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Refinery Process Design

(Lecture Notes)

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Acknowledgements

From my mature childhood, I always had a very passionate desire to venture into slightly newer areas and explore my professional capabilities in these fields. This passionate desire fructified after I met Dr. T. D. Singh (Founding Director, Bhaktivedanta Institute), a grand visionary for the synthesis of science-spirituality. His exemplary life, dedication to the mission day and night always inspired me to set difficult to address targets in life and achieve them. Apart from my fundamental interests in science-spirituality which is my first exploration away from process engineering professional life, Refinery Process Design is my second intellectual offshoot.

The foremost acknowledgement for the Lecture Notes goes to the Almighty God, who provided me the desired intelligence and attributes for delivering the Lecture Notes.

Next acknowledgement for the Lecture Notes goes to the Department of Chemical Engineering, IIT Guwahati which has been my professional father and mother in which I am there for the past six years enjoying my every moment in personal as well as professional activity. IIT Guwahati has truly inspired me to write the Lecture notes due to variety of reasons. To name a few, the early heat exchanger network consultancy projects conducted with IOCL Guwahati and BRPL Bongaigaon have been pretty useful to develop insights into the nature of refinery processes and streams. The ideal setting of IIT Guwahati in the midst of petroleum refineries really inspired me to deliver the lecture notes for professional usage too!

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Finally, I wish to acknowledge my mother and father, as they have been the fulcrum of my continued pursuits in chemical engineering practice for the past fifteen odd years. Last but not the least I acknowledge my wife whose patience and support during long hours I dedicated for the development of the lecture notes is very much appreciated.

Date: 16th March 2010

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Preface

To draft the lecture notes in the specialized topic of Refinery Process Design needs expertise in both fields of petroleum refining and process design. The basic urge of extending my process design specific skills to the design of petroleum refineries started in IIT Guwahati, and specifically after thoroughly reading Jones and Pujado (2006) book which I feel should be the primary reference for refinery design. The primary objective of the lecture notes is to provide a very simple and lucid approach for a graduate student to learn upon design principles. While Jones and Pujado (2006) summarized many calculations in the form of Tables, for better explanations I always felt that calculations should be conducted and explained as well. With this sole objective I drafted the Lecture notes.

Consisting of three main chapters, this lecture notes enables a graduate chemical engineer to learn all necessary content for maturing in the field of refinery process design. In the very first chapter, a very introductory issue of relating the subject matter of refinery process design to conventional chemical process design is presented from a philosophical perspective. Thereby, the second chapter presents a detailed discussion of the estimation of essential refinery stream properties. The third chapter attempts to provide a new approach for conducting refinery mass balances using the data base provided by Maples (2000). The fourth chapter deliberates upon the design of crude distillation unit with further elaborates towards the selection of complex distillation reflux ratio, pump around unit duty estimation and diameter calculations.

All in all, the lecture notes did not present some elementary design related chapters pertaining to the design of heat exchanger networks, design of light end units and design of refinery process absorbers. These will be taken up in the subsequent revisions of the Lecture notes.

I hope that this lecture notes in Refinery Process Design will be a useful reference for both students as well as practicing engineers to systematically orient and mature in process refinery specialization.

Date: 16th March 2010

(**Dr. Ramgopal Uppaluri**)
IIT Guwahati

Dedication

To

Dr. T. D. Singh

Founder Director, Bhaktivedanta Institute

Short biography of Dr. T. D. Singh:

Dr. T. D. Singh received his Ph.D. in Physical Organic Chemistry from the University of California at Irvine, USA in 1974. He was a scientist and spiritualist well-known for his pioneering efforts in the synthesis of science and religion for a deeper understanding of life and the universe. He was trained in Vaishnava Vedanta studies from 1970 to 1977 under Srila A. C. Bhaktivedanta Swami and was appointed Director of the Bhaktivedanta Institute in 1974. He organized four major International conferences on science and religion - First and Second World Congress for the Synthesis of Science and Religion (Mumbai, 1986 & Kolkata, 1997), First International Conference on the Study of Consciousness within Science (San Francisco, 1990) and Second International Congress on Life and Its Origin (Rome, 2004). Collectively thousands of prominent scientists and religious leaders including several Nobel Laureates participated. He authored and edited more than a dozen books including What is Matter and What is Life? (1977), Theobiology (1979), Synthesis of Science and Religion: Critical Essays and Dialogues (1987) and Thoughts on Synthesis of Science and Religion (2001). He was also the founder Editor-in-Chief for two journals of the Bhaktivedanta Institute.

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1. Introduction

1.1 Introduction

Many a times, petroleum refinery engineering is taught in the undergraduate as well as graduate education in a technological but not process design perspective. While technological perspective is essential for a basic understanding of the complex refinery processes, a design based perspective is essential to develop a greater insight with respect to the physics of various processes, as design based evaluation procedures enable a successful correlation between fixed and operating costs and associated profits. In other words, a refinery engineer is bestowed with greater levels of confidence in his duties with mastery in the subject of refinery process design.

1.2 Process Design Vs Process Simulation

A process design engineer is bound to learn about the basic knowledge with respect to a simulation problem and its contextual variation with a process design problem. A typical process design problem involves the evaluation of design parameters for a given process conditions. However, on the other hand, a typical process simulation problem involves the evaluation of output variables as a function of input variables and design parameters.

For example, in a conventional binary distillation operation, a process design problem involves the specification of Pressure (P), Temperature (T), product distributions (D, B, x_B, x_D) and feed (F, x_F) to evaluate the number of stages (N), rectifying, stripping and feed stages (N_R, N_S, N_F), rectifying and stripping section column diameters (D_R and D_S), condenser and re-boiler duties (Q_C and Q_R) etc. On the other hand, a process simulation problem for a binary distillation column involves the specification of process design parameters (N_R, N_S, N_F) and feed (F, x_F) to evaluate the product distributions (D, x_D, B, x_B). For a binary distillation system, adopting a process design procedure or process simulation procedure is very easy. Typically, Mc-Cabe Thiele diagram is adopted for process design calculations and a system of equations whose total number does not exceed 20 are solved for process simulation. For either case, generating a solution for binary distillation is not difficult.

For multi-component systems involving more than 3 to 4 components, adopting a graphical procedure is ruled out for the purpose of process design calculations. A short cut method for the design of multi-component distillation columns is to adopt Fenske, Underwood and Gilliland (FUG) method. Alternatively, for the rigorous design of multi-component distillation systems, a process design problem is solved as a process simulation problem with an assumed set of process design parameters. Eventually commercial process simulators such as HYSYS or ASPEN PLUS or CHEMCAD or PRO-II are used to match the obtained product distributions with desired product distributions. Based on a trail and error

S.No.	Parameters in Chemical Process Design	Parameters in Refinery Process Design	Desired conversions
1	Molar/Mass feed and product flow rates	Volumetric (bbl) feed and product flow rates	- Barrel to kg (using average °API) - Determine average °API of the stream
2	Feed Mole fraction	True Boiling Point (TBP) curve of the feed	- Convert TBP data to Volume average boiling points for various pseudo components
3	Desired product mole fractions	Product TBP curves/ASTM Gaps	- Convert TBP to ASTM and vice-versa
4	Column diameter as function of vapor & Liquid flow rates (mol or kg/hr)	Vapor & Liquid flow rates (mol or kg/hr)	- Convert TBP data to Molecular weight
5	Energy balance may or may not be critical in design calculations	Energy Balance is by far an important equation to solve	- Evaluate enthalpy (Btu/lb) of various streams

Table 1.1 : Analogies between chemical and refinery process design

approach that gets improved with the experience of the process design engineer with simulation software, process design parameters are set. **In summary, it is always a fact that a multi-component process design problem is very often solved as a process simulation problem, as it necessarily eases the solution approach.**

1.3 Analogies between Chemical & Refinery Process design

In a larger sense, a refinery process design procedure shall mimic similar procedures adopted for chemical process design. However, a chemical engineer shall first become conversant with the analogies associated with chemical and refinery process design. Table 1.1 summarizes these analogies in the larger context. It can be observed in the table that refinery process design calculations can be easily carried out as conventional chemical process design calculations when various properties such as VABP, TBP to ASTM conversion and vice-versa, molecular weight, °API and enthalpy have to be evaluated for various crude feeds and refinery intermediate and product streams. Once these properties are estimated, then knowledge in conventional chemical process design can be used to judiciously conduct the process design calculations.

There are few other properties that are also required to aid process design calculations as these properties are largely related to the product specifications. A few illustrative examples are presented below to convey that the evaluation of these additional properties is mandatory in refinery process design:

a) Sulfur content in various intermediate as well as product streams is always considered as an important parameter to monitor in the entire refinery. Therefore, evaluation of sulfur content in various streams in the refinery complex is essential.
b) The viscosity of certain heavier products emanating from the fuel oil pool such as heavy fuel oil and bunker oil is always considered as a severe product constraint. Therefore, wherever necessary viscosity of process and product streams needs to be evaluated at convenience. For such cases, the evaluation of viscosity from blended intermediate stream viscosities is required.
c) In similarity to the case evaluated for the above option (i.e., (b)), flash point and pour point need to be evaluated for a blended stream mixture.
d) Equilibrium flash vaporization (EFV) curve plays a critical role in the design of distillation columns that are charged with partially vaporized feed. Therefore, the determination of EFV from TBP is necessary.

1.4 Refinery Property Estimation

The following list summarizes the refinery process and product properties required for process design calculations:

1. True boiling point curve (always given) of crude and intermediate/product streams
2. Volume (Mean) average boiling point (VABP)
3. Mean average boiling point (MEABP)
4. Weight average boiling point (WABP)
5. Molal average boiling point (MABP)
6. Characterization factor (K)
7. Vapor pressure (VP) at any temperature
8. Viscosity at any temperature
9. Average sulfur content (wt %)
10. ^{o}API
11. Enthalpy (Btu/lb)
12. ASTM conversion to TBP and vice-versa
13. Viscosity of a mixture using viscosity index
14. Flash point of a mixture using flash point index
15. Pour point of a mixture using pour point index
16. Equilibrium flash vaporization curve

For many of these properties, the starting point by large is the crude assay which consists of a TBP, sulfur and ^{o}API curve using which properties of various streams are estimated. In the next chapter, we address the evaluation of these important properties using several correlations available in various literatures.

2. Estimation of Refinery Stream Properties

2.1 Introduction

In this chapter, we present various correlations available for the property estimation of refinery crude, intermediate and product streams. Most of these correlations are available in Maxwell (1950), API Technical Hand book (1997) etc. Amongst several correlations available, the most relevant and easy to use correlations are summarized so as to estimate refinery stream properties with ease.

The determination of various stream properties in a refinery process needs to first obtain the crude assay. Crude assay consists of a summary of various refinery properties. For refinery design purposes, the most critical charts that are to be known from a given crude assay are TBP curve, ^{o}API curve and Sulfur content curve. These curves are always provided along with a variation in the vol % composition. Often, it is difficult to obtain the trends in the curves for the 100 % volume range of the crude. Often crude assay associated to residue section of the crude is not presented. In summary, a typical crude assay consists of the following information:

a) TBP curve (< 100 % volume range)
b) ^{o}API curve (< 100 % volume range)
c) Sulfur content (wt %) curve (< 100 % volume range)
d) Average ^{o}API of the crude
e) Average sulfur content of the crude (wt %)

An illustrative example is presented below for a typical crude assay for Saudi heavy crude oil whose crude assay is presented by Jones and Pujado (2006). A complete TBP assay of the Saudi heavy crude oil is obtained from elsewhere. Figures 2.1, 2.2 and 2.3 summarize the TBP, ^{o}API and sulfur content curves with respect to the cumulative volume %. Amongst these, the ^{o}API and % sulfur content graphs have been extrapolated to project values till a volume % of 100. This has been carried out to ensure that the evaluated values are in comparable range with the average ^{o}API and sulfur content reported for the crude in the literature. These values correspond to about 28.2 ^{o}API and 2.84 wt % respectively.

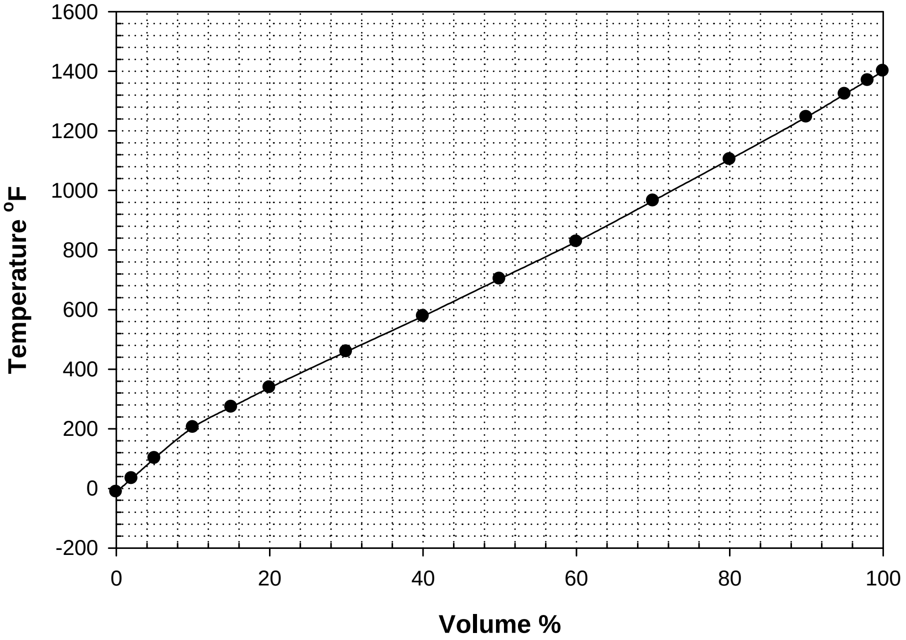


Figure 2.1: TBP Curve of Saudi heavy crude oil.

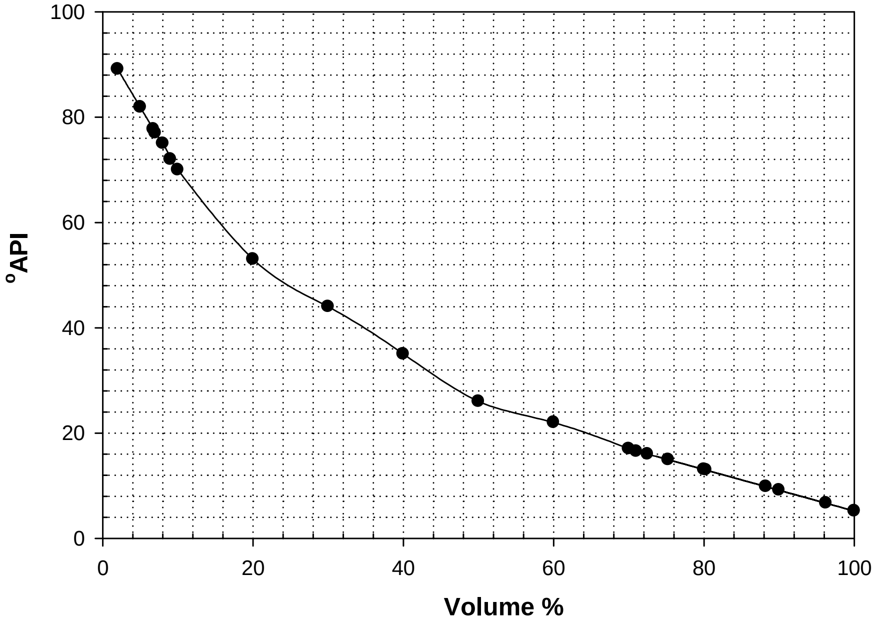


Figure 2.2: ° API Curve of Saudi heavy crude oil.

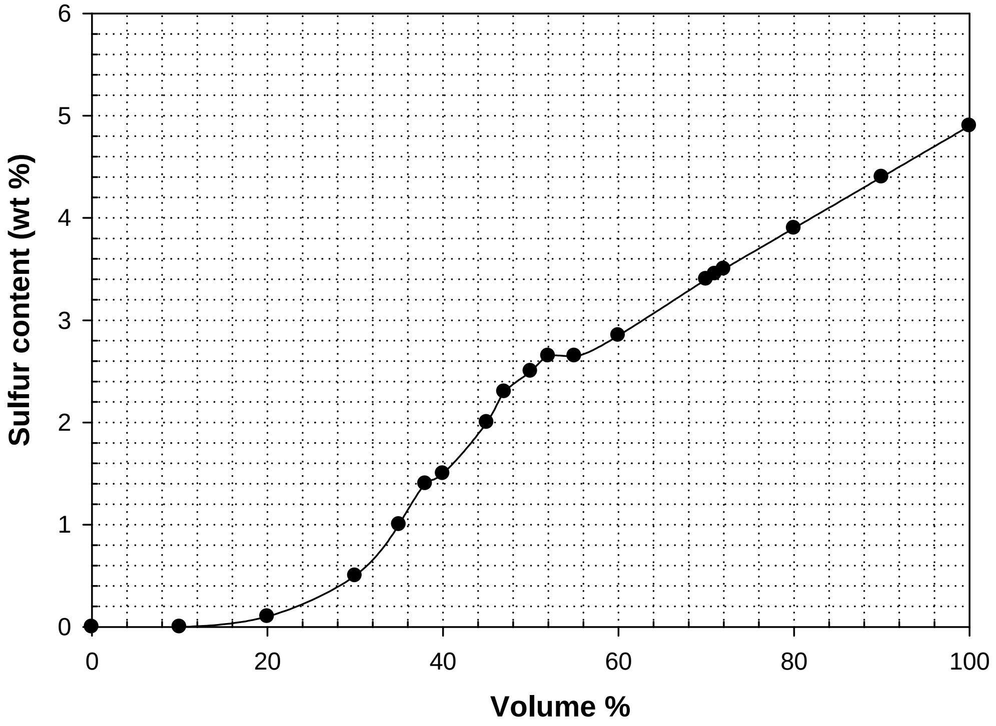


Figure 2.3: Sulfur content assay for heavy Saudi crude oil.

2.2 Estimation of average temperatures

The volume average boiling point (VABP) of crude and crude fractions is estimated using the expressions:

$$VABP = \frac{T_{20}+T_{50}+T_{80}}{3} \quad (1)$$

$$VABP = \frac{T_0+4T_{50}+T_{100}}{6} \quad (2)$$

The mean average boiling point (MEABP), molal average boiling point (MABP) and weight average boiling point (WABP) is evaluated using the graphical correlations presented in Maxwell (1950). These are summarized in Tables 2.1 – 2.3 for both crudes and crude fractions. Where applicable, extrapolation of provided data trends is adopted to obtain the temperatures if graphical correlations are not presented beyond a particular range in the calculations. We next present an example to illustrate the estimation of the average temperatures for Saudi heavy crude oil.

S_{TBP} °F/vol%	Differentials to be added to VABP for evaluating MEABP for various values of VABP				
	200 °F	300 °F	400 °F	500 °F	600+ °F
0.8	0.092				
1	-0.946				
1.2	-2.157	-0.211			
1.4	-3.411	-0.674			
1.6	-4.977	-2.043			
1.8	-6.370	-3.324	-0.237		
2	-8.210	-4.821	-1.366		
2.2	-9.853	-6.171	-2.607	-0.738	-0.832
2.4	-11.664	-7.709	-3.696	-1.635	-1.633
2.6	-13.525	-9.125	-5.037	-2.490	-2.349
2.8	-15.411	-11.074	-6.476	-3.393	-3.267
3	-17.171	-12.564	-7.763	-4.231	-4.369
3.2	-19.428	-14.336	-9.238	-5.174	-5.250
3.4	-21.479	-16.119	-10.731	-6.177	-6.277
3.6	-23.378	-18.015	-12.184	-7.313	-7.182
3.8	-25.665	-19.904	-13.958	-8.525	-8.306
4	-27.863	-21.882	-15.558	-9.794	-9.621
4.2	-30.137	-23.866	-17.390	-11.497	-10.623
4.4	-32.447	-26.153	-19.135	-13.157	-12.039
4.6	-34.585	-28.059	-21.121	-14.815	-13.083
4.8	-37.252	-30.509	-23.220	-16.502	-14.627
5	-39.713	-32.746	-25.268	-18.452	-16.232
5.2		-34.931	-27.320	-20.564	-17.263
5.4		-37.337	-29.452	-24.538	-18.956
5.6		-39.988	-31.934	-31.482	-20.741
5.8		-42.004	-34.231	-34.134	-22.164
6		-45.231	-36.770	-36.644	-23.843
6.2			-39.325	-39.224	-25.438
6.4			-41.766	-41.605	-27.517
6.6			-44.287	-44.536	-29.529
6.8			-47.027	-47.392	-31.390
7			-50.345	-50.612	-33.442
7.2				-53.459	-35.723
7.4				-56.580	-37.580
7.6				-59.714	-39.817
7.8					-42.204
8					-44.253
8.2					-46.637
8.4					-48.916
8.6					-51.513
8.8					-53.958
9					-56.606
9.2					-59.260
9.4					-62.025
9.6					-64.491
9.8					-67.463
10					-70.073

Table 2.1: Tabulated Maxwell's correlation data for the estimation of mean average boiling point (Adapted from Maxwell (1950)).

S_{TBP} °F/vol%	Differentials to be added to VABP for evaluating weight average boiling point			
	200 °F	300 °F	400 °F	500 °F
1	0.385			
1.2	0.871			
1.4	1.562	0.4252		
1.6	1.731	0.6859		
1.8	2.385	1.1478		
2	2.860	1.4331	0.428	
2.2	3.372	1.9128	0.869	
2.4	3.801	2.3559	1.364	0.084
2.6	4.603	2.7724	1.770	0.705
2.8	5.096	3.3327	2.241	1.149
3	5.288	3.9349	2.623	1.356
3.2	5.992	4.3339	3.063	1.598
3.4	6.471	4.8098	3.556	2.094
3.6	7.242	5.2685	3.979	2.690
3.8	7.874	5.9560	4.527	3.164
4	8.392	6.5961	5.281	3.528
4.2	9.103	7.1566	5.850	3.977
4.4	9.822	7.7964	6.378	4.707
4.6	10.406	8.5729	7.051	5.431
4.8	11.149	9.3378	7.698	5.997
5	12.309	9.8381	8.530	6.699
5.2		10.5521	9.450	7.525
5.4		11.3710	10.120	8.340
5.6		12.3187	10.843	9.355
5.8		13.0148	11.766	10.155
6		13.7419	12.818	11.036
6.2			13.806	12.062
6.4			14.651	12.825
6.6			15.504	13.641
6.8			16.619	14.862
7			17.854	16.076
7.2				17.383
7.4				18.437
7.6				19.485
7.8				20.817
8				21.718
8.2				23.035
8.4				24.437
8.6				25.806
8.8				27.236
9				28.498
9.2				29.729
9.4				31.481
9.6				32.713
9.8				34.373
10				35.656

Table 2.2: Tabulated Maxwell's correlation data for the estimation of weight average boiling point (Adapted from Maxwell (1950)).

S_{TBP} ^{o}F/vol%	Differentials to be added to VABP for		
	200 ^{o}F	300 ^{o}F	400 ^{o}F
1.2	-0.198		
1.4	-1.528		
1.6	-4.137	-0.808	-0.808
1.8	-6.243	-2.103	-2.103
2	-7.736	-3.865	-3.865
2.2	-9.942	-5.853	-5.853
2.4	-12.190	-7.855	-7.855
2.6	-14.412	-10.056	-10.056
2.8	-16.720	-12.286	-12.286
3	-19.183	-14.513	-14.513
3.2	-22.225	-16.907	-16.907
3.4	-24.354	-19.439	-19.439
3.6	-27.572	-22.195	-22.195
3.8	-30.467	-25.215	-25.215
4	-33.302	-27.810	-27.810
4.2	-36.718	-30.893	-30.893
4.4	-40.485	-33.998	-33.998
4.6	-43.955	-37.582	-37.582
4.8	-47.289	-40.966	-40.966
5	-51.589	-44.487	-44.487
5.2	-55.380	-48.548	-48.548
5.4	-59.634	-52.525	-52.525
5.6	-63.603	-56.417	-56.417
5.8	-68.993	-60.733	-60.733
6	-73.650	-65.524	-65.524
6.2		-70.237	-70.237
6.4		-75.367	-75.367
6.6		-80.412	-80.412
6.8		-86.654	-86.654
7		-91.941	-91.941
7.2		-98.218	-98.218
7.4		-104.912	-104.912

Table 2.3: Tabulated Maxwell's correlation data for the estimation of molal average boiling point (Adapted from Maxwell (1950)).

Q 2.1: Using the TBP presented in Figure 2.1 and graphical correlation data (Table 2.1 – 2.3), estimate the average temperatures of Saudi heavy crude oil.

Solution:

a. Volume average boiling point

$T_v = t_{20}+t_{50}+t_{80}/3 = (338+703+1104)/3 = 715\ ^0F$

b. Mean average boiling point (from extrapolation)

$\text{Slope} = t_{70}-t_{10} / 60 = 965-205/60 = 12.67$

From Table 2.1

$\Delta T\ (400,12.67) = -116.15\ ^0F$
$\Delta T\ (500,12.67) = -102.48\ ^0F$
$\Delta T\ (715,12.67) = -73.0895^0F$ (from extrapolation)

Mean average boiling point = $T_v + \Delta T = 715 - 73.0895 = 641.91\ ^0F$

c. Weight average boiling point (from extrapolation)

$\text{Slope} = t_{70}-t_{10} / 60 = 965-205/60 = 12.67$

From Table 2.1

$\Delta T\ (400,12.67) = 43.515\ ^0F$
$\Delta T\ (500,12.67) = 41.515\ ^0F$
$\Delta T\ (715,12.67) = 37.215\ ^0F$ (from extrapolation)

Weight average boiling point = $T_v + \Delta T = 715 + 37.215 = 752.21\ ^0F$

d. Molal average boiling point (from extrapolation)

$\text{Slope} = t_{70}-t_{10} / 60 = 965-205/60 = 12.67$

From Table 2.2

$\Delta T\ (400,12.67) = -212.96\ ^0F$
$\Delta T\ (600,12.67) = -191.62\ ^0F$
$\Delta T\ (715,12.67) = -179.3495\ ^0F$ (from extrapolation)

Molal average boiling point = $T_v + \Delta T = 715 - 179.3495 = 535.65\ ^0F$

2.3 Estimation of average °API and sulfur content

2.3.1 Psuedo-component concept

Many a times for refinery products the average °API needs to be estimated from the °API curve of the crude and TBP of the crude/product. The estimation of °API is detrimental to evaluate the mass balances from volume balances. Since oil processing is usually reported in terms of barrels (bbl), the average °API of the stream needs to be estimated to convert the volumetric flow rate to the mass flow rate. In a similar context, sulfur balances across the refinery is very important to design and operate various desulfurization sub-processes in the refinery complex. Henceforth, the average sulfur content of both crude and its CDU products needs to be evaluated so as to aid further calculations in the downstream units associated to the crude distillation unit.

In order to estimate the average °API and sulfur content of a crude/product stream, it is essential to characterize the TBP curve of the crude/product using the concept of psuedocomponents. It is well known that a refinery process stream could not represented using a set of 50 – 100 components, as crude oil constitutes about a million compounds or even more. Therefore, to aid refinery calculations, the psuedocomponent concept is being used. According to the conception of the psuedocomponent representation of the crude stream, a crude oil is characterized to be a constituent of a maximum of 20 – 30 psuedocomponents whose average properties can be used to represent the TBP, °API and % sulfur content of the streams. A pseudo-component in a typical TBP is defined as a component that can represent the average mid volume boiling point (and its average properties such as °API and % sulfur

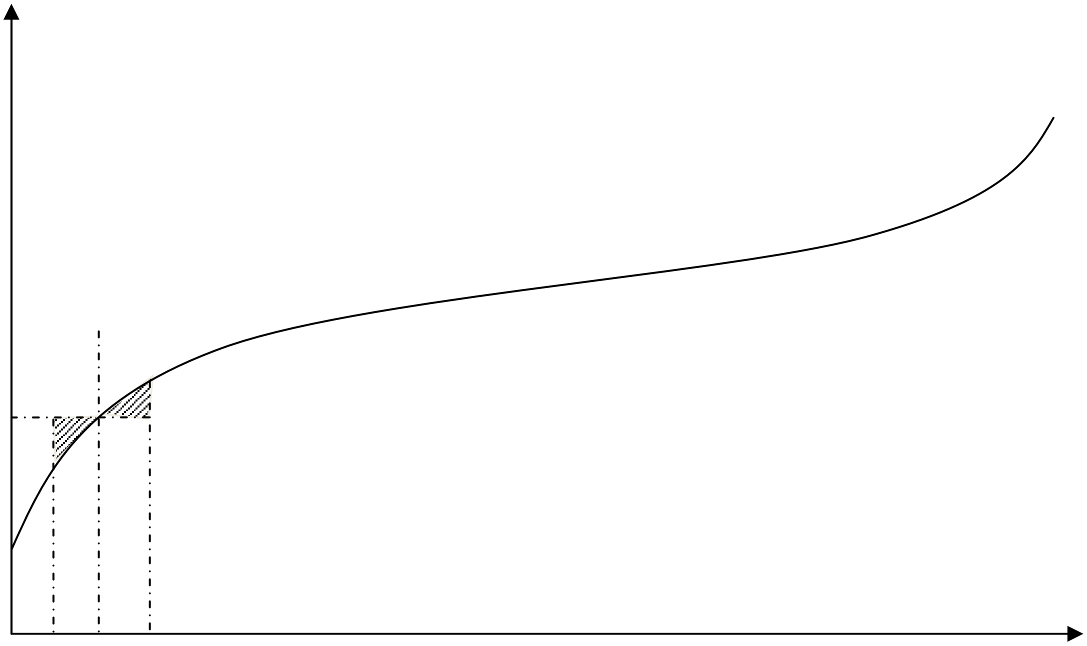


Figure 2.4: Illustration for the concept of pseudo-component.

content). Therefore, in a typical TBP, a pseudo-component is chosen such that within a given range of volume %, the pseudo-component shall provide equal areas of area under and above the curves (Figure 2.4). In a truly mathematical sense, this is possible, if the chosen area for the volume cuts corresponds to a straight line, with the fact that for a straight line exactly at the mid-point, the area above the straight line and below the straight line need to be essential equal. However, since huge number of straight segments are required to represent a non-linear curve, the calculation procedure is bound to be tedious. Henceforth, typically a crude/product stream is represented with no more than 20 – 30 psuedo- components.

The temperature corresponding to the pseudo-component to represent a section of the crude volume on the TBP is termed as mid boiling point (MBP) and the corresponding volume as mid volume (MV). Therefore, in summary, a graphical representation of the TBP is converted to a tabulated data comprising of psuueudo-component number, section temperature range, section volume range, MBP and MV. A similar projection of mid volume with the ^{o}API and sulfur content curves also provides the mid volume ^{o}API and %sulfur content properties. These can be then used to estimate the average ^{o}API and % sulfur content of the crude/refinery product.

We next present an illustration with respect to the pseudo-components chosen to represent the heavy Saudi crude oil.

Q 2.2: Using the TBP, ^{o}API and % sulfur content presented in Figure 2.1 and pseudo-component concept, summarize a table to represent the TBP, ^{o}API and % sulfur content in terms of the chosen pseudo-components.

Solution:

The following temperature and volume bands have been chosen to represent the TBP in terms of the pseudo-components.

Psuedo-component No.	Range	Differential volume (%)	Cumulative volume (%)
1	-12-80	4	0 - 4
2	80-120	2	4 - 6
3	120-160	1.5	6 – 7.5
4	160-180	1.5	7.5 – 9
5	180-220	2	9 – 11
6	220-260	3	11 – 14
7	260-300	3	14 – 17
8	300-360	5	17 – 22
9	360-400	3	22 – 25
10	400-460	5	25 – 30
11	460-520	5.1	30 – 35.1
12	520-580	4.9	35.1 – 40

13	580-620	3	40 – 43
14	620-660	3	43 – 46
15	660-760	8.5	46 – 54.5
16	760-860	7.5	54.5 – 62
17	860-900	3.5	62 – 65.5
18	900-940	2.5	65.5 – 68
19	940-980	3	68 – 71
20	980-1020	3	71 – 74
21	1020-1060	2.5	74 - 76.5
22	1060-1160	7.5	76.5 – 84
23	1160-1280	8.5	84 – 92.5
24	1280-1400	7.5	92.5 -100

For these chosen temperature zones, the pseudo-component mid boiling point and mid volume (that provide equal areas of upper and lower triangles illustrated in Figure 2.4) evaluated using the TBP curve are summarized as:

Psuedo-component No.	**Mid Boiling point**	**Mid volume**
1	35	2
2	100	5
3	140	6.75
4	170	8
5	200	9.75
6	240	12.5
7	280	15.5
8	330	19.5
9	380	23.5
10	430	27.5
11	490	32.5
12	550	37.5
13	600	41.75
14	640	45
15	710	50.75
16	810	58.5
17	880	63.75
18	920	66.75
19	960	69.5
20	1000	72.5
21	1040	75.25
22	1110	80.25
23	1220	88.25
24	1340	96.25

The mid volume (MV) data when projected on the oAPI and % sulfur content curves (Figures 1.2 and 1.3) would provide mid °API and mid % sulfur content for corresponding pseudo-components. These are summarized as

Psuedo-component No.	Mid volume	Mid °API	% sulfur content
1	2	89.1	0.641432
2	5	81.9	0.663074
3	6.75	77.7	0.676386
4	8	75	0.68523
5	9.75	69	0.705736
6	12.5	64.5	0.721939
7	15.5	59	0.742782
8	19.5	53	0.766938
9	23.5	49	0.783934
10	27.5	45	0.8017
11	32.5	40.5	0.822674
12	37.5	35.5	0.847305
13	41.75	33	0.860182
14	45	30	0.876161
15	50.75	25.5	0.901274
16	58.5	22	0.921824
17	63.75	19	0.940199
18	66.75	18	0.946488
19	69.5	17	0.952862
20	72.5	16	0.959322
21	75.25	15	0.96587
22	80.25	13	0.979239
23	88.25	10	1
24	96.25	6.5	1.025362

2.3.2 Estimation of average °API and % sulfur content

The average °API of a refinery crude/product stream is estimated using the expression:

$$\text{av. S.G.} = \frac{\sum_{i=1}^{N_{PC}}[A_i][B_i]}{\sum_{i=1}^{N_{PC}}[A_i]} \quad (3)$$

$$\text{av. \%sulfur content} = \frac{\sum_{i=1}^{N_{PC}}[A_i][B_i][C_i]}{\sum_{i=1}^{N_{PC}}[A_i][B_i]} \quad (4)$$

where $[A_i]$, $[B_i]$ and $[C_i]$ correspond to the pseudo-component (i) differential volume, mid volume S.G. for pseudo-component (i) and mid sulfur content (wt %) for pseudo-component (i) respectively.

An illustrative example is presented to evaluate the average °API and % sulfur content of a crude stream.

Q 2.3: Estimate the average °API and % sulfur content of heavy Saudi Crude oil and compare it with the value provided in the literature.

Solution:

Psuedo-component Number	**Differential Volume %** $[A_i]$	**Mid volume °API**	**Mid volume S.G** $[B_i]$	**Wt. factor** $[A_i][B_i]$	**Mid volume sulfur wt %** $[C_i]$	**Sulfur wt. factor** $[A_i][B_i][C_i]$
1	4	89.1	0.641432	2.56573	0	0
2	2	81.9	0.663074	1.326148	0	0
3	1.5	77.7	0.676386	1.014579	0	0
4	1.5	75	0.68523	1.027845	0	0
5	2	69	0.705736	1.411471	0	0
6	3	64.5	0.721939	2.165816	0	0
7	3	59	0.742782	2.228346	0	0
8	5	53	0.766938	3.834688	0.05	0.191734
9	3	49	0.783934	2.351801	0.2	0.47036
10	5	45	0.8017	4.008499	0.35	1.402975
11	5.1	40.5	0.822674	4.19564	0.7	2.936948
12	4.9	35.5	0.847305	4.151796	1.35	5.604925
13	3	33	0.860182	2.580547	1.7	4.38693
14	3	30	0.876161	2.628483	2	5.256966
15	8.5	25.5	0.901274	7.660828	2.6	19.91815
16	7.5	22	0.921824	6.913681	2.75	19.01262
17	3.5	19	0.940199	3.290698	3.05	10.03663
18	2.5	18	0.946488	2.366221	3.2	7.571907
19	3	17	0.952862	2.858586	3.35	9.576263
20	3	16	0.959322	2.877966	3.55	10.21678
21	2.5	15	0.96587	2.414676	3.7	8.934301
22	7.5	13	0.979239	7.344291	3.9	28.64273
23	8.5	10	1	8.5	4.3	36.55
24	7.5	6.5	1.025362	7.690217	4.7	36.14402
Sum	**100**	**-**	**-**	**87.58476**	**-**	**207.045**

$$\text{av. S.G.} = \frac{\sum_{i=1}^{N_{PC}}[A_i]\,[B_i]}{\sum_{i=1}^{N_{PC}}[A_i]} = \frac{87.58476}{100} = 0.8758$$

$$\text{av. API} = \frac{141.5}{0.874} - 131.5 = 30.0577$$

$$\text{av. \%sulfur content} = \frac{\sum_{i=1}^{N_{PC}}[A_i][B_i][C_i]}{\sum_{i=1}^{N_{PC}}[A_i][B_i]} = \frac{207.045}{87.58476} = 2.6393$$

The literature values of average ^{o}API and average sulfur content (wt %) are 28.2 ^{o}API and 2.84 wt % respectively.

