الخلاصة

التقطير الجوي للنفط الخام

التقطير الجوي للنفط الخام يتضمن وحدات المعالجة الاولى والاساسية في المصفى. تستخدم هذه الوحدات لفصل النفط الخام بعملية التقطير الى مشتقاته، تبعا لاختلاف درجات الغليان تحت ضغط جوي. بصورة عامة يتضمن الفصل الاول مقدمة عن اهم الخواص المميزة للنفط الخام و معلومات حول مكوناته ومنتجاته. الفصل الثاني يحتوي وصف للعملية. الفصل الثالث يتناول توازن المادة و ايجاد كميات المنتجات. الفصل الرابع يتناول توازن الطاقة وايجاد درجات حرارة المنتجات. الفصل الخامس يتناول تصميم معدة مستخدمة في العملية.

جامعة ديالى

كلية الهندسة

قسم الهندسة الكيمياوية

" تقطير النفط الخام"

مشروع مقدم إلى كلية الهندسة / جامعة ديالى – كجزء من متطلبات نيل درجة البكالوريوس في الهندسة الكيمياوية

أعداد الطالبات:

إيلاف سامي صادق

غفران مظهر فرهود

اسماء زاهد كامل

بأشراف:

الاستاذ علي علوان

ه/ رجب ١٤٣٧ م/أيار ٢٠١٦

MINISTRY OF HIGHER EDUCATION
AND SCIENTIFIC RESEARCH
DIYALA UNIVERSITY
COLLEGE OF ENGINEERING
CHEMICAL ENGINEERING
DEPARTMENT

Atmospheric Distillation of Crude Oil

A Project Submitted To The Department Of Chemical Engineering /College Of Engineering In Partial Fulfilment Of The Requirement For The Degree Of B.SC. In The Chemical Engineering.

By
Elaff Sami
Ghofran modher
Asmaa Zahid

SUP:

Ali Alwan

Rajab 1432

May 2010

بسم الله الرحمن الرحيم

﴿يُؤْتِي الْحِكْمَةَ مَن يَشَاءُ وَمَن يُؤْتَ الْحِكْمَةَ فَقَدْ أُوتِيَ خَيْراً كَثِيراً وَمَا يَذَّكَّرُ إِلَّا أُوْلُواْ الأَلْبَابِ﴾

(١٢٦٩ البقرة)

الاهداء

إلى كل من أضاء بعلمه عقل غيره

أو هدى بالجواب الصحيح حيرة سائليه

واظهر بسماحته تواضع العلماء

وبرحابته سماحة العارفين ,,,

إلى من علمني النجاح والصبر .. أبي

والى من تتسابق الكلمات لتخرج معبرة عن مكنون ذاتها

من علمتني وعانت الصعاب لأصل إلى ما أنا فيه ..أمي

ثم إلى كل من علمني حرفا أصبح سنا برقه يضيء الطريق أمامي ..

إلى الذين بذلوا كل جهد وعطاء لكي أصل إلى هذه اللحظة أساتذتي الكرام ولاسيما
الأستاذ الفاضل علي علوان إليكم اهدي هذا العمل..

الحمد لله رب العالمين والصلاة والسلام على سيد المرسلين محمد ﷺ

الشكر والتنزيه لله العلي القدير الذي منّ علينا بنعمة الأيمان والعلم

ويشرفنا التقدم بالشكر والامتنان الجزيل إلى الأستاذ "علي علوان" لتفضله بالأشراف على هذا البحث وتوجيهاته القيمة في كل خطوة نخطوها وكل درب نسلكه.

كما نتقدم بالشكر لرئيس قسم الهندسة الكيمياوية د.انيس وجميع أساتذتنا الافاضل على كل شي. والى كل من شد ازرنا وساندنا وساعدنا .

الباحثين

إيلاف سامي

غفران مظهر

أسماء زاهد

SUPERVISORS CERTIFICATION

We certify that this project entitled "**distilation of crude oil**", was prepared under our supervision at Chemical engineering department/ college of engineering by (**Ghofran modher farhod & Elaff Sami sadek & Asmaa zahid kamil**) as a partial fulfilment of requirements for the degree of B.Sc. in Chemical engineering**.**

Signature

Name: Ali alwan

Date: 10 / 5 / 2016

In view of the available recommendations, I forward this project for debate by the examining committee.

Signature:

Name : Anees Abdullah

(Head of the department)

Date : 10 / 5 / 2016

Abstract:

Atmospheric distillation of crude oil includes the first major processing units in the refinery. They are used to separate the crude oils by distillation into its fractions according to boiling point under atmospheric pressure. In chapter one an introduction on crude oil most important characteristics and information about its contents and products. Chapter two contains the description of the process. Chapter three includes the material balance and finding the amount of the products. In chapter four the energy balance and find the temperature of each product. In chapter five the design of selected equipment.

Nomenclature

API	American petroleum Institute
B.P	Boiling point
CN	Cetane number
cp	Specific heat capacity
CW	cooling water
D	Tower diameter
D_b	Bundle diameter
D_s	Shell diameter
EFV	Equilibrium flash vaporization
K	Characterization Factors
λ	Latent heat
L	Tube length
L_B	Baffle spacing
lbmol	Pound mole
M	Mass flowrate
M.wt	Molecular weight
μ	Viscosity
ON	octane number
ρ_{water}	water density

P	pressure
P	percent removed by stripping
P.P	Partial pressure
p_t	Pitch distance
Q	volumeric flowrate
Q	Heat transferred
S	pound steam per hour
Sp.gr	Specific gravity
Ta	temperature of air
T_B	molar average boiling point , ^oR
TBP	True poiling point
T_f	Feed plate temperature
T_p	Plate temperature
T_s	Steam teperature
Δt	Feed less bottom temperature
V	volume
W	pound per hour

Contents:

Chapter One : Introduction

Chapter Two : The Method Of Processing

Chapter Three : Material Balance

Chapter Four : Energy Balance

Chapter Five : Equipment Energy

1.1: Introduction:

Petroleum, in one form or another, has been used since ancient times, and is now important across society, including in economy, politics and technology. The rise in importance was mostly due to the invention of the internal combustion engine, the rise in commercial aviation and the increasing use of plastic, besides that and the more important that it is a non-renewable source of energy. Crude oil, once it has been extracted from underneath of the bedrock, it has to be processed to be useful. Crude petroleum as it produced from the field is a relatively low value material as it is compared with its products. Since, in its native state, it is rarely usable directly. However it consists of a large number of compounds, it can be refined and further processed into many number of products whose value is many times that of the original oil.[4]

Crude oil comes from different parts of the world and has different Physical and chemical characteristics. The crude oil specification properties shape the range of products that are produced by the processes.[1]

It is a naturally occurring, complex liquid mixture made up of a vast number of hydrocarbon compounds that consist mainly of carbon and hydrogen in differing proportions. In addition, small amounts of organic compounds containing sulfur, oxygen, nitrogen and metals such as vanadium, nickel, iron and copper are also present. Hydrogen to carbon ratios affect the physical properties of crude oil. As the hydrogen to carbon ratio decreases, the gravity and boiling point of the hydrocarbon compounds increases. Moreover, the higher the hydrogen to carbon ratio of the feedstock, the higher its value is to a refinery, because less hydrogen is required.[1]

1.2: Composition Of The Crude Oil:

The composition of crude oil, on an elemental basis, falls within certain ranges regardless of its origin. Table 1.1 shows that carbon and hydrogen contents vary within narrow ranges. For this reason, crude oil is not classified on the basis of carbon content. Despite their low concentrations, impurities such as sulfur, nitrogen, oxygen and metals are undesirable because they cause corrosion and catalyst poisoning during refining process.[1]

Table 1-1 Elemental composition of crude oils.

Element	**Composition (wt%)**
Carbon	83.0–87.0
Hydrogen	10.0–14.0
Sulfur	0.05–6.0
Nitrogen	0.1–0.2
Oxygen	0.05–2.0
Ni	<120 ppm
V	<1200 ppm

The crude oil compounds are classified to the following:

1.2.1: Hydrocarbons:

1.2.1.1: Paraffins:

The paraffins series of hydrocarbons is characterized by the rule that the carbon atoms are connected by a single bond and the other bonds are saturated with hydrogen atoms. The general formula for paraffins is C_2H_{2n+2}.

The simplest paraffin is methane, CH_4, followed by the series of ethane, propane, normal and isobutane, normal, iso-, and neopentane,…etc. When the number of carbon atoms in the molecule is greater than three, several hydrocarbons may exist which contain the same number of carbon and hydrogen atoms but have different structures. This is because carbon is capable not only of chain formation, but also of forming single or double branched chains which give rise to isomers that have significantly different properties. For example, the motor octane number of n-octane is -17 and that of isooctane (2,2,4-trimethyl pentane) is 100.

The number of possible isomers increases in geometric progression as the number of carbon atoms increases. There are two paraffin isomers for butane, three for pentane and, by the time the number of carbon atoms has increased to eight, the number of isomers raise to 17 structural isomers. Crude oil contains molecules with up to 70 carbon atoms, and the number of possible paraffinic hydrocarbons is very high.[2]

1.2.1.2: Naphthenes (Cycloparaffines):

Naphthenes, also known as cycloalkanes, are saturated hydrocarbons that have at least one ring of carbon atoms. They have the general formula C_nH_{2n}. A common example is cyclohexane (C_6H_{12}).

The boiling point and densities of naphthenes are higher than those of alkanes having the same number of carbon atoms. Naphthenes commonly present in crude oil are rings with five or six carbon atoms. These rings usually have alkyl substituents attached to them. Mutli-ring naphthenes are present in the heavier parts of the crude oil.[1]

1.2.1.3: Aromatics:

Aromatics are unsaturated cyclic compounds composed of one or more benzene rings. The benzene ring has three double bonds with unique electron arrangements that make it quite stable.

Crude oils from various origins contain different types of aromatic compounds in different concentrations. Light petroleum fractions contain mono-aromatics, which have one benzene ring with one or more of the hydrogen atoms substituted by another atom or alkyl groups. Examples of these compounds are toluene and xylene. Together with benzene, such compounds are important petrochemical feedstocks, and their presence in gasoline increases the octane number.

More complex aromatic compounds consist of a number of ‘‘fused’’ benzene rings. These are known as "polynuclear aromatic compounds". They are found in the heavy petroleum cuts, and their presence is undesirable because they cause catalyst deactivation and coke deposition during processing, besides causing environmental problems when they are present in diesel and fuel oils. The heaviest portion of the crude oil contains asphaltenes, which are condensed polynuclear aromatic compounds of complex structure.[1]

1.2.2: Non-Hydrocarbons:

1.2.2.1: Sulfur Compounds:

The Sulfur content of crude oils varies from less than 0.05 to more than 10 wt% but generally falls in the range 1–4 wt%. Crude oil with less than 1 wt % sulfur is referred to as low sulfur or sweet, and that with more than 1 wt% sulfur is referred to as high sulfur or sour.

Crude oils contain sulfur heteroatoms in the form of elemental sulfur S, dissolved hydrogen sulfide H_2S, carbonyl sulfide COS, inorganic forms and most importantly organic forms, in which sulfur atoms are positioned within the organic hydrocarbon molecules and these compounds are mercaptans , sulfides and thiophenes.

Sulfur containing constituents of crude oils vary from simple mercaptans, also known as thiols, to sulfides and polycyclic sulfides. Mercaptans are made of an alkyl chain with –SH group at the end (R–SH). In sulfides and disulfides, the sulfur atom replaces one or two carbon atoms in the chain (R–S–R') or (R–S–S–R'). These compounds are often present in light fractions. Sulfides and disulfides may also be cyclic or aromatic. Thiophenes are polynuclear aromatic compounds in which the sulfur atom replaces one or more carbon atoms in the aromatic ring. They are normally present in heavier fractions.[1]

1.2.2.2: Oxygen Compounds:

The oxygen content of crude oil is usually less than 2 wt%. A remarkable high oxygen content indicates that the oil has suffered long exposure to the atmosphere. Oxygen in crude oil can occur in a variety of forms. These include alcohols, ethers, carboxylic acids, phenolic compounds, ketones, and esters. The presence of such compounds causes the crude to be acidic with consequent processing problems such as corrosion.[1]

1.2.2.3: Nitrogen Compounds:

Crude oils contain very low amounts of nitrogen compounds. In general, the more asphaltic the oil, the higher its nitrogen content. Nitrogen compounds are more stable than sulfur compounds and therefore are harder to remove. Even though they are present at very low concentrations, nitrogen compounds have great significance in refinery operations. They can be responsible for the poisoning of a cracking catalyst, and they also contribute to gum formation in finished products.

The nitrogen compounds in crude oils may be classified as basic or non-basic. The greater part of the nitrogen in crude oils is the non-basic nitrogen compounds.[1]

1.2.2.4: Metallic Compounds:

Metallic compounds exist in all crude oil types in very small amounts. Their concentration must be reduced to avoid operational problems and to prevent them from contaminating the products. Metals affect many upgrading processes. They cause poisoning to the catalysts used for hydroprocessing and cracking. Even minute amounts of metals (iron, nickel and vanadium) in the feedstock to the catalytic cracker affect the activity of the catalyst and result in increased gas and coke formation and reduced gasoline yields.[1]

1.3: Classification Of Crude Oil:

1.3.1: API Gravity:

The gravity of crude oil determines its price commercially. It is generally expressed as API gravity defined as:

$$API = \frac{141.5}{Sp.gr} - 131.5 \qquad \text{\_\_\_ (1.1)}$$

where sp.gr is the specific gravity defined as the density of the crude oil relative to the density of water both at 15.6 C (60F). The API gravity can range from 8.5 for very heavy crudes to 44 for light crudes, see table (1.2) .[1]

Table 1.2 Classification of crude depending on API gravity[1].

QUALITY CLASSIFICATIONS	**API Gravity Range**
Light Crude	API > 38
Medium Crude	38 > API > 29
Heavy Crude	29 > API > 8.5
Very heavy crude	API < 8.5

1.3.2: Characterization Factor:

There are several correlations between yield and the aromaticity and paraffinicity of crude oils, but the most widely used is "Watson characterization factor",(K),where :

$$K = (T_B)^{1/3}/sp.gr \qquad \text{\_\_\_(1.2)}$$

Where,

T_B: molar average boiling point, ^oR
Sp.gr: specific gravity at 60 ^oF

The Watson characterization factor ranges from less than 10 for highly aromatic material to almost 15 for highly paraffinic compounds. Crude oils show a narrower range of K and vary from 10.5 for highly naphthenic crude to 12.9 for a highly paraffinic crude base. [2]

1.4: Properties Of Crude Oil:

1.4.1: Physical Properties:

1.4.1.1: Specific Gravity:

Specific gravity for liquid oils is defined as:

$$Sp.gr_{oil} = \rho_{oil}/\rho_{water} \qquad \text{\_\_\_\_\_ (1.3)}$$

Where, where ρ_{oil} is the oil density and ρ_{water} is the water density.

