These set the gap between the temperature at 5% ASTM for the heavier fractions and the temperature at 95% ASTM for the lighter fraction. These specifications must be transformed into ASTM end point specification before continuing with a design.

Material Balance [9]

The material balance is calculable as a function of the required characteristics of the fractions and also the quantity of crude that may be vaporized at the outlet conditions of the furnace. It's necessary to estimate the number of actual trays within the column and also the operating pressures of the column so as to set the conditions at the furnace outlet.

Operating Pressures

They are estimated by assuming a pressure of 0.5 to 1.0 psig in the final accumulator, i.e., the vessel which serves as a suction for the compressor which feeds the gas recovery plant. Ordinarily, this will be the only accumulator in the system although units have been designed using a primary condenser to generate reflux followed by a secondary exchanger for condensing overhead product. In this latter case, this final drum also serves as a suction drum for the recovery plant feed compressor: this drum will operate at a pressure of 0.5 to 1.0 psig. For pressure drop through the condensers, use 2.5 psig per shell which will also take care of piping.

Heat Balance

Other than heat incidental to the removal steam, the only heat to the column is added by the charge furnace. For many cases, the furnace outlet temperature is assumed to be 700°F. The furnace outlet temperature mustn't exceed this temperature while not specific

reasons to the contrary. Some refiners limit furnace outlet temperatures to 650°F for kerosene and jet fuel productions.

Bottom product is assumed to depart the column at a temperature 90°F under the temperature of the flash zone. The other fractions are withdrawn from the column at draw tray temperatures.

Draw tray temperature assumptions are checked by calculating the bubble point temperature of the unstrapped product at the partial pressure of the product vapour in the total vapour leaving the draw tray. This is done by doing a heat balance around the draw tray and calculating the quantity of internal reflux required to absorb the excess heat. This initially needs to do a heat balance at the first tray below the draw tray so as to establish circulating reflux requirements.

For steam strippers, the temperature of the stripped material go away from the bottom of the stripper is assumed to be at least 90°F lower in temperature than the corresponding draw tray temperature. For re boiled strippers, this temperature is assumed to be 30°F beyond the corresponding draw tray temperature.

The temperature of the liquid on the top tray is calculated as mentioned earlier by doing a heat balance. On this tray however, rather than taking the bubble point temperature, the dew point temperature is adjusted to the partial pressure of the condensable hydrocarbons in the total vapour depart from the top tray.

Crude Oil Evaluation

The true boiling point (TBP) experimental data for Narrow fractions of 36.3 API gravity crude oil is collected for design of atmospheric crude distillation column shown in TABLE 1 [1]. The volume percent distilled against temperature data of TBP and equilibrium flash vaporization (EFV) was tabulated in TABLE 2 and plotted in figure 2.

Determination of EFV Curve [5]

The TBP data is converted to EFV data by using Edmister method. The phase diagram is plotted to find out the temperatures above atmospheric pressure by using Edmister method as shown in figure 3.

Cut no.	TBP cut point °C	CUMMULATIVE % on crude		Cut gravity OAPI	Molecular wt. cal	Cummulative Mole per 100 bbl
		Weight	Volume			
1		0.59	1		42.6	0.41
2	14	3.12	4.7	114.1	58.1	13.25
3	42	5.52	8.04	93	72	23.49
4	63	6.93	9.7	81.1	85	28.03
5	78	8.53	11.63	71	87	33.45
6	93	10.56	13.98	62.9	94	39.82
7	100	12.31	15.96	58.2	96	45.2
8	111	14.18	18.03	54.6	99	50.77
9	125	15.96	20.01	55.1	104	55.82
10	132	17.89	22.15	54.3	113	80.86
11	142	19.72	24.13	50	116	65.51
12	149	21.16	26.67	48.7	119	69.08
13	158	23.04	27.69	48.8	125	79.52
14	168	24.95	29.73	47.8	132	77.79
15	177	26.29	31.15	46	136	80.7
16	189	28.26	33.23	45.7	142	84.7
17	194	30.13	35.18	43.1	148	88.52
18	204	31.35	36.45	42.3	154	90.85
19	218	33.28	38.44	41.1	152	94.36
20	225	35.24	40.44	39.9	167	97.83
21	238	37.24	42.45	37.2	176	101.17
22	249	39.28	44.48	36.1	186	104.4
23	258	41.5	46.69	35.2	193	104.79
24	264	42.24	47.42	34.7	198	108.89
25	271	44.27	49.42	34.1	202	111.85
26	284	46.38	51.49	23.4	210	114.81
27	296	48.48	53.55	33.3	221	117.51
28	305	50.81	55.65	33.5	230	120.27
29	315	52.63	57.64	33.2	244	122.71
30	324	54.84	59.79	31.7	250	125.32
31	329	55.8	60.72	30.3	258	126.42
32	343	57.94	62.76	29.1	254	128.81
33	354	60.08	64.8	28.3	276	131.1
34	368	62.33	66.94	27.9	287	133.41
35	371	63.36	67.82	27.7	298	134.42
36	381	65.42	69.87	27.5	-	-
37	392	67.48	71.82	27.4	-	-
38	400	69.58	73.78	26.9	-	-