2.4 Characterization factor

The refinery stream characterization factor (K) is required for a number of property evaluations such as molecular weight, enthalpy etc. Therefore, graphical correlations to evaluate characterization factor are mandatory to evaluate the characterization factor. Maxwell (1950) presents a graphical correlation between crude characterization factor (K), ^{o}API and mean average boiling point (MEABP). The correlation is presented in Table 2.4 which can be used to estimate the characterization factor for both crude and other refinery process streams.

Q 2.4: Estimate the characterization factor for Saudi heavy crude oil.

Solution:

Mean average boiling point = 641.9105 ^{0}F; ^{0}API = 30.3834770

From Table 2.4, characterization factor is 11.75.

VABP	Characterization factor for various values of API																		
°F	90°	85°	80°	75°	70°	65°	60°	55°	50°	45°	40°	35°	30°	25°	20°	15°	10°	5°	0°
100	12.889	12.607	12.302	12.018	11.711	11.414	11.121	10.863	10.571	10.249	9.951	9.687	9.399						
120	13.040	12.755	12.451	12.157	11.859	11.564	11.255	10.983	10.695	10.382	10.080	9.802	9.504	9.202					
140	13.183	12.902	12.586	12.295	11.988	11.693	11.388	11.113	10.822	10.502	10.191	9.925	9.622	9.314					
160	13.331	13.042	12.725	12.426	12.125	11.831	11.519	11.236	10.938	10.627	10.319	10.042	9.728	9.419					
180	13.465	13.177	12.855	12.550	12.252	11.955	11.641	11.355	11.062	10.748	10.432	10.148	9.836	9.527	9.199				
200	13.600	13.314	12.995	12.686	12.391	12.086	11.759	11.474	11.181	10.859	10.541	10.257	9.937	9.631	9.317				
220	13.735	13.447	13.130	12.819	12.514	12.205	11.879	11.587	11.264	10.969	10.645	10.360	10.036	9.727	9.419				
240	13.863	13.577	13.255	12.941	12.624	12.330	11.998	11.692	11.396	11.079	10.757	10.466	10.133	9.826	9.515				
260	13.982	13.703	13.382	13.069	12.744	12.440	12.118	11.796	11.501	11.186	10.858	10.564	10.230	9.919	9.612	9.188			
280		13.828	13.504	13.187	12.854	12.555	12.221	11.904	11.601	11.286	10.948	10.655	10.325	10.003	9.703	9.301			
300		13.946	13.622	13.308	12.968	12.663	12.334	12.014	11.697	11.383	11.048	10.748	10.414	10.102	9.789	9.402			
320			13.742	13.428	13.082	12.780	12.444	12.117	11.792	11.481	11.138	10.840	10.497	10.193	9.879	9.486			
340			13.854	13.535	13.192	12.883	12.553	12.219	11.897	11.577	11.230	10.927	10.589	10.273	9.966	9.575	9.239		
360			13.959	13.650	13.298	12.977	12.655	12.320	11.992	11.671	11.326	11.018	10.677	10.353	10.047	9.666	9.324		
380				13.749	13.401	13.084	12.756	12.418	12.083	11.762	11.410	11.104	10.761	10.431	10.126	9.754	9.406		
400				13.854	13.514	13.181	12.862	12.514	12.183	11.855	11.498	11.184	10.843	10.524	10.206	9.833	9.489		
420				13.964	13.614	13.281	12.961	12.612	12.273	11.946	11.582	11.273	10.924	10.608	10.288	9.915	9.573	9.183	
440					13.724	13.386	13.056	12.712	12.364	12.030	11.665	11.358	11.002	10.682	10.365	9.996	9.649	9.283	
460					13.821	13.490	13.156	12.809	12.460	12.112	11.760	11.436	11.079	10.758	10.436	10.072	9.726	9.365	
480					13.929	13.585	13.259	12.899	12.547	12.202	11.842	11.519	11.158	10.835	10.506	10.145	9.798	9.444	
500						13.677	13.347	12.988	12.635	12.292	11.929	11.595	11.236	10.912	10.579	10.212	9.867	9.521	
520						13.769	13.452	13.081	12.725	12.375	12.013	11.670	11.313	10.988	10.655	10.280	9.934	9.591	9.231
540						13.861	13.533	13.173	12.806	12.460	12.087	11.752	11.392	11.063	10.724	10.355	9.995	9.665	9.303
560						13.956	13.628	13.261	12.903	12.543	12.162	11.830	11.470	11.140	10.788	10.425	10.068	9.736	9.371
580							13.717	13.348	12.985	12.627	12.256	11.902	11.541	11.218	10.863	10.489	10.137	9.810	9.426

VABP °F	Characterization factor for various values of API																		
	90°	**85°**	**80°**	**75°**	**70°**	**65°**	**60°**	**55°**	**50°**	**45°**	**40°**	**35°**	**30°**	**25°**	**20°**	**15°**	**10°**	**5°**	**0°**
600							13.816	13.437	13.074	12.711	12.331	11.989	11.611	11.294	10.937	10.552	10.196	9.876	9.495
620									13.160	12.789	12.404	12.063	11.688	11.368	10.995	10.618	10.256	9.929	9.549
640									13.248	12.873	12.482	12.132	11.763	11.442	11.060	10.684	10.322	9.992	9.611
660									13.331	12.952	12.558	12.208	11.835	11.509	11.128	10.744	10.384	10.054	9.674
680									13.415	13.030	12.644	12.289	11.908	11.570	11.192	10.804	10.443	10.113	9.728
700									13.481	13.112	12.718	12.362	11.984	11.642	11.256	10.865	10.498	10.173	9.781
720										13.194	12.787	12.428	12.058	11.715	11.322	10.925	10.557	10.224	9.839
740										13.258	12.863	12.505	12.125	11.774	11.384	10.988	10.613	10.280	9.891
760										13.344	12.935	12.574	12.192	11.839	11.442	11.050	10.670	10.337	9.940
780										13.417	13.004	12.640	12.260	11.907	11.504	11.111	10.728	10.389	9.990
800										13.492	13.076	12.710	12.329	11.966	11.569	11.170	10.786	10.444	10.039
820											13.145	12.781	12.391	12.021	11.631	11.224	10.845	10.491	10.092
840											13.208	12.850	12.456	12.081	11.687	11.279	10.901	10.546	10.145
860											13.276	12.916	12.526	12.145	11.741	11.341	10.955	10.599	10.193
880											13.352	12.980	12.591	12.205	11.797	11.404	11.015	10.645	10.241
900											13.415	13.047	12.648	12.262	11.858	11.455	11.067	10.696	10.295
920											13.471	13.114	12.708	12.323	11.922	11.516	11.115	10.742	10.341
940												13.181	12.771	12.384	11.983	11.573	11.169	10.792	10.388
960												13.245	12.832	12.440	12.035	11.625	11.224	10.841	10.440
980												13.306	12.891	12.494	12.080	11.685	11.286	10.893	10.487
1000														12.548	12.141	11.739	11.328	10.898	10.520

Table 2.4: Characterization factor data table (Developed from correlation presented in Maxwell (1950)).

MEABP (^{o}F)	Molecular weight at various values of API									
	90	**80**	**70**	**60**	**50**	**40**	**30**	**20**	**10**	**0**
120	82.591									
140	89.244	85.349	81.759							
160	95.864	92.216	87.806	83.893						
180	103.115	99.154	94.260	90.304	86.542	81.845				
200	110.109	106.048	100.836	96.991	92.552	87.991	83.332			
220	117.449	113.544	108.021	103.568	98.771	93.936	89.166	84.078		
240		121.106	115.309	110.680	105.248	100.249	95.244	89.326		
260		129.009	122.453	117.796	112.231	107.030	101.472	95.365		
280		137.114	130.197	125.128	118.714	113.839	108.131	101.448		
300			138.353	132.882	126.099	120.332	114.305	107.771		
320			146.115	140.272	133.574	127.402	120.794	113.873		
330			150.445	144.292	137.140	130.788	124.406	117.517		
340			154.371	148.429	141.234	134.569	127.900	120.713		
360				156.558	149.198	142.287	134.598	127.437		
380				164.892	156.783	149.607	142.029	134.712		
400				173.686	165.470	157.656	149.630	141.480		
420				182.652	174.068	165.958	157.072	148.654		
440				191.989	182.885	174.128	164.891	155.945		
460					192.032	182.646	172.888	163.835		
480					201.319	191.174	181.139	171.254	161.127	
500					211.605	200.591	189.715	179.456	169.110	
520					222.386	210.830	199.031	187.685	176.365	
540					232.797	221.373	208.761	196.571	184.444	
560						232.206	219.068	206.019	192.562	
580						244.280	229.410	216.033	201.881	
600						256.867	241.053	226.386	211.664	
620						270.206	252.796	237.329	221.746	
640						284.579	265.535	249.244	231.901	
660						299.640	279.937	262.088	243.147	
680						314.428	294.088	274.070	254.362	
700						330.525	308.992	286.731	266.284	
720						346.094	324.137	302.034	279.799	256.350
740						362.792	339.010	317.470	293.587	269.871
760						378.761	356.110	331.730	307.285	282.406
780						396.112	371.811	347.474	321.642	295.480
800						414.225	388.517	362.886	336.861	308.755
820						431.710	404.926	378.024	350.271	322.104
840						448.263	422.817	394.631	366.309	336.478
860						467.186	439.000	411.105	381.136	350.167
880						485.253	456.069	426.951	397.625	365.202
900						504.444	473.298	443.368	412.016	380.935
920						521.642	491.813	460.804	427.960	396.645
940						540.431	508.462	477.082	443.855	413.040
960							526.999	493.824	460.293	428.276
980							543.180	511.031	476.876	444.464
1000							561.388	528.021	492.335	459.824
1020							579.708	545.854	510.006	474.981
1030							589.010	554.911	518.720	484.254
1040							597.408	563.222	526.663	492.425
1060								580.517	544.863	510.209
1080								599.227	561.993	526.436
1100									579.496	542.104
1120									596.069	558.957
1140										573.783

Table 2.5: Molecular weight data table (Developed from correlation presented in Maxwell (1950)).

2.5 Molecular weight

The molecular weight of a stream is by far an important property using which mass flow rates can be conveniently expressed in terms of molar flow rates. Typically molar flow rates are required to estimate vapor and liquid flow rates within distillation columns. A graphical correlation between molecular weight of a stream, its characterization factor (K) and MEABP presented by Maxwell (1950) in data format in Table 2.5. An illustrative example is presented below to elaborate upon the procedure involved for molecular weight estimation.

Q 2.5: Estimate the molecular weight of heavy Saudi crude oil.

Solution:

^{0}API = 30.05, Mean average b.pt=641.91 ^{o}F

From Table 2.5, molecular weight of the Saudi heavy crude oil = 266.

2.6 Viscosity

The correlations presented in API Technical Handbook (1997) are used to estimate the viscosity (in Centistokes) of crude/petroleum fractions. The correlations are presented as follows:

a) The viscosity of a petroleum fraction at 100 ^{o}F is evaluated as the sum of two viscosities namely correlated and reference i.e.,
$\mu_{100} = \mu_{ref} + \mu_{cor}$
where the correlated viscosity is estimated using the expressions
$\mu_{cor} = 10^{\lfloor A_1 + A_2 K \rfloor}$
$A_1 = 34.9310 - 0.084387T + 6.73513 \times 10^{-5}T^2 - 1.01394 \times 10^{-8}T^3$
$A_2 = -2.92649 + 6.98405 \times 10^{-3}T - 5.09947 \times 10^{-6}T^2 + 7.49378 \times 10^{-10}T^3$
$K = \sqrt[3]{T}[S.G.]$
And the reference viscosity is estimated using the expressions
$\mu_{ref} = 10^{[-1.35579 + 8.16059 \times 10^{-4}T + 8.38505 \times 10^{-7}T^2]}$

b) The viscosity of a petroleum fraction at 210 ^{o}F is estimated using the viscosity at 100 ^{o}F with the following correlation
$\mu_{210} = 10^{\lfloor -1.92353 + 2.41071 \times 10^{-4}T + 0.5113 \log(T \times \mu_{100}) \rfloor}$

In the above expressions, T refers to the mean average boiling point (MEABP) expressed in ^{o}R.
Next, the estimation of viscosity is presented for the heavy Saudi Crude oil.

Q 2.6: Estimate the viscosity of heavy Saudi crude oil using viscosity correlations presented in API Technical data book and compare the obtained viscosity with that available in the literature.

Solution:

At 100^{o}F, the viscosity is estimated using the following expressions:
Mean average boiling point (MEABP)T=641.9^{o}F=1101.5^{0}R.

S. G. of crude = 0.8758
^{0}API = 30.05
$K = \sqrt[3]{T}[S.G.] = \sqrt[3]{1102} \times 0.874 = 11.81$

$$A_1 = 34.9310 - 0.084387T + 6.73513 \times 10^{-5}T^2 - 1.01394 \times 10^{-8}T^3 = 5.6842$$

$$A_2 = -2.92649 + 6.98405 \times 10^{-3}T - 5.09947 \times 10^{-6}T^2 + 7.49378 \times 10^{-10}T^3 = -0.4193$$

$$\mu_{cor} = 10^{\lfloor A_1 + A_2 K \rfloor} = 5.6325\ \text{Cst}$$

$$\mu_{ref} = 10^{\left[-1.35579 + 8.16059 \times 10^{-4}T + 8.38505 \times 10^{-7}T^2\right]} = 3.6364\ \text{Cst}$$

$$\mu_{100} = \mu_{ref} + \mu_{cor} = 9.269\ \text{Cst}$$

$$\mu_{210} = 10^{\lfloor -1.92353 + 2.41071 \times 10^{-4}T + 0.5113\log(T \times \mu_{100}) \rfloor} = 2.465\ Cst$$

The literature viscosity values are significantly different from those obtained using viscosity correlation. The literature viscosity at 100 °C is 18.9 which is almost twice that obtained from the correlation. The predicted viscosity matches with that reported in the literature for medium heavy Saudi crude oil.

2.7 Enthalpy

The estimation of petroleum fraction enthalpy is required for good number of reasons. Without knowing the enthalpy of petroleum vapor/liquid fractions, it is not possible to estimate the condenser and re-boiler duties and conduct energy balances. Maxwell (1950) presents about 13 graphs that can be conveniently used to estimate the enthalpy of vapor/liquid fractions as functions of refinery stream characterization factor and system temperature (in °F). These graphical correlations are tabulated and presented in Tables 2.6 – 2.17. Using these tables, the enthalpy of refinery process streams is estimated using interpolation and extrapolation procedures. Next, an illustrative example is presented for the estimation of enthalpy of Saudi heavy crude oil.

Q 2.7: Estimate the enthalpy of saturated Saudi heavy crude oil liquid at its volume average boiling point.

Solution:

Volume average boiling point = 715 °F

MeABP	600		800	
K	11	12	11	12
Enthalpy of liquid stream (Btu/lb) at 715 °F	400	427	385	406

Saudi Crude MEABP = 642 ^{o}F

K value for Saudi Crude = 11.75 ^{o}F

Liquid stream enthalpy at 600 ^{o}F and 11.75 ^{o}F = 400 + (11.75-11) x (427-400) = 420.25 ^{o}F

Liquid stream enthalpy at 800 ^{o}F and 11.75 ^{o}F = 385 + (11.75 – 11) x (406 – 385) = 400.75 ^{o}F

Crude oil enthalpy at 715 ^{o}F whose MEABP = 642 ^{o}F = 420.25 +42/200*(420.25-400.75) = 424.395 ^{o}F.

T (^{o}F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0.49	0.41	0.33	0.24	0.16	0.07
20	8.12	8.34	8.56	8.78	9.00	9.22
40	17.98	18.29	18.60	18.91	19.22	19.54
60	27.90	28.32	28.74	29.16	29.58	30.00
80	37.59	38.15	38.71	39.27	39.82	40.38
100	47.38	48.16	48.94	49.71	50.49	51.27
120	57.74	58.76	59.79	60.81	61.83	62.85
140	68.52	69.61	70.70	71.79	72.88	73.96
160	78.81	80.25	81.69	83.13	84.56	86.00
180	88.86	90.77	92.68	94.60	96.51	98.42
200	99.60	101.54	103.47	105.40	107.33	109.27
220	110.37	112.46	114.55	116.64	118.73	120.82
240	122.05	124.37	126.70	129.03	131.36	133.68
260	133.69	136.21	138.72	141.23	143.75	146.26
280	144.74	147.54	150.35	153.15	155.95	158.75
300	156.43	159.66	162.90	166.13	169.36	172.59
320	168.98	172.47	175.95	179.43	182.91	186.39
340	180.43	184.26	188.09	191.92	195.75	199.59
360	192.85	197.02	201.18	205.34	209.51	213.67
380	207.06	211.34	215.62	219.91	224.19	228.47
400	220.58	225.00	229.42	233.84	238.26	242.67
420	234.17	238.93	243.69	248.45	253.21	257.97
440	247.78	253.05	258.32	263.59	268.86	274.13
460	262.39	268.04	273.68	279.33	284.98	290.62
480	277.45	283.60	289.75	295.91	302.06	308.21
500	293.22	300.01	306.80	313.59	320.38	327.17

Table 2.6: Hydrocarbon liquid enthalpy data for MEABP = 200^{o}F and K = 11 – 12.

T (^{o}F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	182.29	179.55	176.81	174.07	171.32	168.58
20	187.88	185.37	182.86	180.36	177.85	175.34
40	194.67	192.23	189.79	187.36	184.92	182.48
60	201.76	199.36	196.97	194.57	192.17	189.78
80	208.69	206.47	204.26	202.04	199.82	197.60
100	215.52	213.65	211.79	209.92	208.05	206.18
120	222.35	220.76	219.16	217.56	215.97	214.37
140	229.24	227.77	226.29	224.82	223.35	221.88
160	236.45	235.14	233.83	232.52	231.21	229.90
180	243.96	242.89	241.81	240.74	239.67	238.59
200	251.70	250.84	249.97	249.11	248.25	247.38
220	260.16	259.65	259.13	258.62	258.11	257.60
240	267.77	267.64	267.51	267.38	267.25	267.12
260	275.72	275.81	275.89	275.98	276.06	276.15
280	284.69	284.87	285.06	285.24	285.42	285.61
300	293.23	293.64	294.06	294.47	294.89	295.30
320	302.34	302.96	303.57	304.18	304.80	305.41
340	311.03	312.03	313.03	314.03	315.02	316.02
360	320.06	321.42	322.77	324.13	325.48	326.84
380	329.97	331.47	332.97	334.46	335.96	337.46
400	339.71	341.41	343.10	344.80	346.49	348.19
420	349.78	351.83	353.88	355.92	357.97	360.02
440	360.60	362.80	365.00	367.20	369.39	371.59
460	370.59	373.18	375.77	378.36	380.94	383.53
480	381.21	384.16	387.12	390.07	393.03	395.98
500	392.13	395.17	398.21	401.25	404.29	407.34
520	404.19	407.12	410.06	412.99	415.92	418.86
540	414.90	418.19	421.48	424.77	428.06	431.35
560	424.77	428.72	432.67	436.63	440.58	444.53
580	437.34	441.54	445.74	449.93	454.13	458.33
600	447.85	452.59	457.34	462.09	466.83	471.58
620	459.48	464.45	469.42	474.39	479.35	484.32
640	470.59	476.04	481.50	486.95	492.40	497.86
660	481.81	488.00	494.19	500.38	506.57	512.76
680	494.36	500.86	507.36	513.85	520.35	526.84
700	507.41	513.71	520.02	526.33	532.64	538.95
720	519.67	526.24	532.81	539.38	545.95	552.52
740	532.34	539.21	546.07	552.93	559.80	566.66
760	545.20	552.41	559.62	566.84	574.05	581.26
780	557.96	565.55	573.14	580.74	588.33	595.92
800	571.08	578.89	586.70	594.52	602.33	610.15
820	584.25	592.40	600.55	608.70	616.85	625.00
840	597.11	605.50	613.88	622.27	630.65	639.03
860	609.73	618.42	627.10	635.78	644.46	653.14
880	622.88	631.97	641.05	650.13	659.21	668.29
900	637.09	646.43	655.76	665.09	674.43	683.76
920	651.13	660.64	670.14	679.65	689.16	698.67
940	664.71	674.57	684.44	694.30	704.17	714.04
960	678.71	689.03	699.34	709.66	719.97	730.29
980	692.89	703.63	714.38	725.12	735.87	746.61
1000	706.48	717.70	728.92	740.14	751.35	762.57
1020	720.64	732.32	743.99	755.67	767.34	779.02
1040	734.09	746.28	758.46	770.64	782.83	795.01
1060	748.40	760.93	773.46	785.99	798.52	811.05
1080	763.68	776.50	789.33	802.16	814.99	827.81
1100	777.76	791.17	804.57	817.97	831.37	844.77
1120	791.62	805.53	819.44	833.36	847.27	861.18

Table 2.7: Hydrocarbon vapor enthalpy data for MEABP = 200^{o}F and K = 11 – 12.

T (^{o}F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	175.25	172.59	169.93	167.28	164.62	161.96
20	180.93	178.59	176.25	173.90	171.56	169.22
40	186.86	184.86	182.86	180.86	178.86	176.86
60	192.94	191.25	189.56	187.88	186.19	184.51
80	199.18	197.80	196.42	195.04	193.66	192.29
100	205.84	204.73	203.62	202.51	201.41	200.30
120	213.09	212.06	211.04	210.01	208.98	207.96
140	220.02	219.20	218.39	217.57	216.75	215.93
160	227.41	226.82	226.24	225.66	225.08	224.50
180	235.52	235.07	234.62	234.17	233.71	233.26
200	243.15	242.92	242.70	242.48	242.26	242.03
220	250.07	250.32	250.57	250.82	251.07	251.32
240	258.30	258.73	259.17	259.60	260.04	260.47
260	266.78	267.32	267.86	268.40	268.94	269.48
280	275.06	275.86	276.66	277.46	278.26	279.06
300	284.19	285.19	286.20	287.20	288.20	289.20
320	292.73	293.93	295.12	296.31	297.50	298.69
340	302.24	303.59	304.94	306.30	307.65	309.00
360	312.58	314.02	315.46	316.90	318.34	319.77
380	322.63	324.18	325.73	327.27	328.82	330.37
400	332.11	333.90	335.68	337.46	339.24	341.03
420	341.54	343.57	345.59	347.61	349.63	351.65
440	351.73	353.86	355.99	358.11	360.24	362.37
460	362.34	364.62	366.89	369.17	371.44	373.72
480	373.42	375.94	378.47	380.99	383.52	386.04
500	384.62	387.29	389.97	392.65	395.32	398.00
520	395.29	398.41	401.53	404.65	407.77	410.89
540	406.19	409.58	412.97	416.36	419.75	423.14
560	416.87	420.52	424.17	427.82	431.47	435.12
580	428.01	431.96	435.91	439.86	443.81	447.76
600	439.69	443.76	447.83	451.90	455.97	460.04
620	451.80	456.19	460.57	464.96	469.35	473.74
640	463.34	468.06	472.77	477.49	482.21	486.93
660	474.93	479.97	485.02	490.06	495.10	500.14
680	487.16	492.61	498.05	503.50	508.95	514.39
700	499.64	505.47	511.30	517.13	522.96	528.79
720	512.30	518.29	524.29	530.29	536.29	542.29
740	524.92	531.22	537.52	543.82	550.12	556.42
760	537.61	544.28	550.96	557.63	564.31	570.98
780	550.17	557.13	564.09	571.05	578.01	584.96
800	562.83	570.00	577.18	584.35	591.52	598.69
820	576.66	584.05	591.43	598.81	606.19	613.57
840	589.01	596.92	604.84	612.76	620.67	628.59
860	601.61	609.89	618.16	626.44	634.71	642.99
880	615.56	623.98	632.41	640.83	649.25	657.68
900	628.88	637.84	646.80	655.76	664.72	673.68
920	641.84	651.41	660.98	670.55	680.12	689.70
940	655.92	665.68	675.43	685.19	694.94	704.70
960	669.92	679.95	689.98	700.02	710.05	720.08
980	683.56	694.03	704.50	714.97	725.43	735.90
1000	697.91	708.57	719.23	729.89	740.55	751.21
1020	711.62	722.83	734.04	745.26	756.47	767.68
1040	725.70	737.23	748.76	760.29	771.82	783.35
1060	740.84	752.50	764.16	775.82	787.48	799.13
1080	755.66	767.71	779.75	791.80	803.85	815.89
1100	769.42	782.09	794.76	807.44	820.11	832.78
1120	784.49	797.57	810.64	823.71	836.78	849.85
1140	798.91	812.31	825.71	839.11	852.51	865.91

Table 2.8: Hydrocarbon vapor enthalpy data for MEABP = 300^{o}F and K = 11 – 12.

T (^{o}F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0.20	0.24	0.29	0.33	0.37	0.41
20	7.75	8.17	8.60	9.03	9.45	9.88
40	16.77	17.34	17.91	18.48	19.05	19.62
60	25.55	26.36	27.16	27.97	28.77	29.57
80	34.35	35.31	36.28	37.24	38.20	39.17
100	43.94	44.85	45.76	46.67	47.58	48.48
120	52.97	54.19	55.40	56.62	57.83	59.05
140	62.57	64.05	65.52	67.00	68.47	69.94
160	72.21	73.96	75.71	77.46	79.21	80.96
180	81.92	83.86	85.80	87.73	89.67	91.61
200	92.43	94.35	96.27	98.19	100.11	102.03
220	102.56	104.89	107.23	109.56	111.89	114.23
240	113.48	115.99	118.49	121.00	123.50	126.01
260	124.52	127.09	129.65	132.21	134.77	137.33
280	135.07	137.91	140.75	143.59	146.42	149.26
300	146.03	149.19	152.36	155.52	158.68	161.85
320	157.82	161.19	164.55	167.92	171.28	174.65
340	169.58	173.29	177.00	180.71	184.42	188.13
360	181.80	185.77	189.75	193.72	197.69	201.67
380	194.68	198.70	202.72	206.75	210.77	214.79
400	207.56	211.67	215.77	219.87	223.98	228.08
420	220.69	224.98	229.28	233.58	237.88	242.18
440	234.11	238.57	243.04	247.50	251.96	256.43
460	246.65	251.52	256.40	261.28	266.16	271.04
480	261.11	266.38	271.66	276.93	282.21	287.48
500	274.84	280.16	285.47	290.78	296.10	301.41
520	288.40	293.89	299.37	304.85	310.33	315.81
540	304.29	310.08	315.88	321.67	327.46	333.25
560	319.00	325.04	331.08	337.12	343.16	349.20
580	335.13	341.51	347.89	354.27	360.66	367.04
600	350.68	357.81	364.93	372.06	379.19	386.31

Table 2.9: Hydrocarbon liquid enthalpy data for MEABP = 300^{o}F and K = 11 – 12.

T (^{o}F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	167.62	165.22	162.82	160.42	158.02	155.62
20	174.02	171.96	169.91	167.86	165.81	163.76
40	179.94	178.13	176.32	174.51	172.69	170.88
60	186.39	184.72	183.06	181.40	179.74	178.08
80	192.50	191.12	189.73	188.35	186.96	185.58
100	199.48	198.21	196.95	195.69	194.42	193.16
120	206.21	205.32	204.44	203.55	202.66	201.78
140	213.27	212.35	211.44	210.53	209.61	208.70
160	221.95	220.89	219.83	218.77	217.71	216.65
180	228.82	228.13	227.43	226.74	226.04	225.35
200	236.82	236.41	236.00	235.58	235.17	234.76
220	244.23	243.93	243.62	243.32	243.01	242.71
240	252.48	252.21	251.93	251.65	251.37	251.10
260	261.24	261.07	260.90	260.73	260.57	260.40
280	270.09	270.06	270.02	269.99	269.95	269.92
300	278.61	278.96	279.30	279.65	279.99	280.33
320	288.11	288.36	288.61	288.85	289.10	289.35
340	297.17	297.81	298.44	299.07	299.71	300.34
360	306.84	307.54	308.24	308.95	309.65	310.36
380	316.15	316.96	317.76	318.57	319.37	320.18
400	325.50	326.72	327.93	329.15	330.37	331.58
420	336.12	337.18	338.23	339.28	340.34	341.39
440	347.06	348.41	349.77	351.12	352.47	353.82
460	356.74	358.66	360.58	362.50	364.42	366.35
480	367.24	369.30	371.35	373.41	375.46	377.52
500	378.94	380.98	383.02	385.05	387.09	389.13
520	389.90	392.33	394.75	397.17	399.59	402.02
540	400.52	403.20	405.88	408.57	411.25	413.93
560	410.40	413.67	416.95	420.22	423.50	426.77
580	422.91	426.22	429.53	432.84	436.15	439.47
600	433.45	437.00	440.55	444.10	447.65	451.20
620	445.20	448.77	452.34	455.91	459.47	463.04
640	457.32	461.21	465.09	468.98	472.86	476.75
660	469.16	473.55	477.95	482.34	486.74	491.14
680	481.18	486.07	490.96	495.85	500.74	505.63
700	492.79	497.98	503.17	508.36	513.54	518.73
720	506.17	511.35	516.53	521.71	526.89	532.06
740	517.71	523.50	529.30	535.09	540.89	546.69
760	530.52	536.53	542.54	548.55	554.56	560.57
780	543.76	549.79	555.82	561.85	567.88	573.92
800	558.15	564.11	570.07	576.04	582.00	587.96
820	570.49	576.97	583.46	589.94	596.43	602.92
840	583.01	589.92	596.82	603.72	610.62	617.53
860	596.52	603.73	610.93	618.14	625.35	632.55
880	610.01	617.44	624.87	632.30	639.74	647.17
900	622.48	630.53	638.57	646.61	654.65	662.69
920	636.94	645.05	653.15	661.26	669.36	677.46
940	649.30	657.97	666.64	675.31	683.97	692.64
960	662.75	671.74	680.73	689.72	698.72	707.71
980	677.17	686.61	696.05	705.49	714.92	724.36
1000	690.46	700.27	710.08	719.88	729.69	739.50
1020	705.18	715.34	725.51	735.67	745.84	756.00
1040	720.34	730.57	740.79	751.01	761.24	771.46
1060	735.34	745.63	755.93	766.23	776.53	786.83
1080	749.35	760.34	771.34	782.33	793.32	804.32
1100	763.82	775.04	786.25	797.47	808.69	819.90
1120	778.70	790.46	802.22	813.99	825.75	837.52
1140	793.54	805.70	817.85	830.00	842.15	854.30

Table 2.10: Hydrocarbon vapor enthalpy data for MEABP = 400^{o}F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	7.70	7.90	8.10	8.30	8.50	8.70
40	15.46	15.93	16.39	16.86	17.32	17.79
60	24.16	24.88	25.59	26.31	27.02	27.74
80	31.58	32.54	33.50	34.46	35.42	36.39
100	39.75	40.86	41.97	43.08	44.18	45.29
120	49.31	50.57	51.83	53.09	54.35	55.62
140	58.42	59.89	61.36	62.84	64.31	65.78
160	67.37	69.02	70.67	72.31	73.96	75.60
180	77.24	78.95	80.65	82.36	84.07	85.78
200	86.99	88.87	90.75	92.62	94.50	96.38
220	96.32	98.37	100.41	102.46	104.51	106.56
240	107.95	109.93	111.91	113.89	115.86	117.84
260	118.58	120.66	122.74	124.82	126.91	128.99
280	129.36	131.66	133.96	136.26	138.56	140.86
300	140.21	142.89	145.56	148.23	150.91	153.58
320	151.32	154.05	156.78	159.51	162.24	164.97
340	163.26	166.17	169.08	171.99	174.90	177.81
360	175.32	178.44	181.56	184.68	187.80	190.92
380	187.16	190.23	193.30	196.37	199.45	202.52
400	200.18	203.39	206.60	209.82	213.03	216.25
420	213.59	216.77	219.95	223.13	226.31	229.50
440	225.94	229.45	232.95	236.46	239.97	243.48
460	238.23	242.03	245.83	249.63	253.43	257.23
480	251.00	255.12	259.23	263.35	267.46	271.57
500	264.20	268.52	272.83	277.15	281.47	285.79
520	277.65	282.44	287.23	292.02	296.81	301.60
540	291.31	296.15	300.98	305.81	310.65	315.48
560	305.20	310.51	315.82	321.13	326.44	331.75
580	320.08	325.59	331.10	336.62	342.13	347.64
600	334.63	340.57	346.51	352.45	358.38	364.32
620	349.72	356.08	362.45	368.81	375.17	381.53
640	365.41	372.19	378.96	385.73	392.50	399.27
660	380.81	387.91	395.02	402.12	409.22	416.32
680	397.78	404.95	412.13	419.31	426.48	433.66

Table 2.11: Hydrocarbon liquid enthalpy data for MEABP = 400°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	165.62	162.14	158.65	155.17	151.69	148.20
20	171.50	168.21	164.92	161.62	158.33	155.04
40	177.35	174.38	171.42	168.45	165.49	162.52
60	184.01	181.21	178.40	175.60	172.79	169.98
80	190.03	187.57	185.10	182.63	180.17	177.70
100	197.11	194.84	192.58	190.31	188.04	185.77
120	203.34	201.23	199.12	197.01	194.90	192.79
140	210.66	208.55	206.45	204.34	202.23	200.13
160	217.84	216.02	214.20	212.39	210.57	208.76
180	225.78	224.08	222.39	220.69	218.99	217.30
200	233.16	231.55	229.94	228.33	226.71	225.10
220	241.18	239.96	238.73	237.51	236.29	235.06
240	249.49	248.36	247.22	246.09	244.96	243.82
260	257.24	256.31	255.39	254.46	253.54	252.61
280	265.65	265.00	264.34	263.69	263.03	262.38
300	275.16	274.42	273.67	272.93	272.19	271.45
320	285.10	284.30	283.50	282.70	281.90	281.11
340	294.83	293.99	293.16	292.32	291.49	290.65
360	304.42	303.73	303.04	302.35	301.65	300.96
380	313.71	313.39	313.08	312.77	312.46	312.15
400	323.59	323.49	323.39	323.30	323.20	323.11
420	332.69	333.01	333.32	333.64	333.95	334.26
440	342.76	343.11	343.46	343.81	344.16	344.50
460	352.66	353.38	354.09	354.80	355.52	356.23
480	364.52	365.34	366.17	366.99	367.82	368.64
500	375.06	376.31	377.57	378.82	380.08	381.33
520	385.70	387.30	388.90	390.50	392.10	393.70
540	397.51	399.10	400.70	402.30	403.89	405.49
560	409.14	410.90	412.66	414.42	416.18	417.94
580	419.94	421.95	423.96	425.96	427.97	429.98
600	431.32	433.56	435.80	438.05	440.29	442.53
620	443.43	445.72	448.01	450.30	452.59	454.88
640	455.04	457.57	460.09	462.62	465.14	467.67
660	467.10	469.93	472.76	475.59	478.42	481.25
680	477.88	481.32	484.75	488.19	491.62	495.05
700	490.28	493.91	497.55	501.18	504.81	508.45
720	502.85	506.65	510.45	514.24	518.04	521.83
740	514.89	519.07	523.24	527.42	531.60	535.77
760	527.83	532.18	536.53	540.87	545.22	549.57
780	539.95	544.71	549.47	554.23	558.99	563.75
800	553.32	558.47	563.62	568.77	573.91	579.06
820	566.16	571.68	577.19	582.71	588.22	593.73
840	578.50	584.25	590.00	595.75	601.50	607.25
860	592.13	597.94	603.74	609.55	615.35	621.16
880	606.12	612.25	618.39	624.52	630.66	636.79
900	618.17	625.11	632.04	638.98	645.91	652.85
920	632.49	639.38	646.27	653.16	660.05	666.94
940	645.97	653.16	660.35	667.54	674.72	681.91
960	660.26	667.76	675.27	682.77	690.27	697.77
980	675.02	682.68	690.34	698.00	705.66	713.32
1000	689.02	696.98	704.95	712.91	720.88	728.84
1020	704.02	712.26	720.50	728.74	736.97	745.21
1040	718.15	726.74	735.33	743.93	752.52	761.11
1060	731.92	741.08	750.24	759.41	768.57	777.74
1080	748.06	757.47	766.88	776.28	785.69	795.09
1100	762.00	772.00	781.99	791.98	801.98	811.97
1120	775.97	786.43	796.89	807.35	817.81	828.27
1140	790.71	801.52	812.34	823.15	833.97	844.78
1160	806.26	817.36	828.45	839.55	850.65	861.74

Table 2.12: Hydrocarbon vapor enthalpy data for MEABP = 500°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	8.34	8.55	8.76	8.97	9.18	9.40
40	16.33	16.42	16.51	16.60	16.70	16.79
60	24.17	24.34	24.51	24.69	24.86	25.03
80	32.51	32.93	33.34	33.76	34.18	34.59
100	40.09	40.74	41.40	42.06	42.71	43.37
120	48.80	49.68	50.56	51.44	52.33	53.21
140	58.68	59.62	60.57	61.51	62.46	63.41
160	67.31	68.39	69.48	70.57	71.65	72.74
180	77.26	78.24	79.22	80.20	81.18	82.16
200	86.56	87.86	89.15	90.45	91.75	93.04
220	95.50	97.15	98.80	100.45	102.10	103.76
240	106.01	107.79	109.58	111.36	113.15	114.93
260	117.04	118.85	120.65	122.45	124.25	126.05
280	127.04	129.10	131.16	133.22	135.28	137.33
300	137.82	140.09	142.36	144.63	146.90	149.17
320	149.33	151.47	153.61	155.74	157.88	160.02
340	159.02	161.68	164.34	167.00	169.66	172.32
360	171.17	173.89	176.62	179.35	182.07	184.80
380	182.97	185.66	188.34	191.03	193.72	196.40
400	194.53	197.33	200.13	202.93	205.73	208.53
420	205.53	208.75	211.97	215.19	218.41	221.63
440	218.36	221.56	224.75	227.94	231.14	234.33
460	230.10	233.59	237.09	240.58	244.08	247.57
480	242.83	246.37	249.92	253.47	257.02	260.57
500	255.38	259.05	262.73	266.40	270.07	273.74
520	268.39	272.46	276.54	280.62	284.70	288.77
540	283.06	287.09	291.11	295.14	299.17	303.19
560	295.46	300.25	305.04	309.82	314.61	319.40
580	310.38	315.43	320.48	325.53	330.59	335.64
600	325.04	330.17	335.31	340.45	345.59	350.73
620	339.59	345.05	350.51	355.96	361.42	366.88
640	353.23	358.89	364.56	370.22	375.88	381.55
660	367.88	373.80	379.71	385.62	391.54	397.45
680	383.10	389.41	395.73	402.04	408.36	414.68
700	398.75	405.14	411.53	417.93	424.32	430.71
720	413.92	420.81	427.69	434.58	441.46	448.35
740	430.95	438.00	445.05	452.10	459.15	466.20
760	446.71	454.06	461.41	468.76	476.11	483.46
780	463.72	471.16	478.61	486.06	493.51	500.95
800	479.61	487.70	495.78	503.87	511.96	520.05

Table 2.13: Hydrocarbon liquid enthalpy data for MEABP = 500°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	7.92	8.06	8.20	8.33	8.47	8.61
40	15.09	15.52	15.94	16.36	16.78	17.20
60	23.22	23.76	24.30	24.85	25.39	25.93
80	32.36	32.87	33.39	33.91	34.42	34.94
100	39.91	40.75	41.59	42.43	43.27	44.11
120	48.39	49.37	50.34	51.32	52.30	53.27
140	56.84	57.94	59.05	60.15	61.25	62.36
160	65.97	67.13	68.29	69.45	70.61	71.76
180	75.95	77.08	78.20	79.33	80.45	81.57
200	85.29	86.55	87.80	89.05	90.31	91.56
220	94.60	96.04	97.48	98.92	100.36	101.80
240	104.68	106.13	107.58	109.03	110.48	111.93
260	114.30	115.96	117.62	119.28	120.94	122.61
280	123.68	125.66	127.65	129.63	131.62	133.61
300	134.60	136.55	138.51	140.46	142.42	144.37
320	145.15	147.36	149.56	151.76	153.96	156.16
340	156.17	158.55	160.93	163.31	165.69	168.07
360	167.38	169.90	172.42	174.94	177.47	179.99
380	178.42	181.16	183.89	186.63	189.37	192.11
400	189.95	192.82	195.69	198.56	201.43	204.30
420	201.97	204.94	207.91	210.88	213.85	216.82
440	214.53	217.58	220.64	223.69	226.75	229.80
460	226.92	230.02	233.13	236.23	239.34	242.44
480	239.57	242.96	246.36	249.76	253.15	256.55
500	252.55	255.81	259.06	262.32	265.58	268.84
520	264.31	267.69	271.07	274.44	277.82	281.20
540	277.65	281.20	284.75	288.30	291.85	295.40
560	290.79	294.62	298.46	302.29	306.13	309.96
580	303.45	307.57	311.68	315.80	319.92	324.03
600	317.22	321.17	325.11	329.05	333.00	336.94
620	331.49	335.58	339.67	343.76	347.85	351.95
640	345.88	350.23	354.59	358.94	363.30	367.65
660	359.94	364.60	369.25	373.90	378.56	383.21
680	374.04	379.04	384.04	389.04	394.04	399.04
700	388.85	394.13	399.42	404.70	409.98	415.26
720	403.65	409.19	414.74	420.28	425.83	431.37
740	419.07	424.95	430.83	436.72	442.60	448.48
760	434.54	440.80	447.07	453.34	459.61	465.88
780	449.97	456.58	463.19	469.79	476.40	483.00
800	466.08	472.93	479.78	486.63	493.48	500.33
820	481.86	489.12	496.37	503.63	510.88	518.14
840	497.97	505.71	513.46	521.20	528.95	536.69
860	515.04	523.28	531.53	539.77	548.01	556.26
880	530.48	539.65	548.81	557.97	567.14	576.30
900	548.65	558.02	567.40	576.77	586.15	595.52

Table 2.14: Hydrocarbon liquid enthalpy data for MEABP = 600 °F and K = 11 – 12.

