PROCESS DESIGN OF ATMOSPHERIC CRUDE DISTILLATION COLUMN

THESIS

Submitted in Partial Fulfillment of the Requirements

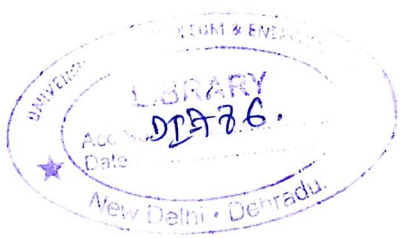
for the Award of the Degree of

MASTER OF TECHNOLOGY

(REFINING & PETROCHEMICAL ENGINEERING)

Submitted by

SREEDHAR KAMBHAM



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By

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May, 2010

DECLARATION

I hereby declare that the work which is being presented in the thesis entitled "PROCESS DESIGN OF ATMOSPHERIC CRUDE DISTILLATION COLUMN" by "SREEDHAR KAMBHAM" in partial fulfillment of requirements for the award of degree of M. Tech. (Refining & Petrochemical Engineering) submitted in the Department of Chemical Engineering at UNIVERSITY OF PETROLEUM & ENERGY STUDIES, DEHRADUN is an authentic record of my own work carried out during a period from 1ST NOV, 2010 to 31ST APR, 2010 under the supervision of Prof. K.V. Rao. The matter presented in this thesis has not been submitted by me in any other University / Institute for the award of M. Tech Degree.

Due to the confidentiality of data, the original name, location and identity of the data has been changed.

Signature of the Student

This is to certify that the above statement made by the candidate is correct to the best of My /our knowledge

Signature of the SUPERVISOR

CERTIFICATE

This is to certify that the work contained in this thesis titled "PROCESS DESIGN OF ATMOSPHERIC CRUDE DISTILLATION COLUMN" has been carried out by Mr. SREEDHAR KAMBHAM, under my supervision and has not been submitted elsewhere for a degree.

KVIC

Signature of the Supervisor

Date: 11-5-2010

ABSTRACT

The main objective of this project is to design a Atmospheric distillation column for crude oils. Crude oil is considered as complex mixture because it is made up of large no of discrete hydrocarbons, each component being present in relatively small amount. So the design is done on an empirical basis.

The method followed here is Watkins procedure. A Type R system which is a rectifying section employing subcooled reflux is chosen. Here the no of trays required for separation of petroleum fractions is assumed and over head accumulator pressure is assumed this decides the flash zone pressure. Based on product specifications and flash zone conditions we can estimate the material balance of the tower. Draw temperatures are assumed using graphical relation and this is estimated by calculating the bubble point temperature of the product at the partial pressure of the product at the draw tray this is done by making heat balance to the tower by assuming linear temperature drop between draw trays. To calculate the properties of petroleum fractions analytical correlations are used.

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NOTATION

T_r	Reduced temperature .
T_{pC}	Pseudo critical temperature of the fraction is in degree rankine.
P_r,	Reduced pressure.
P_{pc}	Pseudo critical pressure of the fraction in Psia.
M	Molecular weight of the fraction.
ω	Accentric factor.
K_W	Watson characterization factor.
T	The boiling point to the corresponding pressure.
T_b	Average boiling point in Kelvin.
SG	Specific gravity.
H_L	Enthalpy of liquid petroleum fractions in Btu/lb.
H	Enthalpy of petroleum fractions in Btu/lb.
h	Enthalpy of Liquid in Btu/lb.
Q	Total heat content in Btu/hr.
VABP	Volumetric average boiling point in ^{0}F.
SL	Slope of the line.
a,b,c	are constants.
V	Vapour flow rate in lb/hr.
L	Liquid flow rate in lb/hr.
P	Total pressure in Psia.
P_{HC}	Partial pressure of hydrocarbon in Psia.
K	Equilibrium ratio.
$\sum D$	sum of distillates.
D	distillate flow rate in lb/hr.
F	Feed flow rate in lb/hr.
W	Bottoms flow rate in lb/hr.
Rc	Reflux flow rate in lb/hr.

SUBSCRIPTS

v	Vapour
L	Liquid
NBP	Normal boiling point
S	Steam
i	component
Avg	average
FO	Furnace outlet
FZ	Flash zone
O	Over flash

1.0 INTRODUCTION

The design and operation of crude oil distillation units are still done almost exclusively on an empirical basis. This is because crude oil and its products are made up of a large number of discrete hydrocarbons, each compound being present in relatively small amount.

Crude oil is a complex mixture of several hundreds of hydrocarbon compounds ranging from simple ethane to very large molecules containing 200 or more carbon atoms. Other than the pure hydrocarbons, the crude contains inorganic impurities in the form of organic compound of sulphur, nitrogen and oxygen. Traces of organo – metallic compounds of iron, nickel, copper, vanadium, arsenic etc. are also present. Crude oil boils over a wide range of 25 – 800^0C. Crude oil being a complex mixture needs to be refined into several products, which are tailored to certain specifications depending upon the end use. The principal products of petroleum refining are:

a) Low boiling products: methane, ethane, propane, butane & iso- butane, LPG.
b) Naphthas: light naphtha, heavy naphtha, gasoline or petrol, aviation gasoline.
c) Light distillates: Kerosene, jet fuels, Aviation turbine Fuel (ATF).
d) Heavy distillates: Light gas oil (LGO), Heavy Gas oil (HGO), Diesel.
e) Vaccum Gas Oils
f) Lubricating oils &waxes.
g) Residues: fuel oils, coke, asphalt, carbon black etc.

1.1 Crude distillation unit (CDU):

It is the core of any petroleum refinery. The CDU differs from the conventional distillation column due to the presence of pump-arounds and side –strippers inter linked to the column. The CDU also lacks a conventional reboiler at the bottom of the main column. The feed stock or the crude oil is desalted and heated to its maximum allowable temperature in a furnace before entering the main column at the flash zone. The flash zone is the area in the CDU where the vaporised feed is allowed to separate from the un-vaporised liquid feed. The vapours rise up through the column and condense on the trays, up on contact with the cold reflux streams coming down. Inter – phase mass transfer takes place primarily on the column trays.

Super heated steam enters the main column from the bottom. Steam serves to strip the liquid feed of its low boiling components in the stripping section. The bottom liquid is withdrawn from the main columns long residue. The main purpose of the stem is to reduce the partial pressure of the hydrocarbons in the vapour phase there by reducing the temperature required for a given amount of vaporisation.

The overhead vapours from the top tray of the column are cooled in a series of heat exchangers where they are partially condensed. The cooled stream enters the reflux drum, where the non- condensable gases are separated and sent to flare. Water is separated out by virtue of its density and immiscibility with the organic product. Water is drained off from the reflux drum. Apart of the liquid is returned to the top tray as reflux and the rest is withdrawn as distillate product.

There are a number of side draws in the tower from which the various products like naphtha; Kerosene, gasoil etc are withdrawn. The products are separated according to their boiling ranges. The side – draw products contain some amount of low boiling materials which are stripped off by superheated steam in the side-strippers. The side strippers are generally plate columns in to which steam is injected below the bottom tray and the stripped vapours enriched with low boiling materials are returned from the top of the side stripper to the main column at a location above the withdrawal tray. The desired product stream, free from the low boiling materials, is drawn from the bottom of the side stripper.

1.2 Various ways of separating the products:

The design method is based on Watkins method. There are three ways of producing distillate products from CDU. They are

Type U: It designates that the side draw products are unrefluxed even though there is reflux at the top of the column.

Disadvantages:

(i). The vapour and liquid traffic throughout the column increases markedly from the bottom to the top stages.

(ii). Sizing a tower based on conditions at the top tray results in a much larger tower than required for either a Type A or Type R.

(iii). This type of unit is uneconomical since heat is removed only in the final condenser.

Type A: It designates that heavy liquid is withdrawn from the tray above the draw tray and returned to the column three trays above the draw tray. This type is called a pump around reflux. The removal of heat in the side cooler results in a smaller requirement for overhead reflux.

Advantages:

Stabilizing vapour and liquid flow throughout the section where it is used.

Disadvantages:

in this three trays used for heat removal must be considered as only one tray for fractionation purposed. This is because the pump around liquid is foreign to the zone in which it is introduced. This liquid is good for heat removal purposes but offers no mass transfer advantages.

Type R: A rectifying section employing sub cooled reflux is designated Type R. This is a more efficient method for recovering heat from the system because it can be used at every section of the column.

Advantages: It has advantage of being true reflux which contributes to fractionation on all trays. A fairly uniform vapour and liquid traffic exists in all sections of the tower.

Disadvantages: This method requires higher capital cost. This method incurs a large stripping cost.

1.3 Characterization of crude

The crude oil contains continuous chemical species like light gas to heavy components like asphaltenes, therefore component by component analysis practically impossible therefore there are standard methods for characterizing the crude. The following the method to characterize the crude by fractionization

1. True boiling point method (TBP)
2. ASTM distillation
3. Equilibrium flash vaporization(EFV)

1.3.1 TBP analysis

This is a batch distillation which consists of a large number trays and a high reflux to distillate ratio so that it gives percentage volume of crude oil distilled at a particular temperature on a temperature- volume yield curve. TBP analysis is only done for crude oil not for petroleum products.

1.3.2 ASTM distillation

In ASTM distillation, 100 ml of crude is distilled at a uniform rate of 5 ml per minute. The vapour from the crude is condensed in a brass tube, cooled by ice. The temperature of vapour at which first drop of condensate falls from the condenser is noted as the initial boiling point and the vapour temperature is also noted as each 10% is collected. When 95% is distilled, the burner flame is increased and the maximum temperature is the end point. Same method can be used analysis petroleum fractions.

1.3.3 Equilibrium flash vaporization

High boiling materials like bright stock and asphalt cannot be completely vaporized in true boiling-point apparatus even at very low pressures. The highest vapour temperature obtained in a true-boiling point apparatus is 538°C. Lubricating oil will be having higher temperature than this.

1.3.4 Mid percent curves.

The property of each drop of oil goes on changing as we advance from the first drop of distillate to the last drop. In distillation, compounds are arranged according to their boiling points. The properties of each fraction of an oil are the averages of the properties of all drops and from the succeeding one, then the drop, distilling at the centre of the fraction has the same property as the average of all drops.

In the refinery there are two terms used to discuss the product composition and the degree of separation between the adjacent fractions. ASTM boiling range defines the general composition of the fraction and is usually one of the key specifications for most distillates from both the atmospheric tower and the vacuum tower.

Second term, 5-95 Gap, defines the relative degree of separation between the adjacent fractions. It is determined by subtracting the 95 volume percent of ASTM temperature fraction from the 5 volume percent adjacent heavy fraction.

Packie defined the degree of difficulty of separation as the difference between the ASTM 50 volume percent temperatures of the fractions under study.

1.3.5 TBP cut points

The TBP range is used to define the crude oil property as whole, when it comes to the product we need on a particular temperature. The TBP cut point temperature is the temperature at which the middle at which 100% lighter and 0% heavier product is distilled.

1.4 Present work:

In the current work, a Type R method is used. This work is based on Watkins method where he followed graphical procedure but in this work analytical procedures are used. The crude oil is fractionated in to overhead vapour, light naphtha, heavy naphtha, light distillate, heavy distillate, gas oil and reduced crude. The design procedure is discussed in Chapter 2. The calculations is discussed in Chapter 3. The results are discussed in chapter 4. The conclusion is discussed in chapter 5.

2.0 DESIGN PROCEDURE

In the separation of chemically similar materials, it is generally true that the greater the difference in boiling points of individual fractions, the easier it is to separate the fractions. The sharpness of separation usually is defined as the gap or overlap of the boiling ranges of adjacent fractions.

2.1. Product specifications:

Generally the process designer must estimate the material balance on the basis of the specifications given for the desired products.

2.1.1 ASTM end points: one method for setting product specifications is to state the maximum allowable end points for the fractions. The heaviest distillate product called AGO, is excluded from this specifications since it is withdrawn from the column to provide adequate fractionation between the other distillates and the reduced crude. These product specifications are shown table 3.4.

2.2.Material Balance:

The material balance is estimated as a function of the desired characteristics of the fractions and the amount of crude that can be vaporized at the outlet conditions of the furnace. It is necessary to estimate the number of actual trays in the column and the operating pressures of the column in order to set the conditions at the furnace outlet.

Number of trays: The following trays are recommended

Tray requirements in atmospheric crude towers	
separation	**Number of trays**
Light naphtha to Heavy Naphtha	6 to 8
Heavy Naphtha to Light Distillate	6 to 8
LightDistillate to Heavy Distillate	4 to 6
Heavy Distillate to Atmospheric gas oil	4 to 6
Flash zone to first draw tray	3 to 4
steam and reboiled stripping sections	4

Operating pressures are estimated by assuming a pressure of 0.5 to 1.0 Psig in the final accumulator. Assume a pressure drop through the condenser and piping is 5 Psig. Assume a

pressure drop per plate is 0.15 Psig. Assume a pressure drop between the furnace outlet and the entrance to the tower flash zone is 5 Psig.