39	414	71.68	75.75	26.3	-	-
40	429	73.78	77.72	25.8	-	-
41	438	75.87	78.69	25.4	-	-
42	447	77.98	81.66	25.1	-	-
43	466	80.27	83.78	24.2	-	-
44	492	82.65	85.97	23.1	-	-
45	Residuum	99.64	100.14	11.4	-	-

Table 1: Experimental TBP data for Narrow fractions of 36.3 API gravity crude oil.

Volume% Distilled	0	10	20	30	50	60	70	80	90
TBP Temperature (°C)	8	62	125	173	277	329	381	441	527
EFV Temperature (°C)	59	89	124	164	234	268	301	334	367

Table 2: TBP and EFV data obtained for Narrow fractions of 36.3 API gravity crude oil.

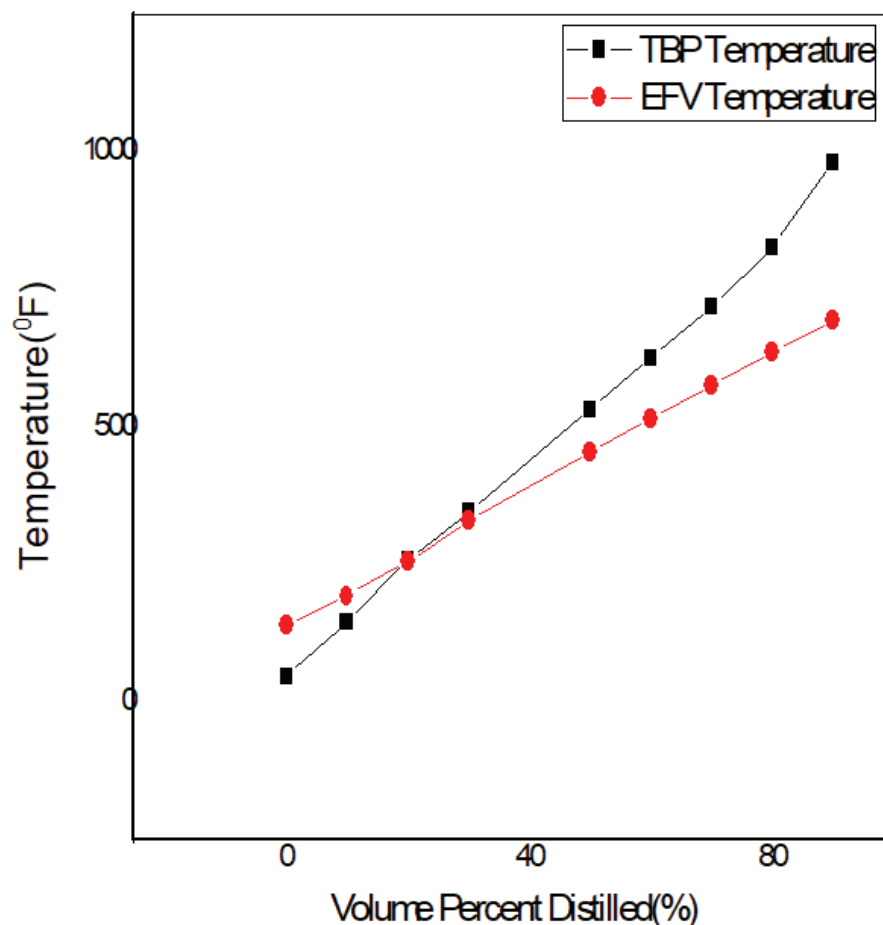


Figure 2: The plot of Temperature against Volume percent distilled.