T (^{o}F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	159.95	156.45	152.95	149.45	145.95	142.45
20	166.78	163.39	160.00	156.61	153.22	149.83
40	172.72	169.52	166.33	163.13	159.93	156.73
60	178.50	175.62	172.73	169.84	166.96	164.07
80	184.91	182.23	179.55	176.87	174.19	171.51
100	192.18	189.47	186.76	184.05	181.34	178.63
120	199.43	196.84	194.24	191.65	189.06	186.47
140	206.08	203.70	201.31	198.93	196.55	194.17
160	213.33	211.14	208.95	206.75	204.56	202.37
180	221.11	219.07	217.04	215.00	212.96	210.93
200	228.41	226.52	224.63	222.74	220.85	218.96
220	236.70	234.92	233.14	231.37	229.59	227.81
240	244.40	242.89	241.38	239.87	238.36	236.85
260	252.48	251.10	249.72	248.34	246.96	245.58
280	261.25	259.93	258.62	257.31	256.00	254.69
300	269.74	268.74	267.74	266.74	265.74	264.74
320	279.44	278.43	277.42	276.41	275.40	274.39
340	288.20	287.18	286.16	285.14	284.12	283.11
360	297.15	296.45	295.75	295.05	294.35	293.65
380	306.77	306.44	306.11	305.78	305.45	305.12
400	315.66	315.65	315.63	315.61	315.60	315.58
420	326.10	326.28	326.47	326.65	326.84	327.02
440	336.49	336.54	336.60	336.65	336.70	336.76
460	346.54	346.85	347.15	347.46	347.77	348.07
480	357.46	358.19	358.93	359.67	360.40	361.14
500	369.03	369.66	370.28	370.91	371.53	372.16
520	380.07	380.92	381.78	382.63	383.49	384.34
540	390.63	391.82	393.00	394.19	395.38	396.56
560	401.89	403.19	404.49	405.79	407.09	408.39
580	413.01	414.50	416.00	417.50	418.99	420.49
600	423.35	425.15	426.95	428.75	430.54	432.34
620	436.02	437.74	439.47	441.19	442.92	444.65
640	448.66	450.51	452.35	454.20	456.05	457.90
660	459.97	462.29	464.61	466.94	469.26	471.58
680	471.32	474.02	476.72	479.43	482.13	484.83
700	484.30	487.00	489.70	492.40	495.09	497.79
720	497.51	500.36	503.20	506.05	508.89	511.73
740	509.77	513.11	516.44	519.78	523.12	526.46
760	522.35	526.05	529.75	533.44	537.14	540.84
780	535.36	539.25	543.15	547.05	550.94	554.84
800	547.58	551.93	556.29	560.64	565.00	569.35
820	561.05	565.64	570.23	574.83	579.42	584.01
840	574.27	578.95	583.62	588.30	592.97	597.65
860	587.07	592.07	597.07	602.07	607.06	612.06
880	600.32	605.69	611.07	616.45	621.82	627.20
900	613.87	619.47	625.08	630.68	636.29	641.89
920	626.79	632.93	639.07	645.20	651.34	657.48
940	640.17	646.58	652.99	659.40	665.80	672.21
960	654.83	661.39	667.96	674.52	681.09	687.65
980	669.29	676.16	683.02	689.89	696.75	703.62
1000	682.24	689.52	696.80	704.08	711.36	718.64
1020	696.85	704.36	711.87	719.37	726.88	734.39
1040	711.12	718.94	726.75	734.56	742.38	750.19
1060	725.03	733.23	741.43	749.63	757.83	766.03
1080	739.54	748.10	756.65	765.21	773.76	782.31
1100	753.82	762.86	771.90	780.93	789.97	799.00
1120	768.44	777.86	787.28	796.71	806.13	815.55
1140	783.69	793.25	802.80	812.36	821.91	831.47

Table 2.15: Hydrocarbon vapor enthalpy data for MEABP = 600^{o}F and K = 11 – 12.

T (^{o}F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	7.74	7.81	7.87	7.94	8.00	8.07
40	15.25	15.41	15.56	15.72	15.87	16.02
60	22.46	22.76	23.06	23.35	23.65	23.95
80	30.81	31.15	31.48	31.82	32.16	32.50
100	38.70	39.22	39.73	40.25	40.76	41.27
120	46.46	47.15	47.84	48.53	49.22	49.91
140	54.20	55.07	55.95	56.82	57.69	58.57
160	63.97	64.75	65.53	66.31	67.08	67.86
180	72.44	73.29	74.13	74.98	75.83	76.67
200	81.27	82.28	83.30	84.31	85.32	86.34
220	89.89	91.19	92.48	93.78	95.08	96.38
240	100.40	101.50	102.59	103.69	104.78	105.88
260	110.69	111.64	112.58	113.52	114.46	115.41
280	120.32	121.50	122.68	123.87	125.05	126.23
300	130.81	132.05	133.29	134.53	135.77	137.01
320	140.10	141.58	143.05	144.52	145.99	147.47
340	151.40	152.88	154.36	155.84	157.32	158.80
360	161.59	163.42	165.25	167.09	168.92	170.75
380	172.03	173.82	175.61	177.40	179.20	180.99
400	183.51	185.38	187.24	189.10	190.97	192.83
420	195.52	197.35	199.17	201.00	202.83	204.65
440	205.96	208.14	210.33	212.51	214.70	216.89
460	218.32	220.59	222.85	225.12	227.38	229.65
480	230.42	232.78	235.14	237.50	239.87	242.23
500	242.71	245.17	247.62	250.08	252.54	255.00
520	254.97	257.53	260.09	262.65	265.21	267.77
540	267.83	270.54	273.26	275.98	278.70	281.42
560	280.18	283.23	286.27	289.31	292.35	295.40
580	293.20	296.48	299.76	303.05	306.33	309.61
600	305.91	309.51	313.12	316.72	320.33	323.93
620	318.91	322.70	326.49	330.28	334.07	337.86
640	332.82	336.78	340.74	344.70	348.66	352.63
660	346.82	350.78	354.74	358.69	362.65	366.60
680	360.05	364.35	368.66	372.97	377.28	381.59
700	373.72	378.30	382.88	387.46	392.04	396.63
720	387.57	392.24	396.91	401.57	406.24	410.91
740	402.54	407.32	412.11	416.90	421.69	426.48
760	417.13	422.11	427.09	432.07	437.05	442.04
780	431.65	436.84	442.03	447.23	452.42	457.61
800	447.21	452.47	457.73	462.98	468.24	473.50
820	461.39	466.85	472.30	477.75	483.20	488.66
840	477.79	483.40	489.00	494.60	500.20	505.81
860	493.26	499.13	504.99	510.86	516.72	522.58
880	509.35	515.19	521.03	526.87	532.72	538.56
900	524.08	530.32	536.56	542.81	549.05	555.29
920	540.55	546.99	553.43	559.87	566.32	572.76
940	557.03	563.65	570.27	576.89	583.51	590.13
960	573.65	580.53	587.41	594.30	601.18	608.06
980	589.75	597.06	604.37	611.67	618.98	626.29
1000	606.56	614.20	621.83	629.47	637.11	644.75
1020	623.47	631.27	639.07	646.87	654.66	662.46
1040	641.90	650.25	658.59	666.94	675.29	683.64
1060	659.93	668.58	677.23	685.87	694.52	703.17
1080	677.10	686.52	695.95	705.38	714.81	724.23
1100	695.64	705.56	715.48	725.40	735.32	745.24

Table 2.16: Hydrocarbon liquid enthalpy data for MEABP = 800^{o}F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	147.65	143.99	140.33	136.67	133.01	129.35
20	152.66	149.20	145.74	142.29	138.83	135.38
40	159.17	155.88	152.58	149.29	145.99	142.70
60	165.75	162.48	159.20	155.92	152.64	149.36
80	171.83	168.66	165.49	162.32	159.15	155.98
100	178.70	175.74	172.77	169.81	166.85	163.89
120	185.20	182.35	179.49	176.64	173.79	170.93
140	192.28	189.64	187.00	184.36	181.72	179.07
160	199.87	197.35	194.84	192.32	189.81	187.29
180	207.11	204.67	202.24	199.80	197.37	194.93
200	215.73	213.29	210.85	208.41	205.97	203.53
220	223.77	221.47	219.17	216.87	214.57	212.27
240	231.59	229.44	227.30	225.16	223.01	220.87
260	239.35	237.48	235.61	233.74	231.87	230.00
280	248.35	246.60	244.85	243.10	241.35	239.60
300	257.59	255.83	254.08	252.32	250.57	248.81
320	266.19	264.51	262.82	261.13	259.45	257.76
340	275.31	273.71	272.10	270.49	268.88	267.27
360	285.00	283.47	281.93	280.39	278.85	277.31
380	294.05	292.81	291.58	290.34	289.10	287.87
400	303.97	302.92	301.88	300.84	299.79	298.75
420	312.78	311.94	311.10	310.26	309.41	308.57
440	323.67	322.70	321.72	320.75	319.77	318.80
460	334.13	333.45	332.78	332.10	331.43	330.76
480	344.19	343.63	343.07	342.51	341.95	341.39
500	356.14	355.67	355.20	354.73	354.26	353.79
520	367.42	367.63	367.84	368.05	368.26	368.47
540	377.61	377.95	378.28	378.62	378.95	379.29
560	388.30	388.94	389.58	390.22	390.86	391.50
580	401.02	401.48	401.95	402.42	402.89	403.36
600	410.89	411.83	412.77	413.71	414.65	415.59
620	422.64	423.84	425.05	426.25	427.45	428.66
640	434.21	435.63	437.04	438.45	439.86	441.28
660	445.97	447.57	449.17	450.77	452.37	453.97
680	458.16	459.94	461.73	463.51	465.29	467.08
700	470.66	472.55	474.45	476.34	478.23	480.13
720	482.53	484.77	487.00	489.23	491.46	493.69
740	495.57	497.91	500.26	502.61	504.95	507.30
760	508.08	510.65	513.22	515.79	518.36	520.93
780	520.96	523.72	526.49	529.25	532.01	534.77
800	534.29	537.22	540.15	543.08	546.02	548.95
820	547.30	550.54	553.77	557.00	560.24	563.47
840	560.83	564.13	567.42	570.72	574.01	577.31
860	574.74	578.08	581.41	584.75	588.09	591.43
880	587.02	590.80	594.59	598.37	602.16	605.94
900	600.30	604.39	608.49	612.59	616.68	620.78
920	613.85	618.38	622.91	627.45	631.98	636.52
940	628.26	633.00	637.74	642.49	647.23	651.97
960	642.25	647.20	652.16	657.12	662.08	667.04
980	655.83	661.14	666.45	671.76	677.07	682.38
1000	670.38	675.96	681.54	687.13	692.71	698.29
1020	684.46	690.48	696.51	702.54	708.56	714.59
1040	698.46	704.76	711.06	717.36	723.66	729.96
1060	712.70	719.35	726.00	732.65	739.29	745.94
1080	727.97	734.95	741.94	748.92	755.91	762.89
1100	741.39	748.93	756.48	764.02	771.56	779.11
1120	756.85	764.55	772.26	779.97	787.67	795.38
1140	771.12	779.40	787.67	795.95	804.23	812.50
1160	787.50	796.00	804.50	813.00	821.50	830.01

Table 2.17: Hydrocarbon vapor enthalpy data for MEABP = 800°F and K = 11 – 12.

T (^{O}F)	C4H10	C5H12	100	150	200	250	300	350	400	450	500	550	600	700	800	900	1000	1100	1200
50	1.47	0.38	0.36	0.12	0.04	0.10													
75	2.32	0.65	0.61	0.22	0.07	0.23	0.01												
100	3.51	1.07	1.00	0.39	0.14	0.47	0.01												
125	5.20	1.64	1.53	0.65	0.25	0.90	0.03	0.01											
150	7.25	2.46	2.32	0.99	0.42	0.16	0.06	0.02	0.01										
175	9.85	3.51	3.35	1.50	0.66	0.27	0.10	0.04	0.01										
200	13.10	4.96	4.56	2.22	0.99	0.44	0.17	0.07	0.02	0.01									
225	17.15	6.78	6.50	3.15	1.48	0.67	0.29	0.12	0.04	0.02	0.01								
250	21.95	9.01	8.56	4.34	2.11	1.01	0.45	0.19	0.08	0.03	0.01								
275	27.91	12.02	11.40	5.82	3.02	1.47	0.68	0.31	0.13	0.05	0.02	0.01							
300	34.90	15.20	14.40	7.78	4.05	2.08	1.00	0.47	0.21	0.09	0.04	0.01							
325	42.40	19.30	18.80	10.05	5.45	2.81	1.43	0.70	0.33	0.14	0.06	0.02	0.01						
350	51.80	24.30	23.20	13.10	7.41	3.80	2.02	1.02	0.48	0.23	0.10	0.04	0.02						
375	60.50	30.20	28.90	16.40	9.42	5.15	2.81	1.45	0.71	0.35	0.16	0.07	0.03						
400	72.10	37.10	35.02	20.70	12.02	6.76	3.68	1.94	1.01	0.51	0.25	0.11	0.05	0.01					
425	83.20	43.60	41.80	25.20	14.90	8.55	4.76	2.66	1.39	0.74	0.36	0.17	0.08	0.01					
450	95.30	51.20	49.10	30.80	18.30	10.70	6.22	3.48	1.86	1.01	0.51	0.25	0.12	0.02					
475		60.00	57.90	36.50	22.20	13.20	7.83	4.51	2.44	1.41	0.72	0.36	0.18	0.04	0.01				
500		70.10	67.40	43.20	27.30	16.30	9.80	5.77	3.21	1.84	1.02	0.52	0.26	0.06	0.01				
525		80.20	77.50	50.10	32.50	19.90	12.40	7.42	4.24	2.48	1.23	0.74	0.38	0.09	0.02				
550		92.00	88.70	59.10	39.10	24.10	15.20	9.23	5. 42	3.24	1.81	1.01	0.53	0.14	0.03	0.01			
575			100.00	69.20	45.20	29.20	18.60	11.60	6.97	4.20	2.42	1.35	0.74	0.20	0.05	0.01			
600				79.70	52.70	34.10	22.30	14.00	8.48	5.22	3.11	1.79	1.01	0.29	0.08	0.02			
650				100.00	71.20	47.60	31.60	20.60	13.00	8.20	5.04	3.05	1.72	0.56	0.17	0.05	0.01		
700					90.00	62.20	42.10	28.40	18.70	12.20	7.65	4.68	2.83	0.98	0.33	0.10	0.03	0.01	
750						79.10	55.20	38.10	25.90	17.60	11.20	7.15	4.24	1.62	0.61	0.20	0.06	0.01	
800						96.70	70.00	49.70	33.50	23.50	15.30	10.70	6.48	2.52	1.02	0.37	0.11	0.03	0.01
850							86.30	63.80	44.10	31.80	20.80	14.80	9.25	3.76	1.61	0.62	0.20	0.06	0.02
900								78.50	57.20	41.70	28.10	19.10	12.60	5.58	2.43	1.01	0.36	0.12	0.03
950								96.10	71.90	53.50	37.10	26.20	17.90	8.05	3.69	1.59	0.62	0.22	0.07
1000									87.40	68.10	47.90	34.80	23.20	11.30	5.36	2.45	1.02	0.39	0.13
1050										83.90	59.60	44.50	30.80	15.20	7.70	3.63	1.59	0.64	0.24
1100										100.00	73.80	56.10	39.10	20.40	10.80	5.22	2.32	0.99	0.42
1150											89.90	68.50	50.20	26.30	14.10	7.36	3.43	1.52	0.65
1200												82.00	62.40	32.40	18.40	9.90	4.86	2.21	1.01

Table 2.18: Vapor pressure data for hydrocarbons.

2.8 Vapor pressure

For pure components and mixtures of known chemical speices and compositions, Antoine expression can be used to conveniently estimate the saturated and partial vapor pressures in gaseous/vapor mixtures. For crude as well as petroleum fractions therefore, correlations are necessary to evaluate vapor pressure. Table 2.18 presents the correlation presented by Maxwell (1950) to relate vapor pressure with temperature and normal boiling point of the hydrocarbon. For crude/petroleum fractions, the normal boiling point is estimated as the volume average boiling point (VABP). Next we present an illustrative example to estimate the vapor pressure of the crude stream at a chosen temperature. The vapor pressure curves are of prominent usage in the flash zone calculations of the CDU. Therefore, we present another illustrative example to convey the efficacy of the vapor pressure plots. **For interpolation purposes, it can be assumed that both axes data agree to a linear fit for a log-log plot.**

Q 2.8: Estimate the vapor pressure of heavy Saudi crude oil at 500 ^oF.

Solution:

$T_v = t_{20}+t_{50}+t_{80}/3 = (338+703+1104)/3= 715\ ^0F$

From Table 2.18, at 500 ^oF, vapor pressure = 0.05 atm.

Q 2.9: Estimate the boiling point of the crude stream at 2.5 atm. The normal boiling point is 700 ^oF

Solution:

From Table 2.18, for column corresponding to 700 ^oF, the vapor pressure of the crude stream at 800 ^oF is 2.52 atm. Therefore, the boiling point of the crude stream is 800 ^oF at 2.5 atm.

2.9 Estimation of Product TBP from crude TBP

One of the important features of refinery process design is to estimate the crude distillation product properties (TBP) for a given assay of the crude oil. In this section, we elaborately present the procedures that need to be followed to evaluate the product TBP properties from feed properties.

A CDU essentially products 5 products namely gas to naphtha (product 1), kerosene (product 2), light gas oil i.e., LGO (product 3), heavy gas oil i.e., HGO (product 4) and atmospheric residue (product 5). An important feature of these products is that they are identified to be distinct based on the cut points of the crude TBP. Associated cut points for these products are summarized as follows:

a) Naptha : Gas to 375^{0}F
b) Kerosene: 375^{0}F- 480^{0}F
c) LGO: 480^{0}F- 610^{0}F
d) HGO: 610^{0}F-680^{0}F
e) Residue: 680^{o}F +

A		B		C		D		E	
T (°F)	Δ (°F)	T (°F)	Δ (°F)	T (°F)	Δ (°F)	T (°F)	Δ (°F)	T (°F)	Δ (°F)
126.4	-41.1	413.2	-18.9	476.5	-7.0	470.5	2.6	504.8	-13.3
316.4	-2.1	421.7	-20.2	453.4	-6.1	482.5	1.7	208.6	-6.6
324.1	-3.6	432.0	-21.6	460.3	-7.4	494.5	0.8	222.3	-8.0
334.4	-5.4	442.3	-22.3	471.4	-9.0	513.4	-1.1	236.0	-8.9
343.0	-7.0	455.1	-21.6	480.8	-10.5	531.3	-3.4	258.2	-11.0
351.5	-8.4	466.3	-20.3	490.2	-11.9	545.9	-4.8	280.5	-12.2
360.1	-10.0	474.8	-19.1	498.8	-13.5	559.6	-7.1	307.9	-14.7
369.5	-11.5	486.8	-17.5	515.1	-16.3	572.4	-9.0	328.4	-15.9
378.1	-13.0	495.4	-16.7	521.9	-17.9	589.6	-12.2	345.5	-17.3
385.8	-14.4	504.8	-15.6	531.3	-19.7	601.5	-14.3	367.0	-19.3
395.2	-15.9	516.8	-14.8	543.3	-21.9	612.7	-16.4	386.6	-20.7
404.6	-17.5	355.8	-4.5	550.2	-23.9	622.1	-18.5	406.3	-22.4
		363.5	-5.6	559.6	-25.6	628.9	-20.4	426.9	-23.5
		371.2	-6.1	565.6	-27.7	637.5	-22.0	446.6	-24.9
		380.7	-7.3	576.7	-30.5	646.1	-23.7	466.3	-26.6
		390.1	-8.2	584.4	-32.3	652.9	-25.7	486.0	-28.2
		396.9	-8.9	593.0	-34.2	659.8	-27.2	504.8	-29.5
		402.9	-9.6	639.2	-40.5	666.6	-29.2	522.8	-30.7
		409.8	-10.5			672.6	-30.7	544.2	-32.6
		417.5	-11.2			682.0	-32.7	565.6	-34.2
		428.6	-12.8			688.9	-35.1	585.3	-35.8
		437.2	-13.8					602.4	-37.0
		444.9	-15.1					620.4	-38.6
		455.1	-16.8					648.6	-40.6
		454.3	-17.0					662.3	-41.6
		464.6	-18.2					677.7	-42.8
		474.8	-20.3					690.6	-43.9
		481.7	-21.7						
		488.5	-23.3						
		496.2	-25.4						
		504.8	-27.5						
		511.6	-29.3						
		519.3	-31.2						
		524.5	-32.6						
		532.2	-34.4						
		545.9	-37.7						
		553.6	-39.8						
		560.4	-41.7						
		569.0	-43.9						
		578.4	-45.4						
		587.0	-45.8						
		599.0	-45.4						
		615.2	-42.1						
		628.1	-37.9						
		640.9	-35.1						
		652.1	-33.0						

Table 2.19: End point correlation data presented by Good, Connel et. al. Data sets represent fractions whose cut point starts at 200 °F TBP or lower (Set A); 300 °F (Set B); 400 °F (Set C); 500 °F (Set D); 90% vol temperature of the cut Vs. 90 % vol TBP cut for all fractions (Set E).

The first step in the estimation of product properties is to evaluate the TBP of the products namely naphtha, kerosene, LGO, HGO and residue. Often crude TBP is provided to cover the cut points specifically till HGO and residue is often ignored. However, using a volumetric balance over the crude and crude assay, the residue TBP can be obtained for a given TBP of the crude and accurately obtained TBP of the other four products (naphtha, kerosene, LGO and HGO). The residue volumetric balance is often ignored due to the fact that the empirical correlations do not satisfy the volumetric balance law. The conceptual procedure to evaluate the product TBPs from crude TBP is summarized as follows:

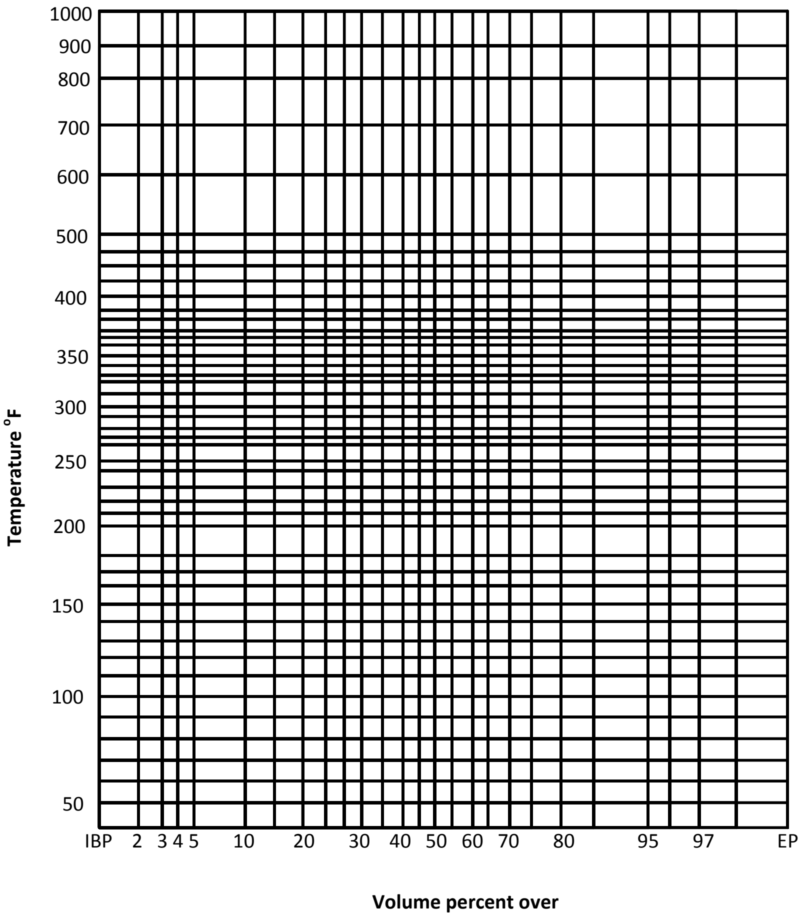


Figure 2.5: Probability chart developed by Thrift for estimating ASTM temperatures from any two known values of ASTM temperatures.

Segment of Distillation Curve, Volume Percent												**ASTM 50 % to TBP 50 %**	
0 to 10 %		10 to 30 %		30 – 50 %		50 – 70%		70 – 90%		90 - 100%		ASTM 50 % Temp (°F)	TBP 50 % Temp (°F)
ASTM ΔT (°F)	TBP ΔT (°F)	ASTM ΔT (°F)	TBP ΔT (°F)	ASTM ΔT (°F)	TBP ΔT (°F)	ASTM ΔT (°F)	TBP ΔT (°F)	ASTM ΔT (°F)	TBP ΔT (°F)	ASTM ΔT (°F)	TBP ΔT (°F)		
0.27	0.27	0.27	0.55	0.27	0.27	0.55	1.09	0.28	0.82	0.55	1.09	101.70	-9.87
1.37	4.37	2.74	4.64	1.93	4.89	3.58	8.16	4.13	7.34	6.32	7.61	141.50	-8.75
3.01	8.19	5.21	11.74	5.78	13.05	3.30	7.07	12.10	17.66	14.29	16.30	199.15	-7.62
6.03	13.92	10.14	20.74	10.45	18.21	7.97	12.50	17.87	25.00	22.26	26.08	258.17	-5.66
8.49	19.65	16.71	31.93	13.75	23.92	20.89	31.52	23.09	30.43	30.23	33.41	310.33	-4.26
11.23	25.11	21.10	38.76	18.70	30.44	29.14	42.12	28.31	36.95	36.00	39.66	359.75	-2.32
14.52	30.29	26.03	46.13	22.55	36.69	33.54	46.74	32.98	41.30	41.77	44.82	416.04	-0.10
18.08	36.57	29.59	51.04	27.22	42.94	38.21	52.99	39.03	48.09	46.16	50.25	464.10	2.11
22.19	42.85	33.15	56.50	31.62	48.10	45.35	59.50	43.97	54.07	51.11	54.87	501.18	4.05
27.12	49.95	38.36	62.24	36.02	53.81	53.32	67.65	51.12	61.40	55.23	60.31	553.37	7.35
30.96	55.14	42.74	66.61	40.14	58.70	58.82	73.36	59.36	69.00	59.08	63.84	595.95	10.37
35.07	60.33	46.85	72.07	44.53	63.04	64.86	79.88	64.30	73.08	65.12	70.90	631.67	13.38
39.73	66.61	50.69	75.90	48.93	67.12	71.45	85.31	69.52	78.51	69.52	75.24	685.26	18.31
43.56	71.79	54.52	80.26	55.52	73.91	77.22	90.47	76.12	85.30	73.37	81.50	725.13	22.96
47.12	76.44	58.90	84.09	60.47	78.79	87.11	98.88	83.26	91.27	78.04	88.83	755.39	27.33
50.69	81.35	63.01	88.19	65.14	83.41	92.88	104.59	90.95	98.88	82.99	96.44	787.03	32.51
56.16	87.90	68.49	92.29	69.81	87.49	99.20	110.02	96.45	102.68	85.74	101.88	818.68	38.23
61.10	93.09	72.05	95.84	74.48	91.83	105.79	116.54	98.92	105.94	89.32	108.40	847.60	44.50
66.03	99.65	75.34	98.57	80.25	96.99	113.76	123.60	104.69	111.10	92.34	114.38	869.64	49.67
70.41	105.65	75.34	98.84	85.47	102.42	120.35	129.84	111.01	117.35	94.82	119.82	897.23	58.37
75.62	111.39	90.14	111.68	91.24	106.77	125.30	134.73	118.70	124.68	97.02	125.80		
81.37	117.94	94.25	115.23	95.36	110.84	131.07	140.17	125.57	131.47				
85.21	123.40	97.81	117.42	103.60	118.17	135.46	145.05	130.24	135.81				
89.04	128.04	102.47	122.06	108.54	122.52	140.68	149.40	137.11	143.96				
		106.85	125.62	114.59	127.95	144.81	154.56	142.60	148.85				
		110.41	128.90	120.63	133.38	150.58	161.08	147.00	153.47				
		114.79	132.45	127.23	139.90	157.72	167.33	150.30	156.73				
		119.45	137.09	131.35	143.43	162.39	173.85	153.87	160.80				
		124.93	141.74	137.12	149.68	167.62	182.00	156.34	163.79				
		129.32	146.11	143.44	156.47	172.29	187.98	159.92	168.41				
		133.15	150.20	148.11	160.81	175.31	191.78	163.76	171.67				
		138.90	155.39	153.05	166.52	179.43	196.13	167.61	177.65				
		144.38	160.58	158.27	173.04	178.06	198.31	170.63	182.00				
		150.14	167.41	162.67	177.93			175.03	187.43				
		155.34	173.96	166.79	183.09			178.33	191.51				
		160.82	180.52	169.54	186.89								
		166.03	187.07	174.22	192.87								
		172.60	194.99	176.97	197.22								
		179.45	203.45	179.71	200.21								

Table 2.20: ASTM-TBP correlation data from Edmister method.

a) For a given crude TBP cut point for the desired product, Good et. al presents correlations to evaluate the ASTM end point. The correlations are applicable for both light and middle distillate products. In other words, Good et al. correlations can be used to estimate the ASTM end points of all products other than the atmospheric residue.
b) It can be fairly assumed that the 50 % TBP of the crude matches with the 50 % TBP point of the products. Eventually, the 50 % TBP can be converted to obtain 50 % ASTM point for the products (other than the residue).
c) Thrift developed a probability chart on which all points connecting IBP, ASTM 10 %, ASTM 20 % …… ASTM end point shall lie on a straight line. In other words, knowing any two ASTM temperatures at any two volume % values, the entire ASTM data for the product can be estimated.
d) The ASTM temperatures of the product can be converted to the TBP using Edmister correlation.

Table 2.19 illustrates the end point correlation developed by Good, Connel et. al. The Edmister correlation useful to convert ASTM to TBP and vice-versa is summarized in Table 2.20. Figure 2.5 illustrates the probability curves developed by Thrift. **These two tables and figure are the most important tools for the evaluation of refinery product TBPs from crude TBPs.**

Q 2.10: For the heavy Saudi crude oil whose TBP was provided previously, using end point correlation provided by Good et. al. and assuming the 50 % TBP of product matches with the corresponding 50 % TBP on the crude, estimate the TBPs of all products other than residue. In the calculation procedure, estimate the accurate end point for the corresponding TBP cut points using interpolation method. Subsequently, plot the TBPs to obtain the pseudo-components volume % for all products to verify the law of volumetric balance. Eventually comment upon the efficacy of the calculation procedure.