Both densities of oil and water are at the same standard temperature and pressure conditions, which are 1 atm (14.7 psia) and 15.6 ^{o}C (60 ^{o}F). Since under the same conditions, most petroleum fractions are lighter than water, i.e. $\rho_{oil} \leq 1$. [1]

1.4.1.2: Viscosity:

The resistance to flow or the pumpability of the crude oil or petroleum fraction is indicated by the viscosity. More viscous oils create a greater pressure drop when they flow in pipes. Viscosity measurement is expressed in terms of kinematic viscosity in centiStocks (cSt). [1]

1.4.1.3: Molecular Weight:

Most crude oils and petroleum fractions have average molecular weights from 100 to 500. Although, there are several methods for measuring the molecular weight, the most suitable method is that based on freezing point depression.[1]

1.4.2: Thermal Properties:

1.4.2.1: Pour Point:

The pour point is defined as the lowest temperature at which the sample will flow. It indicates how easy or difficult it is to pump the oil, especially in cold weather. It also indicates the aromaticity or the paraffinity of the crude oil or the fraction. A lower pour point means that the paraffin content is low.[1]

1.4.2.2: Flash Point:

The flash point of a liquid hydrocarbon is the lowest temperature at which sufficient vapors are produced above the liquid such that spontaneous ignition will occur if a spark is present. It is an important specification for gasoline and naphtha related to safety in storage and transport in high temperature environments. Flash point indicates the fire and explosion potential of a fuel. A low flash point fuel is a higher fire hazard.[1]

1.4.2.3: Distillation Range:

The distillation range of the crude gives an indication of the quantities of the various products present. The most useful type of distillation is known as a true boiling (TBP) distillation and generally refers to a distillation performed in equipement that accomplishes a reasonable degree of fractionation.

TBP distillations characterize the volatility of petroleum fractions and crude oils. The boiling point distribution of crude oil (boiling point versus volume or mass percent distilled) is obtained through a batch distillation test. The distillation apparatus has 15 -18 theoretical plates with a 5:1 reflux ratio. For boiling points below 340 ^{o}C (644 ^{o}F) the distillation is performed at atmospheric pressure. The residue is distilled under vacuum (1–10 mm Hg). The boiling points under vacuum are converted to normal boiling points. The distillation continues to a normal boiling point of 535^{o}C (995 ^{o}F). This test allows for the collection of sample cuts at different boiling point ranges. These cuts can be subjected to physical and chemical measurements.[2][1]

1.4.2.4: Specific Heat:

The specific heat of material is defined as the quantity of heat required to raise the temperature of a unit weight of material through a temperature difference of one degree. The specific heat of crude oil or oil fractions is nearly linear function of temperature and it is also dependent on the specific gravity.[3]

Usually the specific heat is obtained from charts such as figure (5-1) and figure (5-2) in reference [3].

1.5: Products Of petroleum:

There are specifications for over 2000 individual refinery products. Intermediate feed stocks can be processed in various units to produce different blend products depending on market demand. The principal crude oil products fall into the following classes:

1.5.1: Liquefied Petroleum Gas (LPG):

Liquified petroleum gas is a group of hydrocarbon-based gases derived from crude oil refining or natural gas fractionation. They include ethane, ethylene, propane, propylene, normal butane, butylene, isobutane and isobutylene. For convenience of transportation, these gases are liquefied through pressurization.[1]

1.5.2: Gasoline:

Gasoline is classified by octane ratings into three grades: Regular, Midgrade and Premium.

- Regular gasoline: Gasoline having an antiknock index (octane number) rang $88 > ON \geq 85$.
- Mid-grade gasoline: Gasoline having octane rating, $90 \geq ON \geq 88$.
- Premium gasoline: Gasoline having octane rating greater than 90.[1]

1.5.3: Kerosene:

Kerosene is a light petroleum distillate that is used in space heaters, cook stoves and water heaters and which is suitable for use as a light source. This product also upgrade and used by the jet engines. [1]

1.5.4: Jet Fuel:

This category comprises both gasoline and kerosene and meets specifications for use in aviation turbine power units. [1]

1.5.5: Diesel Fuel:

The quality of diesel fuels can be expressed as cetane number or cetane index, which expresses the ignition property of diesel fuel. The cetane number (CN) is expressed in terms of the volume percent of cetane ($C_{16}H_{34}$) which has high ignition (CN = 100) in a mixture with alpha-methyl-naphthalene ($C_{11}H_{10}$) which has low ignition quality (CN = 0).

Diesel fuel includes No.1 diesel (Super-diesel) which has cetane number of 45 and it is used in high speed engines, trucks and buses. No. 2 diesel has 40 cetane number. Railroad diesel fuels are similar to the heavier automotive diesel fuels, but have higher boiling ranges and lower cetane numbers (CN = 30). [1]

1.5.6: Fuel Oil:

The fuel oils are mainly used in space heating and thus the market is quite high specially in cold climates. [1]

1.5.7: Residual Fuel Oil:

It is mainly composed of vacuum residue. Critical specifications are viscosity and sulfur content. Low sulfur residues are in more demand in the market. [1]

1.5.8: Lube Oil:

Lubricants are based on the viscosity index. Paraffinic and naphthenic lubricants have a finished viscosity index of more than 75. [1]

1.5.9: Asphalt:

Asphalt is an important product in the construction industry and comprises up to 20% of products. It can be produced only from crude containing asphaltenic material. [1]

1.5.10: Petroleum Coke:

Carbon compounds formed from thermal conversion of petroleum containing resins and asphaltenes are called petroleum cokes. Fuel grade coke contains about 85% carbon and 4% hydrogen. The balance is made up of sulfur, nitrogen, oxygen, vanadium and nickel. [1]

2.1: Introduction:

Crude distillation unit (CDU) is at the front-end of the refinery, also known as topping unit, or atmospheric distillation unit. It receives high flow rates hence its size and operating cost are the largest in the refinery. The crude stills are are used to separate the crude oils by distillation into fractions according to boiling point so that each of the processing units following will have feedstocks that meet their particular specifications. The inside of the crude still is divided at intervals by horizontal trays the temperature gradually reduces towards the top so, that each tray is a little cooler the one below.[1][2]

The capacity of the CDU ranges from 10,000 barrels per day (BPD) or 1400 metric tons per day (tpd) to 400,000 BPD (56,000 metric tpd). The unit produces raw products which have to be processed in downstream unit to produce products of certain specifications. This involves the removal of undesirable components like sulfur, nitrogen and metal compounds, and limiting the aromatic contents.[1]

Typical products from the unit are:

1. Gases.
2. Light straight run naphtha (also called gasoline or light naphtha).
3. Heavy gasoline (also called military jet fuel or naphtha).
4. Kerosene (also called light distillate or jet fuel).
5. Middle distillates called diesel or light gas oil (LGO).
6. Heavy distillates called atmospheric gas oil (AGO) or heavy gas oil (HGO).
7. Crude column bottoms called atmospheric residue or topped crude.[1]

2.2: Description of The process :

The crude oil should first be desalted , since it contains salt in the form of dissolved salt in the tiny droplet of water which forms a water-in oil emulsion. This water cannot be separated efficiently by gravity or through mechanical means, so it is desalted and separated through electrostatic water separation. The trend toward running heavier crude oils has increased the importance of efficient desalting of crudes. Until recently, the criterion for desalting crude oils was 10 lb salt/1000 bbl (expressed as NaCl) or more, but now many companies desalt all crude oils. Reduced equipment fouling and corrosion and longer catalyst life provide justification for this additional treatment. It involves mixing the crude with dilution water (3–10 vol%) through a mixing valve at temperatures from 200 to 300°F (90 to 150°C). Both the ratio of the water to oil and the temperature of operation are functions of the density of the oil.[1][2]

The crude is further heated in product heat exchangers. The preheating of the crude using the hot products cools down the products to the desired temperature for pumping to the storage tanks. This is essential for the economics of the unit in terms of energy conservation and utilization. After desalting, the crude oil is pumped through a series of heat exchangers and its temperature raised to about 550°F (288°C) by heat exchange with product and reflux streams. It is then further heated to about 750°F (399°C) in a furnace (i.e., direct-fired heater or ‘‘pipe-still’’) and charged to the flash zone of the atmospheric fractionators. The furnace discharge temperature is sufficiently high [650 to 750°F (343 to 399°C)] to cause vaporization of all products withdrawn above the flash zone plus about 10 to 20% of the bottoms product. This10 to 20% ‘‘over-flash’’ allows some fractionation to occur on the trays just above the flash zone by providing internal reflux in excess of the sidestream withdrawals.[1][2]

Reflux is provided by condensing the tower overhead vapors and returning a portion of the liquid to the top of the tower, and by pump-around streams lower in the tower. Each of the sidestream products removed from the tower decreases the amount of reflux below the point of drawoff. Maximum reflux and fractionation is obtained by removing all heat at the top of the tower, but this results in an inverted cone-type liquid loading which requires a very large diameter at the top of the tower. To reduce the top diameter of the tower and even the liquid loading over the length of the tower, intermediate heat-removal streams are used to generate reflux below the sidestream removal points. To accomplish this, liquid is removed from the tower, cooled by a heat exchanger, and returned to the tower or, alternatively, a portion of the cooled sidestream may be returned to the tower. This cold stream condenses more of the vapors coming up the lower and thereby increases the reflux below that point. The energy efficiency of the distillation operation is also improved by using pump-around reflux. If sufficient reflux were produced in the overhead condenser to provide for all sidestream drawoffs as well as the required reflux, all of the heat energy would be exchanged at the bubble-point temperature of the overhead stream. By using pump-around reflux at lower points in the column, the heat transfer temperatures are higher and a higher fraction of the heat energy can be recovered by preheating the feed. Although crude towers do not normally use reboilers, several trays are generally incorporated below the flash zone and steam is introduced below the bottom tray to strip any remaining gas oil from the liquid in the flash zone and to produce a high-flash-point bottoms. The steam reduces the partial pressure of the hydrocarbons and thus lowers the required vaporization temperature.[2]

The atmospheric fractionator normally contains 30 to 50 fractionation trays. Separation of the complex mixtures in crude oils is relatively easy and generally five to eight trays are needed for each sidestream product plus the same number above and below the feed plate. Thus, a crude oil atmospheric fractionation tower with four liquid sidestream drawoffs will require from 30 to 42 trays.[2]

The liquid sidestream withdrawn from the tower will contain low-boiling components which lower the flashpoint, because the lighter products pass through the heavier products and are in equilibrium with them on every tray. These ‘‘light ends’’ are stripped from each sidestream in a separate small stripping tower containing four to ten trays with steam introduced under the bottom tray. The steam and stripped light ends are vented back into the of the atmospheric fractionator above the corresponding side-draw tray.[2]

The overhead condenser on the atmospheric tower condenses the pentane and heavier fraction of the vapors that passes out of the top of the tower. This is the light gasoline portion of the overhead, containing some propane and butanes and essentially all of the higher-boiling components in the tower overhead vapor. Some of this condensate is returned to the top of the tower as reflux, and the remainder is sent to the stabilization section of the refinery gas plant where the butanes and propane are separated from the (C5-180°F) LSR gasoline.[2]

The crude flow rate to the CDU determines the capacity of the whole refinery. The capacity of the column is limited by the vapor flow rate with a velocity between 2.5 and 3.5 ft/s (0.76 and 1.07 m/s). The vapor flow rate increases as the vapors rise from the flash zone to the overhead. To keep the vapor velocity within the limits mentioned above, the pump-arounds, which are installed at several points along the column, extract heat from the column. This results in condensing the rising vapors and reducing the vapor velocity.[1]

To expand crude capacity, the most used technique is to introduce a pre-flash column before the crude heater. The crude oil after preheating in the hot products and pump-around heat exchangers is flashed into a column where the lightest products are removed. The bottoms from the pre-flash column are introduced into the crude heater and then to the crude column. The amounts of the light ends in the crude are now less, and this reduces the vapor loading up the column. Although the unit throughput is increased, the furnace duty is not increased, since the crude rate going to the furnace is not affected due to the removal of the light ends. Pre-flash columns are also introduced in the original design of the CDU when the crude oil is light, and when it contains a lot of light ends in the naphtha range.[1]

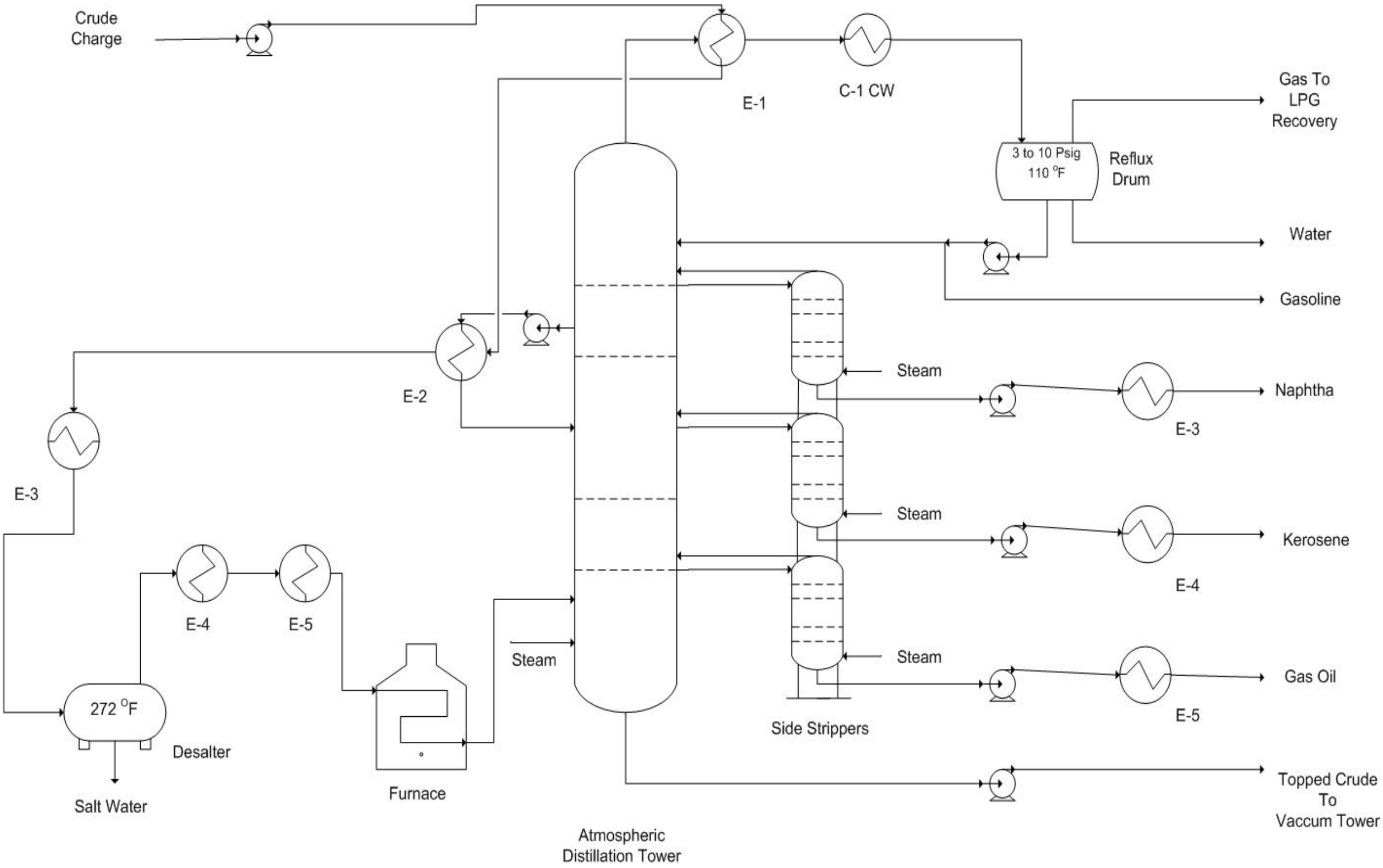


Fig.1: Process flow diagram of atmospheric distillation unit. [2]