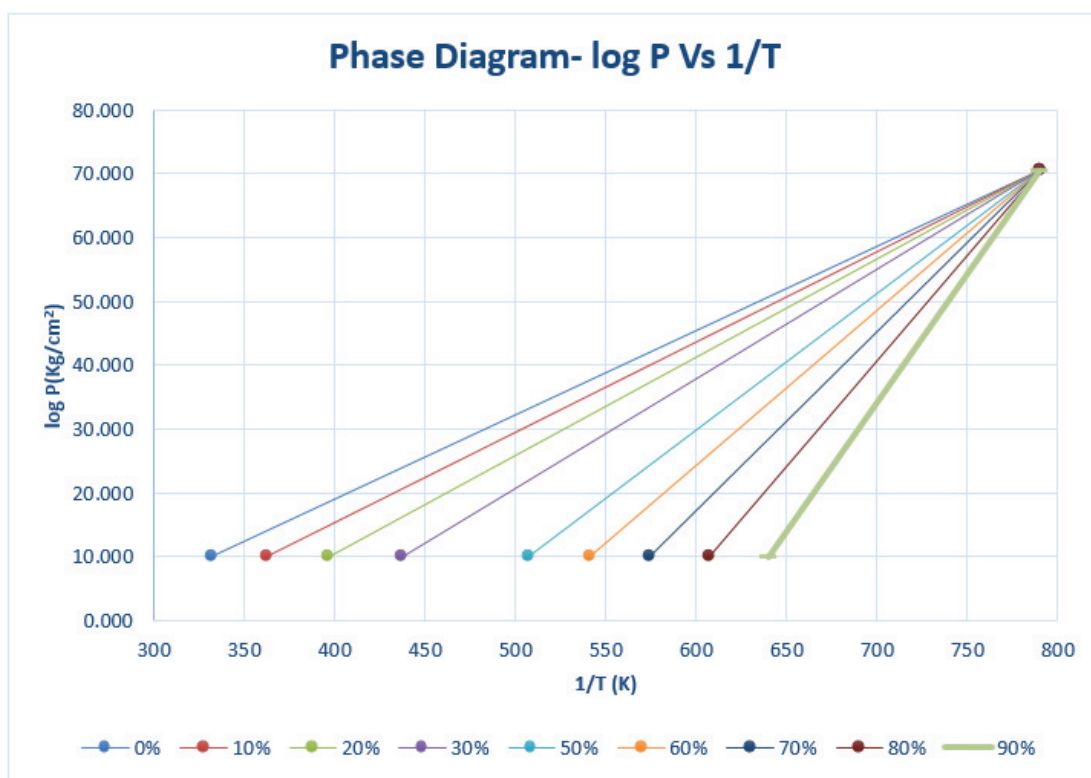


Figure 3: Phase diagram for a crude oil.

Product Distillations [4]

By a trial-and-error procedure, the whole crude is cut into fractions satisfying the boiling range and (5-95) gap criteria shown in TABLE 3. The TBP temperatures of the fractions are calculated by using chart ASTM temperature versus TBP temperature. The various products and their specifications are shown in TABLE 4. The proper allocation and TBP distillations are shown in figure 4.

Fractionation is the distinguishing between 5% ASTM curve of Heavy cut and the 95% point on the ASTM curve of a lighter cut of two adjacent side product.

(595) Gap = $(t_{5H} - T_{95L})$ ASTM

5H=5% OF THE HEAVIER FRACTION, 95L= 95% OF THE LIGHTER FRACTION

TBP CUT POINT = $(T_{OH} + T_{100L})/2$, TBP Overlap= $[T_{100L} - T_{OH}]$

The data is summarized as shown in TABLE 8.