Product specifications determine the cut points between fractions. These temperatures, in turn, determine the quantity of each fraction which will be produced. The method to calculate the material is as follows:

(i). Convert the TBP curve of the whole crude into an equilibrium flash vaporization (EFV) curve for the pressure at the charge furnace outlet.

To do this conversion following procedure has done.

a. First convert.TBP data in to ASTMD86 data.

Riazi and Daubert methods for interconversion of various distillation data are based on generalized correlation for property estimation of hydrocarbons in the form of

$$ASTMD86 = (1/a)^{1/b} (TBP)^{1/b} \quad (1)$$

Where a, b are correlation parameters specific for each conversion type and each vol% on distillation curve. Correlation constants for above equation is given in table 2.1.

Table 2.1. Correlation constants for Eq. (1)

vol%	0	10	30	50	70	90	95
a	0.9177	0.5564	0.7617	0.9013	0.8821	0.9552	0.8177
b	1.0019	1.09	1.0425	1.0176	1.0226	1.011	1.0355

b. Then convert ASTMD86 data in to EFV data. This has done using the following correlation.

$$EFV = a\,(ASTMD86)^b\,(SG)^c \quad (2)$$

$$SG = a(T_{10})^b T_{50}{}^c \quad (3)$$

Where a, b, c are constants given in table 2.2 and 2.3.

Table 2.2. Correlation constants for Eq. (2)

vol%	0	10	30	50	70	90	100
a	2.9747	1.4459	0.8506	3.268	8.2873	10.6266	7.9952
b	0.8466	0.9511	1.0315	0.8274	0.6871	0.6529	0.6949
c	0.4209	0.1287	0.087	0.6214	0.934	1.1025	1.0737

Table 2.3. Correlation constants for Eq.(3)

Distillation type	T10 °C	T50 °C	SG	a	b	c
ASTMD86	35-295	60-365	.7-1.0	0.08342	0.10731	0.26288
TBP	10-295	55-320	.67-.97	0.10431	0.1255	0.20862
EFV	79-350	105-365	.74-.91	0.09138	-0.0153	0.36844

What we calculate the EFV temperatures using above correlations are at atmospheric pressure. To convert this temperatures to the conditions of furnace outlet pressure the following correlation is used. The method widely used in industry is the correlation developed for petroleum fractions by Maxwell and Bonnel and is given by

$$T = \frac{T_b'}{748.1*Q - T_b'*(0.3816*Q - 0.00051606)} \quad (4)$$

Where $T_b' = T_b$-1.3889*F*$(K_W - 12)\log_{10}(\frac{P}{760})$ (5)

$$Q = \frac{6.412631 - 0.989679*\log_{10} P}{2770.085 - 36\log_{10} P} \; where\; P > 760\; mmHg \quad (6)$$

$$Q = \frac{5.994296 - 0.972546*\log_{10} P}{2663.129 - 95.76\log P} \; where\; 2 \le P \le 760\; mmHg \quad (7)$$

$$Q = \frac{6.761560 - 0.987672*\log_{10} P}{3000.538 - 43\log_{10} P} \; where\; P > 2\; mmHg \quad (8)$$

F= 0 when ($T_b < 367K or\ when\ K_W\ is\ not\ available$

F = -3.2985 + 0.009T_b

$K_W = \frac{[(1.8T_b)]\ ^((1/3))}{SG}$ Where $T_b$is normal boiling point in Kelvin

(ii). Calculate the heat content for both vapour and liquid phases of the charge at the furnace outlet temperature and pressure.

To calculate heat content we have to know enthalpy of liquid and vapour and this is calculated by using method given by API-TDB.

Region 1: Liquid phase

Where $T_r \le 0.8$, $P_r \le 1.0$

$$H_L = A_1[T - 259.7] + A_2[T^2 - 259.7^2] + A_3[T^3 - 259.7^3] \quad (9)$$

H_L = Enthalphy of liquid petroleum fraction, $^{BTU}/_{lb}$

$$A_1 = 10^{-3} * \left[-1171.26 + (23.722 + 24.907 * SG) * K_w + \left(\frac{1149.82 - 46.535K_w}{SG}\right)\right] \quad (10)$$

$$A_2 = 10^{-6} * \left[(1.0 + 0.82463 * K_w)\left(56.086 - \frac{13.817}{SG}\right)\right] \quad (11)$$

$$A_3 = -10^{-9}\left[(1.0 + 0.82463 * K_w)\left(9.6757 - \frac{2.3653}{SG}\right)\right] \quad (12)$$

Region II: Vapor phase (or) liquid phase with

$T_r \ge 0.8$, $P_r \ge 1.0$

$$H = H_L + B_1[T - 0.8T_{PC}] + B_2[T^2 - 0.64 * {T_{PC}}^2] + B_3[T^3 - 0.512{T_{PC}}^3] + \frac{RT_{PC}}{M}\left[4.507 + 5.26 * \omega - \left(\frac{\bar{H}^0 - \bar{H}}{RT_{PC}}\right)\right] \quad (13)$$

$$B_1 = 10^{-3}\left[-356.44 + 29.72K_w + B_4\left(295.02 - \frac{248.46}{SG}\right)\right] \quad (14)$$

$$B_2 = 10^{-6}\left[-146.24 + (77.62 - 2.772K_w)K_w - B_4\left(301.42 - \frac{253.87}{SG}\right)\right] \quad (15)$$

$$B_3 = 10^{-9}[-56.487 - 2.95 * B_3] \quad (16)$$

$$B_4 = [(^{12.8}/_{K_w} - 10)\,(1 - {}^{10}/_{K_w})\,(SG - .885)\,(SG - .7) * 10^2\,]^2 \quad (17)$$

For 10<K<12

With .7<SG<.885

$$(\bar{H}^0 - \bar{H})/_{RT_{P_C}} = -P_r[(0.1445 + 0.073\omega) - (0.66 - 0.92\omega)T_r^{-1} - (0.4155 + 1.50\omega)T_r^{-2} - (0.0484 + 0.388\omega)T_r^{-3} - 0.0657\omega T_r^{-8}] \quad (18)$$

Prediction of Molecular weight: the molecular weight is calculated by using Riazi-Daubert method.The correlation is

$$M = 42.965[\exp(2.097 * 10^{-4} * T_{avg} - 7.78712 * \mathrm{SG} + 2.08476 * 10^{-3} * T_{avg} * \mathrm{SG})] * T_{avg}^{1.26007} * SG^{4.98308} \quad (19)$$

Where T_b is average boiling point in degree Kelvin.

This correlation can be applied to hydrocarbons with molecular weight ranging from 70 to 700 .which is linearly equivalent to boiling point range of 300-850 k and API Gravity range of 14.4-93.

- **Prediction of Pseudo critical temperature of the fraction :** Pseudo critical temperature is calculated by using correlation given in API-TDB.
- $T_{pc} = 10.6443[\exp(-5.1747 * 10^{-4} * T_{avg} - 0.54444 * SG + 3.5995 * 10^{-4} * T_{avg} * SG)] * T_{avg}^{0.81067} * SG^{0.53691}$ (20)
- where T_{avg}, T_{pc} *is in rankine.*
- **Prediction of Pseudo critical pressure of the fraction:** Pseudo critical pressure is calculated by using correlation given in API-TDB.
- $P_{pc} = 6.162 * 10^6[\exp(-4.725 * 10^{-3} * T_{avg} - 4.8014 * SG + 3.1939 * 10^{-3} * T_{avg} * SG] * T_{avg}^{-0.4844} * SG^{4.0846}$ (21)
- where T_{avg} *is in rankine and* P_{pc} *is in psia.*
- **Prediction of accentric factor :**
- Accentric factor is calculated by using following correlation
- For $T_R \leq 0.8$, the correlation given by Kesler and Lee is used.
- $\omega = \frac{-\ln P_c - 5.92714 + \frac{6.09648}{T_R} + 1.28862 \ln T_R - 0.169347 T_R^6}{15.2518 - \frac{15.6875}{T_R} - 13.4721 \ln T_R + 0.43577 T_R^6}$ (22)

For $T_R > 0.8$, the correlation given by Edmister and Lee is used

$$\omega = \frac{3}{7}\left[\frac{\log P_c}{T_R^{-1}-1.0}\right] - 1.0 \text{ where } P_c \text{ is in atmospheres} \quad (23)$$

(iii). Repeat the preparation of the EFV curve for the total pressure at the flash zone inlet.

(iv). By trail-and-error procedure which requires assuming a new temperature at the flash zone entrance, determining the new percentage vaporization and then calculating the heat content until satisfying the criterion $Q_{FO}=Q_{FZ}$.

The temperature drop between the furnace outlet and the flash zone inlet will usually be on the order of 4 to 6^0F.

(v). Set the value for overflash. This liquid returns to the flash zone from the tray immediately above it and mixes with the feed flash liquid. The sum of these two liquids is the feed to the top tray of the bottoms stripping section.

(vi). Set a stripping steam rate-10 pounds per barrel of net bottoms is recommended. From Figure 3.9, find the volume percentage of the feed to the stripping section which will be vaporized. From this difference, calculate the volume of net bottoms.

(vii). The total vapour leaving the flash zone is the sum of the feed flash vapour and the stripout.

(viii). Calculate the hydrocarbon partial pressure in the total vapour leaving the flash zone as P_{HC} . Find the 14.7 Psia EFV temperatures for the indicated volume percentage vaporization and convert this to the pressure level of P_{HC}. This is the flash-zone temperature.

(ix). Assume that the temperature of the net bottoms liquid, W, is 30^0F lower than the flash zone temperature. Calculate the heat quantity leaving the system in this stream.

(x). Calculate the heat input to the system from the stripping steam.

(xi). Total distillate yield is the total hydrocarbon vapour leaving the flash zone minus the over flash.

(xii). The volumetric yields of the distillate fractions are determined by converting their ASTM specifications into TBP initial and endpoints using Figure 3.1-3.2. The TBP cut points on either side of a fraction establish the volumetric yield of that fraction.

(xiii). The ASTM, TBP and EFV data for all fractions are determined using correlations given above.

(xiv). Extrapolate the EFV curves from their initial points to the points representing minus 10 percent to account for the removal of 10 percent front ends from the product streams by the product strippers.

(xv).The separation of the Light naphtha in the condenser-accumulator is established by vapour-liquid equilibrium calculations. The light end analysis of the crude and the TBP estimates for the light naphtha are used.

2.3.Heat Balance

Aside from heat accompanying the stripping steam, the only heat to the column is added by the furnace. For most cases, the furnace outlet temperature is assumed to be 700^0F. The furnace outlet temperature should not exceed this temperature without specific reasons.

Draw tray temperatures are assumed using Figure 3.11. The conditions for using Figure 3.11 since it correlates draw tray temperature against the EFV temperature of the product on the tray. Gas oil and bottoms are assumed to be steam stripped while all other products are reboiled stripped while all other products are reboil stripped. For side draw products, bubble point temperatures for the unstrapped liquid are used. For the over head product, the dew point temperature is used.

Draw tray temperature assumptions are checked by calculating the bubble-point temperature of the unstripped product at the partial pressure of the product vapour in the total vapour leaving the draw tray. This is done by making a heat balance around the draw tray and calculating the amount of internal reflux required to absorb the excess heat. This first requires making a heat balance at the first tray below the draw tray in order to establish circulating reflux requirements. Side strippers are calculated in different ways for the steam stripper than for the reboiled stripper. The steam rate is arbitrarily set at 10 pounds per barrel of stripper net bottoms. For distillate fractions, this will result in a stripping rate of between 6 and 10 volume percent of the gross stripper charge.

For steam strippers, the temperature of the stripped material leaving the bottom of the stripper is assumed to be 30^0F lower in temperature than the corresponding draw tray temperature. For reboiled strippers, this temperature is assumed to be 30^0F higher than the corresponding draw tray temperature.

The temperature of the liquid on the top tray is calculated by making a heat balance on the tray. On this tray however, instead of using the bubble point temperature , the dew point

temperature is adjusted to the partial pressure of the condensable hydrocarbons in the total vapour leaving the top tray.

2.3.1.First sidestream product Draw tray:

The heat and material balance relationships is shown in Fig2.1.

Balance above Tray(D_1-1)-Envelope II

(i). Calculate the reflux heat at tray (D_1-1).

(ii). Calculate the heat removal capability of the cooled pumpback reflux.

(iii). Calculate the amount of cool pumpback reflux which is required to absorb the reflux heat at tray(D_1-1). Calculate the heat content of this stream as it reenters the tower.