Basis : 35,000 bbl/day

Crude petroleum analysis (p.329)[3]

API=36

S%=0.2

Density of water at 60 °F [5] =998.6 kg/m^3

(=350.0186 lb/bbl)

Density of crude oil:

$$\text{Sp.gr}=141.5/(\text{API}+131.5)$$

$$=141.5/(36+131.5) = 0.8448$$

Density of crude oil= sp.gr (60°F/60°F) * density of water

=0.8448 * 350.0186

=295.6957 lb/bbl

The concentration of salt =10 lb NaCl /1000 bbl oil

NaCl density [6]: sp.gr at (17.5°C) = 2.163

Density= 2.163 * 998.1805

=756.7723 lb/bbl

$$\text{Mass flowrate } \mathbf{M} \text{ of NaCl} = \frac{10\ \text{lb}}{1000\ \text{bbl}} * 35{,}000 \frac{\text{bbl}}{\text{day}} * \frac{1\ \text{day}}{24\ \text{hr}}$$

=14.5833 lb/hr

$$\text{Volumetric flowrate } \mathbf{Q} \text{ of Nacl} = 14.5833 \frac{\text{lb}}{\text{hr}} * \frac{1}{756.7723\ (\frac{\text{lb}}{\text{bbl}})} * \frac{24\ \text{hr}}{1\ \text{day}}$$

=0.4625 bbl/day

Crude oil volumetric flowrate **Q** =35,000 -0.4625

=34,999.5375 bbl/day

Crude oil mass flowrate **M** = $34{,}999.5375\frac{\text{bbl}}{\text{day}} * 295.6957\frac{\text{lb}}{\text{bbl}}$

=10,349,212.74 lb/day

=431,217.1975 lb/hr

Mass flowrate of **sufur** =$\frac{0.2\text{lb sulfur}}{100\text{ lb oil}} * 10{,}349{,}212.74\ \frac{\text{lb}}{\text{day}}$

=20,698.4255 lb/day

=862.4344 lb/hr

Mass flowrate of feed =Mass flowrate of crude + mass flowrate of salt

= 431,217.1975+ 14.5833

=431,231.7808 lb/hr

3.1: Material Balance on Desalter :

From table (p.35)[2], for API=36 → wash water vol% = 5.8

Temperature at desalter =272 °F

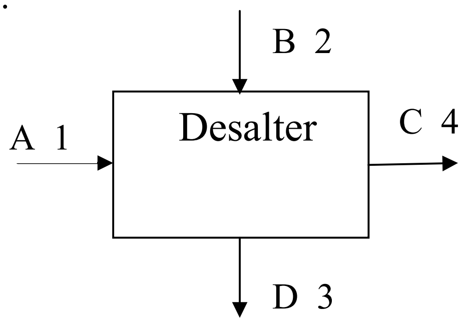
For one stage desalter , the efficiency= 95% [2].

A = crude oil + salt

B = wash water

C = crude oil after desalting

D = brine solution



ρ_{water}(at 272 °F)[5] =326.5922 lb/bbl

A =431,231.7808 lb/hr

$$B = \frac{5.8\ bbl}{100\ bbl} * 35{,}000\ \frac{bbl}{day} * 326.5922\ \frac{lb}{bbl} * \frac{1\ day}{24\ hr}$$

=27,624.2569 lb/hr

NaCl out in brine solution =14.5833 $\frac{lb}{hr}$ * 0.95 =13.8541 lb/hr

NaCl remaining with in crude =14.5833 $\frac{lb}{hr}$ * (1- 0.95) =0.7292 lb/hr

C = 431,217.1975 + 0.7292 = 431,217.9267 lb/hr

D =27,624.2569 +13.8541 =27,638.111 lb/hr

Input =Output

A+B =C+D

A+B = 431,231.7808 + 27,624.2569 = 458,856.0377 lb/hr

C+D = 431,217.9267 + 27,638.111 = 458,856.0377 lb/hr

Table 3-1 Material balance on desalter.

Stream no. / Component	Input		Output	
	1	2	3	4
Crude Oil	430,354.7631	——	——	430,354.7631
Sulfur	862.4344	——	——	862.4344
Salt	14.5833	——	13.8541	0.7292
Water	——	27,624.2569	27,624.2569	——
Total	431,231.7808	27,624.2569	27,638.111	431,217.9267

3.2 : Material Balance on Distillation Tower :

3.2.1: Spesification of Crude Oil and its Products :

1- TBP curve data from (p.409)[2] :

Table 3-2: TBP curve data.

TBP (^oF)	Vol %	API
122	0.5	79.7
167	1.7	78.1
212	3.3	64.5
257	6	57.7
302	9.1	53.5
347	13	50.4
392	17.7	47.8
437	23.4	45.2
482	31.4	42.3
527	42.1	40

We must correct the temperature that obtained at 40 mmHg into temperature at pressure of 760 mmHg (atmospheric pressure) by using Fig(3.6)[2] :

Table 3-3 Temperature correction to atmospheric pressure :

Temp. at 40 mmHg	Temp. at 760 mmHg	Vol%	API
392	583	47.1	36
437	637	57.1	32
482	687	64.9	32.5
527	739	71.9	30
572	793	78.4	27.7

We estimate the shape of TBP curve by extrapolation to 1007 °F:

Table 3-4 TBP curve extrapolation.

Temp. at 760 mmHg	Vol%
959	99.2
965	100

2- ASTM curve:

This curve obtained using the following relation [1],

$$\text{ASTM Temp.= (TBP Temp./a)\^{}}\left(\frac{1}{b}\right) \qquad \text{\_\_\_(2.1)}$$

Table 3-5 ASTM curve data.

Vol%	a	b	TBP	ASTM
0	0.9167	1.0019	86	93.01
10	05277	1.09	311	348.04
30	0.7429	1.0425	474	490.34
50	0.892	1.0176	600	601
70	0.8705	1.0226	725.7	718.51
90	0.949	1.011	884.7	865.41
95	0.8008	1.0355	925	907.03

$$\text{Average volumetric boiling Temp.}=\frac{\text{T10+T30+...+T90}}{5} \qquad \text{\_\_\_(2.2)}$$

$$=574.7639\ ^{o}F = 1034.7639\ ^{o}R$$

Average molal boiling temp. = Average volumetric boiling Temp.-Δt ___(2.3)

Where Δt obtained from Fig(5-4)[3] , ASTM slope(10%-90%)=6.47 → Δt=99 and Average molal boiling point =574.7639 -348.04

=475.7639 °F

=935.4539 °R

3-Base of the Crude Oil :

Key fraction no.1	482 - 527 °F	at 760 mmHg [3]
Key fraction no.2	733 - 779 °F	

Key fraction no.1	42.3 - 38.2 API	from Fig.2
Key fraction no.2	30.7 – 28.4 API	

The cut of API at 504.5 °F=40.1 API

The cut of API at 756 °F =29.5 API

At API =40.1 sp.gr=141.5/(40.1+131.5) = 0.8246

$K=(T_B)^{(1/3)}$ / sp.gr $=(935.4539)^{(1/3)}/0.8246$ =11.86

At API =29.5 sp.gr=141.5/(29.5+131.5) = 0.8789

The characterization factor is (11.28- 11.86) ,the type is(raw 11.5 K stock or Dewaxed 11.75 K stock)[3] ,we use Fig(4.1)[3] to find the viscosity at 210 °F (Fig D)[3],

Table 3-6 Viscosity of the crude.

Viscosity at 100 °F	Viscosity at 210 °F
41	155
47	290
59	630
84	1550

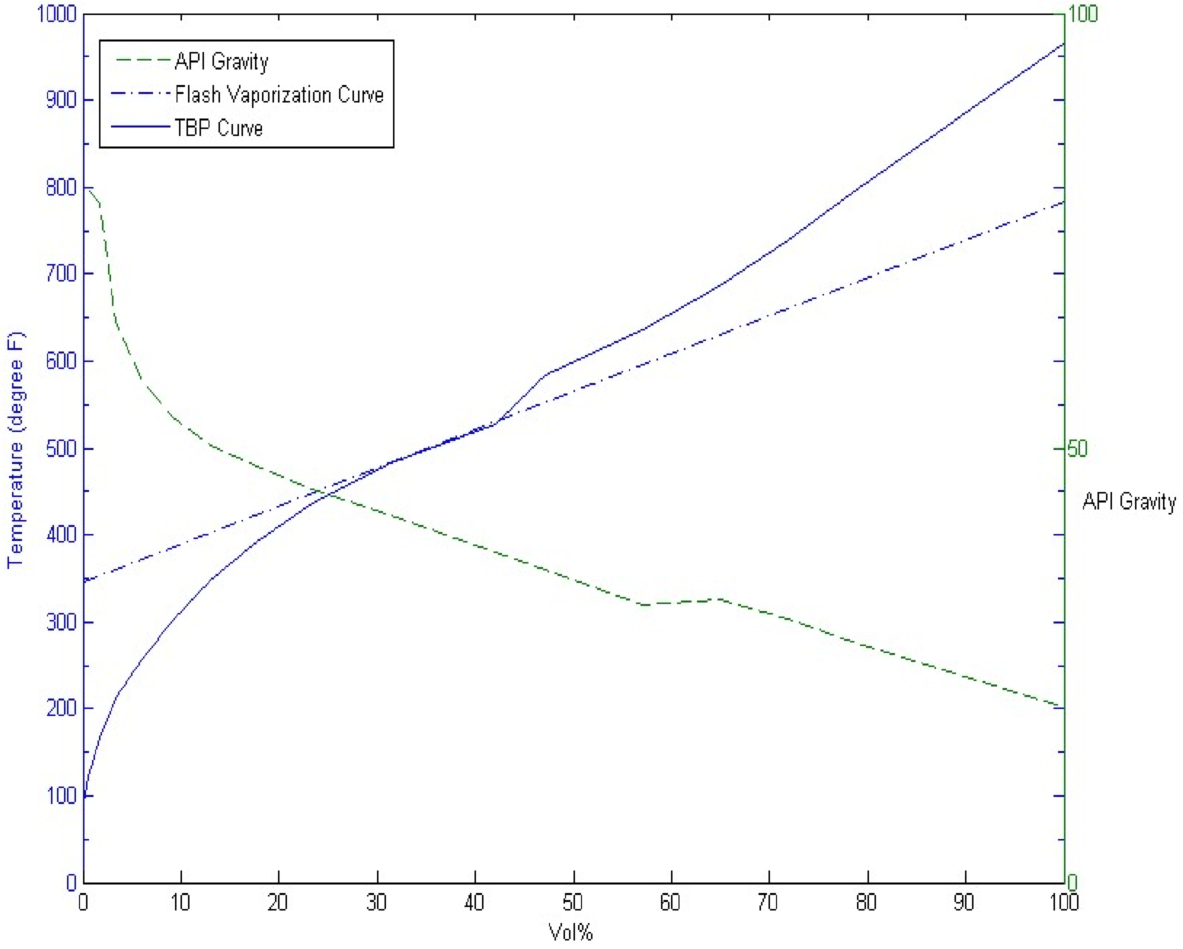


Fig.2:Crude Distillation Curves.

3.2.2 : Crude oil Material Balance on Distillation tower :

The typical cut point temperatures used are (p.38)[2] :

Table 3-7 Typical oil fractions cut point.

Fraction	Boiling range (°F)
Gases	< 90
Light naphtha	90 - 190
Heavy naphtha	190 - 380
Kerosene	380 - 520
Light gas oil	520 – 610
Heavy gas oil	610 – 800
Vaccum gas oil	800 - 1050
Vaccum residue	> 1050

In this order the cut point of products to be withdrawal from the distillation tower are:

Table 3-8 Cut point of the products :

Product	Temperature range (°F)
Gases	<90
Gasoline	90 - 190
Naphtha	190 – 380
Kerosene	380 - 520
Gas oil	520 – 800
Residue	800 - +1050

API and Vol% of oil fractions are obtained from Fig.2,

Table 3-9 : Material balance of crude oil :

Component	Vol %	API	Density (lb/bbl)	Q (bbl/day)	M (lb/hr)
Gases	0.002	80.4997	233.3103	0.7	6.8049
Gasoline	2.598	78.65	235.372	909.2880	8,917.5390
Naphtha	13.9	53.2	267.872	4,864.9357	54,299.1690
Kerosene	23.8	43.4	282.932	8329.8899	98,199.6837
Gas oil	39.1	32.1	302.482	13,684.8192	172,475.4784
Residue	20.6	23.8	318.678	7,209.9047	95,734.9171
Crude oil	100	36	295.6724	34,999.5375	431,217.1975

Column (3) is obtained from appendix "A" [3] ,

Column (5) is obtained as $Q_i = F * Vol\% / 100$

Column (6) is obtained as $M_i = F * Vol\% * \rho_i / (100 * 24)$

Where Q_i : Volumetric flowrate of oil fraction i

F: Volumetric flowrate of crude oil =34,999.5375 bbl/day

Vol% : crude oil fraction percent

ρ_i : Density of oil fraction i

Error=431,217.1975-429,633.5911 =1584.3346 lb/hr

Percentage of crude mass balance error = $\frac{(431,217.1975-429,633.5911)}{431,217.1975} * 100$

=0.37%

5% of salt remaining in the desalted crude (its very little amount) , so assume it will entirely appear in atmospheric residue fraction mainly :

Atmospheric Residue stream (lb/hr) =95,734.9171 + 0.7292

=95,735.6463 lb/hr

Table 3-10 Properties of oil and its fractions.

Component	Spesific Gravity	$T_{50\%}$	Molecular Weight
Gases	0.6674	45	45
Gasoline	0.6732	140	79
Naphtha	0.7661	285	116
Kerosene	0.8090	450	181
Gas oil	0.8649	660	292
Residue	0.9111	882.5	403
Crude oil	0.8448	525.5	-

Column (2) is obtained from eq.(1.1)

Column (4) is obtained from F(5-9)[3]

3.2.3 : Sulfur Material Balance on Distillation tower :

Sulfur content in crude =0.2%

The sulfur content in distributed in oil fractions is estimated from Fig(4-30)[3].

Sulfur in feed = Q * ρ_{oil} * S% /(100 *24) ___(2.4)

$$=34{,}999.5375\frac{\text{bbl}}{\text{day}} * 295.6957\frac{\text{lb}}{\text{bbl}} * \frac{0.2\text{lb sulfur}}{100\text{ lb oil}} *\frac{1\text{day}}{24\text{hr}} =862.4344 \text{ lb/hr}$$

Sulfur in oil fraction = Q_i * ρ_i * S_i% /(100 *24) ___(2.5)

Table 3-10 sulfur material balance

Cut Temp.(^{o}F)	S%	Fraction Stream (lb/hr)	Sulfur mass flowrate (lb/hr)
86 - 90	0	6.8049	0
90 - 190	0	8,917.5390	0
190 - 380	0.025	54,299.1690	13.5748
380 - 520	0.076	98,199.6837	74.6318
520 - 800	0.25	172,475.4784	439.1556
800 - 965	0.35	95,734.9171	335.0722
Total	0.2	431,217.1975	862.4344

3.2.4 : Amount of Steam used for Stripping :

For ordinary conditions the amount of steam used in ranges are listed in table (p.232)[3], and the ratio amount are selected in column (3),table 2-11 ;

Amount of steam required = Qi * (42 gal/ 1 bbl) *(1day/24 hr)
*(lb steam/gal fraction) ___(2.6)

Table 3-11 Amounts of steam used for stripping.