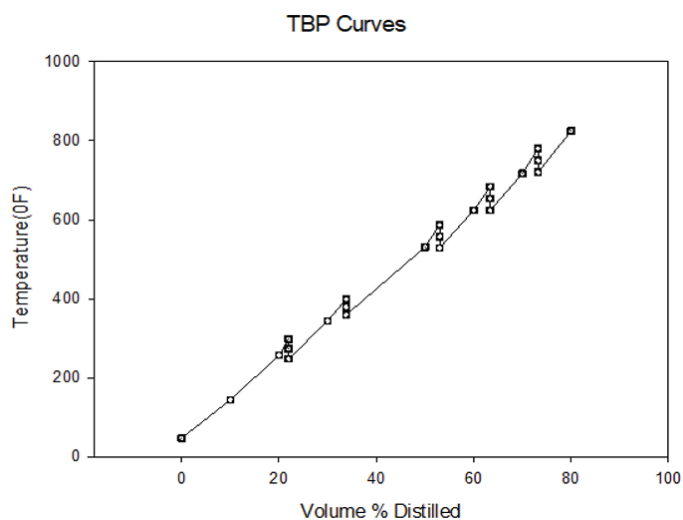


Figure 4: TBP curves for atmospheric tower feed and products yield.

Serial No.	Given		Estimated						
	Stream	T _{100%} ASTM	T _{5-95%} ASTM	T _{100%} TBP	TBP Over- lap	T _{0%} TBP	TBP Cut	Cum. Vol%	Cut Vol%
	Light Naphtha	275	20	298	50	248	273	22	22
	Heavy Naphtha	380	35	399	40	359	379	33.8	11.8
	Light Distillate	560	10	587	58	529	558	53	19.2
	Heavy Distillate	650	5	684	60	624	654	63.4	10.4
	Atm. Gas Oil Residue	735	5	780	60	720	750	73.2	9.8

Table 3: Product Descriptions [1].

Product	Vol %	API	lb/gal	lb/hr (m)	Gal/hr	Mol. Wt. ⁴	Latent heat(Btu/lb)(λ)	Specific Heat ² (C _p)
Light Naphtha	22	77	8.345	93700	11227.7362	86	130	0.56
Heavy Naphtha	11.8	46	8.345	75400	9034.9126	123.3	120	0.64
Light Distillate	19.2	34	8.345	131600	15769.1578	184.9	108	0.69
Heavy Distillate	10.4	30	8.345	73000	8747.3292	302.1	98	0.71
Gas oil	9.8	26.6	8.345	70300	8423.7978	489.8	88	0.74
Residue	26.8	18.6	8.345	209500	25103.6365	-	-	0.77

Table 4: Product specifications [1].

Heat Balance Calculation

Heat Balancet

A heat balance of the tower is computed to determine the amount of heat that must be removed to keep the tower in thermal balance. The crude oil flowrate considered for designing of atmospheric distillation column is 55000 bbl/day (675800 lb/hr) having Characterization Factor 11.57 and specific heat 0.843. The steam flowrate considered for calculation is 6142 lb/hr¹. At 700 °F the light naphtha, heavy naphtha, light distillate, heavy distillate and gas oil are vapors and the reduced crude is a liquid. A sufficient amount of heat must be removed from the vapors at which they are withdrawn to condense the light naphtha, heavy naphtha, light distillate, heavy distillate and gas oil at their withdrawal temperature.

The system mentioned here is operated with hot reflux and circulating reflux. Tower temperature is calculated by considering pressure 780 mmHg at the top. These computations are based on assumptions that the top temperature is measured in the vapor above the top plate. If the temperature was taken in the liquid on the top tray, the temperature would be the same for all types of refluxes. The value of specific heat and latent heat are considered by

referring chart of specific heat against temperature² and latent heat of vaporization against VABP [2] and are reported in TABLE 5.

Sr. No.	Name of Products	Total Heat Content	Mole Fraction
1.	Light Naphtha	1.31 x 10 ⁸ Btu	0.974
2.	Heavy Naphtha	1.25 x 10 ⁸ Btu	0.964
3.	Light Distillate	1.03 x 10 ⁸ Btu	0.945
4.	Heavy Distillate	0.662 x 10 ⁸ Btu	0.88
5.	Gas oil	0.4434 x 10 ⁸ Btu	0.7745

Table 5: Heat balances.

Bottom Temperature [2, 11]

$$\Delta t = \frac{PL}{100h} + \frac{HA(T-T_a)}{Bh} + \frac{S(T-T_s)}{B}$$

Where Δt = feed less bottom temperature, OF

P = percent removed by stripping

L = Latent heat of oil vaporized, Btu/lb

h = Specific heat of oil

H = Transfer rate through insulated wall, Btu/(sq.ft.)hr (OF temperature difference)

A = Area of walls, sq.ft.