Balance above TrayD_1-Envelope III

(i). Calculate the reflux heat at tray D_1. The liquid leaving this draw tray is the sum of the product, D_1, and the pumpback reflux, $R_{c(D1-1)}$ and stripout liquid , $V_{L_{D_1}}$.

(ii). Calculate the heat absorbed by the stripout liquid in passing across the draw tray.

(iii). Calculate the heat removal capability of the internal reflux falling from tray tray (D_1+1).

(iv). Calculate the internal reflux from tray(D_1+1) which is required to absorb the excess heat at tray D_1.

(v). Calculate the mole fraction of product vapours in the total vapours leaving the draw tray, remembering to neglect the presence in the vapour of the product which is to be withdrawn on the next draw tray up in the tower.

(vi). Calculate the hydrocarbon partial pressure of product in the total vapour leaving the draw tray. Convert the atmospheric bubble point of the unstripped liquid on the draw tray to this partial pressure. If this temperature doesnot check the value assumed earlier, repeat the procedure for a new assumed temperature.

(vii). Calculate the heat effect at the sidestream stripper.

(viii). Calculate the reflux induced on tray(D_1-1) as the amount of vapour from tray(D_1-2) which enters and is condensed on tray (D_1-1) for the purpose of converting the subcooled pumpback reflux liquid to its bubble point.

2.3.2. Other sidestream products draw trays:

The remaining sidestream draw trays are calculated by the same procedure as that outlined in the previous step.

2.3.3.Top tray calculations:

The heat and material balance relationships at the top tray are determined by making a balance around EnvelopeIV as shown in figure 2.1.

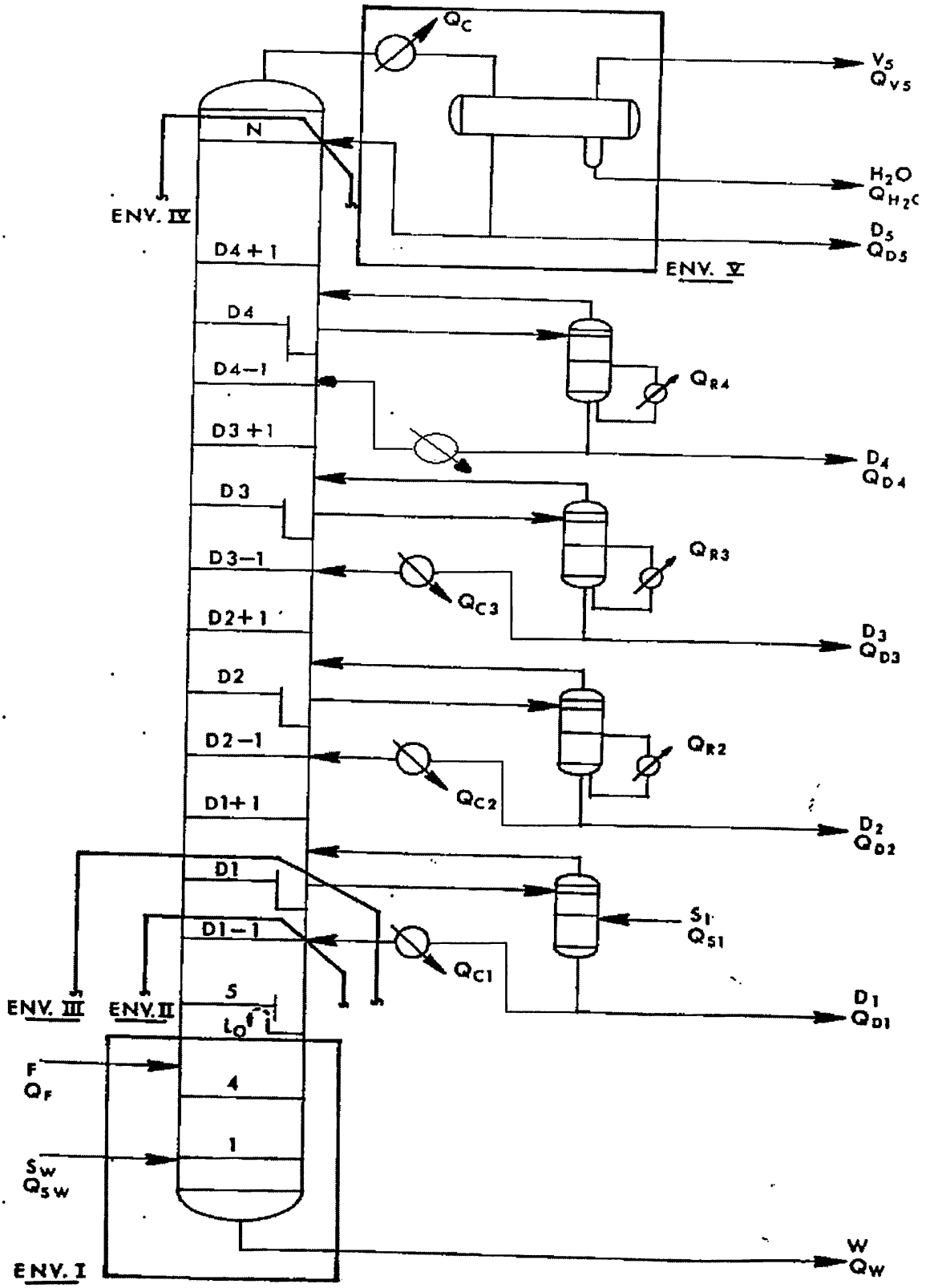

Figure 2.1. Heat and material balance of the tower

3.0 CALCULATIONS

Design basis is -----1hr

The charge capacity will be 55,000 barrels per day (BPSD) of a 36.3^0 API crude oil having an analysis as given below. The crude oil analysis is usually given in the form of a true boiling point curve (TBP) and a light ends analysis and these are given in table3.1. and table3.2 respectively.

Table 3.1. Crude TBP data

Vol %	5	10	20	30	40	50	60	70	80	85
TBP ,°F	62	144	255	344	437	531	623	717	819	897

Table 3.2. Light end Analysis of Crude

Component	C_1	C_2	C_3	iC_4	nC_4	iC_5	nC_5
TNBP, °F	-259	-158	-44	11	31	82	97
Vol%	0.01	0.07	0.92	1.44	2.26	1.9	1.8

Table 3.3. Vol% vs. Wt% curve

Vol%	5	10	20	30	40	50	60	70	80	85
Wt%	3.5	7.5	15.95	25	35	44.5	55	66	76	82

3.1Assumptions

i. Pressure drop between flash zone and top tray will be 4.5 Psig.

ii. Pressure drop between top tray and condenser is 5 Psig.

iii. Linear temperature drop between draw trays.

iv. The bottom residue leaves the column at a temperature 30^0F less than flash zone temperature.

v. The over head accumulator pressure is 15.2 Psig and temperature is 100^0F.

vi. AGO side stream is steam stripped and remaining side stream strippers are reboiled strippers.

vii. The streams leaving from the bottom of the steam stripper will leave 30^0F less than the draw temperature.

viii. The streams leaving from the bottom of the reboiled strippers will leave 30^0F higher than the draw temperature.

3.2 Operating conditions:

The furnace outlet pressure is 29.7 Psia.

For this operating pressure the recommended no of plates is 35

No of plates between flash zone and atmospheric gas oil draw is 3

No of plates between atmospheric gas oil and Heavy distillate is 6

No of plates between Heavy distillate and light distillate is 6

No of plates between Light distillate and Heavy naphtha is 8

No of plates between Heavy naphtha and Light naphtha is 7

No of plates in bottoms stripping section is 5

Table 3.4.Product specifications

Stream	ASTM endpoint,°F	ASTM gap	API
Light Naphtha	275	**20**	77
Heavy Naphtha	380	35	46
Light Distillate	560	10	34
Heavy Distillate	650	5	36
Gas oil	735	5	26.6

From graphical relation between ASTM end point and TBP end point which is shown in Figure 3.1. We estimate TBP endpoint and is shown in Table 3.5.

Table 3.5.Estimated TBP endpoints for the fractions

Stream	TBP end point, °F
Light Naphtha	298
Heavy Distillate	399
Light Distillate	587
Heavy Distillate	684
Gas oil	780

From TBP overlap ($t_{100\ L}$--- t_{0H}) and ASTM gap (t_{5H} --- t_{95L}) graphical relation we estimate TBP initial point as follows. This is shown in Figure3.2.

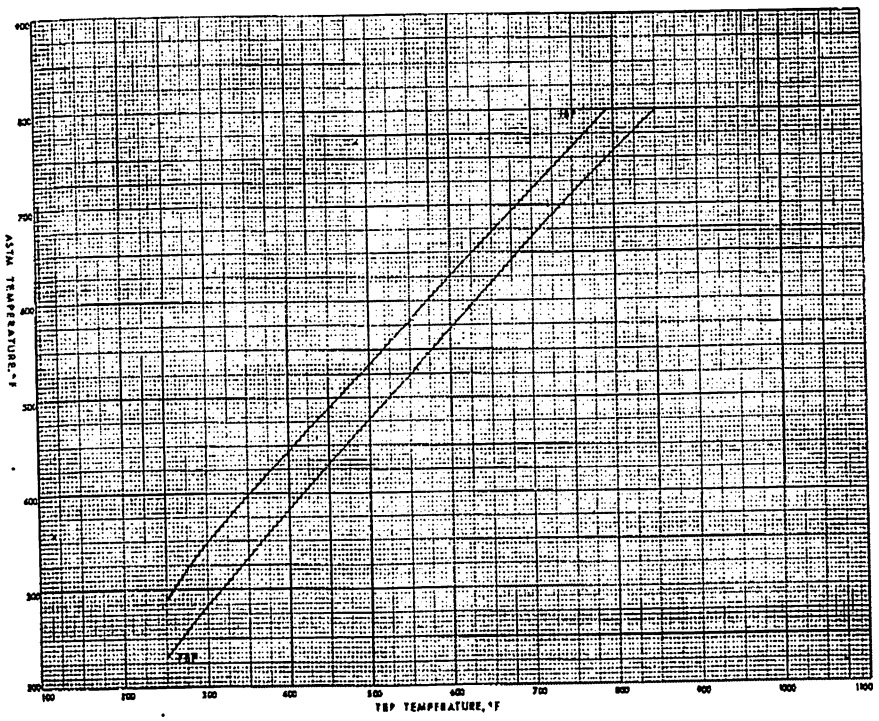

Figure3.1.Relationships between ASTM and TBP initial (0 percent) and final (100 percent) boiling points.

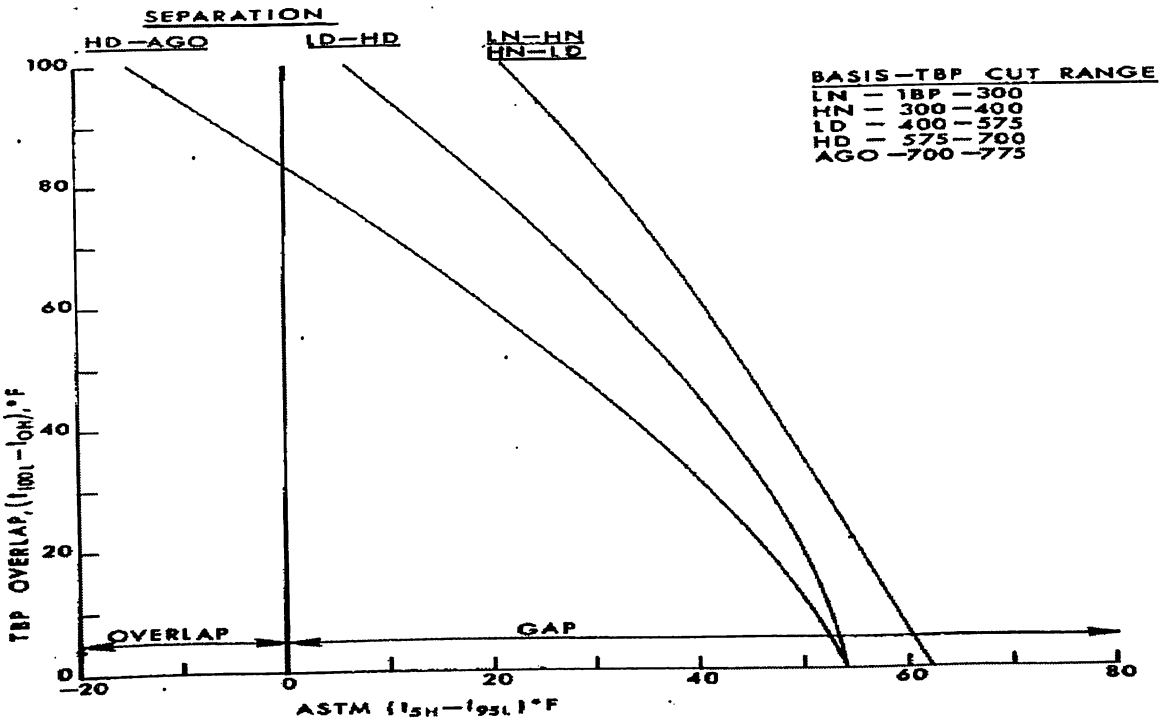

Figure3.2.Relation between TBP overlap ($T_{100L} - T_{0H}$) and ASTM gap ($T_{5H} - T_{95L}$).