Fraction	Fraction flowrate (bbl/day)	Steam amount lbs / gallon of fraction	Steam required (lb/hr)
Naphtha	4,864.9357	0.35	2,979.7731
Kerosene	8329.8899	0.4	5,830.9229
Gas oil	13,684.8192	0.45	10,776.7951
Topped crude	7,209.9047	1.2	15,140.7999
Total			34,728.291

3.2.5 Total Material Balance on Distillation Tower :

Where :

Stream 4= Crude oil after desalting

Stream 5 = Steam for stripping reduced crude

Stream 6 = Steam for stripping naphtha

Stream 7 = Steam for stripping kerosene

Stream 8 = Steam for stripping gas oil

Stream 9 =Gases stream

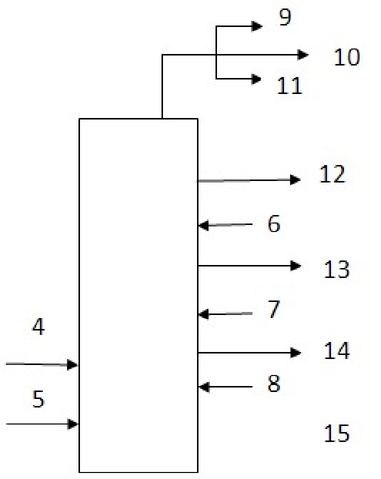
Stream 10 =Gasoline stream

Stream 11=Water stream

Stream 12 =Naphtha stream

Stream 13 =kerosene stream

Stream 14=Gas oil stream



Distillation Tower

Stream 15 =Topped crude stream

Table 3-12 Material balance on distillation tower.

component Stream no.		Oil fraction*	sulfur	Salt	Water	Total
Input	4	430,354.7631	862.4344	0.7292	-	431,217.9267
	5	-	-	-	15,140.7999	15,140.7999
	6	-	-	-	2,979.7731	2,979.7731
	7	-	-	-	5,830.9229	5,830.9229
	8	-	-	-	10,776.7951	10,776.7951
Output	9	6.8049	-	-	-	6.8049
	10	8,917.5390	-	-	-	8,917.5390
	11	-	-	-	34,728.291	34,728.291
	12	54285.5942	13.5748	-	-	54,299.1690
	13	98125.0519	74.6318	-	-	98,199.6837
	14	172036.3228	439.1556	-	-	172,475.4784
	15	95399.8449	335.0722	0.7292	-	95,735.6463

Input = Output

Stream 4+ Stream 5 +Stream 6 +Stream 7 +Stream 8 = Stream 9 + Stream10 + Stream 11+ Stream 12 + Stream 13 +Stream 14+Stream 15

431,217.9267 +15,140.7999 +2,979.7731 +5,830.9229 +10,776.7951 =

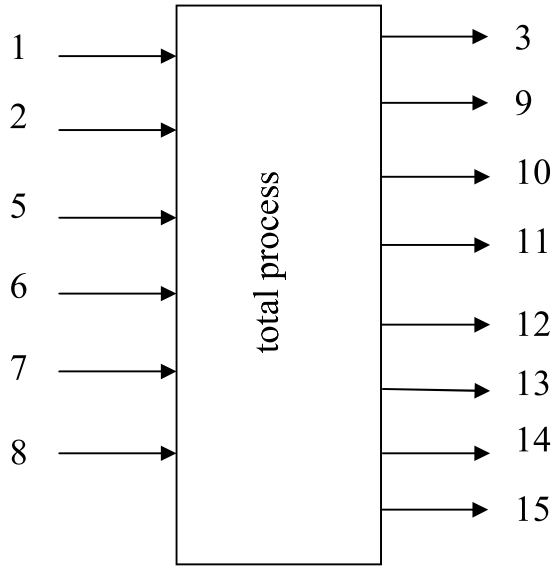
6.8049 +8,917.5390 +34,728.291 +54,299.1690 +98,199.6837 +172,475.4784 +95,734.9171

465,946.2177 = 464,361.8831

Error =465,946.2177 - 464,361.8831=1584.3346 lb/hr

Percentage error = 0.34%

3.3 Overall Material Balance:



Input = Output

Stream 1 +Stream 2 + Stream 5 +Stream 6 +Stream 7 +Stream 8 = Stream 3+ Stream 9 + Stream10 + Stream 11+ Stream 12 + Stream 13 +Stream 14 +Stream 15

431,231.7808 +27,624.2569 +15,140.7999 +2,979.7731 +5,830.9229 +10,776.7951 =27,638.111 +6.8049 +8,917.5390 +34,728.291 +54,299.1690 +98,199.6837 +172,475.4784 +95,734.9171

493,584.3287=491,999.9941 lb/hr

Error= 493,584.3287-491,999.9941 =1584.3346 lb/hr

Percentage error=0.32%

The number of plates estimated from table(16-13)[3] ,the following numbers of plates will be used:

	No. of plates
Gasoline to naphtha	4
Naphtha to kerosene	5
Kerosene to gas oil	7
Inlet to lowest side product	4
Stripping plates	4
Total no. of plates	24

The pressure at the top of the topping tower is seldom more than 4 psig and the pressure per tray ranges from (0.1-0.2) psi [4], thus the pressure for the trays will be ,

Top plate = 4 psig

Naphtha drawoff plate = 4 + 0.15(4) = 4.6 psig

Kerosene drawoff plate = 4 + 0.15(4+5) = 5.35 psig

Gasoline drawoff plate = 4 +0.15(4+5+7) = 6.4 psig

Feed plate = 4 + 0.15(4+5+7+4) = 7 psig

4.1: Equilibrium Flash Vaporization Curves:

4.1.1: Crude oil EFV Curve:

1- TBP slope (10-70) = $\frac{725.5-311}{70-10}$ = 6.91 deg./percent …from Fig.2

2- 50% TBP temp. = 600 °F

3- EFV curve slope = 4.38 …from Fig(4-18)[3]

4- 50% EFV temp.= 600 -35 =565 °F … from Fig(4-19)[3]

100% EFV temp. = 565 + 50 * 4.38 = 784 °F

0% EFV temp. = 565 - 50 * 4.38 = 346 °F

4.1.2: Gasoline EFV Curve:

1- TBP slope (10-70) = $\frac{162.5-102}{70-10}$ = 1.01 deg./percent …from Fig.2

2- 50% TBP temp. = 140 °F

3- EFV curve slope = 0.22 …from Fig(4-18)[3]

4- 50% EFV temp.= 140-11.8 =128.2 °F … from Fig(4-19)[3]

100% EFV temp. = 128.2 + 50 * 0.22 = 139.2 °F

0% EFV temp. = 128.2 - 50 * 0.22 = 117.2 °F

4.1.3: Naphtha EFV Curve:

1- TBP slope (10-70) = $\frac{338-225}{70-10}$ = 1.88 deg./percent …from Fig.2

2- 50% TBP temp. = 285 °F

3- EFV curve slope = 0.54 …from Fig(4-18)[3]

4- 50% EFV temp.= 285 -11.5 =273.5 °F … from Fig(4-19)[3]

100% EFV temp. = 273.5 + 50 * 0.54 = 300.5 °F

0% EFV temp. = 273.5 - 50 * 0.54 = 246.5 °F

4.1.4: Kerosene EFV Curve:

1- TBP slope (10-70) = $\frac{415.5-401}{70-10}$ = 0.24 deg./percent …from Fig.2

2- 50% TBP temp. = 450 °F

3- EFV curve slope = 0.07…from Fig(4-18)[3]

4- 50% EFV temp.= 450 +26 = 476 °F … from Fig(4-19)[3]

100% EFV temp. = 476 + 50 * 0.07 = 479.5 °F

0% EFV temp. = 476 - 50 * 0.07 = 472.5 °F

4.1.5: Gas oil EFV Curve:

1- TBP slope (10-70) = $\frac{705.5-551}{70-10}$ = 2.57 deg./percent …from Fig.2

2- 50% TBP temp. = 660 °F

3- EFV curve slope = 0.93 …from Fig(4-18)[3]

4- 50% EFV temp.= 660 + 1 = 661 °F … from Fig(4-19)[3]

100% EFV temp. = 661 + 50 * 0.93 = 707.5 °F

0% EFV temp. = 661 - 50 * 0.93 = 614.5 °F

4.2: Feed Plate Temperature:

To evaporate all the feed products in the flash zone with 6% of the reduced crude, to provide internal reflux, the temperature required calculated as following, To evaporate all feed with 6% of residue (i.e, 80.64% of feed), from EFV curve Fig.2,

$T_f = 699.2\ ^oF$ (this temp. is to be corrected).

P = 7 psig = 7 + 14.7 = 21.7 psia

Then from Fig(5-27)[3], by the atmospheric B.P = 699.2 ^oF and p = 21.7 psia, then; $T_f = 730\ ^oF$ (this temp. is without using steam).

4.2.1: Feed Plate Temperature (Calculating the effect of steam):

$T_f = 699.2\ ^oF$ (without using steam), P = 21.7 psia

Moles of gases = 6.8049/45 = 0.1512 lbmol / hr

Moles of gasoline = 8917.539/79 = 112.8802 lbmol / hr

Moles of naphtha = 54,299.169/116 = 468.0963 lbmol / hr

Moles of kerosene = 98,199.6837/181.5 = 541.0451 lbmol / hr

Moles of gas oil = 172,475.4784/292 = 590.6694 lbmol / hr

Moles of residue = $(0.06)(7209.9047 * \frac{1}{24} * 318.678)/403$ = 14.2533 lbmol / hr

---

Total moles of hydrocarbons = 1727.0955 lbmol / hr

Moles of steam = 15,140.7999/18 = 841.1556 lbmol / hr (the steam reachs on the feed plate is only that for stripping residue)

Total moles of vapors = 1727.0955 + 841.1556 = 2568.2511 lbmol / hr

p.p of hydrocarbons = (1727.0955/ 2568.2511)* 21.7 = 14.593 psia

For B.P temp. = 699.2 ^oF and P = 14.593 psia , from Fig(5-27)[3] ;

$T_f = 691\ ^oF$

Table 4-1

Component	TBP slope	EFV slope	50% TBP Temp.	50% EFV Temp.	0% EFV Temp.
Crude oil	6.91	4.38	600	565	346
Gasoline	1.01	0.22	140	128.2	117.2
Naphtha	1.88	0.54	285	273.5	246.5
Kerosene	0.24	0.07	450	476	472.5
Gas oil	2.57	0.93	660	661	614.5

4.3: Gas oil drawoff plate temperature:

Assume plate temperature , $T_p = 614.5\ ^{o}F$ (from table (4-1))

Sensible heat = m cp ΔT , cp from Fig(5-2)[3],

Gases 6.8049 (0.678) (691-614.5) = 352.9497 Btu / hr

Gasoline 8917.539 (0.654) (691-614.5) = 446,153.3937 Btu / hr

Naphtha 54,299.169 (0.643) (691-614.5) = 2,670,948.974 Btu / hr

Kerosene 98,199.6837 (0.634) (691-614.5) = 4,762,782.859 Btu / hr

Gas oil 172,475.4784 (0.631) (691-614.5) = 8,325,650.056 Btu / hr

Residue 0.06 (95,734.9171) (0.627) (691-614.5) = 275,518.39 Btu / hr

Steam 15,140.7999 (0.5) (407-691) = -2,149,993.586 Btu / hr

Latent heat = m λ , λ obtained from Fig(5-8)[3] ,

Residue (0.06) (95,734.9171) (87.5) = 502,608.3148 Btu / hr

Gas oil (172,475.4784) (75) = 12,935,660.88 Btu / hr

Total heat = 27,769,682.23 Btu / hr

Internal reflux = $\frac{Total\ heat}{latent\ heat*M.wt}$

$$= \frac{27{,}769{,}682.23}{75 * 292} = 1268.022 \text{ lbmol / hr}$$

Total moles

= moles of internal reflux + moles of hydrocarbons reaching the plate + moles of steam entering the plate + moles of steam coming from lower plates

=1268.022+0.1512+112.8802+468.0963+541.0451+(15,140.7999+10,776.7951) /18

= 3,830.0612 lbmol / hr

Mole fraction of gas oil = $\frac{\text{Internal reflux}}{\text{Total moles}}$

$$= \frac{1268.022}{3{,}830.0612} = 0.331$$

Then , partial pressure of gas oil = mole fraction * total pressure

= 0.331 * (6.4 + 14.7) = 6.9841 psia

For B.P temp. = 614.5 ^{o}F and P = 6.9841 psia , from Fig(5-27)[3] ;

T_p = 563 ^{o}F

4.4: Kerosene drawoff plate temperature:

Assume plate temperature , T_p = 472.5 ^{o}F (from table (4-1))

Sensible heat : cp from Fig(5-2)[3],

Gases 6.8049 (0.645) (691-472.5) = 959.0316 Btu / hr

Gasoline 8917.539 (0.642) (691-472.5) = 1,250,925.618 Btu / hr

Naphtha 54,299.169 (0.639) (691-472.5) = 7,581,331.425 Btu / hr

Kerosene 98,199.6837 (0.634) (691-472.5) = 13,603,503.98 Btu / hr

Gas oil 172,475.4784 (0.598) (691-563) = 13,201,963.02 Btu / hr

Residue 0.06 (95,734.9171) (0.589) (691-563) = 433,058.8122 Btu / hr

Steam (no.5) 15,140.7999 (0.5) (407-472.5) = -495,861.1967 Btu / hr

Steam (no.8) 10,776.7951(0.5) (563-472.5) = 490,344.1771 Btu / hr

Latent heat = m λ , λ obtained from Fig(5-8)[3] ,

Residue (0.06) (95,734.9171) (87.5) = 502,608.3148 Btu / hr

Gas oil (172,475.4784) (96) = 16,557,645.93 Btu / hr

Kerosene (98,199.6837) (103) =10,114,567.42 Btu / hr

Total heat = 63,241,046.53 Btu / hr

Internal reflux of kerosene $= \frac{63,241,046.53}{103 * 181}$ = 3392.214 lbmol / hr

Total moles =

3392.214 + 0.1512 + 112.8802+ 468.0963 + (15,140.7999+10,776.7951+ 5830.9229)/18 = 5737.1483 lbmol / hr

Mole fraction of kerosene = $\frac{3392.214}{5737.1483}$ = 0.591

p.p. = 0.591 * (5.35 + 14.7) = 11.8496 psia

For B.P temp. = 472.5 ^{o}F and P = 11.8496 psia , from Fig(5-27)[3] ;

$T_p = 465\ ^{o}F$

4.5: Naphtha drawoff plate temperature:

Assume plate temperature , $T_p = 246.5\ ^oF$ (from table (4-1))