Solution:

The cut ranges for various products are taken as:

a) Naphtha: Gas to 375 °F
b) Kerosene: 375 °F to 480 °F
c) LGO: 480 – 610 °F
d) HGO: 610 – 680 °F

a) TBP for Naphtha product

Yield vol % = 22.8

TBP end point of cut = 375 °F

From end point correlation, ASTM end point of cut = 375 -11 = 364 °F

50 % of the cut (from Crude TBP) = 22.8/2 = 11.4

TBP 50 % corresponding to 11.4 (from crude TBP) = 225 °F

ASTM 50 % of the cut = 231 °F (From Edmister correlation)

Using probability chart and using ASTM 50 % and ASTM end point and drawing a straight line between these two values, the ASTM temperatures of the Naptha cut are evaluated. These are presented as follows:

Vol %	ASTM		TBP	
	(°F)	ΔF	ΔF	(°F)
IBP	150	34	62	80
10	184	36	47	142
30	210	23	38	189
50	233	-	-	227
70	253	20	31	258
90	290	37	47	305
100	364	74	85	390

b) Kerosene TBP

Cut range: 375 – 480 °F

Yield = 31.6 – 22.8 = 8.8 %

TBP end point of cut = 480 °F

ASTM end point of the cut (from End point correlation) = 480 – 10 = 470 °F

50 % volume of the cut = 22.8 + 8.8/2 = 27.2 %

TBP 50 % corresponding to the cut = 428 °F

ASTM 50 % of the cut (From Edmister correlation) = 427 °F

Using probability chart and using ASTM 50 % and ASTM end point and drawing a straight line between these two values, the ASTM temperatures of the Kerosene cut are evaluated. These are presented as follows:

Vol %	ASTM		TBP	
	(°F)	ΔF	ΔF	(°F)
IBP	393	15	33	357
10	408	12	25	390
30	420	7	13	415
50	427	-	-	428
70	434	7	12	440
90	445	11	16	456
100	470	25	28	484

c) LGO product

Cut range: 480 - 610 ^{o}F

Yield = 42.2 – 31.6 = 10.6 %

TBP end point of cut = 610 ^{o}F

ASTM end point of the cut (from End point correlation) = 610 - 18 = 592 ^{o}F

50 % volume of the cut = 31.6 + 10.6/2 = 37.1 %

TBP 50 % corresponding to the cut = 544^{o}F

ASTM 50 % of the cut (From Edmister correlation) = 537 ^{o}F

Using probability chart and using ASTM 50 % and ASTM end point and drawing a straight line between these two values, the ASTM temperatures of the LGO cut are evaluated. These are presented as follows:

Vol %	ASTM		TBP	
	(^{0}F)	ΔF	ΔF	(^{0}F)
IBP	493	18	38	455
10	511	14	29	493
30	525	12	22	522
50	537	-	-	544
70	549	12	20	564
90	565	16	23	587
100	592	27	31	618

d) HGO product

Cut range: 610 - 680 ^{o}F

Yield = 48 – 42.2 = 5.8 %

TBP end point of cut = 680 ^{o}F

ASTM end point of the cut (from End point correlation) = 680 - 20 = 660 ^{o}F

50 % volume of the cut = 42.2 +5.8/2 = 45.1 %

TBP 50 % corresponding to the cut =644^{o}F

ASTM 50 % of the cut (From Edmister correlation) = 631 ^{o}F

Using probability chart and using ASTM 50 % and ASTM end point and drawing a straight line between these two values, the ASTM temperatures of the HGO cut are evaluated. These are presented as follows:

Vol %	ASTM		TBP	
	(^{0}F)	ΔF	ΔF	(^{0}F)
IBP	602	12	27	582
10	614	9	20	609
30	623	8	15	629
50	631	-	-	644
70	636	5	9	653
90	644	8	12	665
100	660	16	19	684

The obtained TBPs of the products are plotted against various pseudo-component temperature ranges to obtain the pseudo-component volume % contribution in each cut. Subsequently, the pseudo-component break-up in both TBP and evaluated product cuts is presented below along with mid point ^{o}API and midpoint sulfur %:

P.C Range		**Vol%(A)**				**Mid Bpt**	**Mid Vol**	**API**	**S.G(B)**	**Sulfur Wt%(D)**
		Naphtha	**Kerosene**	**LGO**	**HGO**					
1	-12-80	0				35	2	89.1	0.641432	0
2	80-120	5.5				100	5	81.9	0.663074	0
3	120-160	10.5				140	6.75	77.7	0.676386	0
4	160-180	9.5				170	8	75	0.68523	0
5	180-220	20.5				200	9.75	69	0.705736	0
6	220-260	24.5				240	12.5	64.5	0.721939	0
7	260-300	18.3				280	15.5	59	0.742782	0
8	300-360	8.2	0.8			330	19.5	53	0.766938	0.05
9	360-400	3	15.7			380	23.5	49	0.783934	0.2
10	400-460		75.5	1		430	27.5	45	0.8017	0.35
11	460-520		8	27.5		490	32.5	40.5	0.822674	0.7
12	520-580			56.3		550	37.5	35.5	0.847305	1.35
13	580-620			15.2	19.5	600	41.75	33	0.860182	1.7
14	620-660				63.3	640	45	30	0.876161	2
15	660-760				17.2	710	50.75	25.5	0.901274	2.6
16	760-860					810	58.5	22	0.921824	2.75
17	860-900					880	63.75	19	0.940199	3.05
18	900-940					920	66.75	18	0.946488	3.2
19	940-980					960	69.5	17	0.952862	3.35
20	980-1020					1000	72.5	16	0.959322	3.55
21	1020-1060					1040	75.25	15	0.96587	3.7
22	1060-1160					1110	80.25	13	0.979239	3.9
23	1160-1280					1220	88.25	10	1	4.3
24	1280-1400					1340	96.25	6.5	1.025362	4.7
Total		100	100	100	100					

From the above table, it can be observed that the volumetric balance law is violated for Naphtha product itself, what to speak of other high boiling products. This is because for pseudo-component 1, the volume % in naphtha product is zero, where as it is about 4 % in the TBP. Therefore, it is herewith commented that the said procedure is not accurate enough to predict the TBP of the residue product. However, the obtained TBPs can be used to obtain a fair estimate of the product specific gravity and °API.

2.10 Estimation of product specific gravity and sulfur content

Using the obtained TBPs of the products, the pseudo-components distribution summarized in the table can be used along with the mid point °API and mid sulfur content of the crude for each pseudo-component, the product °API and average sulfur content can be estimated using the equations:

$$\text{av. S. G.} = \frac{\sum_{i=1}^{N_{PC}}[A_i][B_i]}{\sum_{i=1}^{N_{PC}}[A_i]} \qquad (5)$$

$$\text{av. \%sulfur content} = \frac{\sum_{i=1}^{N_{PC}}[A_i][B_i][C_i]}{\sum_{i=1}^{N_{PC}}[A_i][B_i]} \qquad (6)$$

where the columns $[A_i]$, $[B_i]$ and $[C_i]$ correspond to pseudo-component volume % in the product, pseudo-component °API from the crude assay and pseudo-component sulfur content from the crude assay.

An illustrative example is presented below to evaluate these average properties.

Q 2.11: For the heavy Saudi crude oil, estimate the average sulfur content and specific gravity of all the products using the TBPs developed for each cut

Solution:

Naphtha product

P.C Range	Naphtha Vol%(A)	S.G(B)	Wt factor (C=A*B)	Sulfur Wt%(D)	Sul factor (E=C*D)
1	0	0.641432	0	0	0
2	5.5	0.663074	3.646907	0	0
3	10.5	0.676386	7.102055	0	0
4	9.5	0.68523	6.509685	0	0
5	20.5	0.705736	14.46758	0	0
6	24.5	0.721939	17.6875	0	0

7	18.3	0.742782	13.59291	0	0
8	8.2	0.766938	6.288889	0.05	0.314444
9	3	0.783934	2.351801	0.2	0.47036
Total	**100**		**71.64733**		**0.784805**

Specific gravity of Naphtha product = 71.64733/100 = 0.716473

This corresponds to the °API of 65.995

Sulfur content of the Naphtha product = 0.784805/71.64733 = 0.010954

Kerosene product

P.C Range	**Kerosene Vol%(A)**	**S.G(B)**	**Wt factor (C=A*B)**	**Sulfur Wt%(D)**	**Sul factor (E=C*D)**
8	0.8	0.766938	0.61355	0.05	0.030678
9	15.7	0.783934	12.30776	0.2	2.461551
10	75.5	0.8017	60.52833	0.35	21.18492
11	8	0.822674	6.581395	0.7	4.606977
Total	**100**		**80.03103**		**28.28412**

Specific gravity of Kerosene product = 80.03103/100 = 0.803103

This corresponds to the °API of 45.30642

Sulfur content of the Kerosene product = 28.28412/80.03103 = 0.353414

LGO product

P.C Range	**LGO Vol%(A)**	**S.G(B)**	**Wt factor (C=A*B)**	**Sulfur Wt%(D)**	**Sul factor (E=C*D)**
10	1	0.8017	0.8017	0.35	0.280595
11	27.5	0.822674	22.62355	0.7	15.83648
12	56.3	0.847305	47.70329	1.35	64.39945
13	15.2	0.860182	13.07477	1.7	22.22711
Total	100		84.20331		102.7436

Specific gravity of LGO product = 0.8420331

This corresponds to the ^{o}API of 36.54565

Sulfur content of the LGO product = 102.7436/84.20331 = 1.220185

HGO product

P.C Range	HGO Vol%(A)	S.G(B)	Wt factor (C=A*B)	Sulfur Wt%(D)	Sul factor (E=C*D)
13	19.5	0.860182	16.77356	1.7	28.51505
14	63.3	0.876161	55.46099	2	110.922
15	17.2	0.901274	15.50191	2.6	40.30497
Total			87.73646		179.742

Specific gravity of HGO product = 0.8773646

This corresponds to the ^{o}API of 29.77845

Sulfur content of the LGO product = 179.742/87.73646 = 2.048658

Evaluated properties summary

Properties	Whole crude	Naphtha	Kerosene	LGO	HGO
Cut volume (%)	100	22.8	8.8	10.6	5.8
S.G	0.8758	0.716473	0.803103	0.8420331	0.8773646
A.P.I	30.05	65.995	45.30642	36.54565	29.77845
S%	2.36393	0.010954	0.353414	1.220185	2.048658

Average specific gravity of the residue = $\frac{100\times0.8758-22.8\times0.7164-8.8\times0.8031-10.6\times0.842-5.8\text{x}0.877}{100-22.8-8.8-10.6-5.8} = 0.96467$

This corresponds to ^{o}API of 15.18

Average sulfur content of the residue

= $\frac{100\times0.8758\times2.6393-22.8\times0.716\times0.01095-8.8\times0.8031\times0.3534-10.6\times0.842\times1.2202-5.8\text{x}0.877\times2.0486}{(100-22.8-8.8-10.6-5.8)\times0.96467}$ = 4.281

All properties summary

Properties	Whole crude	Naphtha	Kerosene	LGO	HGO	Residue
Cut volume (%)	100	22.8	8.8	10.6	5.8	50.16
S.G	0.8758	0.716473	0.803103	0.8420331	0.8773646	0.96467
A.P.I	30.05	65.995	45.30642	36.54565	29.77845	15.18
S%	2.36393	0.010954	0.353414	1.220185	2.048658	4.281

2.10 Estimation of other properties of the products

Using the correlations presented previously in other sections, the other properties of various products namely average volume, mean, weight, molal boiling points, characterization factor, molecular weight, enthalpy, viscosity and vapor pressure can be estimated.

Next, we present an illustrative example to estimate other properties of products.

Q 2.12: For Naphtha, Kerosene, LGO and HGO products of the crude oil, estimate average volume, mean, weight, molal boiling points, characterization factor, molecular weight, saturated liquid enthalpy, viscosity and vapor pressure

Solution:

Naphtha product

I) Calculations for boiling point

a. **Volume average boiling point**

$T_v = t_0+4t_{50}+t_{100}/6 = (80+4*227+390)/6= 229.66^0F$

b. **Mean average boiling point**

Slope = $t_{70}-t_{10} / 60 = 258-142/60= 1.933$

From Maxwell's correlation

$\Delta T(229.66, 1.933) = -5^0F$

Mean average boiling point = $T_v + \Delta T = 229.66 - 5 = 224.66\ ^0F$

c. Weight average boiling point

Slope= t_{70}-t_{10} / 60 = 258-142/60= 1.933

From Maxwell's correlation

Δ T (229.66, 1.933) = 2^0F

Weight average boiling point = $T_v + \Delta T$ =229.66+ 2 = 231.66 ^{0}F

d. Molal average boiling point

Slope = t_{70}-t_{10} / 60 = 258-142/60= 1.933

From Maxwell's correlation

Δ T (229.66, 1.933) = -6^0F

Molal average boiling point = $T_v + \Delta T$ =229.66 – 6= 223.66 ^{0}F

Average temperatures summary

volume	mean	wt	molal
229.66^{0}F	224.66 ^{0}F	231.66 ^{0}F	223.66^{0}F

II) Characterization factor

Mean average boiling point = 224.66 ^{0}F

^{0}API = 65.99514

From Maxwell's correlation = 12.3

III) Molecular weight

^{0}API = 65.99514

Mean average b.pt=224.66 ^{0}F

From Maxwell's correlation, mol.wt=107

IV) Enthalpy

Volume average boiling point=229.66 ^{0}F

Mean average b.pt=224.66 ^{0}F

Characterization factor,K= 12.3

Mean avg B.pt	200		250	
K	11	12	11	12
Enthalpy(Btu/lb)	118	128	110	122

H^L(224.66 ^{0}F,12.3, 229.66 ^{0}F)

From Maxwell's correlations and using interpolation, enthalpy= 127.796 Btu/lb.

V) **Viscosity**

At 100^{o}F

T=mean avg B.pt=224.66 ^{0}F =684.33^{0}R

S. G. of cut= =0.716473

^{0}API = A.P.I=(141.5/S.G)-131.5=65.99514

A_1= 2.701423

A_2= -0.29506

Watson K factor = T^(1/3)/SG=12.29

LOG vcor = A1+A2K= -0.92483

$\nu_{cor=}$0.118897

LOG ν_{ref} = -1.35579+8.16059*10^(-4)T+8.38505* 10^(-7) T^2= -0.40466

$\nu_{ref=}$0.39383

ν_{100} = ν_{ref}+ ν_{cor}=0.512757 centi stokes

Viscosity At 210

B1= -1.92353

B2=2.41071x10^{-4}

B3= 0.5113

LOGv210= B1+B2*T+B3 *LOG(T*(v100)) = -0.45721

V210=0.348972 centi stokes

VI) Vapor pressure

Assume Temperature= 500^{0}F

Normal boiling point= 229.66 ^{0}F

Vapor pressure from Maxwell;s correlation = 22 Atm

Kerosene product

I) Calculations for boiling point

a. Volume average boiling point

$T_v = t_0+4t_{50}+t_{100}/6$ = (357+4*428+484)/6= 425.5^{0}F

b. Mean average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 440-390/60= 0.8333

From Maxwell’s correlation

Δ T (425.5, 0.8333) = 0^{0}F

Mean average boiling point = T_v + Δ T =425.5+0 = 425.5^{0}F

c. Weight average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 440-390/60= 0.8333

From Maxwell’s correlation

Δ T (425.5, 0.8333) = 0^{0}F

Weight average boiling point = T_v + Δ T ==425.5+0 = 425.5^{0}F

d. Molal average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 440-390/60= 0.8333

From pg. No 14 graph in Maxwell’s book

Δ T (425.5, 0.8333) = 0^{0}F

Molal average boiling point = T_v + Δ T =425.5+0 = 425.5^{0}F

Average temperatures summary

volume	mean	wt	molal
425.5^{0}F	425.5^{0}F	425.5^{0}F	425.5^{0}F

II) Characterization factor

Mean average boiling point = 425.5^{0}F

A.P.I=(141.5/S.G)-131.5=45.30642

From Maxwell’s correlation Characterization factor is = 12.00

III) Molecular weight

^{0}API = 45.30642

Mean average b.pt=425.5^{0}F

From Maxwell's correlation, mol.wt = 172

IV) Enthalpy

Volume average boiling point=425.5^{0}F

Mean average b.pt=425.5^{0}F

Characterization factor,K= 12.00

Mean avg B.pt	400		500	
K	11	12	11	12
Enthalpy(Btu/lb)	215	235	210	225

H^L(425.5^{0}F,12.00, 425.5^{0}F)

From Maxwell's correlation and interpolation, Enthalpy= 232.45 Btu/lb

V) Viscosity

At 100^{o}F

T = mean avg B.pt=425.5^{0}F =885.17^{0}R

S. G. of cut= 0.80031

A.P.I=(141.5/S.G)-131.5=45.30642

A_1= 2.386988

A_2= -0.22025

Watson K factor = 11.9973

LOG vcor = A1+A2K= -0.25544

$\nu_{cor=}$0.555346

$\nu_{ref=}$1.055726

$\nu_{100} = \nu_{ref} + \nu_{cor}$=1.611072 Cst

Viscosity At 210

B1= -1.92353

B2=2.41071*10^-4

B3= 0.5113

LOGv210= -0. 09743

V210= 0.799044 Cst

VI) Vapor pressure

Assume Temperature= 500^{0}F

Normal boiling point= 425.5^{0}F

From Maxwell's correlation, Vapor pressure = 2.38 Atm

LGO product

I) Calculations for boiling point

a. Volume average boiling point

$T_v = t_0+4t_{50}+t_{100}/6$ = (455+4*544+618)/6= 541.5^{0}F

b. Mean average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 564-493/60= 1.1833

From Maxwell's correlation, Δ T (541.5, 1.1833) = 0^{0}F

Mean average boiling point = $T_v + \Delta T$ =541.5 +0 = 541.5 ^{0}F

c. Weight average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 564-493/60= 1.1833

From Maxwell's correlation, Δ T (541.5, 1.1833) = 0^{0}F

Weight average boiling point = $T_v + \Delta T$ =541.5 +0 = 541.5 ^{0}F

d.Molal average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 564-493/60= 1.1833

Δ T (541.5, 1.1833) = 0^{0}F

Molal average boiling point = $T_v + \Delta T$ =541.5 +0 = 541.5 ^{0}F

Average temperatures summary

volume	mean	wt	molal
541.5^{0}F	541.5^{0}F	541.5^{0}F	541.5^{0}F

II) Characterization factor

Mean average boiling point = 541.5^{0}F

^{0}API = 36.54565

From Maxwell's correlation, Characterization factor is = 11.85

III) Molecular weight

^{0}API = 36.54565

Mean average b.pt=541.5^{0}F

From Maxwell's correlation, mol.wt=214

IV) Enthalpy

Volume average boiling point=541.5^{0}F

Mean average b.pt=541.5^{0}F

Characterization factor, K= 11.85

Mean avg B.pt	500		600	
K	11	12	11	12
Enthalpy(Btu/lb)	284	305	277	295

H^L(541.5^{0}F,11.85, 541.5^{0}F)

From Maxwell's correlation and interpolation, Enthalpy= 297.886 Btu/lb.

V) Viscosity

At 100^{o}F

T = mean avg B.pt=541.5^{0}F =1001.17^{0}R

S. G. of cut =0.842033

A.P.I=(141.5/S.G)-131.5=36.54565

A_1 = 3.72279

A_2 = -0.29367

Watson K factor = 11.8806

LOG vcor = A1+A2K=0.23385

$\nu_{cor=}$1.713365

LOG ν_{ref} = -1.35579+8.16059*10^(-4)T+8.38505* 10^(-7) T^2= 0.301692

$\nu_{ref=}$2.003051

ν_{100} = ν_{ref}+ ν_{cor}=3.716416 Cst

Viscosity At 210

B1= -1.92353

B2=2.41071*10^-4

B3= 0.5113

LOGv210= B1+B2*T+B3 *LOG(T*(v100)) = 0.143486

V210=1.391509Cst

VI) Vapor pressure

Assume Temperature= 500^{0}F

Normal boiling point= 541.5^{0}F

From Maxwell's correlation, vapor pressure = 0.59~= 0.6 Atm

HGO product

I) Calculations for boiling point

a. Volume average boiling point

$T_v = t_0 + 4t_{50} + t_{100}/6 = (582+4*644+684)/6 = 640.33^0F$

b. Mean average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 653-609/60= 0.7333

From Maxwell's correlation, ΔT (640.33, 0.7333) = -1^0F

Mean average boiling point = $T_v + \Delta T$ =640.33– 1 = 639.33 ^0F

c. Weight average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 653-609/60= 0.7333

From Maxwell's correlation, ΔT (640.33, 0.7333) = 0^0F

Weight average boiling point = $T_v + \Delta T$ =640.33+0= 640.33 ^0F

d. Molal average boiling point

Slope= $t_{70}-t_{10}$ / 60 = 653-609/60= 0.7333

From Maxwell's correlation, ΔT (640.33, 0.7333) = -3^0F

Molal average boiling point = $T_v + \Delta T$ =640.33-3= 637.33 °F

Average temperatures summary

volume	mean	wt	molal
640.33^0F	639.33^0F	640.33^0F	637.33^0F

II) Characterization factor

Mean average boiling point = 639.33 ^0F

^{0}API = 29.77845

From Maxwell's correlation, Characterization factor is = 11.75

III) Molecular weight

^{0}API = 29.77845

Mean average b.pt= 639.33 ^{0}F

From Maxwell's correlation, mol.wt=265

IV) Enthalpy

Volume average boiling point=640.33 ^{0}F

Mean average b.pt= 639.33 ^{0}F

Characterization factor, K= 11.75

Mean avg B.pt	600		800	
K	11	12	11	12
Enthalpy(Btu/lb)	345	368	333	350

H^L= (639.33 ^{0}F ,11.75, 640.33 ^{0}F)

From Maxwell's correlation and interpolation, enthalpy= 359 Btu/lb.

V) Viscosity

At 100^{o}F, T = mean avg B.pt=639.33 ^{0}F =1099^{0}R

S. G. of cut=0.877365

API=(141.5/S.G)-131.5=29.77845

A_1 = 5.625067

A_2 = -0.41546

Watson K factor = 11.762

LOG vcor = A1+A2K=0.738431

$\nu_{cor=}$ 5.475588

LOG ν_{ref} = -1.35579+8.16059*10^(-4)T+8.38505* 10^(-7) T^2= 0.553806

$\nu_{ref=}$3.579365

$\nu_{100} = \nu_{ref} + \nu_{cor}$=9.054953 centi stokes

Viscosity At 210

B1= -1.92353

B2=2.41071*10^-4

B3= 0.5113

LOGv210= B1+B2*T+B3 *LOG(T*(v100)) = 0.385523

V210=2. 429535 centi stokes

VI) Vapor pressure

Assume Temperature= 500^{0}F

Normal boiling point= 640.33 ^{0}F

Vapor pressure from Maxwell's correlation = 0.16 atm.

Properties summary

Properties		Whole crude	Naphtha	Kerosene	LGO	HGO
Vol avg B.Pt (^{0}F)		715	229.6	425.5	541.5	640.3
Mean avg B.Pt (^{0}F)		642	224.6	425.5	541.5	639.3
Wt avg B.Pt (^{0}F)		752.2	231.6	425.5	541.5	640.3
Molal avg B.Pt (^{0}F)		535.6	223.6	425.5	541.5	637.3
Characterization factor(K)		11.75	12.3	12	11.85	11.75
Molecular wt		266	107	172	214	265
Saturated Liquid Enthalpy (Btu/lb)		416.8	127.8	232.45	298	359
Viscosity (Centi-stokes)	100^{0}F	9.269	0.513	1.611	3.716	9.055
	210^{0}F	2.465	0.349	0.799	1.392	2.429
Vapour pressure (atm) at 500 ^{o}F		0.05	22	2.38	0.6	0.16

Blending Index	Viscosity (Cst)	Blending Index	Viscosity (Cst)
139.56	1.05	39.16	10.00
128.20	1.25	37.87	11.85
120.63	1.40	36.57	13.95
109.27	1.64	35.18	17.46
99.81	1.87	33.51	21.35
94.89	2.01	31.75	29.10
88.83	2.23	29.99	38.15
84.67	2.41	28.23	52.41
79.75	2.64	27.58	61.67
76.34	2.81	27.03	70.35
73.31	2.99	26.29	81.50
60.44	4.01	25.27	101.23
50.22	5.51	22.49	204.86
43.41	7.10	20.46	401.96
40.76	7.89	19.54	495.45
38.86	8.39	19.26	592.13
37.73	8.83	18.80	691.37
36.59	9.55	18.61	807.30
35.08	9.89	18.15	892.81
		15.57	2683.03
		14.36	4014.16
		13.81	5025.56
		13.44	5777.70
		13.16	6851.76
		12.89	7636.84
		12.43	8847.94
		12.33	9635.42

Table 2.21: Blending Index and Viscosity correlation data.

2.11 Estimation of blend viscosity

The viscosity of a blend is correlated using the concept of viscosity index. For a chosen viscosity, a specific viscosity index exists which would contribute on a volumetric basis to the blend's viscosity index. Therefore, knowing the viscosity of the various intermediate streams and their volumetric basis of mixing, one can evaluate the viscosity of the blend i.e., the product stream that is prepared by blending the said products in the specified volumetric basis. Maxwell (1950) provided the viscosity blending index as shown in Table 2.21. Two illustrative examples are presented next to demonstrate the utility of blending index correlations.

Q 2.13: Assuming the Saudi crude viscosity to be 9.269 Cst at 100 ^{o}F and that for products namely Naphtha, kerosene, LGO and HGO to be 0.513, 1.611, 3.716 and 9.055 Cst at 100 ^{o}F respectively, determine the viscosity of the residue product using the blending index correlation presented by Maxwell (1950). The volume % corresponding to various product fractions can be assumed as 22.8 for Naphtha, 8.8 % for Kerosene, 10.6 % for LGO and 5.8 % for HGO respectively.

Solution:

For Crude oil, blending index corresponding to 9.269 Cst = 40

For Naphtha, blending index corresponding to 0.513 Cst = 89.5

For Kerosene, blending index corresponding to 1.611 Cst = 61.8

For LGO, blending index corresponding to 3.716 Cst = 49.2

For HGO, blending index corresponding to 9.055 Cst = 40.2

Based on volumetric balance of the blending index, blending index for the residue =

$$= \frac{100\times40-22.8\times89.5-8.8\times61.8-10.6\times49.2-5.8\text{x}40.2}{(100-22.8-8.8-10.6-5.8)} = 12.709$$

From blending index curve, viscosity of the residue = 6900 Cst.

Q 2.14: In a particular refinery operation, the automotive diesel product is produced by blending both gas oil and diesel products generated as intermediate streams in the refinery. The refinery fed gas oil and diesel have been found to have a viscosity of 12 and 3.5 Cst at 100 ^{o}F. The automotive diesel product viscosity has been specified to have the viscosity in the range of 4.5 – 6 Cst at 100 ^{o}F. To obtain this viscosity range, determine the maximum and minimum volume ratio of gasoline to diesel (intermediate) to be blended to obtain the desired product.

Solution:

Let vol % gas oil to blend with diesel by x. Then, vol % of diesel (intermediate) is 100 –x.

For these two streams, vol %, viscosity and blending index are:

Stream	Viscosity	Blending Index	Vol %
Gas oil	12	37.9	x
Diesel	3.5	50	100 – x

a) Minimum volume of gas oil

Viscosity of automotive diesel product = 4.5

Blending index of the product = 47

Therefore, 100 x 47 = (x)(37.9)+ (100 – x) (50)

Solving for x, we get x = 24.8 vol %

b) Maximum volume of gas oil

Viscosity of automotive diesel product = 6

Blending index of the product = 44

Therefore, 100 x 44 = (x)(37.9)+ (100 – x) (50)

Solving for x, we get x = 49.6 vol %

Flash point (^{o}F)	Flash point index
697.74	0.013
696.08	0.020
695.95	0.024
690.10	0.039
688.49	0.057
687.63	0.076
678.55	0.101
674.94	0.143
682.92	0.201
679.94	0.301
675.68	0.407
667.31	0.653
665.72	1.012
657.36	1.985
651.68	4.166
657.38	6.138
646.84	21.718
640.06	38.515
636.58	62.783
637.73	79.484
637.58	104.076
630.40	143.354
630.82	211.203
626.23	290.902
620.40	414.412
619.52	675.510
618.67	1029.350
609.80	1595.330

Table 2.22: Flash point index and flash point correlation data

2.12 Flash point estimation and flash point index for blends

The flash point of a refinery intermediate/product stream is evaluated using the expression:

$$\text{Flash point} = 0.77[\text{ASTM } (5^{\text{o}}\text{F}) - 150]$$

Therefore, knowing the ASTM profile of the products could allow one to determine the flash point of the product. In similarity to the viscosity blending index, flash point index exists as presented by Jones and Pujado (2006). The flash point index is fortunately a straight line which can be used to evaluate the flash point of a blend from the known volumetric distributions and flash point indices of various intermediate streams that contribute towards the blended stream. Table 2.22 summarizes flash point and flash point index correlation data as presented by Jones and Pujado (2006). Next, we present an illustrative example to evaluate the flash point of the residue stream for evaluated crude and product flash points using the flash point equation.

Q 2.15: For the heavy Saudi crude oil and its products, evaluate the flash point by evaluating their ASTM $^{\text{o}}$5 temperatures. Subsequently, evaluate the flash point of the residue product for known flash point values of the crude and other products by using the flash point index correlation provided by Jones and Pujado (2006).

Solution:

To evaluate the Saudi crude oil ASTM $^{\text{o}}$5F, we choose any two points on the crude TBP and evaluate their ASTM values using Edmister correlation. Subsequently, using probability chart, ASTM $^{\text{o}}$5F is evaluated.

For the heavy Saudi crude oil stream, 50 % TBP from TBP assay = 723 $^{\text{o}}$F

For the heavy Saudi crude oil stream, 70 % TBP from TBP assay = 945 $^{\text{o}}$F

From probability chart, ASTM $^{\text{o}}$5F for Saudi crude oil = 304 $^{\text{o}}$F

Flash point of the crude = 0.77 (304 – 150) = 118.58 $^{\text{o}}$F

For the product streams other than the residue, we make use of the previously generated ASTM data in the solved problems. Using them and using probability chart, ASTM $^{\text{o}}$5F of these products is evaluated.

ASTM $^{\text{o}}$5F for Naphtha = 172 $^{\text{o}}$F

ASTM $^{\text{o}}$5F for Kerosene = 404 $^{\text{o}}$F

ASTM $^{\text{o}}$5F for LGO = 505 $^{\text{o}}$F

ASTM $^{\text{o}}$5F for HGO = 609 $^{\text{o}}$F

From these ASTM values, the flash point of products are estimated as Naphtha = 16.94 ^{o}F, Kerosene = 195.58 ^{o}F, LGO = 273.35 ^{o}F, HGO = 353.43 ^{o}F.

For both products and crude, the flash point and flash point index are summarized as follows:

Stream	Volume %	Flash	Flash point index
Naphtha	22.8	16.94	1364
Kerosene	8.8	195.58	23.03
LGO	10.6	273.35	3.47
HGO	5.8	353.43	0.66
Residue	52	Not known	Say x
Crude	100	118.58	313.43

ASTM 50 % = 300 ^{o}F		**ASTM 50 % = 400 ^{o}F**		**ASTM 50 % = 500 ^{o}F**		**ASTM 50 % = 600 ^{o}F**	
Pour point (^{o}F)	Pour point index	Pour point (^{o}F)	Pour point index	Pour point (^{o}F)	Pour point index	Pour point (^{o}F)	Pour point index
-49.92	8.97	-49.62	3.51	-49.80	2.03	-49.98	1.26
-40.06	11.20	-45.84	3.95	-48.32	2.13	-40.29	1.77
-30.20	13.67	-40.25	4.74	-43.56	2.52	-30.09	2.50
-20.18	17.07	-30.23	6.17	-40.43	2.79	-27.63	2.80
-9.99	21.43	-19.87	8.36	-37.47	3.07	-25.00	3.04
-0.29	26.52	-13.13	10.11	-34.68	3.39	-22.20	3.39
9.90	32.98	-9.84	11.28	-32.05	3.66	-19.90	3.69
19.76	40.62	-9.84	11.28	-30.24	3.93	-18.42	3.93
30.27	51.23	-3.43	13.63	-20.21	5.43	-10.03	5.19
40.30	64.62	-0.15	14.99	-15.45	6.44	-4.61	6.37
49.83	77.72	-0.15	14.99	-10.19	7.60	0.16	7.66
-49.93	5.57	3.47	16.63	-7.39	8.39	3.45	8.43
-39.91	7.23	5.77	17.52	-1.80	10.24	8.05	10.19
-30.05	9.25	9.39	19.27	0.01	10.79	9.86	10.89
-25.78	10.27	13.99	22.53	2.97	11.92	12.65	11.97
-20.02	11.95	19.74	26.60	7.07	13.49	16.43	13.74
-14.44	13.78	25.17	31.55	9.87	14.90	19.72	15.25
-10.00	15.44	30.10	35.69	12.99	16.54	21.70	16.61
-7.04	16.73	36.01	42.73	15.13	17.76	24.00	17.83
-4.08	17.88	42.75	52.15	18.09	19.71	26.63	19.70
-0.30	19.84	50.31	63.64	19.90	20.87	30.08	22.50
10.38	26.00			30.09	29.64	40.11	31.96
20.08	33.58			30.09	29.64	40.11	31.96
29.94	42.96			39.62	40.35	49.64	44.33
40.13	55.76			49.82	56.78		
49.99	71.34						

Table 2.23: Pour point and pour point index correlation data.

Upon volumetric balance, flash point of the residue

$$x = \frac{100\times313.43-22.8\times1364-8.8\times23.03-10.6\times3.47-5.8\text{x}0.66}{(100-22.8-8.8-10.6-5.8)} = 0.01$$

From flash point index correlation, residue flash point = 700 ^{o}F.

2.13 Pour point estimation and pour point index for blends

A correlation is presented in API technical data book (1997) to relate the pour point of a petroleum fraction to its MEABP and specific gravity. The correlation is presented as

$$\text{Pour point} = 3.5 \times 10^{-8} \times \text{MEABP}^{5.49} \times 10^{\left(-0.712\text{MEABP}^{0.315}+0.133\times\text{SG}\right)} + 1.4$$

where MEABP and SG correspond to the mean average boiling point (in ^{o}R) and specific gravity (at 60 ^{o}F) for the fraction.

In similarity to the viscosity and flash point, the pour point of a blend can be conveniently expressed using a volumetric balance over the pour point index. Jones and Pujado (2006) presented a correlation between pour point of a fraction and its ASTM (50 ^{o}F) with the pour point index (data presented in Table 2.23) using which the pour point of a blend can be evaluated. Next an illustrative example is presented to evaluate the pour point of crude and crude fractions using the above equation and evaluate the residue pour point using the pour point index correlated by Jones and Pujado (2006).

Q 2.16: For the heavy Saudi crude oil and its products (other than residue), evaluate the pour point using correlation presented in the API technical data book. Subsequently, evaluate the pour point of the residue product for known pour point values of the crude and other products by using the pour point index correlation provided by Jones and Pujado (2006).

Solution:

Using the pour point formula, the pour point of the crude and products other than residue is first evaluated:

Stream	MEABP (^{o}F)	MEABP (^{o}R)	SG	PP (^{o}R)	PP (^{o}F)
Crude	642	1102	0.8758	505.59	45.59
Naphtha	224.6	684.6	0.7165	310.03	-149.97
Kerosene	425.5	885.5	0.8031	419.6	-40.39
LGO	541.5	1001.5	0.842	470.23	10.23
HGO	639.3	1099.3	0.8774	504.36	44.36

The sequential procedure for the calculation of pour point index and associated factors is presented below in a column wise approach:

Stream	**Differential volume (%)** **[A]**	**ASTM 50 %** **[B]**	**ASTM factor** **[A] X [B]**	**Pour point (^oF)**	**Pour point index (from correlation)** **[C]**	**Pour point factor** **[A] X [C]**
Naphtha	22.8	233	5312.4	-149.97	-0.85	-19.38
Kerosene	8.8	427	3757.6	-40.39	1.95	17.16
LGO	10.6	537	5692.2	10.23	9.05	95.93
HGO	5.8	631	3659.8	44.36	32.19	186.7
Residue	52					
Crude	100	723	72300	45.59	25.7	2570

Residue ASTM factor = 72300 – (5312.4 + 3757.6 + 5692.2 + 3659.8) = 53878

ASTM 50 % of the residue = 53878/52 = 1036 ^oF

Pour factor off residue = 2570 – (-19.38 + 17.16 + 95.93 +186.7) = 2288

Pour point index of the residue = 2288/52 = 44

From pour point index correlation (graph),

For pour point index = 44 and ASTM 50 % temperature of 600 ^oF, pour point = 49 ^oF

For pour point index = 44 and ASTM 50 % temperature of 700 ^oF, pour point = 54 ^oF

Upon extrapolation, for residue whose pour point index = 44 and ASTM 50 % temperature is 1036 ^oF, pour point of the residue = 54 + (1036 – 700) x 5/(700 – 600) = 70.8 ^oF

It is interesting to note that from literature, residue pour point = 55.9 ^oF.

Maxwell's correlation 1		**Maxwell's correlation 2** **T_{50} (DRL) < 300 °F**		**Maxwell's correlation 2** **T_{50} (DRL) > 300 °F**		**Maxwell's correlation 3**	
S_{TBP}	S_{FRL}	S_{TBP}	**ΔT_{50} (DRL – FRL)**	S_{TBP}	**ΔT_{50} (DRL – FRL)**	**Percent off**	$\frac{\Delta T50\ (Flash - FRL)}{\Delta T50\ (DRL - FRL)}$
0.02	7.74	0.02	7.74	2.00	0.65	1.08	0.33
0.48	7.71	0.48	7.71	2.46	3.33	3.49	0.37
0.88	7.69	0.88	7.69	2.84	5.62	7.49	0.39
1.34	7.27	1.34	7.27	3.09	8.32	13.47	0.38
1.76	8.02	1.76	8.02	3.39	11.39	18.66	0.35
2.16	9.15	2.16	9.15	3.65	14.47	24.64	0.34
2.58	11.06	2.58	11.06	3.91	19.10	30.63	0.34
3.00	12.19	3.00	12.19	4.17	23.34	35.23	0.34
3.31	14.11	3.31	14.11	4.61	28.73	40.02	0.34
3.67	17.96	3.67	17.96	4.97	32.97	45.21	0.34
3.97	22.20	3.97	22.20	5.25	34.50	50.80	0.34
4.37	27.20	4.37	27.20	5.53	36.03	54.19	0.34
4.85	31.82	4.85	31.82	5.79	36.40	60.19	0.34
5.23	34.50	5.23	34.50	6.13	37.93	66.57	0.34
5.69	36.79	5.69	36.79	6.39	38.30	71.17	0.34
6.19	37.92	6.19	37.92	6.61	39.06	75.36	0.34
6.91	39.04	6.91	39.04	6.95	39.42	79.35	0.35
7.43	39.39	7.43	39.39	7.21	39.41	83.75	0.34
7.87	38.98	7.87	38.98	7.55	39.77	87.54	0.34
8.39	39.33	8.39	39.33	8.05	39.74	91.13	0.34
8.95	39.68	8.95	39.68	8.63	39.31	94.93	0.34
9.54	39.64	9.54	39.64	9.26	40.05	98.92	0.34
10.14	39.60	10.14	39.60	9.78	39.63		
10.74	39.95	10.74	39.95	10.40	39.97		
11.24	39.53	11.24	39.53	10.88	39.94		
11.68	39.89	11.68	39.89	11.44	39.91		
11.98	39.87	11.98	39.87	11.78	39.50		
9.60	6.45			11.78	39.50		
9.88	6.71						
10.18	7.01						

Table 2.24: EFV-TBP correlation data presented by Maxwell (1950).