B = lb of oil per hr

S = lb steam per hr

T = temperature of oil, 0F

Ta = temperature of air, 0F

D = tower diameter

Ts = temperature of incoming steam

$$\Delta t = 1.2 P + 35 \frac{A(T-Ta)}{B} + 0.65 \frac{S(T-Ts)}{B}$$

$$\Delta t = 1.2 \times 63.4 + 35 \frac{166.2(700-87.8)}{675800} + 0.65 \frac{6142(700-298)}{675800}$$

= 83.724⁰F

Bottom temperature = 700 – 83.724 = 616.28⁰F.

Mass Balances across the Atmospheric Distillation Unit (ADU) [8]

The mass balance across the Atmospheric Distillation Column is performed for five different products namely Light Naphtha (LN), Heavy Naphtha (HN), Light Distillate (LD), Heavy Distillate (HD) and Atmospheric gas oil (AGO) shown in figure 5.

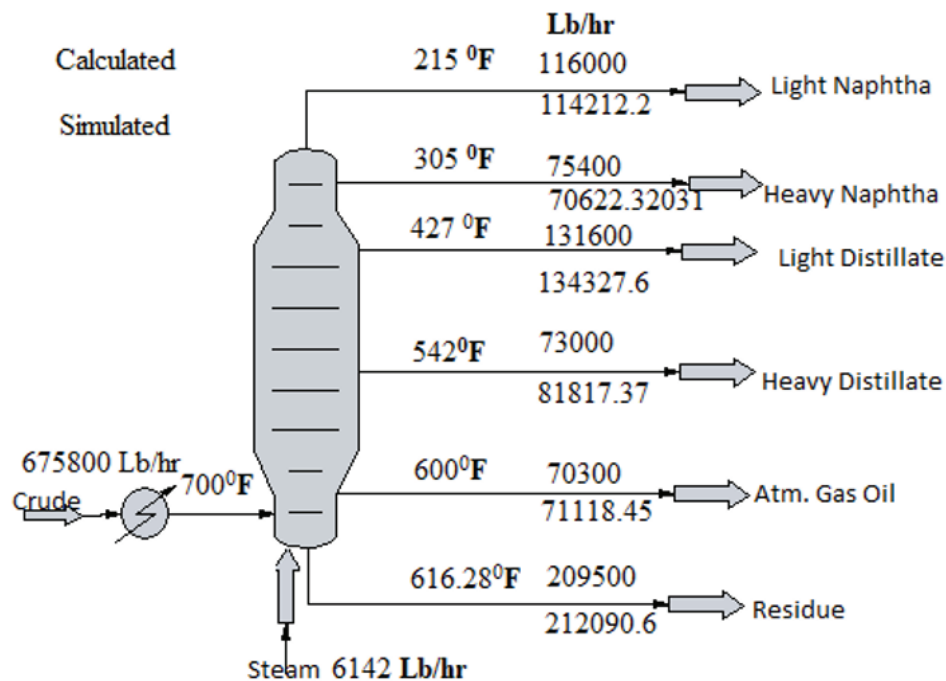


Figure 5: Flowsheet of the Process Feed in ADU [13].

The volumetric balance for the ADU is defined as

$$F_{crude} = F_{gas+LN} + F_{HN} + F_{LD} + F_{HD} + F_{AGO} + F_R \dots\dots\dots (1)$$

Where F refers to the volumetric flow rates of various streams (LN, HN, LD, HD, AGO and R)

The mass balance for the ADU is defined as

$$MF_{crude} = MF_{gas+LN} + MF_{HN} + MF_{LD} + MF_{HD} + MF_{AGO} + MF_R$$

Where MF terms as the mass flow rates associated to the feed and product streams.

Eventually, the mass balance expression can be written in terms of the volumetric balance as

$$SG_{crude} F_{crude} = SG_{gas+LN} F_{gas+LN} + SG_{HN} F_{HN} + SG_{LD} F_{LD} + SG_{HD} F_{HD} + SG_{AGO} F_{AGO} + SG_R F_R \dots\dots\dots (2)$$

Where SG_i refers to the specific gravity of various streams (crude, LN, HN, LD, HD, AGO & R).