Sensible heat : cp from Fig(5-2)[3],

Gases	6.8049 (0.58) (691-246.5) = 1754.3713 Btu / hr
Gasoline	8917.539 (0.575) (691-246.5) = 2,279,211.499 Btu / hr
Naphtha	54,299.169 (0.551) (691-246.5)= 13,298,925.32 Btu / hr
Kerosene	98,199.6837 (0.595) (691-465) = 13,204,911.47 Btu / hr
Gas oil	172,475.4784 (0.619) (691-563) = 13,665,577.1 Btu / hr
Residue	0.06 (95,734.9171) (0.612) (691-563) = 449,969.428 Btu / hr
Steam (no.5)	15,140.7999 (0.5) (407-246.5) = 1,215,049.192 Btu / hr
Steam (no.8)	10,776.7951(0.5) (563-246.5) = 1,705,427.825 Btu / hr
Steam (no.7)	5,830.9229 (0.5) (465-246.5) = 637,028.3268 Btu / hr

Latent heat = m λ , λ obtained from Fig(5-8)[3] ,

Residue	(0.06) (95,734.9171) (87.5) = 502,608.3148 Btu / hr
Gas oil	(172,475.4784) (96) = 16,557,645.93 Btu / hr
Kerosene	(98,199.6837) (100) = 9,819,968.37 Btu / hr
Naphtha	(54,299.169) (139) = 7,547,584.491 Btu / hr

Total heat = 80,885,661.64 Btu / hr

Internal reflux of naphtha $= \frac{80,885,661.64}{139 * 116}$ = 5,016.4762 lbmol / hr

Total moles =

5,016.4762 + 0.1512 + 112.8802+ (34,728.291/18) = 7,058.8571 lbmol / hr

Mole fraction of naphtha = $\frac{5,016.4762}{7,058.8571}$ = 0.7107

p.p. = 0.7107 * (4.6 + 14.7) = 13.717 psia

For B.P temp. = 246.5 °F and P = 13.717 psia , from Fig(5-27)[3] ;

T_p = 242 °F

4.6: Top temperature:

Assume top temperature , T = 139.2 °F (from table (4-1))

Sensible heat : cp from Fig(5-2)[3],

Gases 6.8049 (0.562) (691-139.2) = 2,110.2784 Btu / hr

Gasoline 8917.539 (0.54) (691-139.2) = 2,657,176.931 Btu / hr

Naphtha 54,299.169 (0.551) (691-242) = 13,433,560.11 Btu / hr

Kerosene 98,199.6837 (0.595) (691-465) = 13,204,911.47 Btu / hr

Gas oil 172,475.4784 (0.619) (691-563) = 13,665,577.1 Btu / hr

Residue 0.06 (95,734.9171) (0.612) (691-563) = 449,969.428 Btu / hr

Steam (no.5) 15,140.7999 (0.5) (407-139.2) = 2,027,353.107 Btu / hr

Steam (no.8) 10,776.7951(0.5) (563-139.2) = 2,283,602.882 Btu / hr

Steam (no.7) 5,830.9229 (0.5) (465-139.2) = 949,857.3404 Btu / hr

Steam (no.6) 2979.7731 (0.5) (242-139.2) = 153,160.3373 Btu / hr

Latent heat = m λ , λ obtained from Fig(5-8)[3] ,

Residue (0.06) (95,734.9171) (87.5) = 502,608.3148 Btu / hr

Gas oil (172,475.4784) (96) = 16,557,645.93 Btu / hr

Kerosene (98,199.6837) (100) = 9,819,968.37 Btu / hr

Naphtha (54,299.169) (142) = 7,710,481.998 Btu / hr

Total heat = 83,417,983.6 Btu / hr

Assume temp. of cold reflux = 50 ^{o}F , latent heat of gasoline λ =153 Btu / lb

$$\text{External reflux} = \frac{\text{Total Heat}}{(\lambda+cp(T-T_R))*M.wt}$$

$$= \frac{83{,}417{,}983.6}{(153+0.388\ (139.2-50))*79} = 5628.304 \text{ lbmol / hr}$$

Total moles = moles of gasoline + moles of external reflux + moles of steam

+ moles of gases

= 112.8802 + 5628.304 +(34,728.291/18) + 0.1512

= 7670.6849 lbmol / hr

$$\text{Mole fraction of gasoline} = \frac{112.8802 \quad 5628.304}{7670.6849} = 0.7485$$

Pressure on the top of tower = 4 psig +14.7 = 18.7 psia

p.p. of gasoline = 0.7485 * 18.7 = 13.997 psia

For B.P temp. = 139.2 ^{o}F and P = 13.997 psia , from Fig(5-27)[3] ;

T_p = 133 ^{o}F

<u>Tower diameter :</u>

Sensible heat : cp from Fig(5-2)[3],

Gases 6.8049 (0.55) (691-133) = 2,088.4238 Btu / hr

Gasoline 8917.539 (0.558) (691-133) = 2,776,600.613 Btu / hr

Naphtha 54,299.169 (0.551) (691-242) = 13,433,560.11 Btu / hr

Kerosene 98,199.6837 (0.595) (691-465) = 13,204,911.47 Btu / hr

Gas oil 172,475.4784 (0.619) (691-563) = 13,665,577.1 Btu / hr

Residue 0.06 (95,734.9171) (0.612) (691-563) = 449,969.428 Btu / hr

Steam (no.5) 15,140.7999 (0.5) (407-133) = 2,074,289.586 Btu / hr

Steam (no.8) 10,776.7951(0.5) (563-133) = 2,317,010.947 Btu / hr

Steam (no.7) 5,830.9229 (0.5) (465-133) = 967,933.2014 Btu / hr

Steam (no.6) 2979.7731 (0.5) (242-133) = 162,397.634 Btu / hr

Latent heat = m λ , λ obtained from Fig(5-8)[3] ,
Residue (0.06) (95,734.9171) (87.5) = 502,608.3148 Btu / hr
Gas oil (172,475.4784) (96) = 16,557,645.93 Btu / hr
Kerosene (98,199.6837) (100) = 9,819,968.37 Btu / hr

Naphtha (54,299.169) (142) = 7,710,481.998 Btu / hr

Total heat = 83,645,043.13 Btu / hr

Moles of hot reflux = 83,645,043.13 /(148 * 79) = 7154.0406 lbmol / hr

Moles of gasoline = 112.8802 lbmol / hr

Moles of steam = 34,728.291 / 18 = 1929.3495 lbmol / hr

Moles of gases = 0.1512 lbmol / hr

Total moles = 9196.4215 lbmol / hr

$$\text{Pressure} = (4 + 14.7) * \frac{760\ \text{mmHg}}{14.7\ \text{psia}} = 966.8\ \text{mmHg}$$

Using gases law, approximate volume [3] equal ;

$$V = 9196.4215\ \frac{\text{lbmol}}{\text{hr}} * 379\ \frac{\text{ft}^3\ .\text{hr}}{\text{lbmol}} * \frac{(133+460)\ \text{R}}{520\ \text{R}} * \frac{760\ \text{mmHg}}{966.8\ \text{mmHg}}$$

= 3,124,542.083 ft^3 /hr

Lbs of gasoline = Reflux + gasoline product

= (7154.0406 + 112.8802) * 79

= 574,086.7432 lb/ hr

Lbs of steam = 34,728.291 lb/ hr

Lbs of gases = 6.8049 lb/ hr

Total lbs = 574,086.7432 + 34,728.291+ 6.8049

= 608,821.8391 lb/ hr

Density of vapor ρ_v = 608,821.8391 (lb/ hr) / 3,124,542.083 ft^3 /hr

= 0.195 lb/ ft^3

$w/\ a = k\ (\rho_v\ (\rho_L - \rho_v)^{0.5}$ ___(p.493)[3]

where k at tray spacing of 22 in for perforated tray, is equal 990 from Fig (16-13) [3].

$\rho_L = \text{sp.gr} * \rho_{water}$

ρ_{water} = 62.34 lb/ ft^3 , sp.gr of gasoline at 50 °F (from Fig(5-14)[3])= 0.678

ρ_L = 0.678 * 62.34 = 42.2665 lb/ ft^3

then, $w/\ a = 990(0.195(42.2665 - 0.195))^{0.5}$

= 2835.61 lb/ hr. ft^2

a = 608,821.8391 (lb/hr) / 2835.61 lb/ hr. ft^2 = 214.7058 ft^2

D = $(\frac{4a}{\pi})$^1/2

= $((4 * 214.7058) / \boldsymbol{\pi})^{0.5}$ = 16.534 ft = 5.04 m

Linear velocity = V/ a * 3600

= 3,124,542.083 / (214.7058 * 3600)

= 4.0424 ft/ s this velocity can be reduced using pump around and the actual tower may never worked at its full capacity [3].

4.7: Bottom and Stripping Temperature (Reduced Crude Outlet Temperature):

Given for 6% of extra vaporization of reduced crude oil and 10 ft of tower height below feed plate, calculate reduced crude oil outlet temperature ;

From eq.(16.2)[3],

$$\Delta t = \text{approx. } 1.2P + 35\left(\frac{D}{B}\right)(T-T_a) + 0.65\left(\frac{S}{B}\right)(T-T_s)$$

P = 6% , D =16.534 ft , B = 95,734.9171 lb/ hr , $T=T_f= 691$ °F

$T_a= 86$ °F , $T_s= 407$ °F , S = 15,140.7999 lb/ hr

$$\Delta t = \text{approx. } 1.2 * 6 + 35\left(\frac{16.534}{95{,}734.9171}\right)(691-86) + 0.65\left(\frac{15{,}140.7999}{95{,}734.9171}\right)(691407)$$

= 40.052 °F

The bottom temperature is = 691 - 40.052

= 650.948 °F

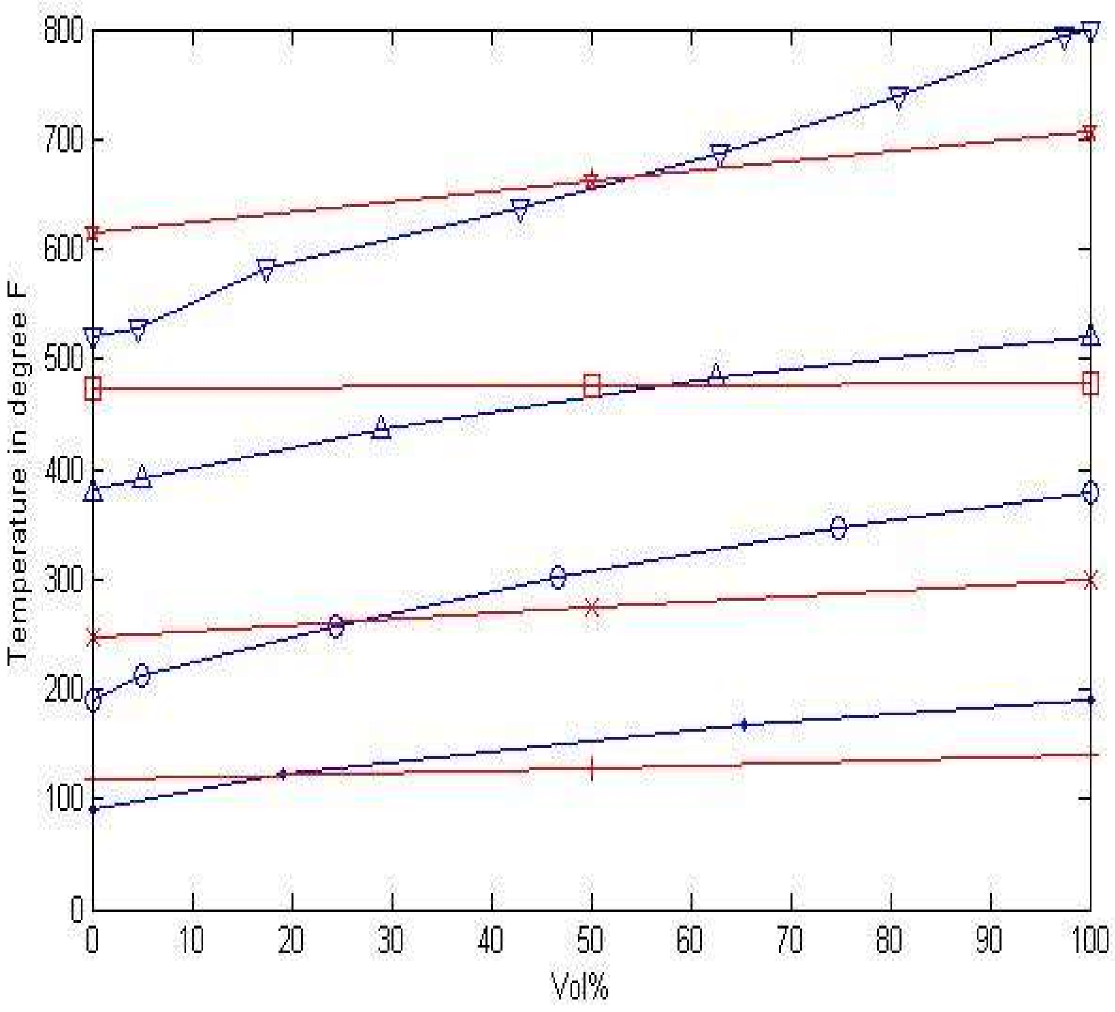


Fig .3 : Distillation Curves of oil Products.

—●— Gasoline TBP curve

—+— Gasoline EFV curve

—○— Naphtha TBP curve

—×— Naphtha EFV curve

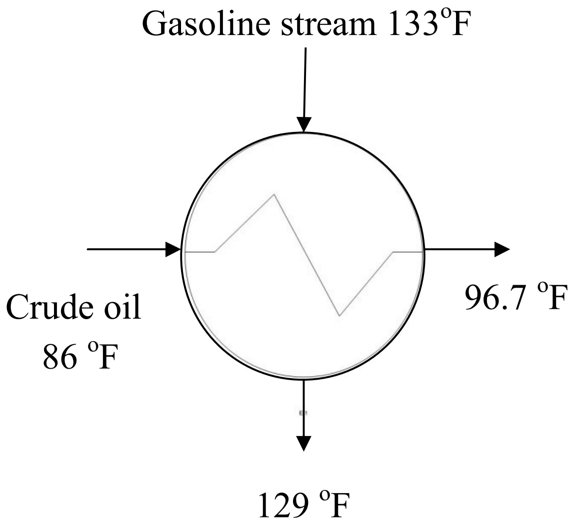
—△— Kerosene TBP curve

—□— Kerosene EFV curve

—▽— Gas oil TBP cruve

—☆— Gas oil EFV curve

5.1: Design of Heat Exchanger (1):



Fixed tube sheet , horizontal heat exchanger.

Heat duty = 3,449,854.246 Btu /hr

Tube Side :

Crude oil 431,231.7808 lb/ hr

Tubes dimensions : carbon steel

d_o= 1.315 in

di = 0.957 in

thickness = 0.179 in

L = 18 ft

Inlet temperature = 86 °F

Shell Side :

Gasoline stream	488,288.6509 lb/ hr
Gasoline	8917.539 + 444,636.016 = 453,553.555 lb/ hr
Steam	34,728.291 lb/ hr
Gases	6.8049 lb/ hr

Inlet temperature = 133 °F

The pounds of reflux and gasoline are 444,636.016 lb/ hr plus 8917.539 lb/ hr or 453,553.555 lb/ hr , also 34,728.291 lb/ hr of steam and 6.8049 lb/ hr of gases .