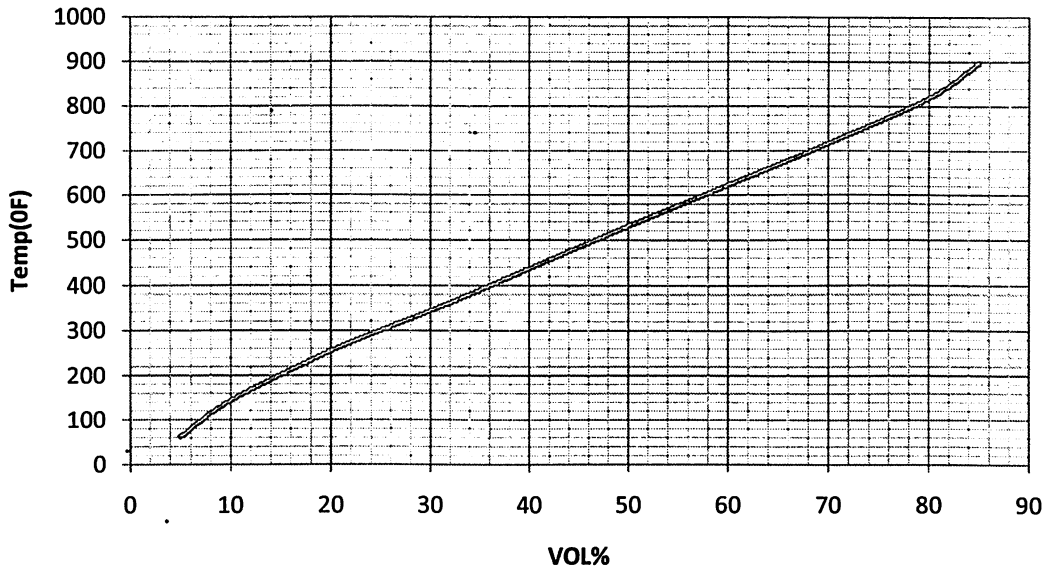

Figure3.3. Crude TBP curve

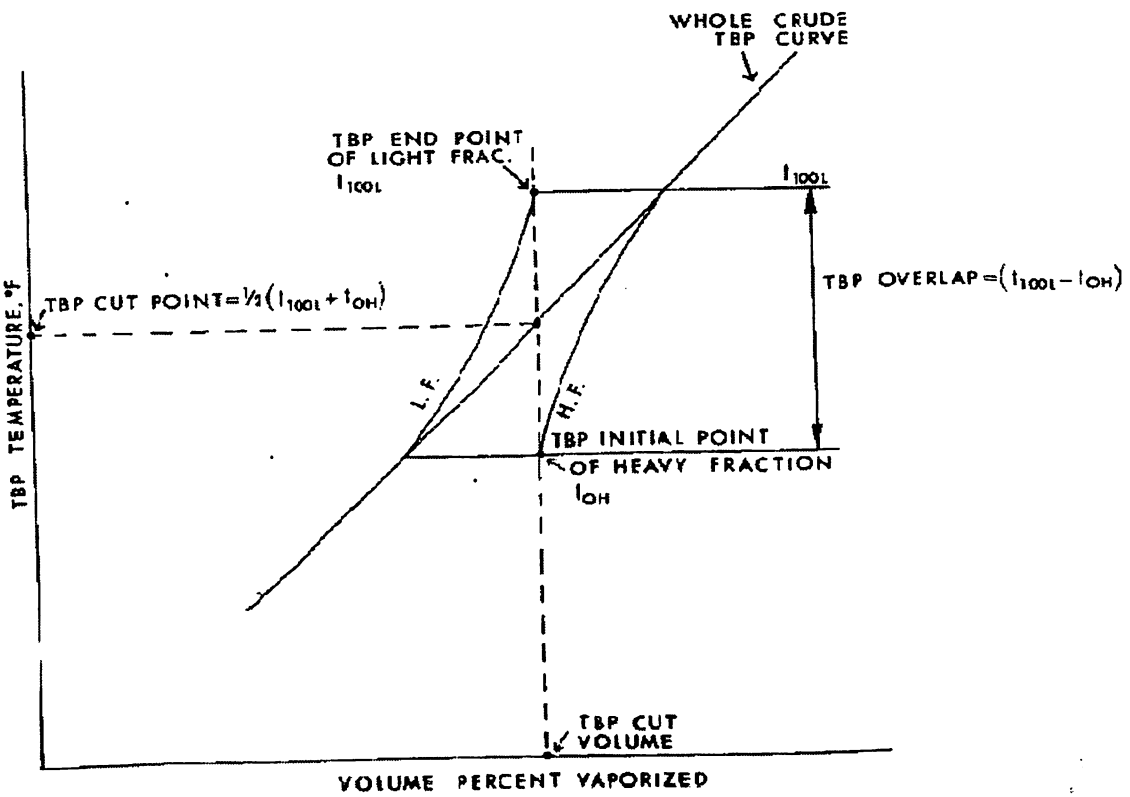

Figure 3.4. Temperature relationships around the cut point between fractions.

3.3 TBP Cut point and Cut volume

For Light Naphtha (LN) and Heavy Naphtha (HN):

Given ASTM gap (t_{5H} --- t_{95L}) = 20 for this from graph TBP overlap ($t_{100\,L}$--- t_{0H}) = 50

Here we know $t_{100\,L}$=298^{0}F

t_{0H} = 298 – 50 =248^{0}F i.e. initial point of Heavy Naphtha is 248^{0}F.

We now calculate cut point of Light Naphtha as follows:

TBP cut point = (1/2)*($t_{100\,L}$ + t_{0H}) = (1/2)*(298 + 248) = 273^{0}F.

From Fig3.3. Cut volume percent = 22%.

For Heavy naphtha (HN) and Light distillate (LD):

Here HN is lighter fraction (L) and LD is heavier fraction (H)

Given ASTM gap (t_{5H} --- t_{95L}) = 35 for this from graph TBP overlap ($t_{100\,L}$--- t_{0H}) =40

Here we know $t_{100\,L}$=399

t_{0H} = 399 – 40 = 359^{0}F i.e. initial point of Light Distillate is 359^{0}F.

We now calculate cut point of Heavy Naphtha as follows:

TBP cut point = (1/2)*($t_{100\,L}$ + t_{0H}) = (1/2)*(399 + 359) = 379^{0}F

From Figure 3.3. Cut volume percent = 11.8%.

For Light Distillate (LD) and Heavy Distillate (HD):

Here LD is lighter fraction (L) and HD is heavier fraction (H)

Given ASTM gap (t_{5H} --- t_{95L}) =10 for this from graph TBP overlap ($t_{100\,L}$--- t_{0H}) =58

Here we know $t_{100\,L}$=587^{0}F

587 – 58 =529^{0}F i.e. initial point of Heavy Distillate is 529^{0}F.

We now calculate cut point of Light Distillate as follows:

TBP cut point = (1/2)*($t_{100\,L}$ + t_{0H}) = (1/2)*(587 + 529) = 558^{0}F

From Figure 3.3. Cut volume percent = 19.2%.

For Heavy Distillate (HD) and Atmospheric gas oil (AGO):

Here HD is lighter fraction (L) and (AGO) is heavier fraction (H)

Given ASTM gap (t_{5H} --- t_{95L}) =5 for this from graph TBP overlap ($t_{100\,L}$--- t_{0H}) = 60

Here we know $t_{100\,L}$=684^{0}F

t_{0H} = 684 – 60 =624^{0}F i.e. initial point of Atmospheric gas oil is 6240F.

We now calculate cut point of Heavy Distillate as follows:

TBP cutpoint = (1/2)*($t_{100\,L}$ + t_{0H}) = (1/2)*(684 + 624) = 654^{0}F

From Figure 3.3. Cut volume percent = 10.4%.

For Atmospheric gas oil (AGO) and Residue crude (RC):

Here AGO is lighter fraction (L) and RC is heavier fraction (H)

Given ASTM gap (t_{5H} --- t_{95L}) =5 for this from graph TBP overlap ($t_{100\,L}$--- t_{0H}) = 60

Here we know $t_{100\,L}$=780^{0}F

t_{0H} =780– 60 =720^{0}F i.e. initial point of Residue crude is 720^{0}F.

We now calculate cut point of Heavy Distillate as follows:

TBP cutpoint = (1/2)*($t_{100\,L}$ + t_{0H}) = (1/2)*(780 + 720) = 750^{0}F

From Figure 3.3. Cut volume percent = 9.8%.

Table3.6. Estimated cut points and cut Vol% of the fractions:

Stream	Boiling range,^{0}F	TBP cut pt,^{0}F	Cut Vol%	Cumulative Vol%
Light Naphtha	0- 298	273	22	22
Heavy Naphtha	248-399	379	11.8	33.8
Light Distillate	359-587	558	19.2	53
Heavy Distillate	529-684	654	10.4	63.4
Gas oil	624-780	750	9.8	73.2

Based on the boiling ranges of the product as shown in Table 3.6 the TBP curve for each product is reproportioned to extend from 0 to 100 volume percent as shown in Table 3.7.

Table 3.7.Product TBP, ASTM and EFV temperatures

Stream	Vol%	TBP,^{0}F	ASTM,^{0}F	EFV,^{0}F
Light Naphtha	0	-65	-3	56
	10	-3	32	70
	30	41	56	83
	50	156	156	150
	70	206	192	168
	90	255	231	186
	100	298	271	204

Stream	Vol%	TBP,^{0}F	ASTM,^{0}F	EFV,^{0}F
Heavy Naphtha	0	248	283	313
	10	277	296	318
	30	304	309	325
	50	326	321	331
	70	345	333	337
	90	370	351	344
	100	399	377	351

Stream	Vol%	TBP,^{0}F	ASTM,^{0}F	EFV,^{0}F
Light Distillate	**0**	**359**	**398**	**447**
	10	395	415	453
	30	432	434	463
	50	467	455	473
	70	504	480	489
	90	542	508	501
	100	587	550	515

Stream	Vol%	TBP,^{0}F	ASTM,^{0}F	EFV,^{0}F
Heavy Distillate	0	529	542	596
	10	557	555	601
	30	587	570	608
	50	606	580	613
	70	628	594	619
	90	655	613	627
	100	684	639	634

Stream	Vol%	TBP,°F	ASTM,°F	EFV,°F
Gas oil	0	624	624	693
	10	650	636	697
	30	680	51	704
	50	700	662	709
	70	718	673	714
	90	745	692	723
	100	780	724	731

The average boiling point of the above streams is calculated by using the correlation given by API-TDB.

Example calculation for light naphtha:

This uses ASTM data of the fractions

$$\text{VABP} = \frac{T_{10}+T_{30}+T_{50}+T_{70}+T_{90}}{5} = \frac{32+56+156+192+231}{5} = 133.4$$

$$\text{SL} = \frac{T_{90}-T_{10}}{80} = \frac{231-32}{80} = 2.4875$$

$$T_{avg} = VABP - \Delta_4$$

$$\ln\Delta_4 = -0.94402 - 0.00865 * (VABP - 32)^{0.6667} + 2.99791 * SL^{0.333}$$

$$= 2.92862$$

$$\Delta_4 = \exp(2.92862) = 18.7$$

$$T_{avg} = VABP - \Delta_4 = 133.4 - 18.7 = 114.698 = 115°\text{F}$$

Table 3.8. Product Average boiling point and Specific gravity

Stream	Avg boiling point, ^{0}F	Specific gravity
Light naphtha	115	0.6786
Heavy Naphtha	319	0.79718
Light Distillate	453	0.85498
Heavy Distillate	580	0.87616
Gas oil	661	0.895

By knowing Average boiling point and specific gravity of the fractions we can estimate average molecular weight, average critical temperature, average critical pressure, and accentric factor by using following correlations and these properties are used to estimate enthalpy and moles of the fractions.

For Light Naphtha:

T_{avg}= 115^{0}F or 319.26^{0}K and SG = 0.6786

1) $M.W = 42.965[\exp(2.097 * 10^{-4} * T_{avg} - 7.78712 * SG + 2.08476 * 10^{-3} * T_{avg} * SG)] * T_{avg}^{1.26007} * SG^{4.98308}$

Where T_{avg} is in Kelvin.