The Specific gravity is calculated by using following formula and reported in TABLE 6.

$$^{\circ}API = 141.5 / (S.G.) - 131.5$$

The mass flow rate of a stream is associated with the volumetric flow rate (barrels per day) using the expression:

$$MP_i = P_i \times 42 \times 8.33 \times SG_i \dots\dots\dots (3)$$

Where S_G terms as the specific gravity of the stream i.

Thereby, the mass balances of different products viz. Light Naphtha (LN), Heavy Naphtha (HN), Light Distillate (LD), Heavy Distillate (HD) and Atmospheric gas oil (AGO) calculated by using equation (3) across the ADU are presented in TABLE 7.

Calculation for Tower Diameter

The tower diameter is determined from the flooding correlation for a chosen tray spacing. K is an empirical constant, depends on tray spacing and can be estimated against the flow parameter based on

mass flow rate of liquid (L) and vapour (V) [11,12]. It is a common practice to have uniform tower diameter in all sections of the column even though the vapour and liquid loadings are expected to be different to minimize the cost of construction. The uniformity in tower diameter may require selecting different tray spacing in different sections of the tower.

Total number of plates used in the given problem are 35.

The Tower diameter specification across the ADU are tabulated in TABLE 8.

Properties	Whole Crude	Light Naphtha	Heavy Naphtha	Light Distillate	Heavy Distillate	Gas Oil	Residue
Cut volume %	100	22	11.8	19.2	10.4	9.8	26.8
A.P.I.	36.3	77	46	34	30	26.6	18.6
S.G.	0.843	0.678	0.797	0.855	0.876	0.895	0.943

Table 6: Properties of Crude fractions.

Sr. No.	Stream	Volume %	Flows (lb/hr)	S.G.	Mass Flow Rate
1	Light Naphtha	22	116000	0.678	27515789.93
2	Heavy Naphtha	11.8	75400	0.797	21024416.87
3	Light Distillate	19.2	131600	0.855	39365547.48
4	Heavy Distillate	10.4	73000	0.876	22372847.28
5	AGO	9.8	70300	0.895	22012666.41
6	Residue	26.8	209500	0.943	69117816.81
	Crude (Total)	100	675800	0.843	199315032.1

Table 7: Mass balances across the ADU.

Sr. No.	Product	Tray Spacing (inch)	K value	Tower Diameter (ft)
1.	Light Naphtha	25.0	1082.3	22.51
2.	Heavy Naphtha	20.1	902.0	22.94
3.	Light Distillate	28.5	1156.0	17.95
4.	Heavy Distillate	18.7	835.3	18.83
5.	Atm. Gas Oil	17.2	774.3	14.55

Table 8: Tower diameter specification across the CDU.

Calculation for Tower Height

The number of trays used in the column are 35 trays, from LN-HN section consists of 8 trays, HN-LD 8 trays, LD-HD 6 trays, HD-AGO 6 trays, AGO-B(R) 3 trays and flashing zone to bottom portion of the column consists of 4 trays [10].

Assume the thickness of the valve type tray is 12 US Std. Gauge (2 mm).

The height of the distillation column is calculated using the formula [11],

$$H = C (N_R - 1) + C_T$$

Where,

H=Height of the column,

C=Plate Spacing (ft),

C_T =Allowance = 4 m at the top = 13.1234 ft

Total height of tower (H) = Height from [(LN-HN) + (HN-LD) + (LD-HD) + (HD-AGO) + (AGO-B) + Flashing zone] + Allowance

$$H = \{2.0833(8-1)\} + \{1.675(8-1)\} + \{2.375(6-1)\} + \{1.55833(6-1)\} + \{1.4333(3-1)\} + \{1.3333(4-1)\} + 13.1234$$

H = 65.97 ft = 20.108 m.

Result and Conclusion

The atmospheric distillation column for Five petroleum product viz. LN, HN, LD, HD, and AGO is designed. The height of tower (H) for crude oil feed rate 675800 lb/hr of a 36.3°API at 700°F is 65.97 ft (20.108 m). The petroleum products LN, HN, LD, HD, and AGO are removed at temperatures 215°F, 305°F, 427°F, 542°F, 600°F and 616.28°F respectively.

The tower diameter is calculated for sieve and valve tray considering tray spacings 25, 20.1, 28.5, 18.7 and 17.2 inch are 22.51, 22.94, 17.95, 18.83 and 14.55 ft for LN, HN, LD, HD, and AGO respectively.

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