Using 36.3 ^oF approach on the hot end of the heat exchanger , the heat absorbed by the crude oil between 86 and 96.7 is ,(using Fig(5-3)[3]) ;

431,231.7808 (74 - 39) = 3,449,854.246 Btu/ hr

By trial the temperature of gasoline outlet is 129 ^oF, and 6.6% of gasoline will be cooled and condensed ; (cp calculated from Fig(5-1) and (5-2)[3])

Sensible heat

453,553.555 (1-0.066)(0.565)(133-129) = 957,378.986 Btu/ hr

453,553.555 (0.066)(0.405)(133-129) = 48,493.9461 Btu/ hr

Latent heat

453,553.555 (0.066) (79) = 2,364,828.236 Btu/ hr

Sensible heat for steam

34,728.291(0.5)(133-129) = 69,456.582 Btu/ hr

Total heat = 3,435,157.79 Btu/ hr

This is sufficient close checking the temperature of 129 ^oF .

5.2: Overall Heat Transfer Coefficient:

$\Theta_1 = T_1 - t_2 = 133 - 96.7 = 36.3$ ^oF

$\Theta_2 = T_2 - t_1 = 129 - 86 = 43^oF$

$\Delta T_{lm} = (\Theta_1 - \Theta_2) / \ln (\Theta_1 / \Theta_2)$

$= (36.3 - 43) / \ln (36.3 /43) = 39.555$ ^oF

For two passes , and for p = 0.228 and R= 0.374 , from Fig(17-4)[3],

F = 0.994

Then $\Delta T_{lm} = F * \Delta T_{lm}$

$= 0.994 * 39.555$

$= 39.32\ ^{o}F$

$Q = U\ A\ \Delta T_{lm}$, let U = 10 ,

$3{,}449{,}854.246 = 10 * A * 39.32$

$A = 8773.79\ ft^2$

Area of one tube $= \boldsymbol{\pi}\ d_o\ L$

$= \boldsymbol{\pi}\ (1.315\ /12) * (17.0157)$

$= 5.8579\ ft^2$

Number of tubes $(N_t) = \dfrac{8773.79}{5.8579} = 1497.777 = 1498$ tube

Tube Side :

$T_{av} = \dfrac{86+96.7}{2} = 91.35\ ^{o}F$

The viscosity is obtained from Fig(4-43)[3] , the viscosity of stream = 6.8 cp and the fouling factor from Fig(17-8)[3], is 3.7.

Tube cross sectional area for one tube $= \pi\ d_i^2\ /4$

$= \pi * (0.957\ /12)^2/4 = 0.005\ ft^2$

Tube per pass $= \dfrac{1498}{2} = 749$ tube

Total flow area $= 749 * 0.005 = 3.745\ ft^2$

Mass velocity = $\frac{3{,}449{,}854.246}{3.745*3600}$ = 31.9857 lb/ ft^2. s

film transfer coefficient h_i is obtained from Fig(17-11)[3] is 13 Btu/ ft^2. hr. °F .

Shell side :

Using square pitch (p_t) ,of 1.25 d_o, p_t = 1.25 *(1.315 /12)

= 0.137 ft

Bundle diameter (D_b) =$d_o\left(\frac{N_t}{k1}\right)^{(\frac{1}{n1})}$

where k_1 and $n_1$are constants obtained from table (12.4)[7]

k_1 = 0.249 , n_1 = 2.207 .

$D_b = 1.315*\left(\frac{1498}{0.249}\right)^{(\frac{1}{2.207})}$

= 67.813 in = 5.65 ft

From Fig(12-10)[7], Bundle clearance = 19 mm = 0.748 in = 0.062 ft

Shell diameter (D_s) = 5.65 + 0.062 = 5.712 ft

Baffle spacing (L_B) = 0.5 * D_s = 0.5 * 5.712 = 2.856 ft

Cross flow area = (p_t - d_o)D_s* L_B /p_t

=(0.137 – 0.1096) (5.712)(2.856) /0.137

= 3.263 ft^2

The mass flowrate of the stream = 2,835.61$\frac{\text{lb}}{\text{ft}^2.\text{hr}}$* 214.7058 ft^2

= 608,821.9135 lb/ hr

The mass flowrate ofstream =608,821.9135 (lb/ hr) / 3.263 ft^2

= 162,569.2693 lb/ ft^2. hr

= 45.1581 lb/ ft^2. s

$$T_{av} = \frac{133+129}{2} = 131\ ^{o}F$$

The viscosity is obtained from Fig(4-43)[3] , the viscosity of stream = 0.416 cp and the fouling factor from Fig(17-9)[3], is 0.75.

The film transfer coefficient for the shell $h_o$is estimated from Fig(17-12)[3], is 135 Btu/ ft^2. hr. ^{o}F .

$$U = \frac{1}{\frac{1 d_o}{h_i d_i} + \frac{1}{h_o} + R_D} \_\_\_ (17\text{-}15)[3]$$

Total fouling factor R_D = 0.0037 + 0.00075 = 0.00445

$$U = \frac{1}{\frac{1.315}{13 * 0.957} + \frac{1}{135} + 0.00445} = 8.5029 Btu/\ ft^2.\ hr.\ ^{o}F$$

5.3: Tube Side Pressure Drop :

$$Re = \frac{G\ d}{\mu} = \frac{31.9857*(0.957\ /12)}{6.8 * 0.00672} = 558.22$$

Friction factor are obtained from Fig(13-2)[3], f = 0.006

For total 2 passes ,the friction losses using eq.(13-1)and eq.(17-16)[3] is,

$$\Delta P_f = \frac{2\ f\ G^2\ L}{144\ \ g\ \ D\ \rho}$$

$$= \frac{2*2*0.006*31.9857^2*17.0157}{144*32.2*\left(\frac{0.957}{12}\right)*(295.6957*0.1781\ )} = 0.0215 \text{ psi}$$

$$\Delta P_h = \frac{0.00032\ n\ G^2}{\rho}$$

$$= \frac{0.00032*\ 2*31.9857^2}{295.6957*0.1781} = 0.0124 \text{ psi}$$

Total pressure drop in tube side = 0.0215 + 0.0124 = 0.0339 psi

5.4: Shell Side Pressure Drop :

$$\Delta P_s = 8\ j\left(\frac{D_s}{de}\right)\left(\frac{L}{L_B}\right)\left(\frac{\rho u^2}{2}\right)$$ ___[7]

$$de = \frac{1.27}{do}\left(p_t^2 - 0.785\ d_o\right)$$ ___[7]

$$= \frac{1.27}{(1.315/12)}(0.173^2 - 0.785\ (1.315/12)^2)$$

$$= 0.1083 \text{ ft}$$

$$Re = \frac{G\ d}{\mu}$$

$$= \frac{(45.1581*0.1083)}{0.416*0.000672} = 17{,}494.499$$

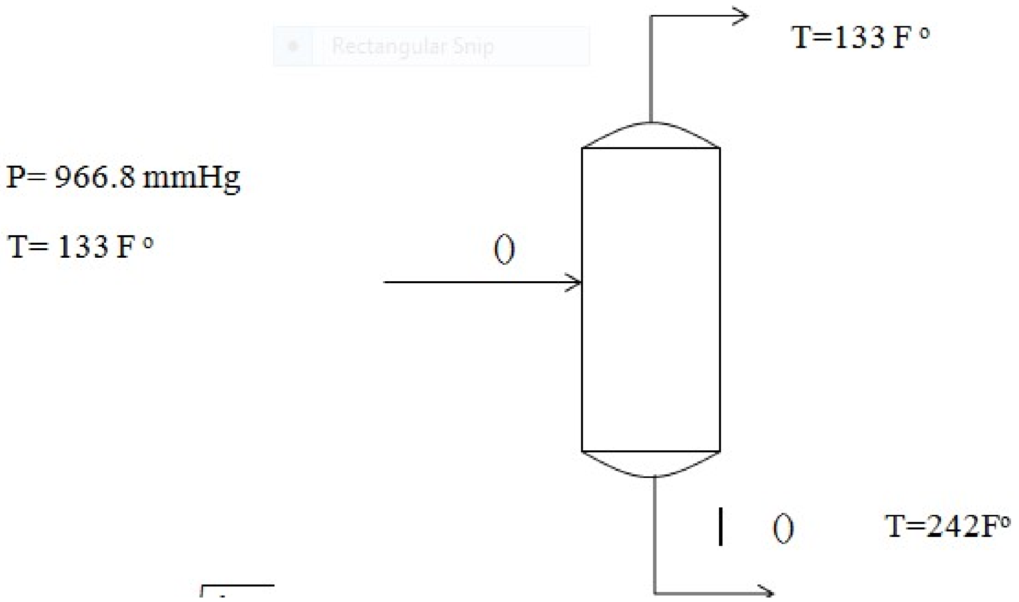
j from Fig(13-2)[3], j =0.0007

$$\Delta P_s = 8\ 0.0007\left(\frac{5.712}{0.1083}\right)\left(\frac{17.0157}{2.856}\right)\left(\frac{42.2665*\left(\frac{35.3146}{2.2046}\right)(1.0684/3.2802)^2}{2}\right)$$

$$= 63.1971 \text{ pa}$$

$$= 0.00917 \text{ psi}$$

Design of stripping



$U_{perm} = k_{drum} \sqrt{(\rho l-\rho v)/\rho v}$

$K_{drum} = \exp[A+B \ln Flv +C (\ln Flv)2 +D (\ln Flv)3 + E (\ln Flv)4]$

$Flv = Wl/(WV) \sqrt{\rho v/\rho l}$

$Flv = 542299.1690/574086.7432 \sqrt{0.0278/0.04787} = 0.7198$

$K_{drum} = \exp [-1.8774 - 0.8145 (0.7198) - 0.1870 (0.7198)2 - 0.0145 (0.7198)3 - 0.001 (0.7198)4]$

$K_{drum} = 0.0768$

$U_{perm} = k_{drum} \sqrt{(\rho l-\rho v)/\rho v}$

$U_{perm} = 0.0768 \sqrt{(0.04787-0.0278)/0.0278}$

$U_{perm} = 0.8496$ ft2/hr

$A_c = (V*(MW)_v)/(U_{perm}*\rho_v*3600)$

$A_c = (4060.438*78)/(0.8496*0.0278*3600)$

$A_c = 3724.823$ ft2

$D = \sqrt{(4A_c/\pi)}$

$D = \sqrt{((4*3724.823)/3.14)}$

$D = 68.86$ ft

$h_v = 0.9144 + 0.5 *D$

$h_v = 35.344$ ft

$h_f = 0.4064 + 0.5 * D$

$h_f = 43.8364$ ft

$h_L = (V_{surge})/((\pi*D^2)/4)$

$L_{flow\ rate} = 54299.169$ Ib/hr ÷ 3600

$= 15.083$ Ib/ s ÷ ρ_l Ib/ft3

$= 15.083$ Ib/ s ÷ 0.04787

$= 315.0825$ ft3/s $*$ 300 s

$V_{surge} = 94524.75$ ft3

$h_L = (94524.75)/((3.14*{68.86}^2)/4)$

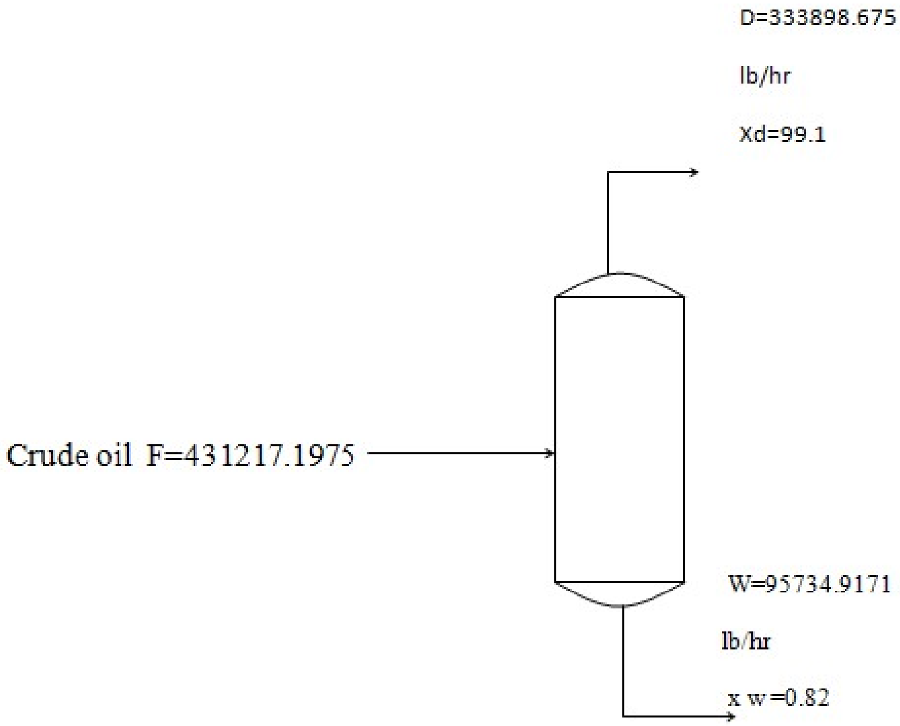
$h_L = 25.395$ ft

$H_{total} = h_v + h_f + h_L$

$H_{total} = 35.344 + 34.8364 + 25.395$

$H = 95.5754$ ft $D = 68.86$ ft

Design of distillation column



Calculation of equilibrium data:

$$In\ ^{0} = A - \frac{B}{T + C}$$

Where:

$p^{0} = vapo\ \ prssure\ (mmHg)$

$$A, B, C = constant$$

$T = Temperature\ (K^{o})$

$$Inp^0{}_W = 16.3872 - \frac{3885.70}{341.3 + 230.170} \Rightarrow p^0{}_W = 14584.54\, mmHg$$

$$Inp^0{}_M = 16.5785 - \frac{3638.27}{341.3 + 239.5} \Rightarrow p^0{}_M = 30159.67\, mmHg$$

$$\alpha_{top} = \frac{p^0{}_M}{p^0{}_W} = \frac{30159.67}{14584.54} = 2.06$$

At Bottom T=110C^o = 383K^o

$$Inp^0{}_W = 16.3872 - \frac{3885.70}{383 + 230.170} \Rightarrow p^0{}_W = 23158.84\, mmHg$$

$$Inp^0{}_M = 16.5785 - \frac{3638.27}{383 + 239.5} \Rightarrow p^0{}_M = 45884.81\, mmHg$$

$$\alpha_{bot} = \frac{p^0{}_M}{p^0{}_W} = \frac{45884.81}{23158.84} = 1.98$$

$\alpha_{av} = (\,\alpha_{top} * \alpha_{bot})^{0.5}$ [10]

$\alpha_{av} = (2.06 * 1.98)^{0.5} = 2.02$

$Y = \frac{\alpha_{av} * X}{1+(\alpha_{av}-1)*X}$ [10]

$Y = \frac{2.02 * X}{1 + 2.02 * X - X}$ $equilibrium\ curve$

At X=0.10

$$Y = \frac{2.02 * 0.1}{1 + 2.02 * 0.1 - 0.1} = 0.183$$

X	Y
0.000	0.000

0.100	0.168
0.200	0.335
0.300	0.464
0.400	0.573
0.500	0.668
0.600	0.751
0.700	0.824
0.800	0.889
0.900	0.947
1.000	1.000

Calculation of q-line

$$q = \frac{\text{heat reguired to vaporized 1K mol of feed}}{\text{molar latent heat of feed}}$$

$$Q = \frac{C_{pmix} * T \text{ feed} + \lambda mix}{\lambda mi}$$

where:

C_{pmix}= average molar heat capacity of feed KJ/Kmol. °C

$\lambda mix =$ Average molar latent heat of feed KJ/Kmol

$$\Delta T = feed\ bubble\ point - feed\ temparature°\mathrm{C}$$

$C_{pmix} = C_{p1} * X_{f1} + C_{p2} * X_{f2} + C_{p3} * X_{f3}$

$C_{pmix} = (87.354*0.0541) + (75.4*0.5744) + (37.19*0.3715)$

$C_{pmix} = 61.85\ KJ/Kmol\ .C^{o}$

$$\lambda mi = \lambda 1 * Xf1 + \lambda 2 * Xf2 + \lambda 3 * Xf3$$

$= (35246.2*0.0541) + (40626*0.5744) + (12187.31*0.3715)$

$= 29769.9\ KJ/Kmol$

$\therefore Feed\ bubble\ point = 102℃$

$q = \frac{61.85*102+2976\ .9}{29769.9}$ =1.21

Slop of q-line $= \frac{q}{q-1} = \frac{1.21}{0.21} = 5.76$

$V_1 = R + D$

ASSUM R / D = 1.3

R = 1.3 D

V_1 =1.3 D +D D = 1.39$kmol/hr$

V_1 = 3.197 $kmol/hr$

R = 1.807 $kmol/hr$

Calculation number of plates:

- Draw the equilibrium data X vs Y.
- Draw the q-line with slope = $\frac{q}{q-1} = \frac{1.21}{0.21} = 5.76$
- Draw the top of operating line from X_D on the diagonal
- $Q = \frac{X_d}{R+1} = \frac{0.903}{1.807 +1} = 0.32$
- Draw the bottom operating line from X_B on the diagonal tp the point of intersection of the operating line and the q-line.
- Starting at X_D step off the number of stages.