2.14 Equilibrium flash vaporization curve

Equilibrium flash vaporization (EFV) curve is an important graph in the design calculations of refinery distillation columns. The data is especially required to predict the over-flash temperatures of the distillation which eventually affect the design characteristics of the distillation columns. Usually, the EFV curve is determined using Maxwell's correlations (1950) and a given TBP assay of the stream.

We now briefly present the procedure adopted for EFV calculation following which an illustrative example is presented

a) For the given TBP assay, determine the volume % and TBP values for 0%, 10 %, 20 %....100%. The data set is designated as column [A].

b) Firstly, evaluate the slope of the distillation reference line (DRL) using the formula

$$\mathrm{S_{DRL}} = \frac{\mathrm{TBP_{70}} - \mathrm{TBP_{10}}}{70 - 10}$$

c) For various volume % values (i.e., x_{DRL} other than the 10 % and 70 %), evaluate the corresponding distillation reference line points using either of the following expressions:

$$\mathrm{y_{DRL}} = \mathrm{TBP_{10}} + \mathrm{S_{DRL}}(\mathrm{x_{DRL}} - 10)$$

$$\mathrm{y_{DRL}} = \mathrm{TBP_{70}} + \mathrm{S_{DRL}}(\mathrm{x_{DRL}} - 70)$$

y_{DRL} values are designated as column [B]

d) Use Maxwell's correlation data presented in Table 2.24 to evaluate column [C] as ordinate for various values of abscissa chosen as values taken in column [A]. Column [C] corresponds to the factor defined as $\frac{\Delta\mathrm{T(Flash-FRL)}}{\Delta\mathrm{T(TBP-FRL)}}$. Here FRL refers to Flash reference line.

e) Evaluate column [D] using the expression [D] = [A] – [B]. Column [D] corresponds to $\Delta\mathrm{T(TBP - FRL)}$

f) Evaluate column [E] using the expression [E] = [C] x [D]. Column [E] corresponds to $\Delta\mathrm{T(Flash - FRL)}$

g) Using Maxwell's correlation data presented in Table 2.24 that relates the slope of FRL (S_{FRL}) to S_{DRL}, evaluate the slope of the flash reference line S_{FRL}.

h) Using Maxwell's correlation presented in Table 2.24 that relates the $\Delta\mathrm{T_{50}(DRL - FRL)}$, determine $\Delta\mathrm{T_{50}(DRL - FRL)}$ for the known S_{DRL}.

i) Using the value of $\Delta\mathrm{T_{50}(DRL - FRL)}$, determine FRL at the volume % corresponding to the value of 50 % using the expression

$$\mathrm{FRL_{50}} = \mathrm{DRL_{50}} + \Delta\mathrm{T_{50}(DRL - FRL)}$$

j) Using S_{FRL} and FRL_{50}, for various volume % values (i.e., x_{FRL} other than the 50 % case), evaluate the corresponding flash reference line (FRL) points using the expression:

$$\mathrm{y_{FRL}} = \mathrm{FRL_{50}} + \mathrm{S_{FRL}}(\mathrm{x_{FRL}} - 50)$$

The data points thus evaluated as designated as column [F].

k) The EFV data designated as column [G] is evaluated using the expression [G] = [E] + [F].

Q 2.17: For the heavy Saudi crude oil, whose TBP assay has been presented previously, determine the EFV curve data using Maxwell's correlation data presented in Table 2.24.

Solution:

Sample calculations:

Slope of the DRL = (965 – 205)/60 = 12.67

Equation to represent DRL is given as y = 12.67x + 78.3 and using this equation, data in column [B] were evaluated.

Slope of the FRL for a corresponding slope of 12.67 for DRL is 9.2

$\Delta T_{50}(\mathrm{DRL} - \mathrm{FRL})$ for S_{DRL} = 12.67 is 40 °F

FRL_{50} = 711.7 – 40 = 671.7 °F

The equation to represent FRL is $y_{\mathrm{FRL}} = 9.2x_{\mathrm{FRL}} + 211.7$.

The evaluation of various columns [A] to [G] to yield the EFV curve data is presented in the following table:

Volume %	TBP data [A]	DRL [B]	$\frac{\Delta T(\mathbf{Flash} - \mathbf{FRL})}{\Delta T(\mathbf{TBP} - \mathbf{FRL})}$	[D] [A] – [B]	[E] [C] X [D]	FRL [F]	EF V
0	-12	78.3	0.24	-90.3	-21.68	211.7	190
10	205	205	0.4	0	0	303.7	303.7
20	338	331.7	0.36	6.3	2.28	395.7	397.9
30	459	458.3	0.34	0.7	0.23	487.7	487.9
40	578	585	0.33	-7	-2.31	579.7	577.4
50	703	711.7	0.33	-8.7	-2.86	671.7	568.8
60	828	838.3	0.33	-10.3	-3.41	763.7	760.3
70	965	965	0.33	0	0	855.7	855.7
80	1104	1091.7	0.33	12.3	4.07	947.7	951.7
90	1246	1218.3	0.33	27.7	9.13	1039.7	1048.8
100	1400	1345	0.33	55	18.15	1131.7	1149.8

2.15 Summary

In this chapter, a brief account of various relevant correlations to yield useful refinery properties such as average temperatures, molecular weight, characterization factor, enthalpy, viscosity, flash and pour points have been presented. Along with these, suitable procedures have been also demonstrated through theory as well example calculations for the estimation of refinery process product properties. The significance of pour point, flash point and viscosity blending indices in the estimation of blend properties has also been presented. Though approximate these correlations are useful to estimate the properties for the process engineer to orient towards the refinery properties. In addition, for an illustrative example, all calculations have been demonstrated for a single crude stream, as sample calculations enable one to develop confidence in the subject matter. One of the major objectives of this section is to highlight the issue that with minimum crude assay data, a large number of useful refinery stream parameters apparent for design calculations can be evaluated using the correlations presented by Maxwell (1950), Jones and Pujado (2006) and in relevant books such as API technical data book (1997). In the next chapter, we attempt to orient the process engineer towards the mass balances across an appropriate refinery complex and associated elementary calculation principles.

3. Refinery Mass Balances

3.1 Introduction

A petroleum refinery mass balance is required for number of purposes. These are:

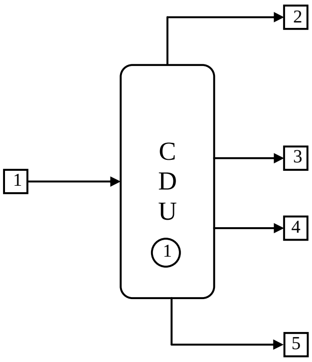
a) To understand the primary processing operations in various sub-processes and units.
b) To estimate the flow-rates of various intermediate streams using which final product flow rates can be estimated.
c) To serve as the initial data for the elementary process design of sub-processes and units.

A petroleum refinery can be conveniently represented using a block diagram that consists of various units and splitters. While units involve physical/chemical processing, the splitters only enable splitting of a stream into various streams of similar composition. Therefore, it can be easily visualized that physical/chemical processes produce streams whose composition is different from that of the feed and splitters produce streams whose composition is equal to the composition of the feed stream. **Whatever may be the case, as far as mass balances are concerned, since mass can neither be created nor destroyed, mass entering a unit shall be equal to the mass leaving the unit.** For instance, in a polymerization reaction of ethylene to yield say $(C_2H_4)_{20}$, 20 moles of ethylene react to produce one mole of the polymerized product. In other words, ethylene entering the reactor is about 20 x 28 = 560 kg/s and if 100 % conversion is assumed, the polymerized product leaving the reactor will be also 560 kg/s (This is due to the fact that 560 kg of the polymer corresponding to one mole of the polymerized product).

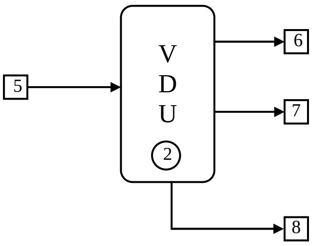
In representing a process flow-sheet, split fractions are conveniently used to identify the distribution of mass (and volumes) in the system. Split fraction associated to a product stream from a unit is defined as the mass flow rate of the product stream divided with the mass flow rate of the feed stream entering the unit. Fundamentally, the petroleum refinery can be regarded to be a combination of physical and chemical processes. In a physical process, the product stream flow rates can be easily evaluated with the knowledge of the associated split fractions. For instance, for the CDU process, the volumetric split fraction of naphtha stream is equal to the corresponding volume % of naphtha cut in the crude TBP. In other words, sum of the volume and mass based split fractions associated for all product streams for the CDU will be equal to 1, as no chemical transformations took place. However, in a chemical process, the associated volumetric split fractions are not bound to be summed up to 1, where as the mass based split fractions shall sum up to 1, as mass can neither be created nor destroyed.

3.2 Refinery Block diagram

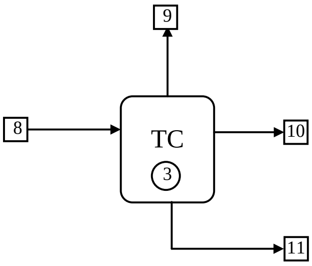
A conventional petroleum block diagram presented by Jones and Pujado (2006) for Kuwait crude is elaborated in this chapter with minor modifications to the flow-sheet. For the matter of convenience, the entire flow-sheet is represented using 22 sub-process units/splitters. Further, all streams are numbered to summarize their significance in various processing steps encountered in various units. Figure 3.1a-u summarizes these 22 sub-processes. A brief account of these units and their functional role is presented as follows:



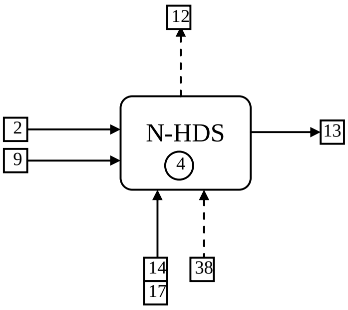
(a) Crude distillation Unit



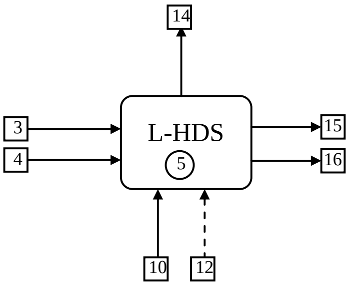
(b) Vacuum distillation unit



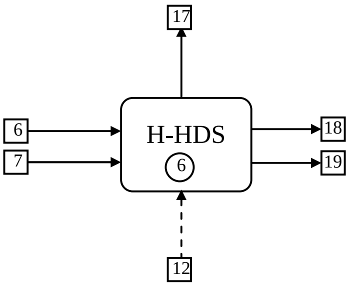
(c)Thermal Cracker



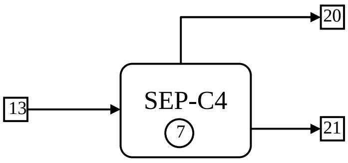
(d) Naphtha HDS



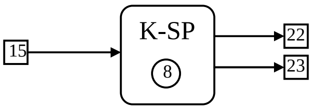
(e) LGO Hydrotreater



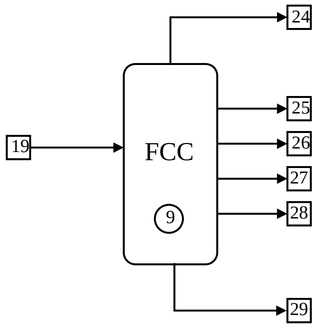
(f) HGO Hydrotreater



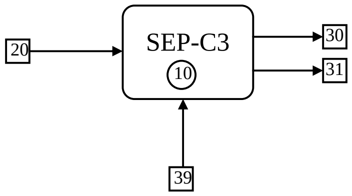
(g) C4 Separator



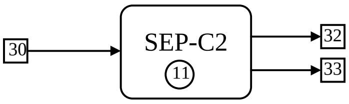
(h) Kerosene Splitter



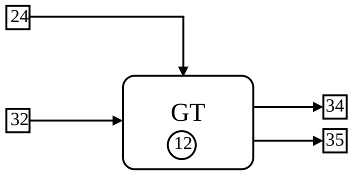
(i) FCC Unit



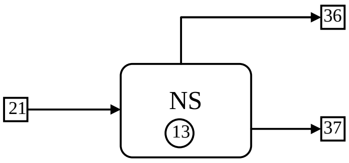
(j) C3 Separator



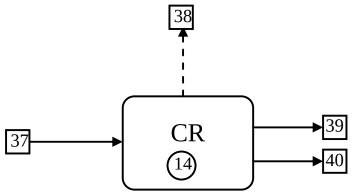
(k) C2 Separator Unit



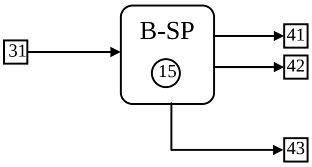
(m) Gas treating Unit



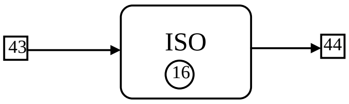
(n) Naphtha Splitter



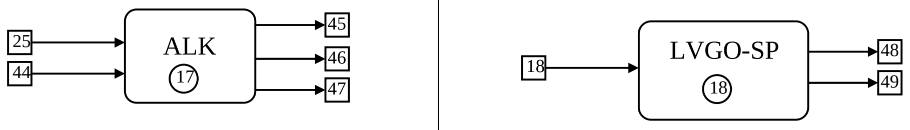
(o) Catalytic Reformer



(p) n-Butane Splitter

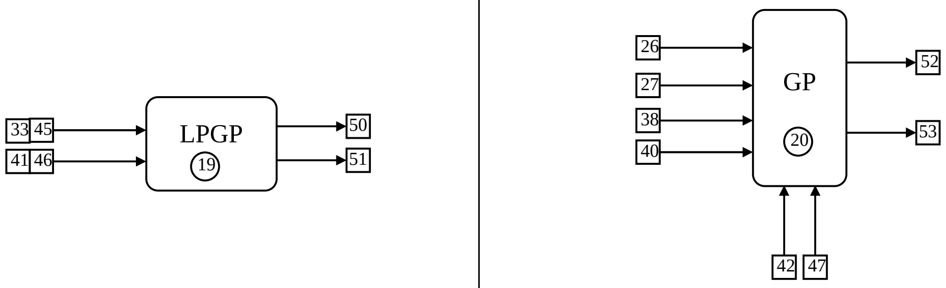


(q) Isomerization Unit



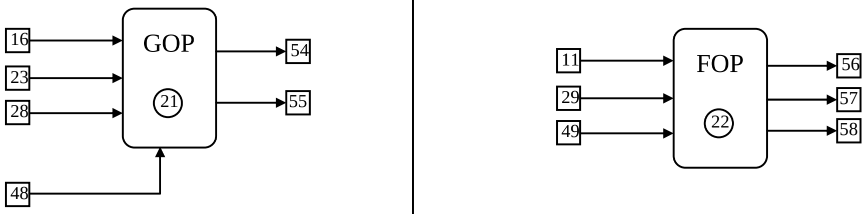
(r)Alkylation Unit

(s) LVGO Splitter



(t) LPG Pool Unit

(u) Gasoline Pool



(v) Gasoil Pool

(w) Fuel Oil Pool

Figure 3.1: Summary of prominent sub-process units in a typical petroleum refinery complex.

a) **Crude distillation unit (CDU):** The unit comprising of an atmospheric distillation column, side strippers, heat exchanger network, feed de-salter and furnace as main process technologies enables the separation of the crude into its various products. Usually, five products are generated from the CDU namely gas + naphtha, kerosene, light gas oil, heavy gas oil and atmospheric residue (Figure 3.1a). In some refinery configurations, terminologies such as gasoline, jet fuel and diesel are used to represent the CDU products which are usually fractions emanating as portions of naphtha, kerosene and gas oil. Amongst the crude distillation products, naphtha, kerosene have higher values than gas oil and residue. On the other hand, modern refineries tend to produce lighter components from the heavy products. **Therefore, reactive transformations (chemical processes) are inevitable to convert the heavy intermediate refinery streams into lighter streams.**

b) Vacuum distillation unit (VDU): The atmospheric residue when processed at lower pressures does not allow decomposition of the atmospheric residue and therefore yields

LVGO, HVGO and vacuum residue (Figure 3.2b). The LVGO and HVGO are eventually subjected to cracking to yield even lighter products. The VDU consists of a main vacuum distillation column supported with side strippers to produce the desired products. Therefore, VDU is also a physical process to obtain the desired products.

- **c)** **Thermal Cracker:** Thermal cracker involves a chemical cracking process followed by the separation using physical principles (boiling point differences) to yield the desired products. Thermal cracking yields naphtha + gas, gasoil and thermal cracked residue (Figure 3.2c). In some petroleum refinery configurations, thermal cracking process is replaced with delayed coking process to yield coke as one of the petroleum refinery products.
- **d)** **Hydrotreaters:** For many refinery crudes such as Arabic and Kuwait crudes, sulfur content in the crude is significantly high. Therefore, the products produced from CDU and VDU consist of significant amount of sulfur. Henceforth, for different products generated from CDU and VDU, sulfur removal is accomplished to remove sulfur as H_2S using Hydrogen. The H_2 required for the hydrotreaters is obtained from the reformer unit where heavy naphtha is subjected to reforming to yield high octane number reforme product and reformer H_2 gas. In due course of process, H_2S is produced. Therefore, in industry, to accomplish sulfur removal from various CDU and VDU products, various hydrotreaters are used. In due course of hydrotreating in some hydrotreaters products lighter than the feed are produced. For instance, in the LVGO/HVGO hydrotreater, desulfurization of LVGO & HVGO (diesel) occurs in two blocked operations and desulfurized naphtha fraction is produced along with the desulfurized gas oil main product (Figure 3.1f). Similarly, for LGO hydrotreating case, along with diesel main product, naphtha and gas to C5 fraction are obtained as other products (Figure 3.1e). Only for kerosene hydrotreater, no lighter product is produced in the hydrotreating operation. It is further interesting to note that naphtha hydrotreater is fed with both light and heavy naphtha as feed which is desulfurized with the reformer off gas. In this process, light ends from the reformer gas are stripped to enhance the purity of hydrogen to about 92 % (Figure 3.1d). Conceptually, hydrotreating is regarded as a combination of chemical and physical processes.
- **e)** **Fluidized catalytic cracker:** The unit is one of the most important units of the modern refinery. The unit enables the successful transformation of desulfurized HVGO to lighter products such as unsaturated light ends, light cracked naphtha, heavy cracked naphtha, cycle oil and slurry (Figure 3.1i). Thereby, the unit is useful to generate more lighter products from a heavier lower value intermediate product stream. Conceptually, for the purpose of modeling, the unit can be regarded as a combination of chemical and physical processes.
- **f)** **Separators:** The gas fractions from various units need consolidated separation and require stage wise separation of the gas fraction. For instance, C4 separator separates the desulfurized naphtha from all saturated light ends greater than or equal to C4s in composition (Figure 3.1g). On the other hand, C3 separator separates butanes (both iso and nbutanes) from the gas fraction (Figure 3.1j). The butanes thus produced are of necessity in isomerization reactions, LPG and gasoline product generation. Similarly, the C2 separator separates the saturated C3 fraction that is required for LPG product generation (Figure 3.1k) and generates the fuel gas + H_2S product as well. All these units are regarded as physical processes for modeling purposes.
- **g)** **Naphtha splitter:** The naphtha splitter unit consisting of a series of distillation columns enables the successful separation of light naphtha and heavy naphtha from the

consolidated naphtha stream obtained from several sub-units of the refinery complex (Figure 3.1n). The naphtha splitter is regarded as a physical process for modeling purposes.

h) **Reformer:** Heavy naphtha which does not have high octane number is subjected to reforming in the reformer unit to obtain reformate product (with high octane number), light ends and reformer gas (hydrogen). Thereby, the unit produces high octane number product that is essential to produce premium grade gasoline as one of the major refinery products. A reformer is regarded as a combination of chemical and physical processes.

i) **Alkylation & Isomerization:** The unsaturated light ends generated from the FCC process are stabilized by alkylation process using iC_4 generated from the C4 separator. The process yields alkylate product which has higher octane number than the feed streams (Figure 3.1r). As isobutane generated from the separator is enough to meet the demand in the alkylation unit, isomerization reaction is carried out in the isomerization unit (Figure 3.1q) to yield the desired make up iC_4.

j) **Gas treating:** The otherwise not useful fuel gas and H_2S stream generated from the C2 separator has significant amount of sulfur. In the gas treating process, H_2S is successfully transformed into sulfur along with the generation of fuel gas (Figure 3.1m). Eventually, in many refineries, some fuel gas is used for furnace applications within the refinery along with fuel oil (another refinery product generated from the fuel oil pool) in the furnace associated to the CDU.

k) **Blending pools:** All refineries need to meet tight product specifications in the form of ASTM temperatures, viscosities, octane numbers, flash point and pour point. To achieve desired products with minimum specifications of these important parameters, blending is carried out. There are four blending pools in a typical refinery. While the LPG pool allows blending of saturated C3s and C4s to generate C3 LPG and C4 LPG, which do not allow much blending of the feed streams with one another (Figure 3.1t). The most important blending pool in the refinery complex is the gasoline pool where in both premium and regular gasoline products are prepared by blending appropriate amounts of n-butane, reformate, light naphtha, alkylate and light cracked naphtha (Figure 3.1u). These two products are by far the most profit making products of the modern refinery and henceforth emphasis is there to maximize their total products while meeting the product specifications. The gasoil pool (Figure 3.1v) produces automotive diesel and heating oil from kerosene (from CDU), LGO, LVGO and slurry. In the fuel oil pool (Figure 3.1w), haring diesel, heavy fuel oil and bunker oil are produced from LVGO, slurry and cracked residue.

l) **Stream splitters:** To facilitate stream splitting, various stream splitters are used in the refinery configuration. A kerosene splitter is used to split kerosene between the kerosene product and the stream that is sent to the gas oil pool (Figure 3.1h). Similarly, butane splitter splits the n-butane stream into butanes entering LPG pool, gasoline pool and isomerization unit (Figure 3.1p). Unlike naphtha splitter, these two splitters facilitate stream distribution and do not have any separation processes built within them.

With these conceptual diagrams to represent the refinery, the refinery block diagram with the complicated interaction of streams is presented in Figure 3.2. A concise summary of stream description is presented in Table 3.1.

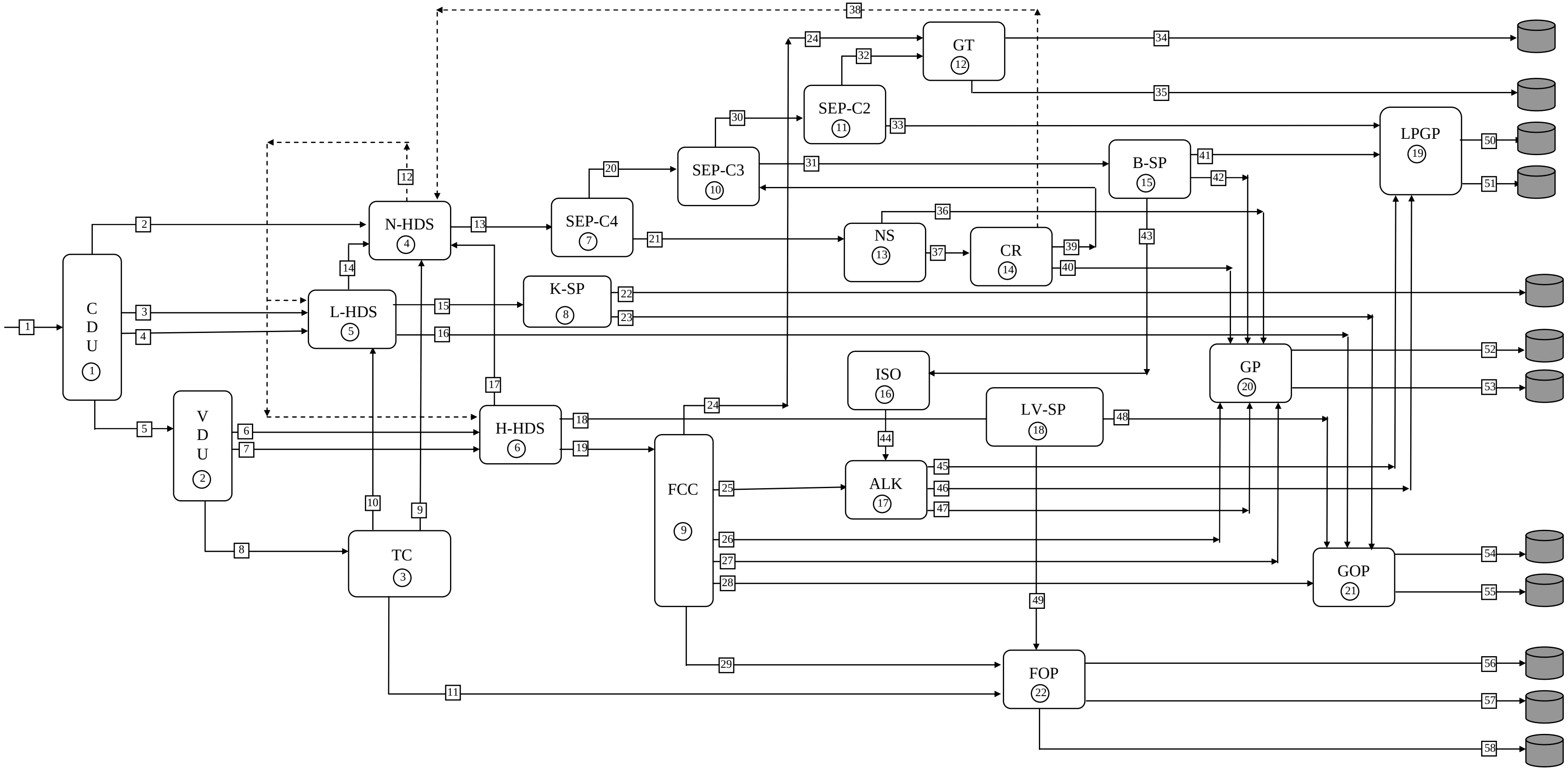


Figure 3.2: Refinery Block Diagram (Dotted lines are for H_2 stream).

Stream	Stream Make	Source	Destination	Functional Role
1	Crude oil	Market	CDU ①	Separation
2	Gas + Naphtha	CDU ①	N-HDS ④	Sulfur removal
3	Kerosene	CDU ①	L-HDS ⑤	Sulfur removal
4	Light Gas oil	CDU ①	L-HDS ⑤	Sulfur removal
5	Atmospheric residue	CDU ①	VDU ②	Separation
6	Light Vacuum Gas oil (LVGO)	VDU ②	H-HDS ⑥	Sulfur removal
7	Heavy Vacuum Gas oil (HVGO)	VDU ②	H-HDS ⑥	Sulfur removal
8	Vacuum residue	VDU ②	TC ③	Cracking
9	Gas + Naphtha	TC ③	N-HDS ④	Sulfur removal
10	Cracked Gas oil	TC ③	L-HDS ⑤	Sulfur removal
11	Cracked residue	TC ③	FOP ㉒	Product Blending
12	Hydrogen	N-HDS ④	L-HDS ⑤ H-HDS ⑥	Hydrodesulfurization of intermediate products
13	Desulfurized Gas + Naphtha	N-HDS ④	SEP-C4 ⑦	Separation of gas (≤ C4) and Naphtha (LN + HN)
14	Desulfurized Gas + Naphtha	L-HDS ⑤	N-HDS ④	Naphtha stabilization (to saturate unsaturates)
15	Desulfurized Kerosene	L-HDS ⑤	K-SP ⑧	Splitting Kerosene for blending pool and product
16	Desulfurized LGO	L-HDS ⑤	GOP ㉑	Togenerate Auto diesel and heating oil products
17	Desulfurized Gas + Naphtha	H-HDS ⑥	N-HDS ④	Naphtha stabilization (to saturate unsaturates)
18	Desulfurized LVGO	H-HDS ⑥	LV-SP ⑱	To by-pass the stream
19	Desulfurized HVGO	H-HDS ⑥	FCC ⑨	To catalytically crack and produce lighter products
20	Saturated light ends	SEP-C4 ⑦	SEP-C3 ⑩	To separate C3s from C4 fraction
21	Desulfurized LN+HN	SEP-C4 ⑦	NS ⑬	To split Light Naphtha (LN) from Heavy Naphtha (HN)
22	Desulfurized Kerosene product	K-SP ⑧	Storage tank	Storage
23	Desulfurized Kerosene	K-SP ⑧	GOP ㉑	To blend and produce auto diesel and heating oil
24	Gaseous FCC product	FCC ⑨	GT ⑫	Sulfur recovery and fuel gas production
25	Unsaturated Light Ends	FCC ⑨	ALK ⑰	Conversion of C_{3-4} to alkylates
26	Light cracked naphtha	FCC ⑨	GP ⑳	Togenerate Premium and regular

				gasoline
27	Heavy cracked naphtha	FCC ⑨	GP ⑳	Togenerate Premium and regular gasoline
28	Cycle oil	FCC ⑨	GOP ㉑	Togenerate Auto diesel and heating oil products
29	Slurry	FCC ⑨	FOP ㉒	Togenerate haring diesel, heavy fuel oil and bunker oil
30	Saturated light ends ($\leq$ C3s)	SEP-C3 ⑩	SEP-C2 ⑪	To separate C3s from the stream
31	C4s (normal and Isobutane mixture)	SEP-C3 ⑩	B-SP ⑮	To by-pass the stream
32	Fuel gas + H_2S	SEP-C2 ⑪	GT ⑫	To recover sulfur and produce Fuel gas
33	C3s	SEP-C2 ⑪	LPGP ⑲	To recover sulfur and produce Fuel gas
34	Fuel gas	GT ⑫	Fuel gas storage tank	Storage
35	Sulfur	GT ⑫	Sulfur storage tank	Storage
36	Desulfurized LN	NS ⑬	GP ⑳	To prepare Premium and Regular gasoline products
37	Desulfurized HN	NS ⑬	CR ⑭	To crack heavy naphtha into lighter products
38	Reformer off-gas	CR ⑭	N-HDS ④	H_2 purication by loss of Light Ends in N-HDS process
39	Cracked Light ends	CR ⑭	SEP-C3 ⑩	To separate $\leq$ C3's from Butanes
40	Reformate	CR ⑭	GP ⑳	To prepare premium and regular gasoline products
41	Normal Butane	B-SP ⑮	LPGP ⑲	To produce C3 LPG and C4 LPG products
42	Normal Butane	B-SP ⑮	GP ⑳	To produce premium and regular gasoline products
43	Normal Butane	B-SP ⑮	ISO ⑯	To convert nC_4 into iC_4
44	Isobutane	ISO ⑯	ALK ⑰	To reactant unsaturates with isobutane and produce alkylates
45	C3s	ALK ⑰	LPGP ⑲	To produce C3 LPG and C4 LPG products
46	C4s	ALK ⑰	LPGP ⑲	To produce C3 LPG and C4 LPG products
47	Alkylate	ALK ⑰	GP ⑳	To produce premium and regular gasoline products
48	Desulfurized LVGO	LV-SP ⑱	GOP ㉑	Togenerate Auto diesel and heating oil products

$\boxed{49}$	Desulfurized LVGO	LV-SP ⑱	FOP ㉒	Togenerate haring diesel, heavy fuel oil and bunker oil
$\boxed{50}$	C3 LPG product	LPGP ⑲	Storage tank	Storage
$\boxed{51}$	C4 LPG product	LPGP ⑲	Storage tank	Storage
$\boxed{52}$	Premium gasoline	GP ⑲	Storage tank	Storage
$\boxed{53}$	Regular gasoline	GP ⑳	Storage tank	Storage
$\boxed{54}$	Auto diesel	GOP ㉑	Storage tank	Storage
$\boxed{55}$	Heating oil	GOP ㉑	Storage tank	Storage
$\boxed{56}$	Haring diesel	FOP ㉒	Storage tank	Storage
$\boxed{57}$	Heavy fuel oil	FOP ㉒	Storage tank	Storage
$\boxed{58}$	Bunker oil	FOP ㉒	Storage tank	Storage

Table 3.1: Summary of streams and their functional role as presented in Figures 3.1 and 3.2.

3.3 Refinery modeling using conceptual black box approach

All refinery processes on a conceptual mode can be classified into physical and chemical processes. For conceptual modeling purposes, split fraction of ith product (with volumetric flow rate P_i) emanating from a process fed with feed F (volumetric flow rate) is defined as:

$$\xi_i^{\,v} = \frac{P_i}{F}$$

a) Physical process: Physical process is a process in which the sum of the volumetric split fractions associated with the products of the system is equal to 1.

$$\sum_i \xi_i^{\,v} = 1$$

b) Chemical process: A chemical process is a process in which the sum of the volumetric split fractions associated with the products of the system is not equal to 1.

$$\sum_i \xi_i^{\,v} \neq 1$$

For both chemical as well as physical processes, the mass split fractions can be defined as

$$\xi_i^{\,m} = \frac{PM_i}{FM}$$

where PM_i and FM are the product and feed mass flow rates

For both chemical as well as physical processes, since mass can neither be created nor destroyed, mass based split fractions associated to the products shall always sum up to 1.

$$\sum_i \xi_i^m = 1$$

In a physical process, the volumetric split fractions of product i can be evaluated from the TBP assay of the crude which indicates the volume assay of the crude. For a chemical process, the volumetric split fractions are not possible to estimate from the feed data itself. Therefore, pilot plant/licensor's data based on detailed experimentation is required to specify these volumetric or mass based split fractions.

3.4 Mass balances across the CDU

A CDU produces five different products namely gas + naphtha (GN), Kerosene (K), Light gas oil (L), heavy gas oil (H) and residue (R).

The steady volumetric balance for the CDU is defined as

$$F_{crude} = F_{GN} + F_K + F_L + F_H + F_R \tag{1}$$

where F refers to the volumetric flow rates of various streams (crude, GN, K, L, H and R).

The mass balance for the CDU is defined as

$$MF_{crude} = MF_{GN} + MF_K + MF_L + MF_H + MF_R$$

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

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

$$MP_i = P_i \times 42 \times 8.33 \times SG_i$$

Where SG_i refers to the specific gravity of the stream i.

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

$$F_{crude} SG_{crude} = F_{GN} SG_{GN} + F_K SG_K + F_L SG_L + F_H SG_H + F_R SG_R \tag{2}$$

Where SG_i refers to the specific gravity of various streams (crude, GN, K, L, H and R).

In the above expression, the specific gravity of the crude is usually given in the crude assay. The specific gravity of the products is evaluated as a function of the product TBP and crude SG assay. Detailed procedures for the same are presented in the earlier chapter. In summary, both equations (1) and (2) can be used to determine two unknowns. Usually, residue flow rate and residue specific gravity is not provided and hence, these can be determined from these two equations as two unknowns.

Since sulfur is an important parameter that needs to be monitored throughout the petroleum refinery, sulfur balance is also usually taken care in the refinery mass balance calculations. Sulfur balance across the CDU is expressed as

$$MF_{crude} SU_{crude} = MF_{GN} SU_{GN} + MF_K SU_K + MF_L SU_L + MF_H SU_H + MF_R SU_R \tag{3}$$

In terms of the volumetric flow rates, the above expression can be reformulated as

$$\begin{aligned} F_{crude} SG_{crude} SU_{crude} &= F_{GN} SG_{GN} SU_{crude} + F_K SG_K SU_K \\ &+ F_L SG_L SU_L + F_H SG_H SU_L + F_R SG_R SU_L \end{aligned} \tag{4}$$

Once the residue flow rate and specific gravity are known from the above expression, residue sulfur content can be evaluated and the mass balance across the CDU can be completed.

We next present an illustrative example for the mass balance across the CDU for the heavy Saudi crude oil.

Q 2.1: Assuming that the refinery processing capacity is 30,000 barrels per day, conduct the mass balances across the CDU processing heavy Saudi crude oil. The crude product properties can be assumed to be those obtained in Chapter 2.

Solution:

The average properties of various products as obtained from their TBP and SG, sulphur assay of the crude is presented below:

Properties	Whole crude	Naphtha	Kerosene	LGO	HGO	Residue
Cut volume (%)	100	22.8	8.8	10.6	5.8	50.16
S.G	0.8758	0.716473	0.803103	0.8420331	0.8773646	0.96467
A.P.I	30.05	65.995	45.30642	36.54565	29.77845	15.18
S%	2.36393	0.010954	0.353414	1.220185	2.048658	4.281

Thereby, the mass balances across the CDU are presented as follows:

Stream	Vol %	Flows (Barrels per day)	S.G.	Mass flow rate (mmlbs/day)	Sulfur content (wt %)	Sulfur flow (lbs/day)
Gas + Naphtha	22.8	6840	0.716	1.715	0.011	0.0002
Kerosene	8.8	2640	0.803	0.742	0.353	0.0026
LGO	10.6	3180	0.842	0.937	1.220	0.0114
HGO	5.8	1740	0.877	0.534	2.049	0.0109
Atmospheric Residue	52	15600	0.965	5.265	4.281	0.2254
Crude (Total)	100	30000	0.8758	9.192	2.639	0.2426

It can be observed that most of the sulphur is present in the atmospheric residue for the Saudi crude.