$=42.965[\exp(2.097 * 10^{-4} * 319.26 - 7.78712 * 0.6786 + 2.08476 * 10^{-3} * 319.26 * 0.6786)] * 319.26^{1.26007} * 0.6786^{4.98308}$

=75.8

2) Pseudo critical temperature is calculated by using following correlation

$T_{pc} = 10.6443[\exp(-5.1747 * 10^{-4} * T_{avg} - 0.54444 * SG + 3.5995 * 10^{-4} * T_{avg} * SG)] * T_{avg}^{0.81067} * SG^{0.53691}$

Where $T_{avg} is\ in\ rankine.$

$=10.6443[\exp(-5.1747 * 10^{-4} * 577.67 - 0.54444 * 0.6786 + 3.5995 * 10^{-4} * 577.67 * 0.6786)] * 577.67^{0.81067} * 0.6786^{0.53691}$

=881.1451^{0}R

3) Pseudo critical pressure is calculated by using following correlation

$P_{pc} = 6.162 * 10^6[\exp(-4.725 * 10^{-3} * T_{avg} - 4.8014 * SG + 3.1939 * 10^{-3} * T_{avg} * SG] * T_{avg}^{-0.4844} * SG^{4.0846}$

Where T_{avg} *is in rankine and* P_{pc} *is in psia.*

$= 6.162 * 10^6[\exp(-4.725 * 10^{-3} * 577.67 - 4.8014 * 0.6786 + 3.1939 * 10^{-3} * 577.67 * 0.6786] * 577.67^{-0.4844} * 0.6786^{4.0846}$

=515.1308 Psia.

4) Accentric factor is calculated by using following correlation

For $T_R \leq 0.8$, the correlation given by Kesler and Lee is used.

$$\omega = \frac{-\ln P_c - 5.92714 + \frac{6.09648}{T_R} + 1.28862 \ln T_R - 0.169347 T_R^6}{15.2518 - \frac{15.6875}{T_R} - 13.4721 \ln T_R + 0.43577 T_R^6}$$

For $T_R > 0.8$, the correlation given by Edmister and Lee is used

$\omega = \frac{3}{7}\left[\frac{\log P_c}{T_R^{-1} - 1.0}\right] - 1.0$ Where P_c is in atmospheres

Since here $T_R = 0.652185$

$$\omega = \frac{-\ln 35.04291 - 5.92714 + \frac{6.09648}{0.652185} + 1.28862 \ln 0.652185 - 0.1693470.652185^6}{15.2518 - \frac{15.6875}{0.652185} - 13.4721 \ln 0.652185 + 0.435770.652185^6}$$

$= 0.232474.$

Table 3.9. Estimated product critical temperature, pressure, accentric factor and molecular weight.

Stream	T_c , °R	P_c , Psia	Accentric factor(ω)	Molecular weight
Light Naphtha	881.14	515.13	0.23474	76
Heavy Naphtha	1126.54	387.17	0.360369	132
Light Distillate	1274	319.53	0.453482	178
Heavy Distillate	1392	249.4062	0.574868	237
Gas oil	1468	219.3262	0.651864	281

3.4 Furnace Calculations:

The furnace outlet temperature is assumed to be 700°F. The furnace outlet pressure is maintained at 29.7 Psia .At this temperature we have to find how much crude is vaporized. To find out this first we have to convert TBP data to EFV curve for 29.7 Psia.

3.4.1 Conversion of TBP data to EFV curve at atmospheric:

First we have to convert TBP data to ASTM and then ASTM to EFV.

ASTMD86 and TBP conversion:

ASTMD86= $(1/a)^{1/b}\ (TBP)^{1/b}$

Table 3.10. Estimated ASTMD86 temperatures from TBP temperatures.

Vol (%)	a	b	TBP , ^{0}K	ASTM D86, ^{0}K
0	0.9177	1.0019	219.11	236.3
10	0.5564	1.0900	335.22	355
30	0.7617	1.0425	446.33	452
50	0.9013	1.0176	550.22	546.3
70	0.8821	1.0226	653.55	640.22
90	0.9552	1.0110	788.556	767
95	0.8177	1.0355	827.44	798.1

$EFV = a * (ASTMD86)^b * SG^c$ **Where a, b, c are constants**

The EFV values are at atmosphere to convert these values to furnace outlet pressure i.e. 29.7 Psia. The procedure is given below

Table 3.11. Estimated EFV temperatures from ASTMD86 temperatures

Vol%	a	b	c	ASTM D86,^{0}K	EFV,^{0}K
0	2.9747	0.8466	0.4209	236	282.8
10	1.4459	0.9511	0.1287	355	376.8
30	0.8506	1.0315	0.0817	452	459.6
50	3.268	0.8274	0.6214	546	540.68
70	8.2873	0.6871	0.934	640	598.7
90	10.6266	0.6529	1.1025	767	673.1
100	7.9952	0.6949	1.0737	827	709

For example the EFV value at 10 vol% and at atmospheric pressure is 376.898^{0}k, to convert this to 29.7 Psia

$$T = \frac{T_b'}{748.1 * Q - T_b' * (0.3816 * Q - 0.00051606)}$$

Where $T_b' = T_b$-1.3889*F*$(K_W - 12)\log_{10}(\frac{P}{760})$

$Q = \frac{6.412631 - 0.989679 * \log_{10} P}{2770.085 - 36\log_{10} P}$ *where* $P > 760\ mmHg$

F= 0 when ($T_b < 367K$ *or when* K_W *is not available*

F = -3.2985 + 0.009T_b

$K_W = \frac{[(1.8T_b)]\ ^((1/3))}{SG}$ Where T_b is normal boiling point in Kelvin

Here P = 1535.5 mm Hg

$$Q = \frac{6.412631 - 0.989679 * \log_{10} 1535.5}{2770.085 - 36\log_{10} 1535.5} = 0.001227$$

F = -3.2985 +0.009*376.898 = 0.093582

$K_W = \frac{[(1.8*376.898)]\ ^((1/3))}{0.843} = 10.422$

$T_b' = 376.8\text{-}1.3889*0.093582*(10.422 - 12)\log_{10}(\frac{1535.5}{760}) = 376.96$

$T = \frac{376.96}{748.1*0.001227 - 376.96*(0.3816*0.001227 - 0.00051606)} = 402.7\ K = 265.3^0$F

The above calculations are repeated for all Vol% at furnace outlet pressure and flash zone pressure and the results are listed below:

Table 3.12. Estimated EFV temperatures at elevated pressures

Vol %	EFV, ^{0}F @14.7Psia	EFV, ^{0}F @24.7Psia	EFV, ^{0}F @29.7Psia
0	49.37	76	86.8
10	218.5	252	265
30	367.6	406	421
50	513.5	555	572
70	618.1	661	679
90	751.9	796	814
100	816.53	860	879

From Fig 3.6. At 700 ^{0}F 72.5 Vol% crude will be vaporized and from Figure3.5. This volume percent is equivalent to 68.2 Wt%. Since the total crude has a specific gravity of 0.843, the density of the vapour and liquid portions is determined by

$(S.G)_V = (S.G)_{crude}$ (Wt frac/Vol frac) $_v = 0.843*(0.682/ 0.725) = 0.793$ or 46.9^0 API

$(S.G)_L = (S.G)_{crude}$ (Wt frac/Vol frac) $_L = 0.843*(1 - 0.682)/ (1 - 0.725) = 0.97481$ or 13.6^0 API.

Since crude charged is 55,000 Barrels per day (BPD)

F = 2291.67 bbl/hr = 675809 lb/hr

$V_{FO} = 0.682*675809 = 460901$lb/hr @ 46.9^0API

$L_{FO} = 675809 - 460901 = 214908$ lb/hr @13.5^0 API

To calculate total heat content of the crude at the furnace out let we need to calculate enthalpy of the vapour and liquid .this calculation as follows:

H_{FO} @700^0F and 46.9^0API =585.8145 Btu/lb

h_{FO} @700^0F and 13.6^0API =449.4791 Btu/lb

$Q_{FO} = V_{FO} H_{FO} + L_{FO} h_{FO}$

$= 460901*585.8145 + 214908*449.4791$

$= 366599143.3$ Btu/hr.

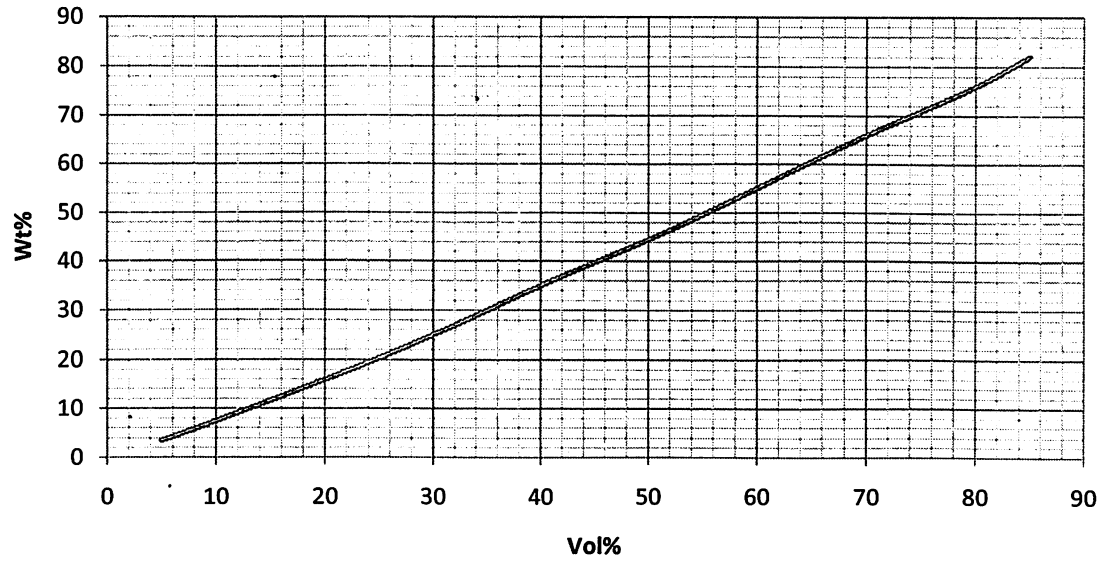

Figure3.5. Vol% vs. Wt% relationship

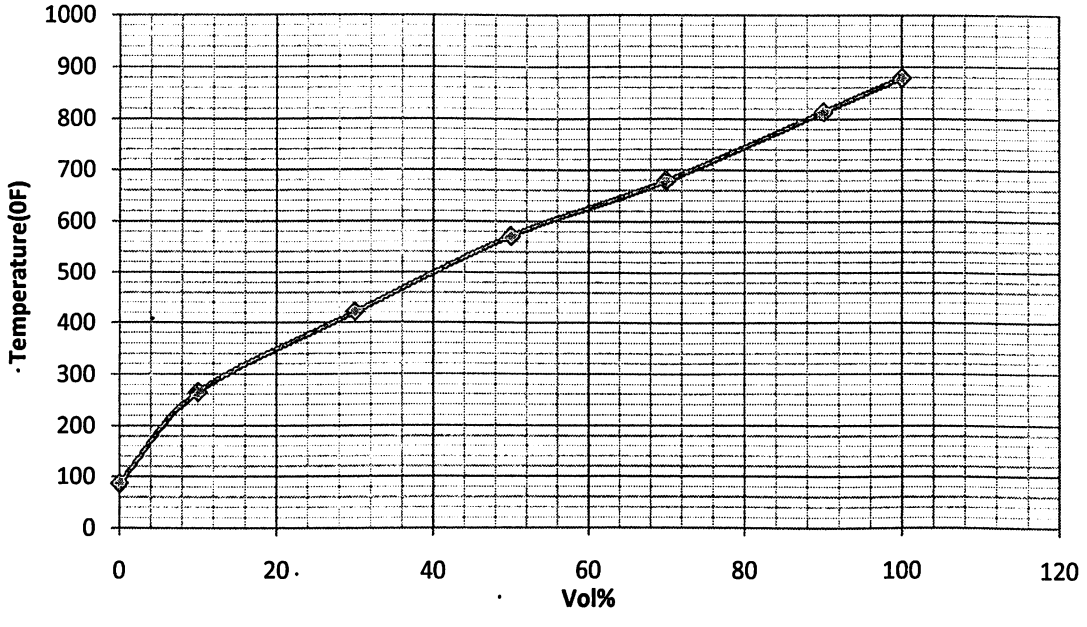

Fig 3.6. EFV temperature (^{0}F)@29.7 Psia vs. Vol%.

3.5 Flash Zone:

Assume the pressure in the flash zone will be 5 Psi less than that at the furnace outlet, i.e. 29.7 – 5.0 = 24.7 Psia. Draw an EFV curve for 24.7 Psia. This is shown in Figure 3.7.