  From plot No. of stage =21
  No. of plate = No. of stages – 1
  =21 – 1 = 20
  E_0= plate efficiency
  Viscosity (μ) = 0.1329
  Relative volatility = 0.0709

$$E_0 = 0.52782 - 0.27511 * \log(\propto \mu) + 0.044923 * [\log(\propto \mu)]^2$$

$$= 0.52782 - 0.27511 * \log(0.09432) + 0.044923 * [\log(0.09432)]^2$$

= 0.8573 =85.7%

Actual Number of stages =$\frac{21}{0.8573}$= 25 stages

COLUMN HEIGHT

The column height is heavily dependent upon the spacing between the sieve trays. In our design, 18 inches were chosen for spacing to provide a reasonable space to ease the accessibility for manual workers to crawl between the plates for maintenance. According to Turton's Distillation Column Design Heuristics (1955), a safety factor of 10% is to be added to the final design height. The column height is determined as follows.

Total Actual number of stages= 25 stage

Safety Factor = 25*(0.1) = 2.5stage

Total Construction stages = 2.5+25+1 ≈ 29 STAGE including reboiler

Column Height = Tray Spacing * (Num. of stages + safety factor)

$$= 18\text{inch}\frac{0.0254\,m}{1\,inch} *(25+2.5)$$

$$=12.57\text{ m}$$

COLUMN DIAMETER

Rectifying (TOP) Section Diameter

The first step is the determination of the **flow parameter** as follows

$$F_P = \frac{Ln}{Vn} * (\frac{\rho v}{\rho L})^{0.5}$$

$$F_P = 0.9958 * (\frac{0.08259}{51.35})^{0.5}$$

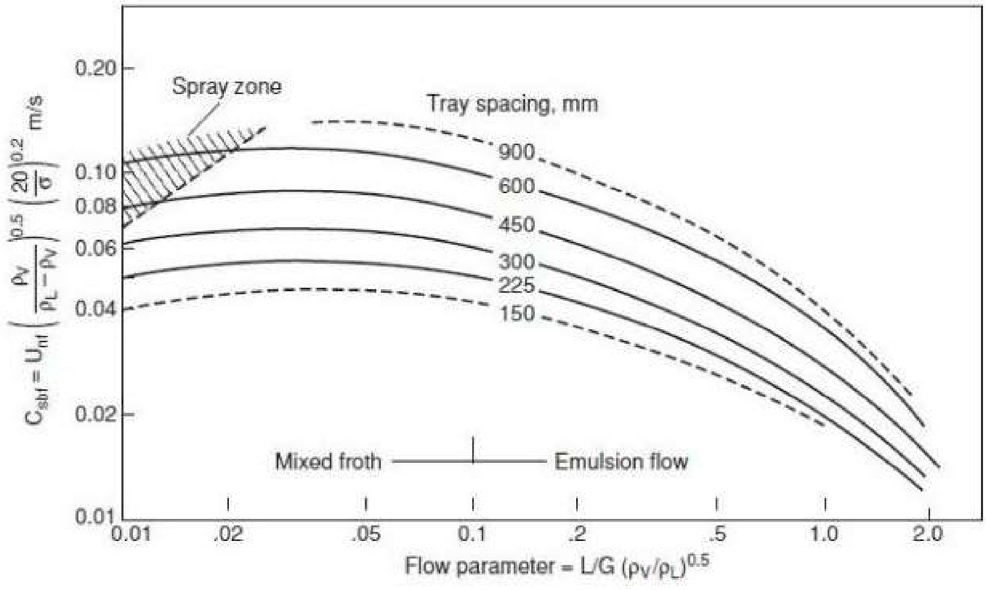
$F_P = 0.03993$

18 inch tray spacing is to be used as moderate average of the capacity factor of flooding. Utilizing a nonlinear regression of the capacity factor chart by Kessler and *Wankat*(1987) as follows

$\text{Log}C_{sb,f} = -\ 1.0262 - 0.63153 * \log F_P - 0.20097(\log F_P)^2$

$\text{Log}C_{sb,f} = -\ 1.0262 - 0.63153 * 0.03993 - 0.20097(0.03993)^2$

$C_{sb,f} = 0.2994\ ft/s$This is correlated by the following chart



Then, the operation velocity is calculated as follows

$$\text{Uflood} = C_{sb,f}\ (\frac{\sigma}{20})^{0.2}\sqrt{\frac{\rho l\quad \rho v}{\rho v}}$$

$$\text{Uflood} = 0.2994\ (\frac{25}{20})^{0.2}\sqrt{\frac{51.35-0.08259}{0.08259}}$$

$$\text{Uflood} = 7.6705\ ft/s$$

From mass balance

$$v = 3.197\frac{kmole}{hr} = 2519.9\ Ibmole/hr$$

According to *Wankat*(1987), the fraction of flooding that is utilized by the operational velocity is ranging between 0.65 and 0.9. *Jones* and *Mellbom*(1982) suggested an average fraction of 0.75.

As for the fraction of cross-sectional area that is available for vapor flow η, *Wankat*(1987) presented a rage of 0.85 and 0.95. An average of η=0.9 is to be used in our design

Diameter sizing of the top section

$$\text{Dia} = \sqrt{\frac{4 * V * R * T}{\pi * \eta * 3600 * P * Faraction * \text{Uflood}}}$$

$$\text{Dia} = \sqrt{\frac{4 * 2519.9 * 1.314 * 349.4}{\pi * 0.9 * 3600 * 1 * 0.75 * 7.6705}}$$

Dia = 8.115 ft = 2.474 m

Stripping (BOTTOM) Section Diameter

$$F_P = \frac{Ln}{Vn} * \left(\frac{\rho v}{\rho L}\right)^{0.5}$$

$$F_P = 7.9039 * \left(\frac{0.04571}{46.77}\right)^{0.5}$$

$F_P = 0.24709$

$\text{Log}C_{sb,f} = -\ 1.0262 - 0.63153 * \log F_P - 0.20097(\log F_P)^2$

$\text{Log}C_{sb,f} = -\ 1.0262 - 0.63153 * 0.24709 - 0.20097(0.24709)^2$

$C_{sb,f} = 0.1929\ ft/s$

Molecular Weight

Rectifying Section: $MW_T = X_M * MW_M + X_F * MW_F + X_W * MW_W$

$MW_T = 0.00017 * 32.042 + 0.395 * 30.02 + 0.6047 * 18$

$MW_T = 22.7479$

Stripping Section: $MW_B = X_m * MW_m + X_W * MW_W$

$MW_B = 0.903 * 32.042 + 0.0961 * 18$

$MW_B = 30.66$

$$\frac{(Dia\ top)}{(Dia\ bot)} = \left(\frac{Vtop}{Vbot}\right)^{0.5} * \left(\frac{\mathrm{MWtop}}{\mathrm{MWbot}}\right)^{0.25} * \left(\frac{Ttop}{Tbot}\right)^{0.25} * \left(\frac{\sigma top}{\sigma bot}\right)^{0.1} * \left(\frac{\mathrm{Csb,top}}{\mathrm{Csb,bot}}\right)^{0.5} * \left(\frac{\rho l\ top}{\rho l\ bot}\right)^{0.25}$$

$$\frac{(Dia\ top)}{(Dia\ bot)} = (0.99)^{0.5} * \left(\frac{30.66}{22.7479}\right)^{0.25} * \left(\frac{341.3}{383}\right)^{0.25} * \left(\frac{37.23}{25}\right)^{0.1} * \left(\frac{0.1929}{0.2944}\right)^{0.5} * \left(\frac{46.77}{51.35}\right)^{0.25}$$

$$\frac{(Dia\ top)}{(Dia\ bot)} = 0.87789$$

Diameter of bottom $= \frac{Dia\ top}{0.87789} = \frac{2.474}{0.87789} = 2.81\ m$

Abot$= \frac{\pi * (Diabot)^2}{4} = \frac{\pi * (2.81)^2}{4} = 6.19\ \mathrm{m}^2$

Columncross-sectionalarea

$A_c = \frac{\pi * (Dia\ top)^2}{4} = \frac{\pi * (2.474)^2}{4} = 4.8$ m

Down comer area

$Ad = (1-\eta)A_c$

$Ad = (1-0.9)\ 4.8$

$Ad = 1.5764\ \mathrm{m}^2$

Value of is chosen 0.1 according to Wankat (1987) as a common standard of the relation between the weir length and diameter.

The $\frac{\text{WEIR LENGTH}}{Dia\ top}$ ratio is provided by Wankat (1987) as 0.726

$I_{weir} = Dia\ top\ ^{*} 0.726 = 2.474 * 0.726$

$I_{weir} = 1.796$ m

Active area

$A_a = A_c - 2A_d = 4.8 - 2*1.5764 = 1.647\ m^2$

Hol area

$A_h = 10\% A_a$ as first trial

$A_h = 0.1 * 1.647 = 0.1647\ m^2$

5-8 location and site considerations:

5-8-1 Select plant site Plant Location:

The concept of choosing plant site:

(1) Find the plant, which established for the first time site.

(2) or plant originates along with other factories belonging to one company.

(3) Selection of location of the company's stores whether these stores of materials processed or semi-processed or raw materials the fact that the storage is complementary to productive activity.

The decision to choose the location of the important and difficult decisions faced by the industrial company because:

(1) the magnitude of employee financial investments in the factories of the newly established or in the old factories size.

(2) This decision is linked to long-term strategies that affect the future of the company and its success to and from these strategies largely processing of raw materials strategy, marketing strategy, and strategic storage.

5-8-2 The goals of the site selection:

Strategic goal depends on the type of industry (goods or services) Strategic objective for companies producing goods is to reduce the following costs:

1. The cost of the site include the purchase of land, the cost of construction and reconstruction, and the cost of hiring workers, and government fees and taxes for that site.

2. The cost of distributing products: include costs associated with the transfer and distribution of products from the company's factories to its

stores, and then from the warehouse to retail outlets within the geographical distribution network of the company

3. The initial cost of materials: the costs associated with a purchase price of raw materials involved in the production process and availability, and the cost of buying the operator of the plant electrical energy such as energy, oil and coal.

5-8-3 Differentiation factors in the selection of the plant site Passes choose:

factory site several steps starting from selecting a country and then select the region and finally choose the region. Vary the importance of each variable of variables that influence the choice of the location of the industry to another, and generally on the company to reduce these factors by excluding factors or variables that do not meet one of the following conditions (Krajewski, Ritzman & Malhotra, 2007: 421-422):

1. have a variable effect is obvious and anisotropic when the trade-off between the number of sites.

2. have a variable degree of obvious importance among other variables to take into account.

5-8-4 Factors influencing the decision to choose the site:

1. labor productivity: This is an important factor for the private service organizations, where the owner is a factor center stage for her. Where it should take into account the low wages and labor productivity (labor productivity = daily production rate / daily wage labor).

2. Proximity to markets proximity to Markets: and also be important for the organization of production in the case of high transport costs for products heavy Kalmakinat perishable and breakable or goods due to production in large quantities.

3. Proximity of processors and resources proximity to suppliers & Resources For raw materials, perishables such as meat and dairy or high transport such as the establishment cement factory near the stone quarries or create a paper plants near the forest parties or canned vegetables produced near the farm costs, but the multiplicity of raw materials for the

plant one makes the possibility of proximity to all of them impossible, so you must choose the site that takes into account the multiplicity primary sources.

4. competitors site Location of Competitors: To find out the current competitors' sites as well as their reaction to the new site Dimension from competitors areas is important, But it may be the proximity of them important in some cases as an opponent selling cars and fast food restaurants, where the strategy seeks to create a so-called critical density Critical Mass and meaning of the existence of several competing companies at a single site attracts customers more than the total number of customers who shopped at the same stores in similar areas . In addition, some enterprises prefer to adopt a strategy of "follow the leader Follow-the-leader".

5. Political risk, values and culture, Political Risk, Values, Culture: Site selection is influenced by local governments look toward intellectual property rights and environmental pollution and unemployment, as well as workers' attitudes toward labor turnover and union workers and absence from work. These values also affect when using the same workers when they decide to find a new location for its facilities productivity is one of the big challenges for the resolutions of international operations and dealing with different cultures about the company's culture such as lack of commitment to employees working hours or not processors obligation to extradite which significantly affect production schedules and delivery and creates corruption administrative.

6. exchange rates and currency risk Exchange Rates & Currency risk Wage rates may seem in some encouraging states to create a new location for one of the companies, however, the exchange rate may deny this option and make the site uneconomic The company may benefit from exchange rates of encouragement for the currency by creating a site in such states or by exporting them.

5-9 SAFETY:

What is the concept of industrial safety?

It is the protection of human hazards and prevent the loss of property of materials, equipment and facilities, and by providing preventive services and procedures and provide precautionary and protective equipment for the purpose of doing business properly and safely without incident.

5.9.1. Industrial safety goals:

1. protect workers from occupational hazards

2. devices and equipment protection.

3. protection of production materials.

The foundations of industrial safety:

1. controls the entry and exit to and from the industrial facility.

2. banning smoking in industrial areas.

3. Prompt reporting of accidents, injuries and fires.

4. Remove or bury oil waste.

5. Remove the dry grass.

6. Provide industrial safety equipment.

Industrial safety controls:

1. fencing of industrial units and isolate warehouses, workshops and offices for operational units.

2. Put the number of emergency exits, preferably one exit (at least) on each side.

3. Provide barriers for all rotating parts.

4. Providing blankets and cushions to dig iron complex valves and distribution pipes.

5. restraints iron status of the existing lanes at high altitudes.