3.5 Mass balances across the VDU

With atmospheric residue as feed, the VDU produces three products namely LVGO, HVGO and vacuum residue. Mass balance expressions for the VDU are presented as follows:

$$F_R = F_{LVGO} + F_{HVGO} + F_{VR} \tag{4}$$

$$F_R SG_R = F_{LVGO} SG_{LVGO} + F_{HVGO} SG_{HVGO} + F_{VR} SG_{VR} \tag{5}$$

$$F_R SG_R SU_R = F_{LVGO} SG_{LVGO} SU_R + F_{HVGO} SG_{HVGO} SU_{HVGO} + F_{VR} SG_{VR} SU_{VR} \tag{6}$$

In these expressions, having known the volumetric flow rate of LVGO, HVGO and their properties (SG and SU), the volumetric flow rate of the vacuum residue and the properties of the vacuum residue (SG & SU) can be estimated.

The volumetric flow rate of the LVGO and HVGO can be obtained from their respective yields that is evaluated using the range of the cut temperatures on the crude TBP assay. The cut temperatures corresponding to these products is taken as follows:

a) LVGO: 650 – 750 °F
b) HVGO: 750 – 930 °F
c) Vacuum Residue: 930 °F

The specific gravity and sulphur content of LVGO and HVGO are evaluated with the TBP curve data of these products and crude SG and sulphur assay curves. The TBP curves of these products is estimated by assuming that their 50% and 70 % TBP data points match with the corresponding crude TBP cuts. This is because the end point correlation was not providing good predictions for 100 % cut temperatures. Eventually, ASTM temperature evaluation is carried out using probability chart and the ASTM to TBP conversion is carried out using Edmister correlation.

Below we illustrate the procedures for mass balance across the VDU for Saudi heavy crude.

Q 2.2: Assuming the atmospheric residue flow rate, specific gravity and sulphur content to be those obtained from the previous solved problem (Q1), conduct the mass balances across the VDU.

Solution

For the VDU unit, the following cut range data is assumed with the fact that the HGO cut for the CDU varied till 680 °F but not 650 °F:

a) LVGO: 680 – 780 °F
b) HVGO: 780 – 930 °F
c) Vacuum Residue: 930 °F +

Evaluation of LVGO ASTM & TBP:

From crude TBP assay, LVGO Cut range volume (680 – 780 °F): 48.2 to 56.2 % = 8 %

TBP 50 % = 730 °F

From Edmister correlation ASTM 50 % = 709 °F

TBP 70 = 750 °F
From Edmister correlation ASTM 70 % = 722 °F

Using probability chart, the ASTM data at other points was obtained which is eventually converted to TBP using the Edmister correlation. Calculations are summarized as follows:

Volume	ASTM		TBP	
%	°F	ΔT (°F)	ΔT (°F)	°F
IBP	656	25	48	628
10	681	19	37	676
30	700	9	17	713
50	709			730
70	722	13	20	750
90	746	24	33	783
EP	784	37	41	824

Evaluation of HVGO ASTM & TBP:

From crude TBP assay, HVGO cut range volume (780 – 930 °F): 56.2 to 67.4 = 11.2 %

TBP 50 % = 852 °F

From Edmister correlation ASTM 50 % = 815 °F

TBP 70 = 882 °F
From Edmister correlation ASTM 70 % = 835 °F

Using probability chart, the ASTM data at other points was obtained which is eventually converted to TBP using the Edmister correlation. Calculations are summarized as follows:

Volume	ASTM		TBP	
%	°F	ΔT (°F)	ΔT (°F)	°F
IBP	760	21	42	747
10	781	20	38	789
30	801	14	25	827
50	815			852
70	835	20	30	882
90	857	22	30	912
EP	900	43	47	959

Using these TBP data, the pseudo-component distribution is evaluated for both these cuts. The same is presented as follows along with the mid-point specific gravity and sulphur content (from crude assay data):

Psuedo Component No.	Vol % LVGO	Vol % HVGO	Mid pt S.G.	Mid pt. sulphur wt %
14	4.2		0.876161	2
15	74.1	1.8	0.901274	2.6
16	21.7	53.6	0.921824	2.75
17		28.1	0.940199	3.05
18		13.9	0.946488	3.2
19		2.6	0.952862	3.35
Total	100	100	-	-

Average specific gravity of LVGO product $= \frac{\sum [A][B]}{\sum [A]} = 0.904679$

Average specific gravity of HVGO product $= \frac{\sum [A][B]}{\sum [A]} = 0.930853$

Average sulphur content of LVGO product $= \frac{\sum [C][D]}{\sum [D]} = 2.6087$

Average sulphur content of HVGO product $= \frac{\sum [C][D]}{\sum [D]} = 2.9121$

From mass balance calculations of the CDU, Atmospheric residue (which is VDU feed) properties are: Specific gravity = 0.96467 and sulphur = 4.281

Normalized volume fractions of the VDU system are:

Feed (Atmospheric residue) = 100 %

LVGO = $= \frac{8.}{52} \times 100 = 15.384\%$

HVGO = $= \frac{11.2.}{52} \times 100 = 21.538\%$

Vacuum residue = 100 – 15.384 – 21.538 = 63.078 %

Average specific gravity of the Vacuum residue product

$$= \frac{100 \times 0.96467 - 15.384 \times 0.904679 - 21.538 \times 0.930853}{63.078} = 0.99084$$

Average sulphur content of the vacuum residue product

$$= \frac{100 \times 0.96467 \times 4.281 - 15.384 \times 0.904679 \times 2.6087 - 21.538 \times 0.930853 \times 2.9121}{100 \times 0.96467 - 15.384 \times 0.904679 - 21.538 \times 0.930853}$$

$$= \frac{318.28457}{62.5002} = 5.09252\%$$

Therefore, mass balances across the VDU are presented as follows:

Stream	Vol %	Flows (Barrels per day)	S.G.	Mass flow rate (mmlbs/day)	Sulfur content (wt %)	Sulfur flow (lbs/day)
LVGO	15.384	2400	0.90468	0.760	2.60876	19816.85
HVGO	21.538	3360	0.93085	1.094	2.91210	31865.41
VR	63.078	9840	0.99084	3.411	5.09252	173710.3
Atmospheric Residue	100	15600	0.96467	5.265	4.281	225394.3

3.6 Mass balances across the Thermal Cracker

The Thermal cracker produces 4 products namely gas with H_2S) (TCG), naphtha (TCN), gas oil (TCGO) and residue (TCR) from the vacuum residue feed (VR).

Volumetric, mass and sulphur balance relationships are presented as follows:

$$F_{VR} = F_{TCG} + F_{TCN} + F_{TCGO} + F_{TCR} \quad (4)$$

$$F_{VR}SG_{VR} = F_{TCG}SG_{TCG} + F_{TCN}SG_{TCN} + F_{TCGO}SG_{TCGO} + F_{TCR}SG_{TCR} \quad (5)$$

$$F_{VR}SG_{VR}SU_{VR} = F_{TCG}SG_{TCG}SU_{TCG} + F_{TCN}SG_{TCN}SU_{TCN} + F_{TCGO}SG_{TCGO}SG_{TCGO} + F_{TCR}SG_{TCR}SU_{TCR} \quad (6)$$

In the above expressions, one unknown in each equation can be obtained with the prior information for all other unknowns.

Since thermal cracker is a chemical process, pilot plant/licensor’s data is required. However, pilot plant data can be difficult to analyze. Below, we present a procedure to extract the desired information for the mass balances across the thermal cracker.

Maples (2000) presented various correlations for the thermal cracker balances. These include API gravity, sulphur content, H_2S wt % in the themal cracker product etc. However, unlike for four different products, Maples (2000) provides data for 5 different products namely Gas, Gasoline, Distillate, Gasoil and Residue. When Maples (2000)’ correlations were studied for the mass balances provided by Jones and Pujado (2006), it was found that the gasoil and residue represented in Maples (2000) matched the cracked residue mass balance data presented by Jones and Pujado (2006). This is possible with the fact that the cut range temperatures for distillate match very much with the gas oil cut range temperatures and for gas oil and residue the cut range temperatures were fairly high. Therefore, firstly, mass balances are conducted assuming five products are generated from the thermal cracker and later they are adjusted to suite the requirements of four different products. The sequential procedure for the evaluation of thermal cracker mass balances is presented as follows.

a) For the feed stream, evaluate mass flow rate (in mmlbs/CD) of the product as well as the sulphur.

b) The weight percent distributions of the products were evaluated using Table 9.1 (Visbreaker database) of Maples (2000) (pg. 125). In this table, the weight percent distribution of various products is given as a function of the thermal cracker feed and its condrason carbon residue (CCR). Fortunately, the relation between CCR with the specific gravity of the thermal cracker feed is presented by Dolomatov *et al* (1988) using the following expression:

$$CCR = 0.0058\exp^{7.8499\times SG}$$

Using Table 9.1, determine the weight percent distribution of products. These are conveniently expressed as w_{TCG}, $w_{TCGA,}w_{TCD}$, w_{TGO}, w_{TCR} which correspond to the weight percents of gas, gasoline, distillate, gas oil and thermal cracked residue products. The sum of these weight percent distributions is 100. Also note the wt % conversion value from Table 9.1 as **many correlations in Maples (2000) are provided as functions of wt % conversion.**

c) Using the plot given in Figure 9.13 convert the wt% data to LV% (liquid volume) data. These are designated as v_{TCG}, $v_{TCGA,}v_{TCD}$, v_{TGO}, v_{TCR}. The sum of these volume % data termed as SLV_{TC} will not be equal to 100 and would be higher than 100. The LV% value signifies that when 100 LV equivalent of feed is taken, SLV_{TC} equivalent thermal cracked products is obtained.

d) Using SLV_{TC}, determine the total product volumetric flow rate (barrels per day) from thermal cracker using the expression:

$$FT_{TC} = \frac{SLV_{TC}F_{VR}}{100}$$

e) The specific gravity of the total products SGT_{TC} is evaluated using the expression:

$$SGT_{TC} = \frac{F_{VR}SG_{VR}}{FT_{TC}}$$

f) Determine the individual flow rates (BPCD) of various products using the expressions:

$$F_{TCG} = \frac{FT_{TC}v_{TCG}}{SLV_{TC}};\ F_{TCN} = \frac{FT_{TC}v_{TCN}}{SLV_{TC}}\text{ etc.,}$$

g) Using Fig. 9.6, 9.7, 9.8 and 9.9 of Maples (2000), determine the °API of gasoline, distillate, gas oil and residue products.

h) The specific gravity of the gas product from the thermal cracker is obtained from the mass balance expression

$$SG_{TCG} = \frac{F_{VR}SG_{VR} - F_{TCN}SG_{TCN} - F_{TCGO}SG_{TCGO} - F_{TCR}SG_{TCR}}{F_{TCG}}$$

i) With product volumetric flow rate and specific gravity known, conduct mass balance calculations for feed and products.

j) Using Fig. 9.12 of Maples (2000), determine the sulphur content of the gasoline, distillate and residue products.

k) Using Fig. 9.1 of Maples (2000), determine the weight percent H_2S (tw_{H2S}) in the gaseous product

l) Determine the sulphur content of the gaseous product using the expression

$$SU_{TCG}=\frac{\frac{tw_{H2S}\times\frac{32}{34}}{FT_{TC}}}{F_{TCG}SG_{TCG}}$$

m) Having known the sulphur content of all products other than the thermal cracked residue, determine its sulphur content using the expression:

$$SU_{TCR}=\frac{F_{VR}SG_{VR}SU_{VR}-(F_{TCG}SG_{TCG}SU_{TCG}+F_{TCN}SG_{TCN}SU_{TCN}+F_{TCGO}SG_{TCGO}SG_{TCGO})}{F_{TCR}SG_{TCR}}$$

n) Having determined the sulphur content of all the products, conduct sulphur mass balance for all products

o) Having obtained mass and sulphur balances across the thermal cracker configuration provided by Maples (2000), adjust the mass, specific gravity and sulphur content of the gas oil and thermal cracker residue of Maples (2000) to the thermal cracker residue of Jones and Pujado (2006).

We present below an illustrative example for the thermal cracker mass balances.

Q 2.3: Assuming the vacuum residue flow rate, specific gravity and sulphur content to be those obtained from the previous solved problem (Q2), conduct the mass balances across the thermal cracker.

Solution

a) Vacuum residue specifications:

Flow rate = 9840 barrels/day
Specific gravity = 0.99085
Mass flow rate = 3.411 mmlbs/day
Sulphur content = 5.09252
Sulphur mass flow rate = 0.173712 mmlbs/day

b) API gravity of the feed = 141.5/0.99085 – 131.5 = 11.3

CCR of the feed = $0.0058\exp^{7.8499\times 0.99085}=13.84$

In Table 9.1 of Maples (2000), the data corresponding to 11.3 API and 13.84 CCR is not available. The closest available data is that corresponding to 12.5 API and 14.5 CCR with a conversion of 11.83 %.

c) The weight percent of products are gas 3.27 %; gasoline 8.56%; distillate 11.5%; gasoil 27.56 % and residue 50.91. The total of these wt% values is 100 %.

d) From Fig. 9.13, the corresponding LV% of these streams is gas 4.827%; gasoline 9.907 %, distillate 12.73%, gasoil 26.423 % and residue 50.573%. The total volume of these LV% i.e., SLV = 104.46%.

Total product flow rate = 104.46/100 x 9840 = 10279 barrels/day

e) After normalization of the LV% data, the volumetric flow rates of various products can be determined as gas 475 barrels/day; gasoline 975 barrels/day; distillate 1225 barrels/day; gasoil 2600 barrels/day and cracked residue 4976 barrels/day.

f) From Fig. 9.6 – 9.9 of Maples (2000), the °API of gasoline, distillate, gas oil and residue products are 57.5, 31.9, 20.1 and 4.1 respectively. These when converted to SG are about 0.7487, 0.866, 0.9334 and 1.0435 respectively.

g) From overall mass balance, mass flow rate of the gas = 3.411 – 1.817 – 0.849 – 0.380 – 0.255 = 0.110 mmlbs/day

h) From mass balances, specific gravity of the gas product
= 0.110 x 1000000/(42 x 8.33) = 0.6646

i) From Fig. 9.12, sulphur content of gasoline, distillate and residue products are 1.24, 2.78 and 5.39 % respectively.
j) From Fig. 9.1, H_2S wt % in the gas product = 0.39 wt %
k) H_2S in gas = 0.39/100 x 3.411 = 0.0133 wt %
l) Sulphur wt % in gas = 0.0133 x 32/34/0.110 = 0.1137 wt fraction or 11.37 wt %
m) For residue, sulphur content = [0.173712 – (0.012521 + 0.003166 + 0.01055 + 0.097925)]/1.817 = 5.390 wt %

Mass balance table for products based on Maples (2000) configuration

Stream	Flow (barrels per day)	°API	SG	Total mmlbs/day	SU (wt %)	Sulphur mmlbs/day
VR Feed	9840.0000	11.3067	0.9909	3.4111	5.0925	0.1737
Gas	474.9768		0.6646	0.1104	11.3375	0.0125
Gasoline	974.8488	57.5000	0.7487	0.2553	1.2400	0.0032
Distillate	1252.6320	31.9000	0.8660	0.3795	2.7800	0.0106
Gasoil	2600.0232	20.1000	0.9334	0.8490	5.8360	0.0495
Crack Res	4976.3832	4.1000	1.0435	1.8168	5.3900	0.0979
Total	10278.8640		0.9485	3.4111	5.0925	0.1737

Mass balance table for products based on Jones and Pujado (2006) configuration

Stream	Flow (bbls/day)	S.G.	Mmlbs/day	SU (wt%)	Sulphur mmlbs/day
Gas	474.9768	0.6646	0.110	11.33	0.0125
Naphtha	974.8488	0.7487	0.256	1.24	0.0031
Gas oil	1252.632	0.866	0.378	2.78	0.0105
Residue	7576.4064	1.0057	2.666	5.5320312	0.147475

3.7 Mass balances across HVGO hydrotreater

The HVGO hydrotreater according to the refinery block diagram produces four different products namely desulfurized naphtha, desulphurized gas oil, light ends with H_2S and vent gas stream that consists of H_2S with other light ends. These products are designated as NH, GOH, LEH, VEH. As expressed by Jones and Pujado (2006), for gases instead of specific gravity, molecular weight is taken as a basis for mass balance calculations as their flow rates are usually expressed in scf/day and not barrels/day. The feed streams to the HVGO hydrotreater are termed as HVGO and HH

Since hydrotreating is a mild chemical process, once again volumetric split fractions based on feed data could not be used in carrying out an approximate calculation of the mass balances across the HVGO hydrotreater.

The mass balance across the hydrotreater is presented as follows:

$$F_{HVGO}SG_{HVGO}M_1 + F_{HH}MW_{HH}M_2 = F_{NH}SG_{NH}M_1 + F_{GOH}SG_{GOH}M_1$$
$$+ F_{LEH}MW_{LEH}M_2 + F_{VEH}SG_{VEH}M_2$$

$$F_{HVGO}SG_{HVGO}M_1SU_{HVGO} + F_{HH}MW_{HH}M_2SU_{HH} = F_{NH}SG_{NH}M_1SU_{HVGO} +$$
$$F_{GOH}SG_{GOH}M_1SU_{GOH} + F_{LEH}MW_{LEH}M_2SU_{LEH} + F_{VEH}SG_{VEH}M_2SU_{VEH}$$

In the above expressions, M_1 and M_2 are multiplication factors to convert the flow terms into mass terms. M_1 is 42 x 8.33 and $M_2 = 1/379$. Also, in the above expressions, only one unknown in each equation can be evaluated with the knowledge of the remaining items.

The following procedure is adopted to sequentially freeze various terms in the expressions and proceed towards mass balance.

a) F_{HH}, SG_{GOH}, SU_{GOH} can be obtained from virgin gas oil data base provided in Table 15.7 of Maples (2000).
b) From Table 15.7 of Maples (2000), obtain the scf H_2 required for barrel of HVGO desulphurization. From this determine the flow rate of H_2 in scf/day.
c) Assume the molecular weight of hydrogen as 11 (due to the presence of other light ends). Since the feed H_2 gas is not having any sulphur in it, SU_{H2} can be taken as zero.
d) Usually 80 – 90 % sulphur removal is assumed by weight and in general, 85 % sulphur loss is assumed from the final product.
e) From data tables/licensor's data sheet, obtain the sulphur content of the desulfurized HVGO product. This is usually about 0.3 – 0.5 wt % (3000 – 5000 ppm).
f) Since amount of sulphur removed is known, the desulfurized gas oil product can be determined from the following sulphur balance expression:

$$F_{GOH} = \frac{F_{HVGO}SG_{HVGO}M_1SU_{HVGO} \times \left(1 - \frac{SR_H}{100}\right)}{SG_{GOH}M_1SU_{GOH}}$$

Where SR_H refers to the sulphur wt % removed (on a total feed basis).

g) From data tables/licensor's data sheet, obtain the sulphur content of the desulfurized naphtha product. This is usually about 80 – 90 ppm.

h) Usually 2 - 3 % of the light ends on a weight basis gets removed from the liquid products (desulphurized Naphtha & gas oil products) and enters the light ends stream mixed with H_2S. Assume that this is about 2 %. In other words, the following expression is used further to solve the mass balances:

$$1-\frac{F_{NH}SG_{NH}M_1+F_{GOH}SG_{GOH}M_1}{F_{HVGO}SG_{HVGO}M_1}=0.02$$

i) Assume the desulfurized naphtha products have a specific gravity that is dependent on the feed specific gravity. A convenient correlation is obtained from data presented by Jones and Pujado (2006) and Prakash in their books. Jones and Pujado (2006) reported that for a HVGO feed SG of 0.9218, the naphtha SG is 0.786 (Table 2.21 of Jones and Pujado (2006) book). Similarly, in Table 2.12 of his book, Prakash reported that for a feed SG of 0.8967, the naphtha density is 0.7716. Using these two data sets, a linear correlation is developed to relate the naphtha specific gravity as function of the feed SG as

$$SG_{NH}=0.573SG_{HVGO}+0.257$$

j) Having obtained all terms in the light ends balance expression other than F_{NH}, determine F_{NH} using this expression. Conduct mass and sulphur balances for both desulphurized naphtha and gas oil products.

k) For the gaseous products, we need to specify their flows (F_{VH}, F_{LEH}), their molecular weights (MW_{VH}, MW_{LEH}) and their sulphur content (SU_{VH}, SU_{MEH}). Amongst these since we have two mass balance expressions, four parameters need to be specified. In this case we specify the F_{LEH} using suitable multiplication factor, assume MW_{VH} and MW_{LEH} from Jones and Pujado (2006) and use the overall mass balance expression to determine the F_{VH}. Eventually, by specifying either SU_{VH}, SU_{MEH}, the other can be obtained from the sulphur mass balance. For this purpose, based on Jones and Pujado (2006), it is assumed that the light ends + H_2S product is 10 % of the feed H_2 on a volume basis. The average molecular weights of the gaseous products are assumed to be similar to those provided by Jones and Pujado (2006) i.e., 28 for H_2S rich stream (vent) and 34 for LE rich stream. Further, either SU_{VEH} or SU_{LEH} is assumed to conduct the sulphur mass balance.

We next present an illustration of the above procedure for the HVGO hydrotreater mass balance in the Saudi heavy crude oil processing example.

Q 2.4: Assuming that the HVGO hydrotreater is fed with HVGO obtained from the VDU along with the evaluated properties in **Q2**, conduct mass balances across the HVGO with the methodology outlined in this section.

Solution:

HVGO SG = 0.93085 (^{o}API = 20)
HVGO flow rate = 3360 bbl/day
HVGO sulphur content = 2.9121
HVGO overall mass flow rate = 1.094 mmlbs/day
HVGO sulphur mass flow rate = 0.031865 mmlbs/day

From Table 15.7 of Maples (2000), the closest data for HVGO hydrotreater is as follows:
Feed API = 21; Feed sulphur wt % = 2.6
^{o}API of desulphurized HVGO product = 24.5. This implies that product SG = 0.907051.

SU_{GOH} = 0.5 (wt %)
SCF/barrel of H_2 required = 210.

Assume that 85 % sulphur removal from the feed on a weight basis. Also, assume SU_{NH} = 0.08 wt %.

$$SG_{NH} = 0.573 \times 0.93085 + 0.257 = 0.790377$$
$$SU_{NH} = 0.08$$

$$F_{GOH} = \frac{3360 \times 0.93085 \times 42 \times 8.33 \times 2.9121 \times (1 - \frac{85}{100})}{0.907051 \times 42 \times 8.33 \times 0.5} = 3012.4 \text{ bbl/day}$$

F_{HH} = 210 x 3360 = 705600 bbl/day

It is assumed that 2 % of total light ends are lost to the gaseous products from feed. In other words,

$$1 - \frac{F_{NH} \times 0.790377 + 3012.4 \times 0.907051}{3360 \times 0.93085} = 0.02$$

From the above expression, F_{NH} = 420.94 bbl/day

Assume F_{LEH} = 0.1 x 705600 = 70560 scf/day

Assume that the average molecular weight of the gases are MW_{VEH} = 28 and MW_{LEH} = 34.

Therefore, using these values, from overall mass balance, once can determine F_{VEH} from overall mass balance

F_{VEH} from overall mass balance = 487746 scf/day. This amounts to about 69.12 volume % of the feed flow rate.

It is further assumed that SU_{VEH} = 67. Eventually, from overall sulphur mass balance, SU_{LEH} = 50.7%. Overall mass and sulphur mass balances are presented in the following table

Stream	Flow (bbl/day or scf/day)	SG or MW	Mass flow Mmlbs/day	Sulphur (wt %)	Sulphur mass flow Mmlbs/day
HVGO	3360	0.93085	1.094242	2.9121	0.031865
H2	705600	11	0.020479	0	0
Total in	-	-	1.114721	-	0.031865
Des. HVGO	3012.41	0.907051	0.955962	0.5	0.00478
Naphtha	420.92	0.790377	0.116394	0.08	9.31E-05
H_2S vent	487746	28	0.036034	66	0.023782
H_2S+LE	70560	34.00	0.00633	50.7	0.003209
Total	-	-	1.114721	-	0.031865

3.8 Mass balances across LVGO hydrotreater

LVGO hydrotreater produces four fractions namely Gas to C5, Naphtha, gas oil and vent gas. Once again two gaseous products and two liquid products are produced from the LVGO hydrotreater. Mass balances across the LVGO hydrotreater are carried out using procedure similar to that presented for the HVGO hydrotreater. The following salient features apply for the same:

a) Since LVGO is lighter than HVGO, not more than 0.5 % by weight of light ends is assumed to enter the gaseous products from the LVGO hydrotreater.
b) Certain values of molecular weights and sulphur content of the gaseous products may have to be adjusted to obtain an acceptable mass balance.

The illustrative example presented below elaborates the above salient features.

Q 2.5: Assuming that the LVGO hydrotreater is fed with the LVGO with properties generated in Q 2.2, conduct the mass balances across the LVGO hydrotreater along with suitable assumptions.

Solution:

Once again Licensor's data is not available and one has to depend on data banks provided by Maples (2000). A detailed sequential procedure is presented as follows for LVGO hydrotreater mass balance.

LVGO SG = 0.90468 (^{o}API = 25)
LVGO flow rate = 2400 bbl/day
LVGO sulphur content = 2.60876
LVGO overall mass flow rate = 0.7596 mmlbs/day
LVGO sulphur mass flow rate = 0.019817 mmlbs/day

From Table 15.7 of Maples (2000), the closest data for LVGO hydrotreater is as follows:
Feed API = 24.6; Feed sulphur wt % = 2.3
^{o}API of desulphurized HVGO product = 27.7. This implies that product SG = 0.888819.
SU_{GOH} = 0.43 (wt %)
SCF/barrel of H_2 required = 240

Assume that 85 % sulphur removal from the feed on a weight basis. Also, assume SU_{NH} = 0.08 wt %.

$$SG_{NL} = 0.573 \times 0.90468 + 0.257 = 0.775382$$
$$SU_{NL} = 0.08$$

$$F_{GOL} = \frac{2400 \times 0.90468 \times 42 \times 8.33 \times 2.60876 \times (1 - \frac{85}{100})}{0.888819 \times 42 \times 8.33 \times 0.43} = 2223.05 \text{ bbl/day}$$

F_{HL} = 240 x 3360 = 720000 bbl/day

It is assumed that 0.5 % of total light ends are lost to the gaseous products from feed. In other words,

$$1-\frac{F_{NL}\times 0.775382+2223\times 0.888819}{2400\times 0.90468}=0.005$$

From the above expression, F_{NL} = 237.92 bbl/day

Assume F_{LEL} = 0.12 x 720000 = 86400 scf/day

Assume that the average molecular weight of the gases are MW_{VEH} = 23 and MW_{LEH} = 48. This data is taken from Jones and Pujado (2006).

Therefore, using these values, from overall mass balance, once can determine F_{VEL} from overall mass balance

F_{VEL} from overall mass balance = 226616 scf/day. This amounts to about 31.47 volume % of the feed flow rate.

It is further assumed that SU_{VEL} = 80. Eventually, from overall sulphur mass balance, SU_{LEL} = 52.9

Overall mass and sulphur mass balances are presented in the following table

Stream	Flow (bbl/day or scf/day)	SG or MW	Mass flow Mmlbs/day	Sulphur (wt %)	Sulphur mass flow Mmlbs/day
LVGO	2400	0.90468	0.759627	2.60876	0.019817
H2	720000	11	0.020897	0	0
Total in	-	-	0.780524	-	0.019817
Des. LVGO	2223.053	0.888819	0.691286	0.43	0.002973
Naphtha	237.9267	0.775382	0.064544	0.08	5.16E-05
H_2S vent	226616.1	23	0.013752	80	0.011002
Gas to C5	86400	48.00	0.010942	52.92	0.005791
Total	-	-	0.780524	-	0.019817

In the mass balance, it is interesting to note that while many parameters evaluated were similar with the calculations reported by Jones and Pujado (2006), the molecular weight of Gas to C5 is quite low (52.92 instead of 80).

3.9 Mass balances across the FCC

The FCC unit according to the refinery block diagram produces six different products namely:

a) Gas to C3
b) C3 to C5
c) C5 to 300 °F (Light Cracked Naphtha)
d) 300 to 420 °F (Heavy Cracked Naphtha)
e) Light cycle oil (LCO)
f) Slurry

For these products, conducting mass balances requires many data such as conversions, liquid volume yield (with respect to the feed), specific gravity and sulphur content. Amongst several products, a microscopic breakdown of C3 to C5 product is very much desired in the FCC operation, as this is the feed that will be sent to alkylation unit and therefore, is desired in the mass balance data. The FCC mass balance therefore is pretty complex even when a macroscopic mass balance is considered.

Maples (2000) summarized FCC data base in Tables 12.1 a – g. In these tables, Maples (2000) provided the following data as a function of feed characterization factor (K), °API, LV% conversion and sulphur content.

a) Gas fraction (wt % only)
b) C3 LV%
c) C3- (Unsaturated) LV%
d) C4- (Unsaturated) LV%
e) nC4
f) iC4
g) Gasoline
h) Light cycle oil
i) Heavy cycle oil

A critical observation of Maples (2000) data and the correlations provided by him in his book for estimation of °API and sulphur content indicates that the gasoline data has not been distributed towards both light and heavy cracked naphtha. Also gas fraction is presented with respect to weight and not volume. Therefore, the macroscopic balances need to be conducted with certain approximations.

Therefore, the following procedure is adopted for the mass balances across the FCC.

a) Assume a LV% conversion of about 70 – 75 %. This is because, at this conversion, appropriately larger quantity of gasoline is produced in the FCC.
b) Carefully match the most relevant data from Maples (2000) Tables 12.1a-g from available feed data including its oAPI and sulfur content. Always choose a data that provides maximum information with respect to the LV% of all 9 products listed above.
c) Convert the gas fraction (wt %) provided by Maples (2000) to LV% using the expression:

$$LV_{GFCC} = WT_{GFCC}\frac{8.1}{4.7}$$

These multiplication factors were obtained from an analysis of mass balance calculations provided by Jones and Pujado (2006).

d) Assume that gasoline LV% gets distributed by a conversion factor to both light cracked naphtha and heavy cracked naphtha. From Jones and Pujado (2006) mass balance calculations, it appears that about 2/3 (66.7%) volume of the gasoline distributes towards light cracked naphtha and the balance towards heavy cracked naphtha.
e) Assume that the heavy cycle oil designated in Maples (2000) data base refers to the slurry in the FCC block diagram considered here
f) Determine the specific gravity of C3 to C5 product with their known compositions and individual specific gravities.
g) Assume gas to C3 product specific gravity to be about 0.5204 (from Jones and Pujado (2006)).
h) Determine the specific gravity of gasoline, LCO and slurry from Fig. 12.11, 12.13 and 12.16 of Maples (2000) respectively.
i) If a feasible mass balances case arises, then fix the specific gravity of the gasoline from the Maples (2000) data base. Else, use data provided by Jones and Pujado (2006) for LCN and HCN specific gravity.
j) Determine the sulphur content of LCO and slurry from Fig. 12.12 of Maples (2000).
k) Determine the sulphur content of gasoline from Fig. 12.8 of Maples (2000).
l) Assume sulphur content of C3 to C5 and light naphtha to be negligible (that is zero)
m) Adjust all sulphur available in the gasoline (from Maples (2000)) to the heavy cracked naphtha product.
n) From an overall mass balance determine the specific gravity of heavy cracked naphtha.
o) From a mass balance of sulphur (overall) determine the sulphur content of the gaseous product.

These procedural steps are illustrated in the following illustrative example for the macroscopic balances of the FCCU.

Q 2.6: Conduct the mass balances across the FCC unit with the assumption that the properties of the feed correspond to the desulphurized HVGO stream in Q4.

Solution

Desulphurized HVGO flow rate = 3012.41 bbl/day
Specific gravity = 0.907051. Feed ^{o}API = 24.5
Feed sulfur content = 0.5 wt %
Feed mass flow rate = 0.955961 mmlbs/day
Feed sulphur mass flow rate = 0.000478

From Maples (2000) Table 12.1c the following data is chosen to be the closest to represent the FCC process:

Feed API = 24.9
%LV Conversion = 72.7
Feed sulphur = 0.6 wt %

Products:
Gas wt% =2.7

C3 = 2.7
C3- = 7.0
iC4 = 4.8
nC4 = 1.2
C4- = 7.3
Gasoline = 58.9
LCO = 21
HCO = 6.2

From these data, the following LV% of the products was obtained:

Gas fraction = 2.7 x 8.1/4.7 = 4.65 %

C3 to C5 = 2.7 + 7 + 4.8 +1.2 + 7.3 = 23 %

Light cracked Naphtha = 0.67 x 58.9 = 39.463 %

Heavy cracked naphtha = (1-0.55) x 58.9 = 19.437 %

Light cracked naphtha = 21

Slurry = 6.2

Total LV% = 113.7532

Since LV% is known for all products, their flow rates can be determined next.

Flow of Gas fraction = 4.65 x 3012.41 = 140.17 bbl/day

Similarly, the flows of other products are
C3 to C5 product = 692.85 bbl/day
LC Naphtha = 1188.79 bbl/day
HC Naphtha = 585.52 bbl/day
LCO = 632.61 bbl/day
Slurry = 186.77 bbl/day
Total products flow rate = 3426.71 bbl/day

Next, we move towards the evaluation of specific gravities for different products. First, we focus on the specific gravities of the C3 to C5 product. For this purpose, the specific gravities presented by Jones and Pujado (2006) (in page 93) for C3 to C4 components are taken. The specific gravity calculation is presented in the following table:

	lbs/gal	Vol %	Flows bbl/day	Lbs	gal
C3	4.22	2.7	81.33507	14415.83	3416.073
C3-	4.34	7	210.8687	38437.15	8856.485
iC4	4.68	4.8	144.5957	28421.73	6073.019
nC4	4.86	1.2	36.14892	7378.718	1518.255
C4-	5	7.3	219.9059	46180.25	9236.049
Total	4.633478	23	692.85	134833.7	29099.88

Specific gravity of the C3 to C5 product = 134833.7/29099.88/42 = 0.55624
From Maple Fig. 12.11, 12.13 and 12.16, API of various products are:

Gasoline = 55.5
LCO = 18
Slurry = 15.5

These when converted to SG were about

SG of Gasoline = 0.756684
SG of LCO = 0.946488
SG of Slurry = 0.962585

From Jones and Pujado (2006) Mass balances, assume LC naphtha SG = 0.793518

In this regard, it can be observed that Jones and Pujado (2006) Naphtha SG is higher than the gasoline SG presented by Maples (2000). Therefore, SG data of gasoline is ignored in future calculations to serve as an important parameter to check.

Proceed for overall mass balance to evaluate the SG of heavy cracked naphtha. From this calculation, the SG of the heavy cracked naphtha = 0.9431. This value is significantly higher than that provided by Jones and Pujado (2006) (SG = 0.9159).

On an average, using these two values of SG, the gasoline naphtha from specific gravity blending calculations is about 0.842 which is about ^{o}API of 36.3. This value is quite significantly different from that provided by Maples (2000) in his correlation (^{O}API = 55.5).

Next we proceed towards the sulphur mass balance. From Maples (2000) Fig. 12.2 and 12.8, sulphur content of

LCO = 0.54
Slurry = 1
Gasoline = 0.06143 (from interpolation of data in the graph).

Since we assume that LCN does not have any sulphur in its product, we evaluate the distribution of sulphur into the HCN according to gasoline sulphur mass balance i.e.,

$$SU_{HCN} = 0.06143 \times \frac{0.193196 + 0.20948}{0.20948} = 0.1663$$

Since in the overall sulphur mass balance, only sulphur wt % in the gas fraction is not known, it is solved to get the sulphur content as 10.57%.

When compared with the data provided by Jones and Pujado (2006), the sulphur content is significantly high (Jones and Pujado (2006) value is about 2.19 wt %).

The overall mass balance for the FCC unit is presented below

Stream	LV%	Flow (bbl/day)	S.G.	Sulphur content (wt %)	Mass flow (mmlbs/day)	Sulphur mass flow (mmlbs/day)
Des HVGO		3012.41	0.907051	0.5	0.955961	0.00478
Gas to C3	4.653191	140.17	0.5204	10.57288	0.025521	0.002698
C3 to C5	23	692.85	0.556239887	0	0.134834	0
LC Naphtha	39.463	1188.79	0.793518	0	0.330031	0
HC Naphtha	19.437	585.52	0.943107451	0.166323	0.193196	0.000321
LCO	21	632.61	0.946488294	0.54	0.20948	0.001131
Slurry	6.2	186.77	0.962585034	1	0.062898	0.000629
Total products	113.7532	3426.71			0.955961	0.00478

3.10 Mass balances across the Diesel hydrotreater

The LGO hydrotreater desulphurizes both diesel and kerosene as two blocked operations. Diesel hydrotreater is fed with LGO generated from both CDU and thermal cracker. Therefore these two streams need to be consolidated first to get the overall feed stream properties (density and sulphur content).

Regarding diesel hydrotreater data base, during calculations it was observed that there is lack of consistency in the data base presented in Table 15.3 of Maples (2000). Therefore, where applicable, necessary adjustments in % sulphur removed are carried out. These adjustments are carried out in order to obtain the diesel product flow rate lower than the consolidated LGO feed stream to the hydrotreater.

All necessary assumptions are illustrated in the following example. In summary, diesel hydrotreater calculations are similar to the LVGO and HVGO hydrotreater calculations.

Q 2.7: Conduct the mass balances across the LGO hydrotreater assuming that it is fed with the LGO streams generated from the CDU and TC units processing heavy Saudi crude oil.

Solution:

Firstly, LGO streams are consolidated and evaluated for the feed stream specific gravity and sulphur content.