Since pressure is decreased some more crude will be vaporized in the flash zone and the amount crude vaporized is determined by trial and error procedure which results in the same

total heat content as at the charge furnace outlet. This will occur normally at a temperature of 4-6^{0}F less than the furnace outlet temperature.

First estimate 4^{0}F less

i.e. $T_{(FZ)1} = 700 - 4 = 696^0F$

From Figure 3.7. At this temperature results in 75.0 Vol% which is equivalent to 70.9 Wt%

$(S.G)_v = (S.G)_{crude}$ (Wt frac/Vol frac)$_v = 0.843*(0.709/ 0.75) = 0.796916$ or 46.05^0 API

$(S.G)_L = (S.G)_{crude}$ (Wt frac/Vol frac)$_L = 0.843*(1 - 0.709)/ (1 - 0.75) = 0.981252$ or 12.7^0 API.

$V_{(FZ)1} = 0.709*675809 = 479148$ lb/hr @46.05^0 API

$L_{(FZ)1} = 675809 - 479148 = 196661$ lb/hr

$H_{(FZ)1}$@696^0F and 46.05^0API =581.9362 Btu/lb

$h_{(FZ)1}$@696^0F and 12.7^0API =444.7861 Btu/lb

therefore $Q_{(FZ)1} = V_{(FZ)1}H_{(FZ)1} + L_{(FZ)1}\ h_{(FZ)1}$

$= 479148*581.9362 + 196661*444.7861$

$= 366305684.9$Btu/hr.

Go for second estimate

$T_{(FZ)1} = 700 - 3 = 697^0F$

From Figure 3.7. At this temperature results in 75.8 Vol% which is equivalent to71.7 Wt%.

$(sp.gr.)_v = (sp.gr.)_{Crude}$ (Wt frac/Vol frac)$_v = 0.843*(0.717/ 0.758) = 0.79740$ or 45.95^0 API

$(sp.gr.)_L = (sp.gr.)_{Crude}$ (Wt frac/Vol frac)$_L = 0.843*(1 - 0.717)/ (1 - 0.758) = 0.98582$ or 12.03^0 API.

$V_{(FZ)1} = 0.717*675809 = 484555$lb/hr @$45.95^0$ API

$L_{(FZ)1} = 675809 - 484555 = 191254$ lb/hr

$H_{(FZ)1}$@697^0F and 45.95^0API =582.0909 Btu/lb

$h_{(FZ)1}$@697^0F and 12.03^0API =443.7955 Btu/lb

therefore $Q_{(FZ)1} = V_{(FZ)1}H_{(FZ)1} + L_{(FZ)1}\ h_{(FZ)1}$

$= 484555*582.0909 + 196661*443.7955$

$= 369332322.9$Btu/hr. This is to high

Therefore 696^0F is OK.

Thus $Q_{Fo} = Q_{FZ} = 366599143.3$ Btu/hr.

From Fig 3.9. The strip out from residuum in a 4th tray section is 1 Vol% of the amount feed when the stem rate is 10 lb/bbl of stripped liquid.

The total feed to the residue stripping section is the sum of the feed flash liquid and the over flash.

Over flash is 2%of the feed L_o = 0.02*2291.67 = 45.8334bbl/hr.

Feed flash liquid = 572.9175bbl/hr.

Therefore stripping section feed = 618.75bbl/hr.

Strip out = 0.01*618.75 = 6.1875bbl/hr.

Net bottoms = W = 618.75 - 6.1875 =612.5625 bbl/hr

Steam is introduced at a rate 10lb/bbl of residue liquid

Therefore steam to tower is = S_W=10*612.6 = 6126 lb/hr =340.3 mol/hr.

Total hydrocarbon vapour leaving flash zone is (F – W+ L_o) = (2291.67 -612.5625 +45.8334) =1724.94 bbl/hr. Which is 75.2 Vol% which is equivalent to 71.0 Wt%.

Thus, total hydrocarbon vapour from flash zone = 1724.94bbl/hr =480272lb/hr @46.3^0 API.

This 75.2 Vol% has average boiling point of 360^{0}F and specific gravity of 0.79591 from these values we can find average molecular weight and is equal to 146.8.

Total hydrocarbon vapour moles leaving flash zone is 3271.6 mol/hr.

Therefore vapour from flash zone = (hydrocarbon+ S_W) = 3271.6 +340.4 =3612 mol/hr

Therefore, the hydrocarbon partial pressure in the flash zone at a total pressure of 24.7 Psia is

P_{HC} = 24.7*(3271.6/3612) = 22.4 Psia.

Using the EFV curve for 14.7 Psia as shown in Figure 3.8. The point for 75.2 Vol % percent vaporized occurs at 658^{0}F. The temperature for same volume percent vaporization for 22.4 Psia is

Here P = 1158.095 mm Hg, T_b =620.9278K

$$Q = \frac{6.412631 - 0.989679 * \log_{10} 1158.095}{2770.085 - 36 \log_{10} 1158.095\ 8} = 0.0012709678$$

F = -3.2985 +0.009*620.927 = 2.289843

$$K_W = \frac{[(1.8*620.927)]\ \hat{}\ ((1/3))}{0.79591} = 13.039$$

$$T_b' = 620.927 - 1.3889*2.289843*(13.039 - 12) \log_{10}(\frac{1158.095}{760}) = 620.322\ ^\circ K$$

$$T = \frac{620.322}{748.1*0.00127096780 - 620.322*(0.3816*0.0012709678 - 0.00051606)} = 639.456\ ^\circ K = 692\ ^0F$$

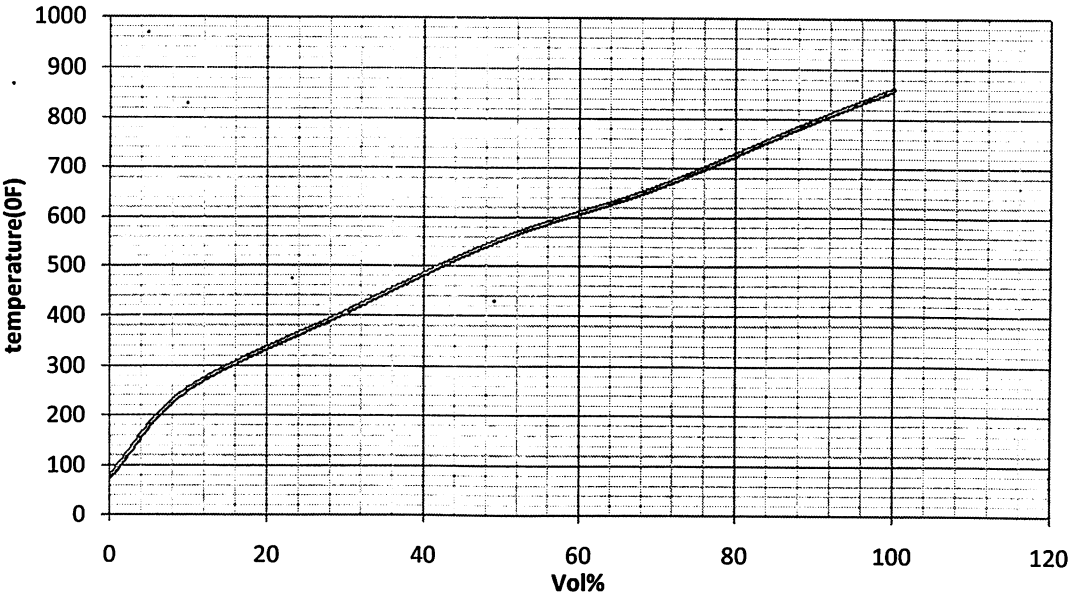

Fig 3.7. EFV temperature (^{0}F)@24.7 Psia vs. Vol%.

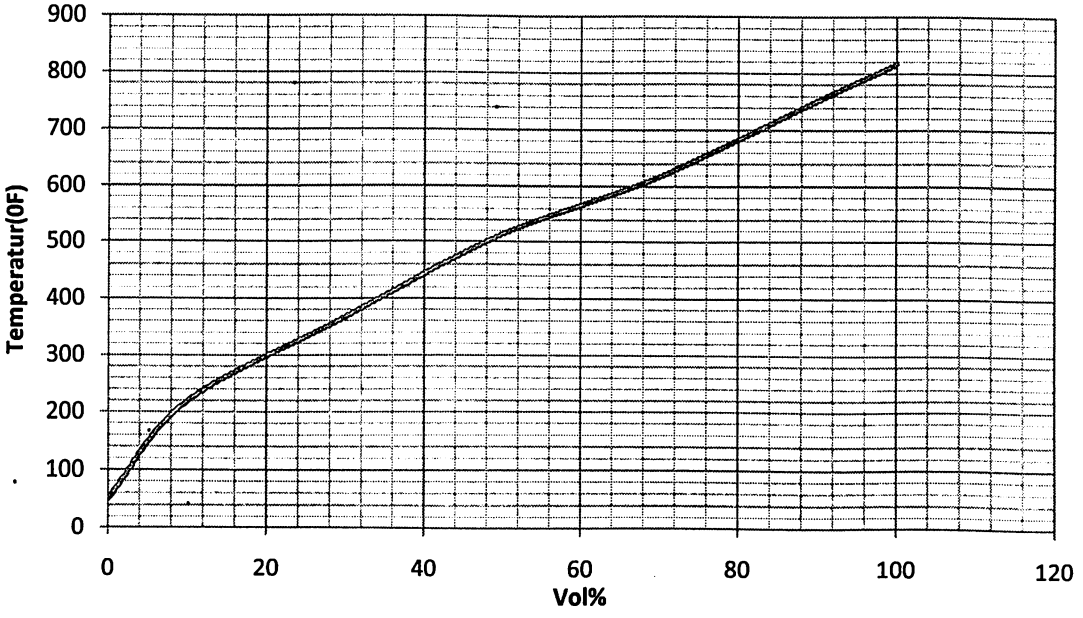

Fig 3.8. EFV temperature (^{0}F)@14.7 Psia vs. Vol%.

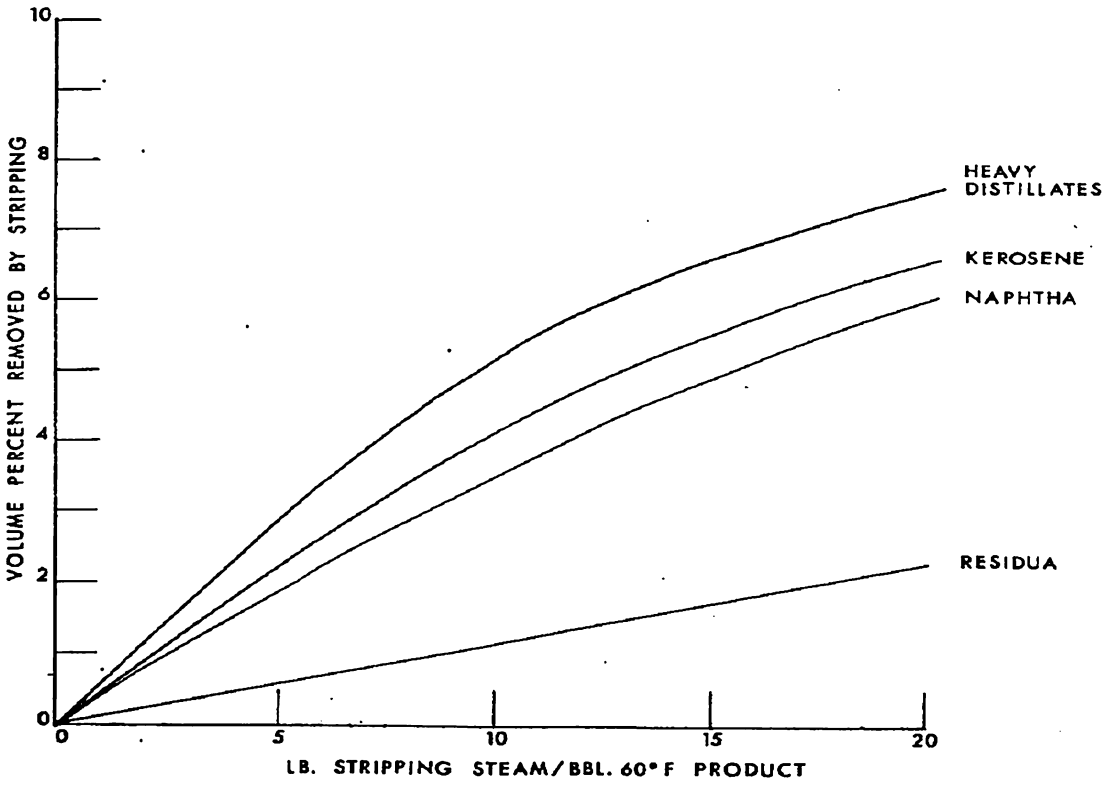

Fig.3.9. steam stripping for petroleum fractions

3.5.1 Flash zone conditions:

Flash zone temperature = 692 ^{0}F.