6. Put corridors above the regular stripping towers.

7. Select the operating safety of emergencies limits.

8. Provide sensing systems to toxic gases and gases inflamed.

Duties of safety personnel:

1. To conduct inspections to detect cases of improper and processed.

2. give a warm and cold work permits.

3. Provide safety devices and tested constantly.

4. train workers on the use of safety devices.

5. Measuring the concentrations of hazardous gases continuously in an industrial facility.

6. Issuing instructions and safety regulations.

7. Put the emergency phones in prominent places.

8. status of persons and equipment warning signs.

Industrial accidents and their consequences The implementation of the industrial business in installations in hazardous conditions punctuated by the occurrence of many incidents which can be identified as follows:

1. fires.

2. explosions.

3. hardware malfunction.

4. Personal Accident.

5. occupational diseases.

That any kind of incidents mentioned above leads to the occurrence of many of the losses, including material and moral ones, and can be summarized as follows:

1. damage and malfunction of equipment and devices.
2. damage to materials produced raw materials.
3. wasting working hours.
4. insurance and compensation costs.

Chemical hazards Chemicals play a major role in the life of individuals and peoples to become the well-being and progress of peoples is measured including findings of the discovery and use of chemicals in the various spheres of life, the use of a weapon of chemicals edged if better use were expressing face shining and beneficial to mankind. If misused these materials, they reveal the ugly face that causes the destruction of humanity and wasted lives of individuals and there is a chemical in the work environment in one of the following images:

- Gases and vapors.
- Dust (organic -gar membership).
- Liquids (acids - alkali - solvents).

It is therefore the chemicals of greatest and most dangerous human faces for many reasons, among them the following:

- Chemicals that take more than one form of it resides on the image (liquid - gas - solid).
- To enter into the human body rapidly and by the ability of (respiratory and digestive and skin contact).

- The impact on the body is interaction with some members of the body and thus affect it adversely.

- The degree of acute effect which results from the body of this article may occur immediately upon entry to the body or occur after a period of time of the discovery.

Some of these materials It has no taste and no smell or color and therefore it is difficult for the human damage or speed detection.

- Effects may occur in some of the equipment work like rust or corrosion and explosion and self-fire Occupational safety and health conditions to be met to protect workers from the risks of chemicals:

1. You must provide precautions to protect workers exposed to the risk of exposure to chemicals used, whether the material in gaseous, liquid or solid and make it within the permissible limits

2. must be performed primary medical examination on workers when they enter the work exposes them to the risks of chemicals to detect any satisfactory phenomenon or underlying condition that affects the workers strongly when exposed to chemical contaminant and retains the outcome of the medical examination worker file to compare the results of these tests.

3. must undergo medical lobe league on workers exposed to the risks of chemicals to detect any occupational disease as a result of early exposure to them and to ensure the continued fitness and medical workers to the nature of the work.
4. You must provide effective for the prevention of harmful.

5. must make the necessary periodic measurements of chemical hazards in the work environment depending on the type of activity sundials and

recorded and compared periodically. To make sure they are within the permissible limits.

6. You must provide personal protective equipment for workers, which are commensurate with the nature of the work that they do and be matched with the technical specifications of exposure.

7. should provide adequate water for washing or bathing for workers after school ends and before they left the workplace to remove the body hanging from a chemical harmful pollutants while providing hygiene equipment (such as soap, towels, etc.).

8. You must provide a special place to switch the workers wear clothes to work or vice versa depending on the nature of the work that these places are far away from the workplace and must provide places to eat workers for the food away from the workplace (exposure) and prevents eat, drink or smoke in the workplace.

9. must be made aware of the dangers workers in the work environment of chemicals and how to protect themselves from them. And a commitment to alerts and warnings issued by the companies producing the chemicals.

10. provide adequate ventilation inside the stores.

11. When you spill any flammable materials on your clothes or any of the parts of your body, it is obligatory for you to use a stream of water on the subject of infection with the speed of disposal of contaminated

clothing and not to approach the flame places open in order to prevent further injury, and the reduction of gravity.

12. hydrochloric acid and nitric and sulfuric acid from the special qualities of liquid chemicals must therefore be taken into account when stored or handled.

13. prevent the entry of non-professionals into the chemical store and censoring places stored is very important.

14. Follow the instructions and handed the receipt of the chemicals in the stomach Bathbadtha records so to combat loss and loss is very important.

15. provide the primary means of firefighting and training on how to use the safety precautions to be followed.

16. You must not use the senses of touch or smell or taste in the identification of chemical substances.

17. ought to keep flammable materials in cool places away from electrical equipment sparks or heat sources.

5-10Economic and costs:

Height h = 95.5754ft

Diameter D = 68.86ft

Pressure p = 966.8mmHg

Material of construction: carbon steel

Cost= 140 $ Fig 6.5b(App.A)
[[55

Pressure factor =5

Material factor =2

Cost = bare cost from fig.*material factor*pressure factor

(Cost =140 * 5* 2= 1400$ in(2004

Cost index in 2004 = 480 [66

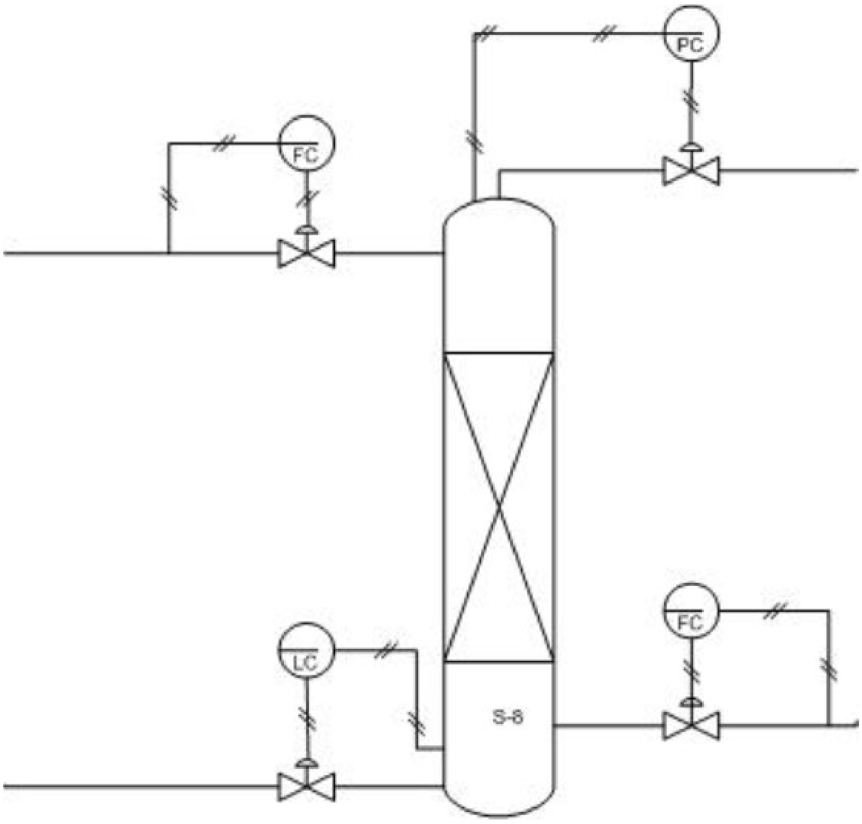
Cost index in 2016 = 1081

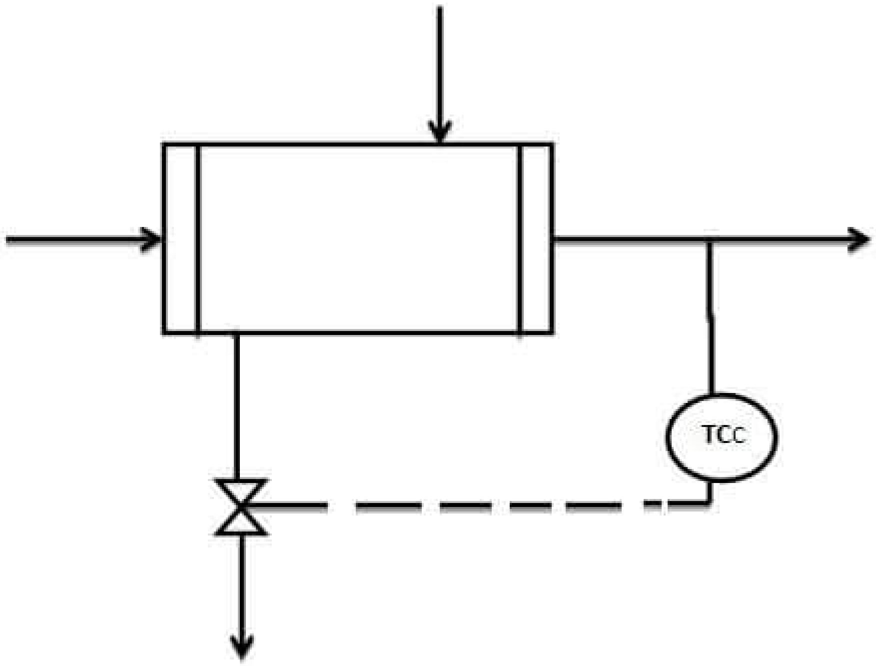
Cost in 2016 = 1400 * (1081/480

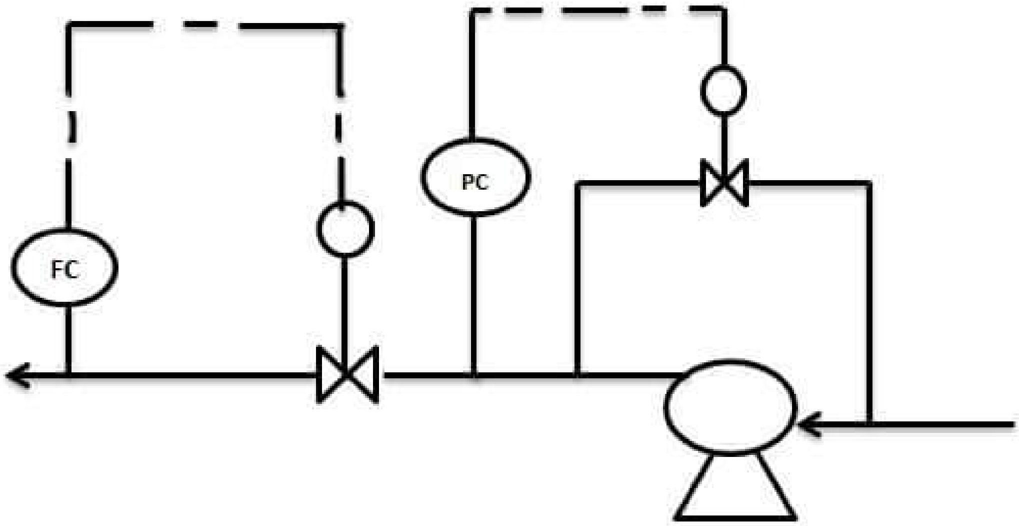
$٣١٥٠=

Control Loop System

For the control ability of the absorber four different loops will be added to the process. The first one will be added to the inlet of the liquid and gas to control the flow rate. The second one will be added to the gas outlet to control the pressure of the absorber. The third one will be added to the liquid outlet to control the level .The fourth one will be added to the liquid outlet to control the .level as in Figure




TCC


PC
FC

References:

[1] Mohamad A. Fahim, Taher A. Al-sahhaf and Amal Elkilani , “Fundementals Of Petroleum Refining”, 1st Edition, 2010.

[2] James and Gary, “Petroleum Refining Technology And Economics”, 2nd Edition, 1984.

[3] W.L. Nelson, “Petroleum Refining Engineering”, 4th Edition, McGRAW-Hill, 1973.

[4] R.N. Watkins, “Petroleum Refinery Distillation”, 2nd Edition, 1979.

[5] J.P.Holman, “ Heat Transfer”,6th Edition.

[6] “Perry’s Chemical Engineering Handbook”,7th Edition.

[7] Coulson and Richardson's, “Chemical Engineering’’, Vol.6, 3rd Edition.

[8] Dr. Ramgopal Uppaluri, “Refinery Process Design (Lecture Notes)” , March 2010.

CRUDE PETROLEUM ANALYSIS

Bureau of Mines Bartlesville Laboratory
Sample 54060

IDENTIFICATION

Bayou des Allemands field
Miocene

Louisiana
Lafourche Parish

GENERAL CHARACTERISTICS

Gravity, specific, 0.845
Gravity, ° API, 36.0
Pour point, ° F., 35
Sulfur, percent, 0.20
Color, brownish green
Viscosity, Saybolt Universal at 100°F., 49 sec.
Nitrogen, percent, 0.040

DISTILLATION, BUREAU OF MINES ROUTINE METHOD

STAGE 1—Distillation at atmospheric pressure, 743 mm. Hg
First drop, 86 ° F.

Fraction No.	Cut temp. ° F.	Percent	Sum, percent	Sp. gr., 60/60° F.	° API, 60° F.	C. I.	Refractive index, n_D at 20° C.	Specific dispersion	S. U. visc., 100° F.	Cloud test, ° F.
1	122	0.5	0.5	0.670	79.7					
2	167	1.2	1.7	.675	78.1	11				
3	212	1.6	3.3	.722	64.5	23	1.39163	137.0		
4	257	2.7	6.0	.748	57.7	26	1.41725	141.7		
5	302	3.1	9.1	.765	53.5	26	1.42648	142.3		
6	347	3.9	13.0	.778	50.4	25	1.43374	140.7		
7	392	4.7	17.7	.789	47.8	24	1.43962	138.0		
8	437	5.7	23.4	.801	45.2	24	1.44529	137.6		
9	482	8.0	31.4	.814	42.3	25	1.45193	137.4		
10	527	10.7	42.1	.825	40.0	26	1.45884	142.9		

STAGE 2—Distillation continued at 40 mm. Hg

Fraction No.	Cut temp. ° F.	Percent	Sum, percent	Sp. gr., 60/60° F.	° API, 60° F.	C. I.	Refractive index, n_D at 20° C.	Specific dispersion	S. U. visc., 100° F.	Cloud test, ° F.
11	392	5.0	47.1	0.845	36.0	31	1.46614	142.6	40	15
12	437	10.0	57.1	.854	32	32	1.46870	139.9	45	30
13	482	7.8	64.9	.863	32.5	33	1.47403	140.4	56	50
14	527	7.0	71.9	.874	30.4	35			81	65
15	572	6.5	78.4	.889	27.7	39			145	85
Residuum		20.8	99.2	.931	20.5					

Carbon residue, Conradson: Residuum, 3.7 percent; crude, 0.8 percent.

APPROXIMATE SUMMARY

	Percent	Sp. gr.	° API	Viscosity
Light gasoline	3.3	0.697	71.5	
Total gasoline and naphtha	17.7	0.759	54.9	
Kerosine distillate	24.4	.816	41.9	
Gas oil	14.2	.850	35.0	
Nonviscous lubricating distillate	14.1	.858-.878	33.4-29.7	50-100
Medium lubricating distillate	8.0	.878-.895	29.7-26.6	100-200
Viscous lubricating distillate	-	-	-	Above 200
Residuum	20.8	.931	20.5	
Distillation loss	.8			

U. S. GOVERNMENT PRINTING OFFICE 16—57835-3