Stream	Flow (bbl/day)	SG	Sulf content (wt%)	Mass flow (mmlbs/day)	Sulf mass (mmlbs/day)
LGO from CDU	3180	0.842	1.22	0.936771	0.011429
LGO from TC	1252.632	0.866	2.78	0.379521	0.010551
Total LGO feed	4432.632	0.848782	1.669788	1.316292	0.021979

LGO consolidated feed ^{o}API = 141.5/0.8488 – 131.5 = 35.2

From Table 15.3 of Maples (2000) (Diesel hydrotreater data base), the following data matches the closest with the feed °API and sulfur content

Feed API	35.9
%S	1.32
Product API	37.9
Product %S	0.21
SCFB H_2	115

Hydrogen feed flow rate = 115 x 4432.6 = 509752.7 scf/day. Assume its molecular weight is 11.

Total feed entering the hydrotreater is presented as follows:

Stream	Flow (bbl/day or scf/day)	SG or MW	Sulf content (wt%)	Mass flow (mmlbs/day)	Sulf mass (mmlbs/day)
Total LGO feed	4432.632	0.848782	1.669788	1.316292	0.021979
H_2 required	509752.7	11	0	0.014795	0
Total Feed to hydrotreater				1.331087	0.021979

From Maples (2000) hydrotreater data base

Diesel product SG = 141.5/(131.5 + 37.9) = 0.8353
Diesel product sulphur content = 0.21

Assume % sulphur removed by weight = 88%. This value is assumed so as to obtain an appropriate flow rate of the diesel product.

With this assumption, the diesel product flow rate = $\dfrac{0.021979\times(1-0.88)\times1000000}{42\times8.33\times0.8487}$

= 4297.7 bbl/day.

For naphtha product, assume its SG is equal to the SG of FCC Naphtha product i.e. SG of Naphtha = 0.7935

Assume sulphur content in naphtha product = 0.08 wt %

Assume that 0.1 wt% of the total LGO feed enters the gaseous products i.e., Gas to C5 and Vent gas.

From this balance, one can evaluate the naphtha flow rate expression as

$$1-\frac{F_{NLGO}\times0.7935+4297\times0.8353}{1.316292}=0.001$$

From this equation, naphtha flow rate = 212.58 bbl/day.

Based on Jones and Pujado (2006) mass balance calculations, it is assumed that Gas to C5 product flow rate (scf/day) to the consolidated LGO feed stream (bbl/day) is about 4.5. Jones and Pujado (2006) value is about 4.9.

Gas to C5 product flow rate = 4.5 x 4297.7 = 19339.65 bbl/day.

Further, assume the molecular weights of Gas to C5 to be 55 and Vent gas to be 37. These values are assumed from Jones and Pujado (2006) mass balances.

Using these values, from overall mass balance expression, one can calculate the Vent gas flow rate as 135380.5 bbl/day.

Further assume the sulphur content of Gas to C5 to be 70% (Jones and Pujado (2006)). Eventually, from overall sulphur mass balance, one can determine the vent gas sulphur content to be 76.5%.

Using these values, the feed and product mass balance is summarized as follows

Stream	Flow (bbl/day or scf/day)	SG or MW	Sulf content (wt%)	Mass flow (mmlbs/day)	Sulf mass (mmlbs/day)
Total LGO feed	4432.632	0.848782	1.669788	1.316292	0.021979
H_2 required	509752.7	11	0	0.014795	0
Total Feed to hydrotreater				1.331087	0.021979
Diesel Product	4297.721	0.835301	0.21	1.255959	0.002638
Naphtha product	212.5844	0.79	0.08	0.059016	4.72E-05
Gas to C5	19946.84	55	70	0.002895	0.002026
Vent gas	135380.5	37	76.53657	0.013217	0.017268
Total				1.331087	0.021979

3.11 Mass balances across the kerosene hydrotreater

Kerosene hydrotreater operates to desulphurize kerosene. Since kerosene itself is a lighter stream, no lighter streams are produced from the hydrotreater. Also, sulphur removal is considered to be about 99 wt % usually. Therefore, kerosene hydrotreater produces highly pure H_2S as product along with desulphurized kerosene product. Losses are possible from strippers within the unit in minute quantities which can be ignored.

Once again Maples (2000) data base is used for the kerosene hydrotreater mass balance to obtain the product ^{o}API.

The illustrative example presents the mass balance across the kerosene hydrotreater in a lucid way.

Q 2.8: Conduct mass balances for the kerosene hydrotreater assuming that the kerosene fed to the hydrotreater is generated from the CDU fed with heavy Saudi crude oil.

Solution:

Kerosene product flow rate (from CDU) = 2640 bbl/day
Kerosene SG = 0.803 (^{o}API = 44.7)
Kerosene sulfur content = 0.353
Kerosene mass flow rate = 0.742 mmlbs/day
Kerosene sulfur mass flow rate = 2618.11 lbs/day

From Maples (2000) data base (Table 15.1), the closest data base is

Feed API = 44.0
Feed %S = 0.4

Product API = 44.8
SCFB H_2= 40

Actual Product API (by adjustment) = 44.7*44.8/44 = 45.5. This corresponds to SG of 0.7994

Assume 99 % sulphur removal.

Therefore, total sulphur in product = 26.18 lbs/day.

Product stream flow rate = 99.8 x 2640 = 2634.72 bbl/day

From sulphur balance, product stream flow rate is = 26.18 x 100/(42 x 8.33 x 0.7994 x 2634.72) = 0.00355%

H_2 required = 40 x 2640 = 105600 scf/day (278.6 lbmol/day)

H_2S generated = 2591.9/32 = 80.99 lbmol/day = 30695 scf/day

Unreacted H_2 = 105600 – 30695 = 74905 scf/day. (In Jones and Pujado (2006) book, all H_2 fed is consumed. However, data base is not conveying the same. We stick to the data base in this section).

Mass balance table is presented as follows for the Kerosene hydrotreater

Stream	Flow	SG or	Sulf	Mass flow	Sulf mass
Kerosene feed	2460	0.803	0.353	0.742	2618.11
H_2 required	105600	11	0	0.030649	0
Total Feed to				0.772649	2618.11
Kerosene	2634.72	0.7994	0.00355	0.736873	26.18
Vent gas (+LE	105600	128.6	7.248	0.035776	2591.2

3.12 Naphtha consolidation

In order to carry out mass balances across the reformer, one needs to consolidate naphtha stream as, the consolidated naphtha stream (inclusive of Gas to C5, Light naphtha and heavy naphtha gathered from various units) is sent to naphtha hydrotreater first whose product splitter feed is fed to naphtha splitter and eventually the heavy naphtha from the naphtha splitter is sent to the catalytic reformer.

Since mass balances now onwards require component wise breakdown of lighter products emanating from various units, it is essential that certain assumptions shall be made. TThese are

a) Since component wise breakdown of lighter components is not available, Jones and Pujado (2006) data is assumed on a volumetric basis and scaled to the overall volumetric flows of the chosen problem.
b) The specific gravity and sulfur content of light naphtha streams from various units is assumed (as they are not known) and heavy naphtha streams from various units are evaluated for their specific gravity and sulfur content.

The following illustrative example summarizes the procedure for naphtha consolidation case of heavy Saudi crude oil refining problem.

Q 2.9: For the refinery block diagram, gather results obtained from Q1-Q7 to consolidate the naphtha and light gas stream data. Eventually, with certain assumptions for the light naphtha SG and sulfur content, evaluate the heavy naphtha properties of various streams.

Solution:

From an analysis of the refinery block diagram, one can identify that gas to C5 steams are emanating from CDU and TC. On the other hand, Naphtha streams are emanating from CDU, TC, HVGO, LVGO and LGO.

A) Consolidated Gas to C5 streams

For Gas to C5 streams emanating from the CDU, Jones and Pujado (2006) volumetric distribution is assumed.

Thus, Gas to C5 stream flow rate for the heavy Saudi crude oil processing case = 855/50000*30000 = 513 BPCD

Assume Gas to C5 stream density = 0.465 (same as Thermal cracker Gas to C5 of the Jones and Pujado (2006) problem).

Assume gas sulfur content to be zero.

Data gathered from the thermal cracker mass balance indicates the following information for Gas to C5 streams:

Flow rate = 474.97 BPCD
SG = 0.6646
SU = 0.1104

Therefore, consolidated Gas to C5 streams are summarized as follows:

Stream	Flow (BPCD)	S.G.	Sulfur content (wt %)
CDU	513	0.465	0
TC	475	0.6646	0.1104
Total	988	0.56092	0.062833

Further, it can be observed from the mass balance of the CDU that the LSR stream consists of both gas + naphtha. Therefore, from overall LSR stream, corresponding CDU Naphtha stream properties are evaluated using the above gas to C5 stream properties i.e.,

Stream	Flow	SG	SU (wt%)
Gas +Naphtha (From CDU MB)	6840	0.716	0.011
Gas	513	0.465	0
CDU Naphtha stream	6327	0.736351	0.011563

From the mass balance tables of TC, LVGO, HVGO and LGO, the following data is summarized for the naphtha stream.

Stream	Flow (BPCD)	SG	SU (wt %)
TC Naphtha	975.3	0.7487	1.24
HVGO Naphtha	420.92	0.7903	0.08
LVGO Naphtha	238	0.7753	0.08
LGO Naphtha	212.6	0.79	0.08

From Jones and Pujado (2006) mass balance tables, the following information is used to evaluate the volume distributions between light and heavy naphtha.

Unit	LN	HN	Vol. fraction LN
CDU Naphtha	5219.5	7100	0.4237
TC Naphtha	436.5	577.5	0.4305
All Hydrotreaters (data missing)	323.5	940.5	0.2559

From these values of volume fractions of LN, the LN yields are evaluated. Subsequently, LN SG and sulfur content are evaluated. These are summarized as follows

LN Source	LN	LN SG	LN SUL (wt%)
CDU	2680.6101	0.698	0.005
TC	419.75789	0.715	0.6
HVGO	107.72755	0.7561	0.04
LVGO	60.912184	0.7412	0.04
LGO	54.411472	0.76521	0.04
Total	3323.4192	0.703923	0.08385

From the values of the total naphtha (LN + HN) summarized previously, and known values of the LN flows, SG and sulfur content, HN streams from various units is evaluated and is presented as follows:

HN Source	HN	HN SG	HN SUL (wt%)
CDU	3646.39	0.764545	0.015968
TC	555.5421	0.774163	1.686617
HVGO	313.1925	0.802064	0.09297
LVGO	177.0878	0.787029	0.092957
LGO	158.1885	0.798527	0.093185
Total	4850.401	0.769998	0.219015

In summary, consolidated naphtha + gas streams are summarized as follows:

Gas to C5 stream	988	0.56092	0.062833
Consolidated LN	3323.4192	0.703923	0.08385
Consolidated HN	4850.401	0.769998	0.219015

The consolidated data is very essential to proceed towards the mass balances associated with the reformer and naphtha hydrotreater.

3.13 Mass balances across the reformer

The catalytic reformer unit is fed with the heavy naphtha generated from the naphtha splitter. The properties of the reformer feed will be similar to the heavy naphtha fraction consolidated from various source units. Since catalytic reformer produces H_2 along with lighter hydrocarbons along with the highly desired reformate product, it is essential to know the product flow rates as well as compositions of the lighter stream from the reformer.

Maples (2000) summarized catalytic reformer data base based on many pilot plant studies. A critical observation of these data sets indicates that only few data sets exists where iC_4 yield is presented. Further, it is important to observe that while reformate product rates are provided in terms of LV%, the lighter hydrocarbons are provided in terms of the wt% of the reformate product.

From the reformer it is important to observe a few concepts:

a) It is important to know upon the compositions of the light end stream from the reformer unit.
b) Hydrogen (Reformer off gas) is also generated in which light ends are present. Their compositions will be different from the light end stream product compositions.

The calculation procedure for the reformer mass balance is simple provided a fairly accurate data base is provided. Always, Maples (2000) data base cannot be also regarded as a absolute data base, for it may not provide necessary balances. Therefore, where applicable, certain assumptions or degree of freedom shall be left as illustrated in the following illustrative example.

Q 2.10: Conduct mass balances across the catalytic reformer assuming that it is fed with the heavy naphtha with properties presented in the solution of Q 2.9.

Solution:

Feed data is presented as follows:

Description	Flow (BPCD)	SG	Sul (wt %)	Mass Mmlbs/d	Sulf mass Mmlbs/d
Ref feed	4850.401	0.769998374	0.219015	1.306657	0.002862

The feed SG corresponds to an API of 52.26.

From Maples (2000) reforming data base, the following data is obtained.

Feed API = 54.5
Product LV% = 80.5
H_2 = 2.3 wt %
Product API = 41

Adjusted API = 41/54.5 x 52.26 = 39.32 which corresponds to reformate SG of 0.8982.

Weight percent yield with respect to reformate product mass flow rate for hydrocarbons is

C1	1.2
C2	2.2
C3	3.1
nC4	3.2
iC4	2.1

RON of the reformate product = 100.

From Maples (2000) correlation presented in Fig. 17.7, 2.3 wt % of H_2 corresponds to 1125 scf/barrel.

Assume H_2 purity in reformer off gas = 70 %.

Reformate product flow rate = 80.5 x 4850.4 = 3904.4 bbl/day. This corresponds to 3904 x 42 x 8.33 /10^6 = 1.1316 mmlbs/day

Off gas flow rate = 1125 x 3904/1e6 = 4.392715 mmscf/day. With MW of 12, the H_2 stream mass flow rate = 0.139083 mmlbs/day.

Lighter product component and total mass flow rates are

C1 = 1.2/100 x 1.1316 = 0.013579 mmlbs/day
C2 = 2.2/100 x 1.1316 = 0.024895 mmlbs/day
C3 = 0.03508 mmlbs/day
nC4 = 0.036211 mmlbs/day
iC4 = 0.026027 mmlbs/day
Total LE = 0.135792 mmlbs/day

The specific gravities of these lighter components are assumed as

C1 = 2.5 lbs/gal
C2 = 3.11 lbs/gal
C3 = 4.23 lbs/gal
nC4 = 4.67 lbs/gal
iC4 = 4.86 lbs/gal

With these flow rates, the barrels/day (volumetric) flow rates of the lighter components and light end product are

C1 = 129.3 bbl/day

C2 = 190.59 bbl/day

C3 = 197.4 bbl/day

iC4 = 132.5 bbl/day

nC4 = 177/4 bbl/day

The total mass flow rate of the products = 1.40675 mmlbs/day. Total mass flow rate of the feed = 1.30657 mmlbs/day.

Since mass can neither be created nor destroyed, the LV% of the reformate is adjusted so as to equate the incoming and outgoing mass balances. After trail and error, LV% was found to be 74.78% and not 80.5% as provided by Jones and Pujado (2006).

Subsequently, mass balance table for the reformer is presented as follows

Description	Flow (BPCD/mmscf/day)	SG/MW	Mass Mmlbs/d
Feed	4850.401	0.769998374	1.306657
Ref. prod.	3627.52	0.828	1.05129
H_2	4.080963	12	0.129213
C1	120.1474	2.5	0.012615
C2	177.0662	3.11	0.023128
C3	183.4402	4.23	0.03259
iC4	123.1194	4.676	0.02418
nC4	164.8113	4.86	0.033641
Light ends	768.5845	3.90807475	0.126155
Total products	4033.788		1.306657

3.14 Mass balances across the naphtha hydrotreater

The mass balance across the hydrotreater needs some assumptions.

a) Certain amount of H_2 is lost in the hydrodesulfurization process. According to Jones and Pujado (2006), it is about 6%. We assume it to be 4 %
b) Reformer off gas purity is about 70%.
c) The reformer off gas looses light ends present in it and these light ends are gained by the debutanizer overheads stream (light end product stream from the naphtha hydrotreater).
d) After hydrotreating, the exiting hydrogen stream is told to be of a purity of 92 %. However, in mass balance calculations this is unlikely to happen. This is because, if one evaluates the amount of light ends lost from the reformer off gas, it should again lead to an appropriate balance of the H_2 from back calculation.
e) The mole % of the gas lost from the light ends and gained by the light end stream product is evaluated based on the following data provided by Jones and Pujado (2006).

Component	LE product stream from Reformer (vol%)	LEs transferred from the off gas to the LE product stream of hydrotreater (vol%)	Multiplication factor
Gas to C3	14.57	3.894	0.26717
C3	33.93	35.74	1.053444
iC4	20.3	22.8	1.12303
nC4	31.18	37.55	1.204259
Total	100	100	-

The following illustrative example summarizes the modified procedure of that presented by Jones and Pujado (2006) for the mass balance across the hydrotreater. Eventually, it can be understood that the lack of appropriate data can lead to difficult to accept circumstances such as the inability to enhance the purity of H_2 in the naphtha hydrotreater which is ofcourse a reality in the industrial processing schemes.

Q 2.11: Conduct the mass balances across the naphtha hydrotreater with the following assumptions for the Saudi heavy crude processing scheme:

a) H_2 consumption = 4 % by volume
b) Reformer off gas purity = 70%
c) Reformer off gas MW = 12
d) H_2 rich product stream MW = 11

Solution:

Feed mass balance is presented as follows:

Stream	Flow (BPCD/mmscf/day)	SG	Mass Mmlbs/day
Gas to C5	988	0.560961538	0.193903
LN	3323.419221	0.703922592	0.818473
HN	4850.400779	0.769998374	1.306657
H2	4.080962985	12	0.129213
Total	9161.82		2.448246

H_2 volume lost = 4 %. This corresponds to 0.1142 mmscf/day (H_2 component only).

H_2 leaving the unit = 4.08096 x 0.7 – 0.1142 = 2.7424 mmscf/day

Assuming no change in product purity, naphtha hydrotreater off gas flow rate = 2.7424/ 0.7 = 3.9129 mmscf/day

Volume of the gas (light ends) absorbed = 4.08096 – 3.9129 = 0.16803 mmscf/day

Moles of gas (light ends) absorbed = 0.16803 / 379 x 1000000 = 443.35 lbs/day.

To evaluate the composition of the light ends in the H_2 rich stream, multiplication factors are used to evaluate the normalized composition. The procedure is summarized as follows:

First, evaluate the volume % composition of the reformer unit light end products and segregate the C1 and C2 products as one component in the mass balance table. This is summarized as shown below

Component	bpcd	lbs/gal	vol%	lbs	Mol wt	Moles
C1	120.1474	2.5	15.6323	12615.48	16	788.4675
C2	177.0662	3.11	23.03795	23128.38	30	770.946
C3	183.4402	4.23	23.86728	32589.99	44	740.6816
iC4	123.1194	4.676	16.01898	24179.67	58	416.8909
nC4	164.8113	4.86	21.44348	33641.28	58	580.0221
Total	768.5845	3.908075		126154.8		

Mole fraction amongst C1 & C2 = (788.4675/(770.946+788.4675) = 0.505618

Average molecular weight of the C1 & C2 together = 0.505618 x 16 + (1-0.05618) x 30 = 22.92 lb/lbmol

Similarly, average SG of the C 1& C2 together = 2.86341

Now, the volumetric multiplication factors are applied to obtain the normalized volume % factors for hydrotreater mass balance calculations.

Component	Flow (BPCD)	Lbs/gal	Lbs	vol%	MF	Vol %	Norm. vol%
C1&C2	297.2136	2.86341	35743.86	38.67025	0.267174	10.3317	13.03062
C3	183.4402	4.23	32589.99	23.86728	1.053444	25.14284	31.71084
iC4	123.1194	4.676	24179.67	16.01898	1.12303	17.9898	22.68923
nC4	164.8113	4.86	33641.28	21.44348	1.204259	25.8235	32.56931

These volume % are used to evaluate the BPCD equivalent of the light end gases transferred to the debutanizer overhead product. The calculations are summarized as follows:

	vol %	vol (mmscf/d)	Mol/D	Mol wt	lbs/D	lbs/gal	BPCD
C1&C2	13.03062267	0.021270996	56.1239987	22.92135	1286.438	2.86341	10.69685
C3	31.71083515	0.051764298	136.581261	44	6009.575	4.201084	34.0591
iC4	22.68923381	0.037037569	97.7244573	58	5668.019	4.676394	28.85831
nC4	32.56930837	0.053165657	140.278778	58	8136.169	4.861332	39.84881
Total	100		430.708494		0.0211	4.427747	113.4631

Eventually, the light end stream consolidated from the feed as well as that absorbed is presented as follows:

Stream	Flow (BPCD)	SG	Mmlbs/day
Gas to C5	988	0.560961538	0.19390291
Abs Les	113.4630715	0.531542216	0.0211002
Total	1101.463072	0.557931018	0.21500311

Eventually, the final products mass balance is presented as follows:

Stream	Flow (BPCD)	SG/MW	Mmlbs/day
Deb overhe	1101.314964	0.557931018	0.2149742
Splitter feed	8173.82	0.741186126	2.11956453
Rich H2 gas	3.917724466	11	0.11370704
Total out			2.44824577

It can be observed that the total products mass is matching with the total feed mass and hence the assumption of no enhancement in purity is justified.

However, according to Jones and Pujado (2006) if the hydrogen purity needs to be enhanced to 0.92 % say, then the compositions of the C1-C4 need to be adjusted to achieve this. The same when run as an optimization problem with necessary constraints yields the following information:

	vol %
C1&C2	30
C3	30
iC4	21
nC4	19
Total	100

Interestingly, the debutanizer product flow rate is about 1294.35 barrels/day, which is significantly lower than what can be obtained. Eventually, it will be difficult to presume that where did a flow of about 1707 (the revised total LE stream flow rate) – 1294.35 = 413.45 bpcd is not matched upon optimization. Therefore, it is obvious that certain that a gross assumption of 92 % hydrogen purity for all circumstances in mass balances across the hydrotreater cannot be assumed.

In summary, we conclude that the mass balances across the naphtha hydrotreater are the most difficult to make. Very accurate data and description of the unit are required to evaluate the appropriate mass balances for an acceptable format.

3.15 Mass balances across the alkylator and isomerizer

The mass balance across the alkylator and isomerizer is obtained with the following procedure:

a) Consolidate the propylene and butylenes streams from the FCCU
b) Use multiplication factors provided by Maples (2000) in Table 19.1 for isobutene/olefin ratio and alkylate/olefin. Eventually, evaluate the RON as well using the same. Here, the ratios are provided for 100% consumption of the individual olefin streams. Therefore, after obtaining necessary volumetric flow rates, the final product flow rates are summed up.

c) Assume the alkylate make up from propylene and butylenes feed according to the table provided by Jones and Pujado (2006).
d) Evaluate the necessary additional iC4 required from the isomerizer
e) For isomerizer, first consolidate the saturated light ends from various streams. Make necessary assumptions based on volumetric distributions provided by Jones and Pujado (2006) in the saturated light ends summary table.
f) For isomerizer, assume 50 % by weight. Eventually, carry out mass balances to obtain the necessary feed flow rate to the isomerizer.

Next we present an illustrative example to evaluate the mass balances across the alkylator and isomerizer.

Q 2.12: Conduct the mass balances across the alkylator and isomerizer for the Saudi heavy crude oil scheme.

Solution:

The consolidated FCCU light end streams are as follows:

	bpcd	lbs/gal	Mass Mmlbs/day
C3	81.33507	4.22	0.014416
C3-	210.8687	4.34	0.038437
iC4	144.5957	4.68	0.028422
nC4	36.14892	4.86	0.007379
C4-	219.9059	5	0.04618
Total	692.8543		0.134834

From Table 19.1 of Maples (2000),

Volume ratio Isobutane/olefin = 1.3 for propylene and 1.13 for butylenes

Volume ratio alkylate/olefin = 1.77 for propylene and 1.74 for butylenes

RON = 90.5 for propylene feed based product and 96 for butylenes based product. Therefore,

Stream	Bbl/day	Product RON
Alkylate from C3-	373.2376	
Alkylate from C4-	382.6363	
Total	755.8739	93.28419

Stream	Flow bbl/day
Isobutane for C3	274.1293
Isobutane for C4	248.4937

The composition of the products (along with the assumed distributions provided by Jones and Pujado (2006)) are as follows:

Propylene based alkylate product

Component	% wt	lbs/gal	vol fac	vol %	Flow (bbl/day)
C5	9	5.25	1.714286	9.894125	51.94415
C6	4	5.53	0.723327	4.174736	23.08629
C7	68	5.73	11.86736	68.49336	392.4669
C8	10	5.89	1.697793	9.798935	57.71573
C9+	9	6.8	1.323529	7.638846	51.94415
Total	100	5.771573	17.3263	100	577.1573

Butylene based alkylate product

Component	% wt	lbs/gal	vol fac	vol %	Flow (bbl/day)
C5	4	5.25	0.761905	4.529847	23.7817
C6	2	5.53	0.361664	2.150244	11.89085
C7	2	5.73	0.34904	2.075192	11.89085
C8	80	5.89	13.58234	80.75279	475.6339
C9+	12	6.8	1.764706	10.49193	71.34509
Total	100	5.945424	16.81966		594.5424

Alkylate product flows are

Stream	BPCD	Lbs/gal	mmlbs/CD
Total alkylate	755.8739	5.859579	0.186022

Total isobutane required = 274.12 + 248.49 = 522.623 barrels/day

Isobutane in the feed stream = 144.5957 barrels/day

Therefore, isobutene make up (to be generated from the isomerizer) = 522.6 – 144.5 = 378.02 mmlbs/day. This corresponds to about 0.074305 mmlbs/day. We next proceed towards the isomerizer mass balance.

The feed to the isomerizer is the nC4 and iC4 streams together obtained from the consolidated saturated light ends. With volume distributions of various components assumed as per Jones and Pujado (2006) for LE streams emanating from CDU and TC, the consolidated light end stream data is presented as follows:

Source	Gas to C3	C3	iC4	nC4
CDU	30.56425	61.7278	67.1215	353.5864
TC	260.9155	86.97183	33.4507	93.66197
Reformer	267.5277	165.1181	110.8222	148.3498
Naphtha HDS	56.124	136.5813	97.72446	140.2788
Total	615.13	450.399	309.12	735.877

Total butanes available are

iC4 = 309.1188 barrels/day (0.060631 mmlbs/day)
nC4 = 735.877 barrels/day (0.150207 mmlbs/day).

In terms of mass fraction, the above stream has 0.28757 wt fraction of iC4 and the balance for nC4. The average density of the stream is 4.803796.

Assuming 50 % wt conversion of nC4 for the feed fed with 0.28757 wt fraction iC4, the product obtained will be 0.28757 iC4 + (1-0.28757)/2 iC4 = 0.64378 wt fraction iC4 and the balance i.e., 0.3562 wt fraction nC4.

In other words, for 100 lbs feed, about 64.378 lbs of isomer product is obtained. For a requirement of 0.074305, feed should be 0.074305 x 64.378/100 = 0.115419 mmlbs/day

The left over iC4 + nC4 stream flow rate = (309.11 x 4.67 + 735.877 x 4.86) x 42 x 10^{-6} - 0.115419 = 0.095419 mmlbs/day.

This corresponds to the following left over stream balance (for usage towards blending pools).

Stream	Flow (barrels/day)	Lbs/gal	Mmlbs/day
iC4	139.8977	4.67	0.02744
nC4	333.0354	4.86	0.067979
Total	472.9331		0.095419

3.16 Mass balances across the gasoline pool

The gasoline pool is byfar the most important pool for mass balances as the products generated from the pool generate far more revenue than any other pool products. Therefore, it is desired to obtain highest product flow rates from these pools. In the gasoline pool, both premium and regular grade gasoline are produced whose octane no. is desired to be atleast 90 and 82 respectively.

For the gasoline pool, it is also desired to obtain the reid vapor pressure of the product to be atleast 7 psia for both the products. For this purpose therefore, butanes (both nC4 and iC4) are added to the streams to obtain the reid vapor pressure of about 7 psia. The reid vapor pressure is evaluated using a bubble point calculation to evaluate the volume % butanes to be added for the feed. Therefore, firstly, after meeting the octant number constraint, bubble point calculation is conducted to obtain the total product flow rate.

Since the TBPs of various streams that are mixed is not available for the RVP calculation, we assume the data provided by Jones and Pujado (2006) for various streams is applicable. However, for octane number calculations, as far as possible, we try to generate data based on the information provided by Maples (2000) and adjusting it for our calculations.

We next present an illustrative example that elaborates upon the mass balance calculations for the gasoline pool.

Q 2.13: Conduct mass balance calculations for the gasoline pool to evaluate the product flow rates for the refinery processing scheme fed with heavy Saudi crude oil.

Solution:

Firstly, we consolidate the streams entering the gasoline pool

	Flow (bbl/day)	RON
Alkylate	755.8739	93.2
HC Nap	585.5221	89
LC Nap	1188.79	94
Reformate	3265.203	100
LSR Nap	3323.419	69

For premium gasoline, the following blending volumes (bbl/day) are assumed to obtain a RON greater than or equal to 90.

Stream	Flow (bbl/day)	Vol %	Ron factor	RON product
Alkylate	755.8739172	22.85041	70447.45	
HC Nap	585.5221317	17.7006	52111.47	
LC Nap	339.3626487	10.25909	31900.09	
Reformate	1070.477546	32.36102	107047.8	
LSR Nap	556.6866367	16.82889	38411.38	
Total	3307.922881		299918.1	90.66661

For various streams such as alkylate, HC naphtha, LC naphtha, Reformate and LSR naphtha, since TBP data are not provided, TBPs provided by Jones and Pujado (2006) are assumed to evaluate the TBP of the blend. From Blend TBP appropriate vol % is evaluated for various pseudo components. The pseudo-component range is the same as provided by Jones and Pujado (2006).

Psuedo-components, vol% distribution and pseudo-component flow rates and total flow rate in the premium gasoline stream are presented as follows:

		LSR		Reformate		Light Cracked Naphtha		Hy Cracked Naphtha		Alkylate		
Comp	Temp ^{o}F	%vol	BPCD	%vol	BPCD	%vol	BPCD	%vol	BPCD	%vol	BPCD	Total
1	45-100	20	111.34	6	64.23	5	16.97					192.53
2	125	15	83.50	4	42.82	5	16.97					143.29
3	150	15	83.50	7	74.93	19	64.48					222.92
4	175	30	167.01	8	85.64	14	47.51			2	15.12	315.27
5	200	10	55.67	8	85.64	15	50.90			8	60.47	252.68
6	250	10	55.67	17	181.98	19	64.48			2	15.12	317.25
7	275			20	214.10	8	27.15			30	226.76	468.01
8	300			12	128.46	5	16.97	5	29.28	45	340.14	514.84
9	350			12	128.46	10	33.94	42	245.92	13	98.26	506.58
10	400			3	32.11			43	251.77			283.89
11	475			3	32.11			10	58.55			90.67
Total			556.69	100	1070.48	100	339.36		585.52	100	755.87	3307.92

Eventually, from the obtained TBP the volume % of various pseudo-components are evaluated. For each pseudo-component based on its mid boiling point, its molecular weight and vapor pressure are evaluated from Maxwell (1950) (Please refer to refinery property estimation for the same). Here, we assume the data provided by Jones and Pujado (2006) to be appropriate to represent the case. Then mole % data is evaluated. Then for butane whose molecular weight is 58, vapor pressure @ 100 ^{o}F is 22 psia, the equilibrium constant (K) is evaluated as vapor pressure divided by the desired RVP i.e., 7. This is carried out for all pseudo-components to evaluate the mole fraction of butane that will agree to the expression:

$$\sum x_i = \sum y_i$$

Where
$y_i = K_i x_i$. The obtained data is presented in the following table:

Comp		lbs/gal	lb/lbmol	VP	vol fr	wt fac	mol fac	mol fr	K=VP/7	mole liq	mol gas
1	75	5.96	72	22	0.0582	0.3469	0.0048	0.0770	3.1429	0.0770	0.2420
2	115	6.03	73	11	0.0433	0.2612	0.0036	0.0572	1.5714	0.0572	0.0899
3	135	6.09	75	6.2	0.0674	0.4104	0.0055	0.0875	0.8857	0.0875	0.0775
4	155	6.17	81	5	0.0953	0.5881	0.0073	0.1160	0.7143	0.1160	0.0829
5	185	6.25	90	2.5	0.0764	0.4774	0.0053	0.0848	0.3571	0.0848	0.0303
6	225	6.39	100	1.8	0.0959	0.6128	0.0061	0.0979	0.2571	0.0979	0.0252
7	260	6.49	110	1.8	0.1415	0.9182	0.0083	0.1334	0.2571	0.1334	0.0343
8	285	6.56	119	1.8	0.1556	1.0210	0.0086	0.1371	0.2571	0.1371	0.0353
9	315	6.64	128	1.8	0.1531	1.0169	0.0079	0.1270	0.2571	0.1270	0.0326
10	375	6.83	148	1.8	0.0858	0.5862	0.0040	0.0633	0.2571	0.0633	0.0163

11	420	6.95	162	1.8	0.0274	0.1905	0.0012	0.0188	0.2571	0.0188	0.0048
12	Butanes			60					8.5714	0.0435	0.3724
										1.0435	1.0435

For a butane mole liquid of 0.0435, the vapor and liquid sum of mole values were found to be equal (about 1.0435). Next, the evaluated mole liquid is back calculated to evaluate the vol% of the butanes to be added to the gasoline. These are summarized as follows:

	Mole liq	wt	vol	vol %
Butane	0.043452	2.520223	0.5245002	3.17745
1	0.077003	5.54424	0.9302415	5.635453
2	0.057187	4.174668	0.6923164	4.19409
3	0.087455	6.559115	1.0770304	6.524708
4	0.116031	9.398514	1.52326	9.227991
5	0.084781	7.630288	1.2208461	7.395951
6	0.097946	9.79457	1.5327965	9.285763
7	0.133411	14.67523	2.2612059	13.69851
8	0.137126	16.31805	2.4875074	15.06945
9	0.126967	16.25179	2.4475584	14.82744
10	0.063299	9.36822	1.3716281	8.309396
11	0.018793	3.044527	0.4380615	2.6538
Total			16.506952	100

From the above table, it can be observed that the vol% of butanes is 3.17745.

Therefore, butanes to be added to the premium gasoline product = 3.17745/(100-3.17745) x 3307.92 = 108.55 bbl/day.

Therefore, total premium gasoline product = 3307.92 + 108.55 = 3416.42 bbl/day.

Next, we proceed for similar calculations for regular gasoline. IN due course of calculations it was observed that the total vapor moles were higher than the total moles liquid. Therefore, no butane is added to the regular gasoline product.

All calculations are summarized as follows:

Firstly, the volumetric flows to produce the desired product and its RON is presented.

Stream	Flow (bbl/day)	Ron factor	RON product
LC Nap	849.4247	79845.92	
Reformate	2194.726	219472.6	
LSR Nap	2766.733	190904.5	
Total	5810.883	490223.1	84.36

Secondly, the volumetric flow rate of the pseudo-components and total flow are summarized:

		LSR		Reformate		Light Cracked Naphtha		
Comp	Temp ^{o}F	%vol	BPCD	%vol	BPCD	%vol	BPCD	Total
1	45-100	20	553.34	6	131.68	5	42.47	727.49
2	125	15	415.01	4	87.79	5	42.47	545.27
3	150	15	415.01	7	153.63	19	161.39	730.03
4	175	30	830.01	8	175.58	14	118.92	1124.51
5	200	10	276.67	8	175.58	15	127.41	579.66
6	250	10	276.67	17	373.10	19	161.39	811.16
7	275			20	438.95	8	67.95	506.90
8	300			12	263.37	5	42.47	305.84
9	350			12	263.37	10	84.94	348.31
10	400			3	65.84			65.84
11	475			3	65.84			65.84
Total			2766.7	100	2194.73	100	849.42	5810.85

Finally, the total mole % liquid and gas are summarized for all components to observe that the mole percent total gas is higher than the mole % total liquid i.e.,

Comp		lbs/gal	lb/lbmol	VP	vol fr	wt fac	mol fac	mol fr	K=VP/7	mole liq	mol gas
1	75	5.96	72	22	0.1252	0.7462	0.0104	0.1458	3.1429	0.1458	0.4583
2	115	6.03	73	11	0.0938	0.5658	0.0078	0.1091	1.5714	0.1091	0.1714
3	135	6.09	75	6.2	0.1256	0.7651	0.0102	0.1435	0.8857	0.1435	0.1271
4	155	6.17	81	5	0.1935	1.1940	0.0147	0.2074	0.7143	0.2074	0.1482
5	185	6.25	90	2.5	0.0998	0.6235	0.0069	0.0975	0.3571	0.0975	0.0348
6	225	6.39	100	1.8	0.1396	0.8920	0.0089	0.1255	0.2571	0.1255	0.0323
7	260	6.49	110	1.8	0.0872	0.5661	0.0051	0.0724	0.2571	0.0724	0.0186
8	285	6.56	119	1.8	0.0526	0.3453	0.0029	0.0408	0.2571	0.0408	0.0105
9	315	6.64	128	1.8	0.0599	0.3980	0.0031	0.0438	0.2571	0.0438	0.0113
10	375	6.83	148	1.8	0.0113	0.0774	0.0005	0.0074	0.2571	0.0074	0.0019
11	420	6.95	162	1.8	0.0113	0.0787	0.0005	0.0068	0.2571	0.0068	0.0018
Total										1	1.0161

Therefore, no extra butanes are added to this product.

Regular grade gasoline product flow rate = 5810.85 bbl/day.

3.17 Mass balances across the LPG, Gasoil and fuel oil pools

The mass balances across the LPG, Gasoil and fuel oil pools is now a relatively easy exercise with all the flow rates, SG and sulfur content of various streams available. Where data is missing, Jones and Pujado (2006) data is assumed. Other data such as pour point are optionally checked to meet the product requirements. All product specifications are chosen to be the same as presented by Jones and Pujado (2006).