Flash zone pressure = 24.7 Psia.

Q_{Fz} = 366599143.3Btu/hr

Therefore sum of distillates =73.2% which is equivalent to 69.0 Wt%

Specific gravity.=0.843*(0.69/0.732) =0.79463 @46.5^{0}API

$\sum D = 466308\ lb/hr$

= 3261mol/hr.

Bottoms residue W = 675809 – 466308 =209501lb/hr

Assume bottoms temperature is 30^{0}F lower than that of the flash zone.

T_W = 692 – 30 = 662^{0}F

Bottoms specific gravity is equal to 0.97511

h_W @ 662^{0}F & 13.6^0 API =422.3539Btu/lb.

Q_W = 209501*422.3539 = 88483564.4Btu/hr.

With stripping steam available at50 Psig saturated (65 Psia & 298 ^{0}F), the heat added is

H_{SW} = 1179 Btu/lb

Q_{SW} = 6126*.1179 =7222554 Btu/hr.

3.6 Separation of Light Naphtha and Gas

Since the over head accumulator is maintained at a temperature100^{0}F and pressure 15.2 Psia, some of the light ends not condensable at these conditions. So we have both vapour and liquid product. To know the vapour and liquid flow rates we made flash calculations and these calculations are shown in Table 3.15.

Since the vapour portion is water saturated. For the conditions of 100^{0}F and 15.2 Psia the vapour pressure of water is calculated as follows:

$\ln P^{sat} = A - \frac{B}{T + C}$ $where P^{sat}$ is in mmHg and T is in °K.

For water A=18.30, B = 3816.44, and C = -46.13

$$\ln P^{sat} = 18.30 - \frac{3816.44}{310.9 - 46.13} = 3.88734$$

$$\Rightarrow P^{sat} = \exp(3.88734) = 48.781 \text{ mmHg}$$

So the hydrocarbon partial pressure is equal to 15.2 – 0.95 = 14.25 Psia.

So at the conditions of P = 14.25Psia and T = 310.9K perform flash calculations

3.6.1 Flash calculations:

$$x_i = \frac{x_{Fi}}{\frac{v}{f} * (k_i - 1) + 1}$$

Where x_{Fi} is composition of ' i' component in feed

And K_i is equilibrium ratio = y_i/x_i

Table 3.13. Composition of Light Naphtha and Gas

Component	$T_{avg,}$ ^{0}F	Vol %	m^3/hr	Density, kg/m^3	MW	M, lb/hr	moles
c1	-259	0.01	0.03636	299.563	16	24.01521	1.50095
c2	-128	0.07	0.254543	376.73	30	211.41	7.04702
c3	-44	0.92	3.345423	505.663	44	3729.465	84.7606
ic4	11	1.44	5.236315	561.5022	58	6482.031	111.759
nc4	31	2.26	8.218106	582.95	58	10561.79	182.099
ic5	82	1.9	6.909	622.972	72	9488.984	131.791
nc5	97	1.8	6.54539	629.4433	72	9082.935	126.152
pseudo compl	130	1.6	5.818128	525.7942	66	6744.238	102.185
pseudo comp2	160	2.5	9.09082	706.4911	87	14159.37	162.751
pseudo comp3	189	2.5	9.09082	639.388	93	12814.51	137.791
pseudo comp4	216	2.5	9.09082	775.27096	100	15537.84	155.378
pseudo comp5	243	2.5	9.09082	672.939	109	13486.94	123.733
pseudo comp6	277	2	7.27266	841.174	113	13486.94	119.353
Total		22	79.99921			115810.5	1446.3

Condition is $\sum y_i - \sum x_i = 1 * 10^{-3}$

Up to pentane we calculate $k_i = \frac{P^{Sat}}{P}$ where P^{Sat} is vapour pressure and is calculated using Antoine equation, $\ln P^{sat} = A - \frac{B}{T+C}$ where P^{sat} is in mmHg and T is in °K. And P is total pressure in the system, here P = 14.25 psia and T=310.9 °K. And for the pseudo components using average boiling points and specific gravities we find average molecular weight, pseudo critical temperature, pseudo critical pressure and accentric factor using correlations which are given.

$$E_J = -\frac{0.545542}{(T_R)_J^3} + \frac{3.1664}{(T_R)_J^2} - \frac{14.9343}{(T_R)_J} + 19.3003 - 7.32633(T_R)_J - 0.561196(T_R)_J^2 + 0.663232(T_R)_J^3$$

The equilibrium constant K_J is given by:

$$if\ -30.0 < E_J(1+\omega_j) < 17.0, \qquad K_J = \frac{(P_C)_J}{P} e^{E(1+\omega_J)}$$

$$if\ E_J(1+\omega_j) \geq 17.0, \qquad K_J = \frac{(P_C)_J}{P} e^{17.0}$$

$$if\ E_J(1+\omega_j) \leq -30.0, \qquad K_J = \frac{(P_C)_J}{P} e^{-30.0}$$

Table 3.14. Antoine constants data

Component	A	B	C
c1	15.2243	897.84	-7.16
c2	15.6637	1511.42	-17.16
c3	15.726	1872.46	-25.16
ic4	15.5381	2032.73	-33.15
nc4	15.6782	2154.9	-34.42
ic5	15.7339	2396.369	-37.705
nc5	15.8333	2477.07	-39.94

Table 3.15. Vapour- liquid equilibrium calculations

Component	X_{fi}	K_i	v/f	x_i	y_i	y_i-x_i
c1	0.0010378	289	0.2585	$1.4*10^{-5}$	0.00397	0.00136
c2	0.0048724	50.3		0.00035	0.01783	
c3	0.058605	13.1		0.0142	0.18598	
ic4	0.0772723	5.0528		0.03774	0.190677	
nc4	0.1259071	3.61		0.07518	0.271409	
ic5	0.091123	1.43798		0.08186	0.1177	
nc5	0.0872237	1.097		0.08509	0.093343	
pseudo compl	0.0706528	0.5702		0.07948	0.045321	
pseudo comp2	0.1125292	0.26353		0.13899	0.0366279	
pseudo comp3	0.0952708	0.154762		0.12191	0.0188665	
pseudo comp4	0.1074314	0.07264		0.14131	0.0102644	
pseudo comp5	0.0855515	0.045552		0.11357	0.005173463	
pseudo comp6	0.0825231	0.016427		0.11066	0.001817786	
Total	1.0000001			1.00034	0.998980049	

Table 3.15. Continued------- Vapour liquid Equilibrium calculations

Component	vapor(mol/hr)	liquid(mol/hr)	vapor(lb/hr)	Liquid(lb/hr)
c1	1.48619	0.0147512	23.779	0.23602
c2	6.666831	0.3801911	200.0049	11.40574
c3	69.53474	15.2258	3059.529	669.936
ic4	71.28863	40.4705	4134.741	2347.29
nc4	101.4715	80.628	5885.348	4676.443
ic5	44.00687	87.7845	3168.494	6320.49
nc5	34.8984	91.25348	2512.684	6570.251
pseudo comp1	16.94437	85.241	1118.328	5625.909
pseudo comp2	13.69407	149.0572	1191.384	12967.98
pseudo comp3	7.053615	130.7368557	655.986	12158.53
pseudo comp4	3.837557	151.54082	383.7557	15154.08
pseudo comp5	1.934199	121.79919	210.8277	13276.11
pseudo comp6	0.679614	118.67383	76.7964	13410.14
Total	373.496586	1072.8066	22621.66	93188.81

From the above calculations V_5 =374 mol/hr,then the total vapour including water is calculated as $P_{HC} = P_t * \frac{V_5}{V_5+V_{water}} \Rightarrow V = V_5 + V_{water} = \frac{15.2}{14.25} * 374 = 399$ mol/hr

Therefore $V_{water} = 399 - 374 = 25$ mol/hr.

Table3.16. Product flow summary

Stream	Mass flow (lb/hr)
Gas, V_5	22621
Light Naphtha, D_5	93189
Heavy Naphtha, D_4	75411
Light Distillate, D_3	131600
Heavy Distillate, D_2	73048
Atmospheric Gas oil, D_2	70314
reduced crude, W	209501

3.7 Draw tray temperatures:

By knowing product EFV temperature we assume draw tray temperatures. The product EFV temperature curves are extended to minus Ten percent. Since the products will be stripped by removing 10 percent of the front ends, this extension is made to estimate the lighter properties existing on the trays. These curves are shown in Fig 3.10. There is a graphical relation between product EFV and draw tray temperature and is shown in Fig 3.11.

From graph we assume draw tray temperatures and is listed below.

Table 3.17. Assumed draw temperatures

Draw tray	Assumed temperature, ^{0}F
Atmospheric gas oil(D_1)	590
Heavy Distillate (D_2)	525
Light Distillate(D_3)	400
Heavy Naphtha(D_4)	316
Light Naphtha(D_5)	215

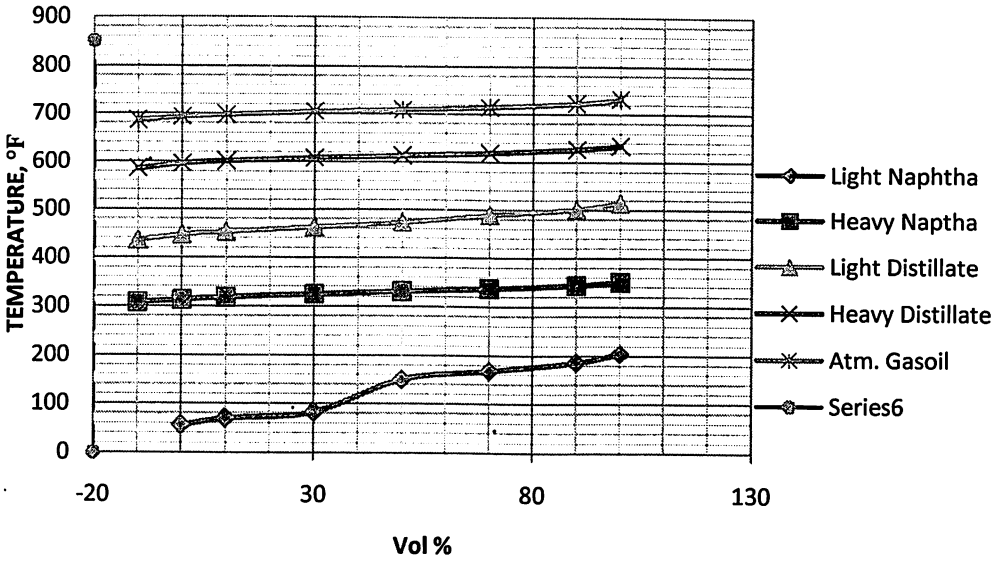

Figure 3.10.Extended EFV curves of the products.

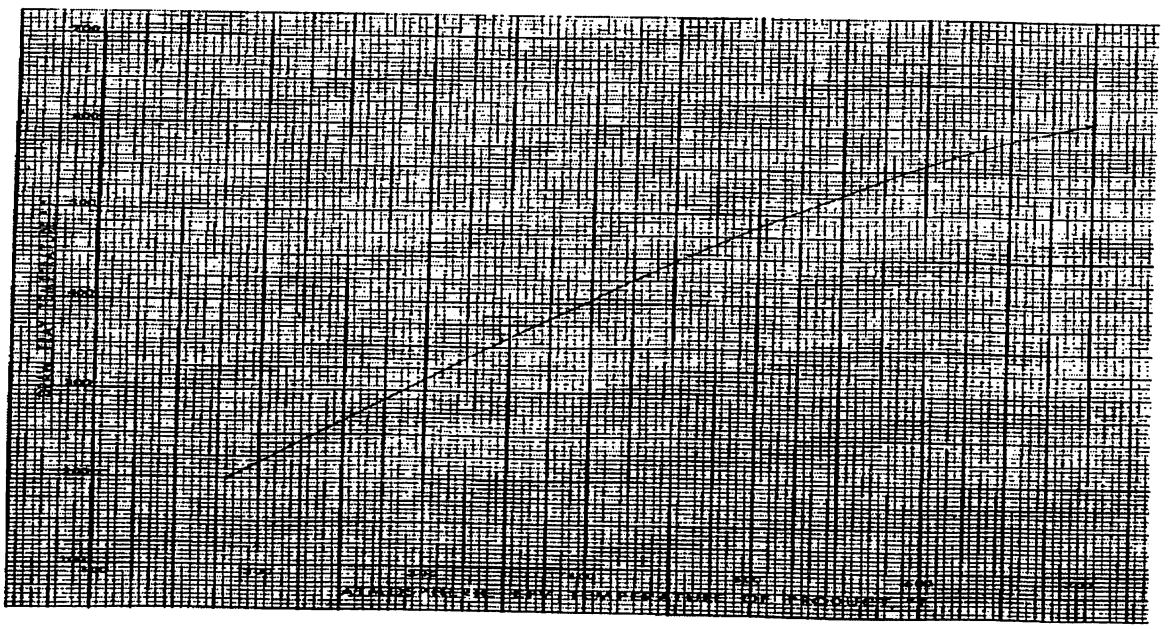

Figure3.11. Draw temperature vs. Atmospheric EFV temperature of the product.

3.8 Heat balance to first Draw tray:

The draw tray is used to with draw gas oil D_1.

The no of trays between flash zone and first draw is 3.

The flash zone temperature is 692 ^{0}F and first side draw temperature is 590 ^{0}F.

Assume linear temperature drop between flash zone and first side draw.

Make a Heat balance about the bottom of the column which includes the tray below the draw tray and cooled reflux is at 400 ^{0}F.

First side draw tray no is 8.

Since three trays are there $\Delta T = \frac{692-590}{3} = 34$ °F

Assume a pressure drop of 0.15 Psia per tray.

Therefore the tray 7 conditions are temperature = 624 ^{0}F.

Pressure =24.4 Psia.

Heat in = Heat out

Heat in to the tray is $Q_{FZ} + Q_{SW} + Q_{L,\,Rc7}$.

Heat out to the tray is $Q_{\sum D} + Q_{V,\,Rc7} + Q_{S7} + Q_W$

Since we know the properties like T_c, P_c, ω which are calculated above and these are used to calculate enthalpy of liquid.

h_{Rc7} @400^{0}F and 26.6 ^{0}API = 269.0556 Btu/lb.

H_{Rc7} @ 624^{0}F and 26.6 ^{0}API = 505.2658 Btu/lb.

$H_{\sum D}$ @624^{0}F and 46.57^{0}API = 533.9957 Btu/lb.

H_{S7} @ 624^{0}F and 24.4 Psia =1347 Btu/lb.

$Q_{FZ} + Q_{SW} + Q_{L,\,Rc7} = Q_{\sum D} + Q_{V,Rc7} + Q_{S7} + Q_W$

366599143.3 + 7222554 + R_{c7}*269.0556 = 466308*533.9957+ R_{c7}*505.2658 +8251722 +

88483564.4 +505.2658 R_{c7} = 118877 lb/hr.

Make second heat balance around the bottom of the column which includes the draw tray (no.8).

To establish flow rates, make a hydrocarbon material balance around the stripper assuming V_{L8} is ten percent of L_8.

$L_8 = R_{c7} + V_{L8} + D_1$

$= 118877 + 0.1 * L_8 + 70314$

L_8 = 210212 lb/hr.

Then the heat balance becomes

Heat in = Heat out

$Q_{FZ} + Q_{SW} + Q_{L,\,Rc7} + Q_{L9} + Q_{V,\,L8} = Q_{(\sum D - D1)} + Q_{L8} + Q_{V,\,L9} + Q_{S8} + Q_W$

h_{L8} @590^{0}F and 26.6^{0}API = 394.5487 Btu/lb

h_{L9} @579^{0}F and 26.6^{0}API = 386.8689 Btu/lb

H_{L9} @590^{0}F and 26.6^{0}API = 484.1689 Btu/lb

$H_{(\sum D - D1)}$ @5900F and 50.13^{0}API = 520.2082Btu/lb

H_{S8}@590^{0}F and 24.15 Psia = 1331 Btu/lb.

366599143.3 + 7222554 +31984522.56 +L_9*386.8689 +21021.2*386.8689 = 395994*520.2082+ 210212*394.5487 +L_9*484.1689 +6126*1331+88483564.4

L_9=291502 lb/hr

=1037 mol/hr

In calculating partial pressure we neglect the moles of D_2.

Total moles = 3314 - 250 – 308 = 2756

Therefore partial pressure of hydrocarbon $P_{HC} = \frac{1037}{1037+2756+341} * 24.2$ =6.07 Psia =313 mmHg.

Using Fig3.8. $T_{0\%}$ EFV temperature at atmospheric pressure is 686^{0}F i.e. bubble point temperature of the unstripped atmospheric gas oil.

This temperature at this partial pressure is 608^{0}F.

3.8.1 Stripping steam:

Steam is used at a rate of 10 pounds per barrel of total stripper bottoms which is

$R_{c7} + D_1$ = 118877 + 70314 = 189191 lb/hr @26.6^0 API or 604.27 bbl/hr.

S_1 = 10*(604.27) =6042.7 lb/hr or 336 mol/hr.

Q_{S1} = 6042.7*1179 = 7124343.3 Btu/hr.

Then the stripper bottoms cooler duty is

Q_{C1} =189191*(373.7168 – 269.0556) = 19800957Btu/hr.

Product heat content is

Q_{D1} =70314*(269.0556) =18918375 Btu/hr.

3.8.2 Tray 7 conditions.

The induced reflux is

$R_{i7} = R_{c7} * \left[\frac{(h_{624} - h_{400})}{(H_{658} - h_{624})}\right] = 118877 * \left[\frac{(418.5832 - 269.0556)}{(525.9133 - 418.5832)}\right]$ =165614 lb/hr.

Therefore liquid loading is

$R_{\sum 7} = R_{c7} + R_{i7}$ = 118877 +165614 = 284491 lb/hr @26.6^0 API.

Vapour loading is

$\sum V = V_{\sum D} + .R_{\sum 7} + S_W$

= 466308 +284491 +6126 =756925 lb/hr

= 3261 +1012 +341 = 4614 mol/hr.

The above calculations are repeated for remaining side draws and these all side streams are reboiled stripped.

Table 3.18. Estimated Draw tray temperatures

Stream unstripped	Temperature, ^{0}F		Pressure, mmHg		Estimated draw temp,°F
	0% EFV	100%EFV	P	P_{HC}	
Light naphtha	56	204	1044	898	213
Heavy Naphtha	308	351	1098	817	318
Light Distillate	436	515	1163	613	416
Heavy Distillate	585	634	1204	393	528
Gas oil	686	731	1253	312	608

Table 3.19. Tray Vapour and Liquid loadings

Tray number	Vapor loading(mol/hr)	Liquid loading(mol/hr)
7	4614	1012
13	5719	1990
19	6071	2639
27	5627	2935
35	6654	4532

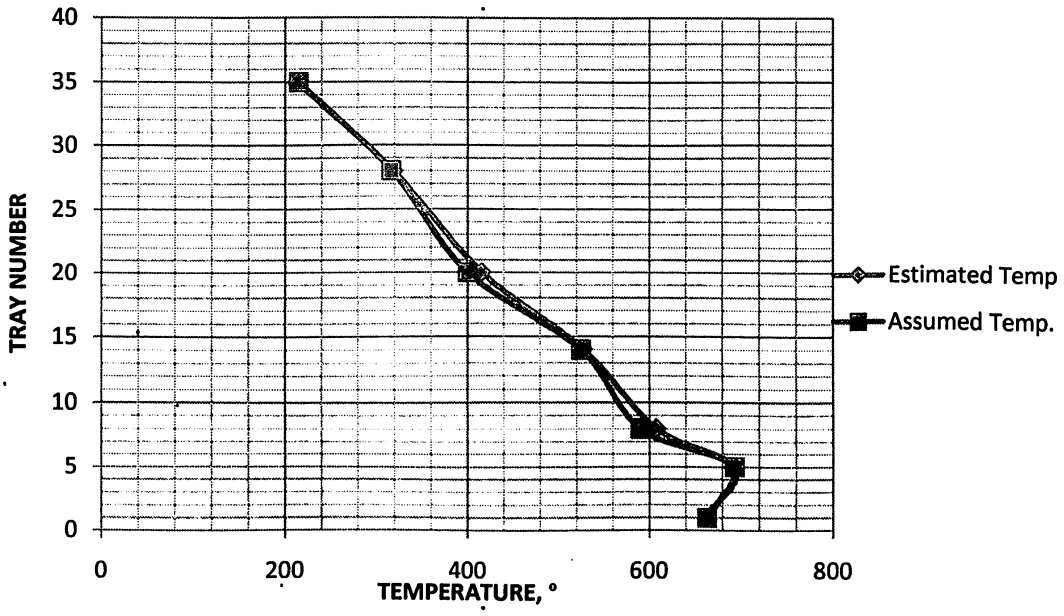

Fig 3.12. Temperature profile in the column.

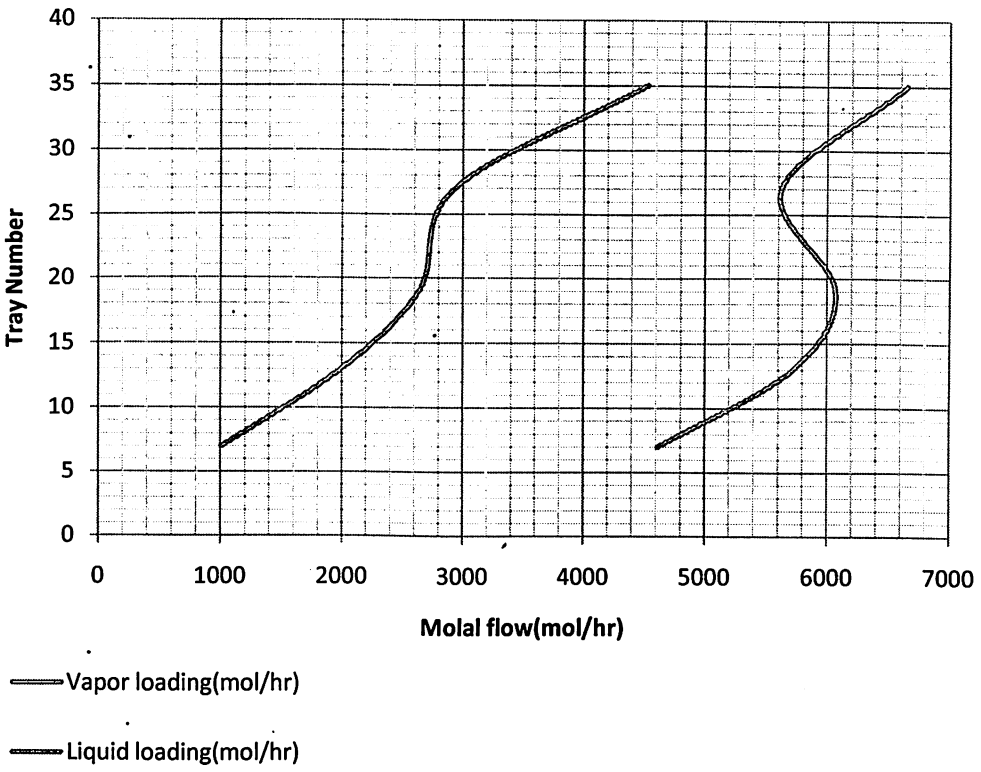

Fig. 3.13. Vapour and Liquid traffic in the column.

4.0 RESULTS AND DISCUSSION:

The design is based on special report on how to design a crude distillation by R.N. Watkins Sugarland, Texas published in 1969, where he followed graphical methods in the design and here we followed analytical methods and there is a slight variation of results . The results are listed below.

By Watkins method the flash zone temperature is 693°F.

By Analytical method the flash zone temperature is 692°F.

The products flow summary as follows

Stream	Mass flow(lb/hr)	
	Watkins method	Analytical method
Gas, V_5	22300	22621
Light Naphtha, D_5	93700	93189
Heavy Naphtha, D_4	75400	75411
LightDistillate, D_3	131600	131600
Heavy Distillate, D_2	73000	73048
Atmospheric gas oil, D_1	70300	70314
Reduced crude, W	209500	209501

The estimated draw temperatures as follows

Stream unstripped	Pressure, mmHg				Estimated draw temp,°F	
	Watkins method		Analytical method		Watkins method	Analytical method
	P	P_{HC}	P	P_{HC}		
Light naphtha	1044	915	1044	898	215	213
Heavy Naphtha	1098	850	1098	817	316	318
Light Distillate	1163	570	1163	613	400	416
Heavy Distillate	1204	340	1204	393	525	528
Gas oil	1253	240	1253	312	590	608

The estimated temperatures from Watkins and analytical methods have some little deviation. In Watkins method he used graphical methods to calculate enthalpy of the fractions which are used in heat balance calculations but in analytical methods due to little error in enthalpy of fractions we have some little deviation in estimated temperatures

5.0 Conclusion:

In Watkins method he used graphical method and due to some error in analytical correlations we have a some little variation between these methods. This is tentative design, it can be further checked in simulation by using simulation packages like ASPEN-HYSIS

REFERNCES:

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