Q 2.14: For the Saudi heavy crude oil processing scheme, conduct the mass balances across the LPG, Gasoil and fuel oil pools to evaluate the final product flow rates. Regard all relevant data from Jones and Pujado (2006) book to be appropriate.

Solution:

LPG pool

C3 from debutanizer	382.4885	Bbl/day
C3 from alkylator	81.33507	Bbl/day
Total C3 LPG	463.8236	
C4 from depropanizer		
Total Butanes	1044.996	Bbl/day
Butanes to Alkylation	472.9331	Bbl/day
Balance Butanes	572.0628	Bbl/day
Butanes to Gasoline Pool	108.5569	Bbl/day
Total C4 LPG	463.5058	Bbl/day

Fuel oil pool

For this pool, assume 50% vol of LVGO enters Marine diesel product. Also, assume, 80% of thermal cracker residue will enter heavy fuel oil product. The balance thermal cracker residue will mix with FCC slurry to produce the high sulfur bunker fuel.

To proceed further for calculations, firstly, the available intermediate streams i.e., LVGO, TC, Slurry are summarized.

Stream	Bbl/day	SG	SUL (wt%)
LVGO	2223.053	0.888819	0.43
TC Res	7576.406	1.0057	2.666
Slurry	186.7694	0.962585	1

The specifications of various products is summarized as follows

Product Stream	SG	Sul	Pour point
Marine Diesel	>0.83	<1%	< 45 °F
Heavy fuel oil	<1	<5	<65
Bunker fuel	<1.05	<6.22	-

To meet these requirements, the following procedure is adopted:

Marine Diesel: First ensure the SG is satisified and then look for Sulfur content. Check that both are satisfied and adjust the TC Residue flow rate accordingly.

Heavy fuel oil: Adjust the LVGO flow rate such that all constraints are satisfied.

Bunker fuel oil: Check whether balance flow rates are such that all product specification constraints are satisfied.

For the chosen example, the following is the jist of the calculations

Marine diesel:

Stream	Flow	SG	SUL (wt%)	Product SUL (wt %)	Product SG
LVGO	1111.526	0.888819	0.43	0.999755	0.915944
TC Res	335.9033	1.0057	2.666		

All conditions are satisfied. Therefore, LVGO flow rate to this blend will be 335.9 bbl/day

Next we check the pour point using pour point blending index data. All other data are assumed from Jones and Pujado (2006).

		vol fr	50 % ASTM	A x B	PP oF	Index	Blend
LVGO	1111.526	0.767931	700	537.5518	15	5	3.839656
TC Res	335.9033	0.232069	920	213.5033	65	22	5.105514
Total		1		751.0551			8.94517

For the blend with a ASTM 50 % of 751 °F and pour point index of 8.94 (assumed as 9), the pour point from pour point index graph presented by Maxwell (1950) (please refer to Refinery Property Estimation section for detailed information) is 28.5 °F which is lower than 45 °F, the desired specification.

Fuel oil: We only check sulfur wt % and SG of the product as pour point will be satisfied (according to Jones and Pujado (2006)).

Flow	Bbl/day	SG	SUL (wt%)	Product SUL (wt%)	Product SG (wt %)
TC Res	6061.125	1.0057	2.666	2.562258	0.999405
LVGO	345	0.888819	0.5		

IT can be observed that the product SG and sulfur content are satisfied.

Next the bunker fuel stream is checked for its product specifications. These are summarized as follows:

Stream	Bbl/day	SG	SUL (wt%)	Product SUL (wt%)	Product SG (wt %)
TC Res	1179.378	1.0057	2.666	2.446716	0.999806
FCC Slurry	186.7694	0.962585	1		

Here as well as product specifications are satisfied. Therefore, all product flow rates are confirmed to be acceptable.

Gas oil Pool

The following constraints are required to be satisfied for the product:

Stream	SG	SUL (wt%)
Auto Diesel	0.83 – 0.86	<0.35
Gas oil	0.84 – 0.86	< 0.3

For these specifications, assume that

a) All left over LVGO will mix with cycle oil to produce Auto diesel
b) Mix appropriate amount of kerosene with Light Cycle oil (FCC product) to produce gas oil.

The product flow rates for these streams are presented as follows for Autodiesel and gas oil. It can be observed that all product specifications are satisfied.

Stream	Bbl/day	SG	SUL (wt%)	Product SUL (wt%)	Product SG (wt %)
LVGO lef	766.5264	0.888819	0.43	0.245092	0.843402
Des LGO	4297.721	0.835301	0.21		

Stream	Bbl/day	SG	SUL (wt%)	Product SUL (wt%)	Product SG (wt %)
LCO	632.61	0.946488	0.54	0.299936	0.874483
Kerosene	606.6722	0.7994	0.00355		

Other than the above product specifications, other specifications such as ASTM 50 %, viscosity needs to be also checked. Since we don't have all these data at hand, we did not check them. We assume that they are satisfied.

Eventually, all products produced are summarized in the following table:

Total C3 LPG	463.8236	BPCD
Total C4 LPG	463.5058	BPCD
Kerosene product	1748.021	BPCD
Premium Gasoline product	3416.48	BPCD
Regular Gasoline product	5810.883	BPCD
Auto Diesel	5064.247	BPCD
Gas oil	1519.305	BPCD
Marine Diesel	1447.43	BPCD
Heavy fuel oil	6406.125	BPCD
Bunker Fuel	1366.147	BPCD

With this the mass balance calculation for the refinery block diagram is completed.

3.18 Summary

The following conclusions are applicable for the refinery mass balances:

a) Refinery mass balances are very important for a refinery engineer. These calculations enable him to understand the complexities involved in the distribution as well as processing of mass across the refinery.
b) At times, it is possible that we violate from the conventional processing schemes and purities (As observed in the case of the hydrotreater). For those circumstances, first hand information shall be obtained from the refinery to correct the mass balances.
c) Significant amount of data (such as intermediate stream TBP, SG and sulfur content) in these calculations is assumed. This shall not be the case for an operating refinery. Therefore, mass balances when conducted for an operating refinery will significantly enhance the confidence levels of the refinery engineer in operation.

All in all refinery mass balances is a fruitful exercise on which a refinery process engineer shall have expertise.

4. Design of Crude Distillation Column

4.1 Introduction

A crude distillation unit (CDU) is very common in petroleum refineries. Crude distillation unit involves complex stream interactions with various sections of the main column that is supplemented with secondary columns. Of late, the utilization of live steam is prominent in petroleum refineries. Before we attempt to deliberate upon various design aspects of crude distillation column, we first orient towards the technological knowhow of the CDU. The following itemized descriptions suits the operation of most modern CDUs.

- Crude oil is pumped from storage and is heated using hot overhead and product side streams using a heat exchanger network (HEN). The HEN enables the crude to achieve a temperature of about 200 - 250 ^{o}F
- Eventually, the pre-heated crude oil is injected to remove salt in a desalter drum which does not remove any organic chlorides and removes dissolved salt.
- Dissolved salt in the crude is removed using electrostatic precipitation as salt water. The salt water is sent to sour water stripper, cleaned and sent to oily waste sewage disposal.
- Eventually crude enters a surge drum and some light ends entrained with water are flashed off the drum. The removed light ends are directly fed to the flash zone of the main column of the CDU.
- A second HEN would energize the crude oil to higher temperatures after which the crude oil enters a furnace and is heated to a temperature that will vaporize distillate products in the crude tower.
- Crude oil is often heated to vaporize about 5 % more than required for the distillate streams. This is called **overflash and this ensures good reflux streams in the tower.**
- The heated crude then enters the fractionation tower in a lower section called flash zone. The unvaporized portion of the crude oil leaves the bottom of the tower via a stream stripper section. The distillate vapors move up.
- Distillate products from the main column are removed from selected trays. These are called **Draw off trays**. The streams are called draw off streams. These streams are steam stripped and sent to storage.
- From the tower top of the main column, full range naphtha (both light and heavy) will leave as a vapor. Eventually, the vapor will be condensed and separated in a phase separator. The separated naphtha product will be partially sent for reflux and the balance sent as reflux stream from the overhead drum.
- In addition, pump around units are included at the LGO draw off and HGO draw off. A pump around involves removing a hot side stream cool it and return it back to the column at a section above the draw off tray. The pump around is an internal condenser that takes out heat of that section and ensures reflux below that section.

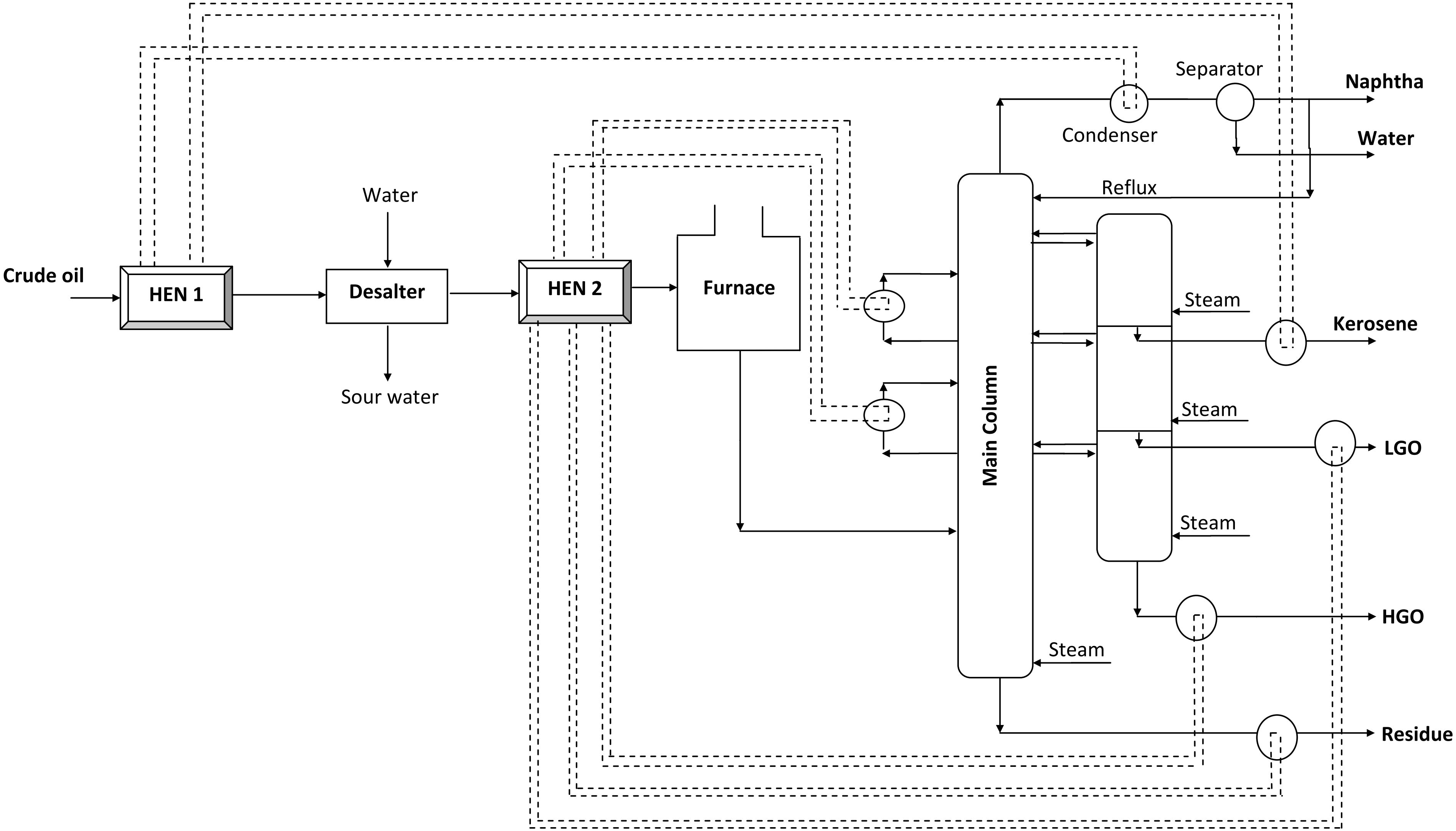


Figure 4.1: A Conceptual diagram of the crude distillation unit (CDU) along with heat exchanger networks (HEN).

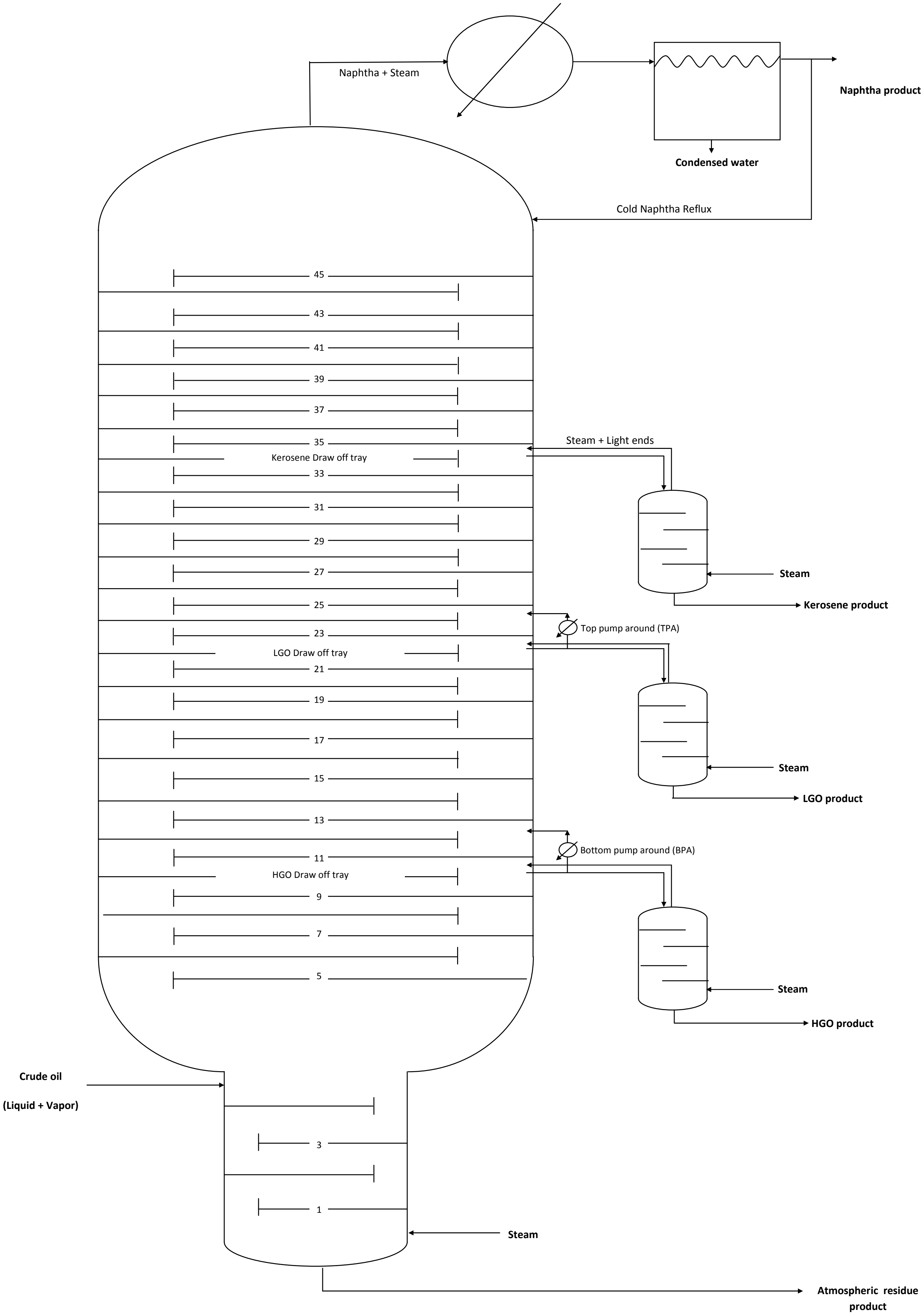


Figure 4.2: Design architecture of main and secondary columns of the CDU.

- Side stream distillate products are kerosene (Jet fuel), light gas oil (diesel) and heavy gas oil. These are stripped free of entrained light ends in separate stripping towers (called secondary columns). About 4 – 6 trays are required in these secondary columns. **The purpose of secondary columns is to strip the side stream distillate products from entrained light ends.** In these columns, steam is injected below the bottom tray which moves up the tower and leaves at the secondary column top along with light ends stripped out. The stripped steam (with light ends) is allowed to enter the main column just above the side stream draw off tray. Often, side stream stripper units (secondary columns) are stacked above one another in a single column (with distributed liquid and vapor traffic) so as to ensure **free flow of side stream product from the draw off tray to the secondary column.**
- In certain circumstances such as kerosene stream sent to jet fuel blending, stripping is effected using a re-boiler.
- The residue product leaves the flash zone at the bottom of the main column. It is stripped free of light ends using steam injected at the bottom of the main column.
- Live steam is pretty useful in the CDU operation. While steam enabled the reduction of installed costs of reboiler from cost perspective, it enables the **flashing of the streams at a reduced partial pressure and therefore contributes significantly for the removal of light ends throughout the main and secondary column.**

A conceptual diagram of the CDU is presented in Figure 4.1.

4.2 Architecture of Main and Secondary Columns

Main and secondary column interaction for the processing of crude oil to produce products is a trivial task. After significant amount of simulation as well as pilot plant based studies, the design architecture of the CDU along with secondary columns evolves. Finer trade offs are associated with the design of main and secondary columns involving effective heat integration and reduction of the total annualized cost. Therefore, it can be easily observed that the optimal design of the main and secondary columns is based on a trial and error approach **in which the design architecture is fixed first and eventually simulation is carried out for the chosen design calculations. Subsequently, further changes in the design architecture is carried out using either trial and error approach or an algorithmic approach to yield the best design architecture.** From literature (Jones and Pujado (2006)), a typical design architecture of the main and secondary columns along with their interaction is presented in Figure 4.2. The typical design architecture The following features summarize the design architecture of the complex distillation arrangement:

a) The main column consists of 45 trays and the secondary columns (side strippers) consist of 4 trays each. Three side strippers are used to strip the light ends from kerosene, LGO and HGO products.
b) The main column has two sections that are distinguished with respect to a flash zone. The flash zone is where the crude oil partially vaporized is fed to the main column. There are about 4 trays below the flash zone and 41 trays above the flash zone of the main column. The bottom most tray (residue stripping tray) is numbered as 1 and the top tower tray is numbered as 45.

Trays 1 to 4 process the atmospheric residue portion of the crude in the section below the flash zone.

c) Trays 5 to 10 (6 trays above the flash zone) process the HGO product portion of the crude. From tray 10, HGO draw off product is taken out (as liquid) and enters the HGO side stripper unit. From tray 10 as well, the liquid stream is drawn and sent to tray 12 via a bottom pump around unit that enables cooling of the liquid stream. The steam + light ends from the HGO side stripper enter tray 11 of the main column.

d) Trays 13 to 22 (10 trays above the HGO processing zone) process the LGO product portion of the crude. From tray 22, LGO draw off product is taken (as liquid) and sent to the LGO side stripper unit. Also, from tray 22, another liquid stream is taken out and sent to tray 24 via a top pump around unit (TPA) that enables cooling of the liquid stream. The steam + light ends from the LGO side stripper enter tray 23 of the main column.

e) Trays 24 to 34 (10 trays above the LGO processing zone) process the kerosene product portion of the crude. From tray 34, the kero draw off stream is taken and sent to the kerosene side stripper unit. The steam + light ends of the kerosene side stripper enter tray 35.

f) Trays 34 to 45 (12 trays above the Kerosene processing zone) process the naphtha product portion of the crude. **It is interesting to note that tray 34 is regarded as a tray processing both LGO as well as naphtha processing zone. This is because there is no pump around associated to the tray 34. Where pump around is associated, that tray is often ignored in counting, as it affects to a large extent the tray hydraulics and contributes less towards the separation of the components.**

g) It is interesting to note that steam enters main column at trays 1, 11, 23, 35 and therefore is present along with the vapor stream along with the hydrocarbons. Therefore, **steam balances throughout the column are very important.**

h) The cold naphtha stream obtained from the phase separator is sent back to the main column as reflux stream.

In summary we have the following important trays that contribute towards the main column hydraulics:

Tray No.	Description
1	Atmospheric Residue draw off
10	HGO Draw off, BPA draw off
11	Vapor from HGO SS enters the tray
12	BPA return stream
22	LGO drawoff, TPA drawoff
23	Vapor from LGO SS enters the tray
24	TPA return stream
34	Kerosene draw off
35	Vapours from Kerosene SS enters the tray
45	Vapours with Naphtha + Gas + stream leave to enter the condenser

4.3 Design aspects of the CDU

In the due course of the design process of the CDU, the following issues need to be taken care (in similarity to a conventional distillation column):

a) **Fix pressures across the column:** Typically the pressures across the column are assumed with approximate assumptions of pressure drops across the reflux drum, condenser and trays. Typically, main column pressures are taken as follows:
   Condenser reflux drum = 5 psig
   Pressure drop across the condensers = 7 psi
   Pressure drop over all trays = 10 psi (@ 0.25 psi per each tray for 40 trays)
   Pressure at flash zone = 22 psig

b) **Conduct mass balances across the CDU as well as flash zone:** The mass balance along with relevant properties such as API, K (characterization factor), molecular weight of various streams is carried out using first principles of refinery property estimation (as presented previously). The mass balances are desired for the overall CDU as well as the flash zone. **This is due to the fact that flash zone mass balance eventually leads to energy balance across the flash zone and determination of residue product temperature. Eventually, all other sections of the main column are solvable.** The rule of thumb for these calculations is to define the product specifications using cut range and the concept of ASTM gap. Typically, the following cut range is assumed
   Upto 375 °F: Gas + Naphtha product
   375 – 480 °F: Kerosene product
   480 – 610 °F: LGO product
   610 – 690 °F: HGO product
   690+ °F: Atmospheric residue.
   For fractionation efficiency, further product specifications are provided in the form of ASTM gaps. The ASTM gap refers to the separation efficiency of two adjacent products. It is defined as the difference in the 5 % ASTM temperature of a heavier product and 95 % ASTM temperature of a lighter product. Typically, the following ASTM gaps are achievable:
   Naphtha – Kerosene: 25 °F
   Kerosene – LGO: -10 °F
   LGO – HGO: -35 °F
   It is further interesting to note that ASTM gaps could be +ve or –ve. A +ve gap indicates 100 % separation of fractions where as a negative gap indicates difficulty to completely separate the fractions. Therefore, it can be seen that naphtha can be fully separated from the kerosene fraction but not for kerosene to get separated from LGO and LGO from HGO fractions.
   These data are used to estimate the TBPs of the products using which the average product properties are estimated. Eventually, both overall mass balance and flash zone mass balance are carried out.

c) **Estimate steam requirements in various sections:** From pilot plant data or correlations, the steam required to produce a required product mass flow rate is available. Using this steam flow rates in various sections of the CDU (main tower bottom, side stream strippers) are estimated.

d) **Determine flash zone temperature:** The flash zone temperature is estimated using the EFV curve of the crude for assumed overflash conditions and partial pressure of the hydrocarbons. The partial pressure concept is extremely important in mass and energy balances carried out in various sections of the CDU as steam enthalpy is a function of the partial pressure of steam that exists in the chosen zone of calculation.

e) **Estimate residue temperature:** Using flash zone temperature and heat balance across the flash zone, estimate the residue temperature.

f) **Estimate draw off temperatures:** Typically correlations exist (such as Packie's correlation) between side stream draw off temperatures and amount of light ends stripped. These correlations are used along with relevant assumptions to estimate the draw off temperatures.

g) **Estimate side stream stripper product temperatures:** The products kerosene, LGO and HGO temperatures are estimated from the energy balances for these products in respective sections of the CDU.

h) **Estimate tower top temperature:** The tower top temperature is estimated using a bubble point calculation across the condenser.

i) **Conduct overall tower energy balance and estimate condenser + BPA + TPA duties:** From overall tower energy balance, total energy loss requirements across the CDU can be estimated.

j) **Estimate condenser duties:** From the top section energy balance (with known top section temperature), estimate the condenser duty. From this estimate the total BPA + TPA heat duty.

k) **Estimate BPA duty:** Using energy balance across the chosen section of the CDU and the concept of fractionation efficiency, estimate the BPA duty. Eventually estimate the TPA duty.

l) **Establish column hydraulics:** At various important trays that were outlined previously where tray hydraulics are prominent, estimate total liquid and vapor flow rates (including steam). These data will be useful for diameter calculations.

m) **Determine column diameter at various sections:** Using estimated vapor and liquid flow rates at various trays, determine the column diameter using flooding correlations.

In the above procedure, the following issues are very relevant to observe:

a) The energy balance across the CDU is very important. We do about six energy balances across the CDU to estimate the tower temperatures, condenser duties, BPA duties etc. Therefore, careful selection of envelope to carry out energy balances is very important.

b) Energy balance requires the estimation of hydrocarbon enthalpy at various instances. From Maxwell (1950) correlations, hydrocarbon enthalpy requires characterization factor and temperature. Eventually, characterization factor estimation requires API value. IN summary, for each stream for which enthalpy data is required, prior knowledge of the API and K values is mandatory. Since K value is an estimate of MEABP, some assumptions are required for the same. Also, API of each intermediate stream is not known; therefore, average properties need to be assumed for calculations.

In the following sections, we deliberate upon each relevant design issue of the CDU to hierarchically estimate required design variables associated to the design and operation of the main and secondary columns of the CDU.

4.4 Mass balances across the CDU and flash zone

To conduct mass balances across the CDU and flash zone, product properties need to be estimated first. While previously we evaluated average product properties in refinery property estimation chapter using end point correlations, the same cannot be used here. This is because the products are more stringently defined using ASTM gaps and the products need to meet these specifications. Therefore, a modified procedure for refinery property estimation is adopted. In this procedure, first the TBP (and ASTM) of naphtha is determined using first principles associated to the TBP of the naphtha product. In other words, we use end point correlation and demister correlation and TBP cut range in the crude TBP assay to evaluate the TBP and ASTM of naphtha. Eventually, using ASTM gap data of naphtha – kerosene, the ASTM data of kerosene is determined using the probability chart. In similarity using ASTM gap data, the ASTM data of LGO and HGO are determined. Eventually, these ASTM data are converted to TBP data using Edmister correlation. Eventually using crude API data, the average API of the products (other than the residue) is estimated. For residue, API is determined from a volumetric balance. Approaches similar to those presented in the refinery property estimation chapter are adopted to estimate other properties such as molecular weight, characterization factor (K), MEABP etc. Below, we present two illustrative examples in sequence to elaborate upon a) the determination of product TBP and ASTM data and b) the average API data.

Q 4.1: For the Ecudaor crude stream whose TBP, sulfur and API assay are presented below, determine the TBP of the products emanating from the CDU unit using the concept of ASTM gaps.

Cumulative volume %	TBP ^oF	^oAPI	Sulfur content (wt %)
0	-30		
1	30		
1.5	60		
2	90	100	0.035
3	120	85	0.035
4	150	80	0.04
6.5	180	72	0.04
9	210	64	0.045
11.5	240	60	0.045
14	270	55	0.05
17.5	310	50	0.055
19	330	48	0.06
21	350	46	0.07
22.5	370	45	0.08

24.5	390	43	0.10
26	410	42	0.12
28	430	41	0.16
30	450	40	0.19
32	470	38	0.25
34	490	37	0.3
36	520	36	0.36
38.5	540	34	0.43
39.5	560	33	0.48
42.5	580	32	0.57
44.5	600	31	0.63
47	620	30	0.71
49	640	29.5	0.77
52	660	29	0.80
55	680	28	0.85
57	700	27	0.87
59	720	26.5	0.89
61	740	26	0.91
62.5	760	25.5	0.93
63.8	780	25	0.96
64.5		24.5	0.98
66		24	1.02
68		23.5	1.06
70		23	1.14

Assume the following cut range of the products on the crude TBP

Naphtha: -30 to 310 ^oF

Kerosene: 310 to 475 ^oF

LGO: 475 to 585 ^oF

HGO: 585 to 680 ^oF

Solution:

I) Naphtha product ASTM & TBP Data

Firstly, we evaluate the TBP data of Naphtha cut. For this we use end point correlation and TBP 50 % of the Naphtha cut to obtain ASTM 50 % and ASTM 100 % data. Using these two data points and using end point correlation, the ASTM data of naphtha cut is obtained. Once again using Edmister correlation we obtain the ASTM of the naphtha product. The hierarchy of these steps along with obtained data are presented as follows.

TBP data of naphtha cut

Cumulative volume %	Differential volume %	TBP ^{o}F
0	0	-30
1	5.714	30
1.5	8.571	60
2	11.428	90
3	17.143	120
4	22.857	150
6.5	37.142	180
9	51.428	210
11.5	65.714	240
14	80	270
17.5	100	310

TBP 50 % of the Naphtha cut on the crude assay = 208 ^{o}F. From Edmister correlation, ASTM 50 % = 215 ^{o}F

Naphtha cut end point = 310 ^{o}F. From end point correlation, ASTM end point = 310 – 1 = 309 ^{o}F

From Probability chart and Edmister correlation, the naphtha product ASTM and TBP are obtained as follows:

ASTM data of Naphtha

Cumulative volume %	TBP ^{o}F
0	152
10	178
30	198
50	215
70	230
90	255
100	309

Product TBP (from Edmister correlation)

Vol %	ΔT_{ASTM} (^{0}F)	ΔT_{TBP} (^{0}F)	T_{TBP} (^{0}F)
0	26	49	92
10	20	37	141
30	17	30	178
50	NR	NR	208

70	15	24	239
90	25	34	273
100	54	60	333

II) Kerosene product TBP & ASTM data

ASTM gap of naphtha –kerosene = 25 °F

From ASTM data of Naphtha product, ASTM 95 % of naphtha = 295 °F

Therefore, ASTM 5 % of kerosene = 295 + 25 = 320 °F

Kerosene product end point on the crude = 385 °F. For this value from end point correlation, ASTM end point of kerosene cut = 456 °F

From Probability chart and Edmister correlation, the kerosene product ASTM and TBP are obtained as follows:

ASTM data of kerosene product

Vol %	T_{ASTM} (^{0}F)
0	305
10	330
30	350
50	367
70	380
90	405
100	456

Kerosene product TBP

Vol %	ΔT_{ASTM} (^{0}F)	ΔT_{TBP} (^{0}F)	T_{TBP} (^{0}F)
0	25	48	249
10	20	38	297
30	17	30	335
50	13	NR	365
70	13	22	387
90	25	34	421
100	51.5	57	478

III) LGO product TBP & ASTM data

ASTM gap of kerosene – LGO = -10 ^oF

From ASTM data of Kerosene product, ASTM 95 % of Kerosene = 445 ^oF

Therefore, ASTM 5 % of LGO = 445 – 10 = 435 ^oF

LGO product end point on the crude = 585 ^oF. For this value from end point correlation, ASTM end point of LGO product = 585 – 10 = 575 ^oF.

ASTM data of LGO (from Probability chart)

Vol %	T_{ASTM} (^0F)
0	458
10	480
30	492
50	505
70	520
90	545
100	575

TBP data of LGO product (from Edmister correlation)

Vol %	ΔT_{ASTM} (^0F)	ΔT_{TBP} (^0F)	T_{TBP} (^0F)
0	22	44	407
10	12	25	451
30	13	24	476
50	NR	NR	500
70	15	24	524
90	25	34	558
100	30	34	592

IV) HGO product TBP & ASTM data

ASTM gap between LGO and HGO products = -35 °F

From LGO ASTM data, ASTM 95 % of LGO product = 573 °F

Therefore, ASTM 5 % of the HGO product = 573 – 35 = 538 °F.

TBP end point of HGO on crude assay = 680 °F. 90% TBP point of HGO on the crude assay = 670 °F

Using the 90% correlation available in the end point correlation data, 90% ASTM point = 627 °F.

From probability chart, HGO product ASTM data is

Vol %	T_{ASTM} (^{0}F)
0	522
10	552
30	571
50	575
70	600
90	627
100	675

From Edmister correlation, the HGO product TBP is

Vol %	ΔT_{ASTM} (^{0}F)	ΔT_{TBP} (^{0}F)	T_{TBP} (^{0}F)
0	29	55	466
10	19	37	521
30	4	8	558
50	-	-	566
70	25	38	604
90	27	37	642
100	48	53	695

Q 4.2: Using the product TBPs estimated in Q1, determine the average product properties (including residue product) such as API, molecular weight, characterization factor. Appropriate pseudo-component selection could be made on the TBP and API assay of the crude.

Solution:

For the crude, the following pseudo-component range along with mid vol, average API for each pseudo-component are summarized based on the TBP and API assay of the crude oil. Sulfur calculations are ignored in this section, as sulfur balance is not going to affect design calculations associated to the CDU.

Psuedo-component No.	TBP range on crude (^oF)	Mid Boiling point on crude ^oF	Mid vol % on crude	Mid ^oAPI
1	60 – 100	75	2.25	90
2	100 – 140	125	3.25	81
3	140 – 180	160	5.25	71
4	180 – 220	195	8.25	64
5	220 – 250	235	11	58
6	250 – 280	265	13.5	53
7	280 – 310	295	16.25	52
8	310 – 330	320	18.5	52
9	330 – 350	340	20.5	46
10	350 – 370	360	22.5	45
11	370 – 390	380	24.5	42
12	390 – 410	400	26.5	41
13	410 – 430	420	28.5	40
14	430 – 450	440	30.5	39
15	450 – 480	470	32.5	37
16	480 -500	490	34.5	37
17	500 – 520	510	36.5	36
18	520 – 540	530	38.5	34
19	540 – 560	550	40.5	33
20	560 – 610	595	44	32
21	610 – 630	620	47.5	31
22	630 – 650	640	49.5	30
23	650 – 670	660	51.5	29
24	670 – 700	685	54	27.5

Based on product TBPs evaluated previously, the corresponding volume % of various pseudo-components in product TBPs is obtained graphically. The same is summarized in the next table.

Psuedo-component No.	Naphtha product differential vol %	Kerosene product differential vol%	LGO product differential vol %	HGO product differential vol %
1	1.8			
2	7.2			
3	20			
4	28			
5	19			
6	15	5.5		
7	3	10		
8	3	11		
9		13		
10		14.5		
11		18		
12		13		
13		7	3	
14		4	5	
15		4	24	
16			17	5
17			17	4.8
18			13	8.3
19			10	14.4
20			10	40.5
21				12
22				7.1
23				3.9
24				4
Total	97*	100	100	100

*For the very first component API gravity is not available and hence volume% is not reported.

For these pseudo-components, their molecular weight as a function of their API and mid boiling point is determined from Maxwell's correlations presented in Refinery Property Estimation chapter. The same is presented in the next table.

Psuedo-component No.	Mid Boiling point on crude ^{o}F	Mid ^{o}API	Molecular weight
1	75	90	70
2	125	81	80
3	160	71	87
4	195	64	94
5	235	58	108

6	265	53	110
7	295	52	115
8	320	52	118
9	340	46	141
10	360	45	144
11	380	42	151
12	400	41	159
13	420	40	165
14	440	39	175
15	470	37	191
16	490	37	197
17	510	36	203
18	530	34	218
19	550	33	230
20	595	32	254
21	620	31	270
22	640	30	283
23	660	29	291
24	685	27.5	300

Since molecular weight needs to be accommodated on a mass basis and since product TBPs do not match with the crude TBPS due to anamolies associated in the volumetric balance pointed earlier, all molecular weight calculations are carried out using crude data only. With the background information ready for calculations, we next present the calculations involved for the estimation of average properties of the products namely API, characterization factor and molecular weight, as only these are required for the design calculations of the CDU.

I) Average Naphtha product properties

i) API and specific gravity

Psuedo-component No.	Vol % distilled [A]	Mid pt API	S.G. [B]
1	1.8	90	0.638826
2	7.2	81	0.665882
3	20	71	0.698765
4	28	64	0.723785
5	19	58	0.746702
6	15	53	0.766938
7	3	52	0.771117
8	3	52	0.771117

Average specific gravity of the naphtha product $= \frac{\sum_{i=1}^{N_{PC}}[A_i][B_i]}{\sum_{i=1}^{N_{PC}}[A_i]} = 0.702993$

ii) Characterization factor (K)

Slope of the Naphtha product TBP = $(T_{70} - T_{10})/60$ = (239-141)/60 = 1.63333

The volume average boiling point $T_{VABP} = (T_0 + 4T_{50} + T_{100})/6$ = (92 + 4x298+333)/6 = 269.5 ^oF

From Maxwell's correlation, MEABP = 269.5 – 3.2 = 266.3 ^oF

Naphtha product characterization factor which is a function of the MEABP and API is obtained from Maxwell's correlation as 12.5.

iii) Molecular weight

Psuedocomponent No.	Vol % on crude	S.G	Weight Factor	Molecular Weight	Mole Factor
1	0.5	0.638826	0.319413	70	0.004563
2	1.5	0.665882	0.998824	80	0.012485
3	2.5	0.698765	1.746914	87	0.020079
4	3.5	0.723785	2.533248	94	0.026949
5	2	0.746702	1.493404	108	0.013828
6	3	0.766938	2.300813	110	0.020916
7	2.5	0.771117	1.927793	115	0.016763
8	2	0.771117	1.542234	118	0.01307
Total			12.86264		0.128655

Average molecular weight = 12.86264/0.128655 = 99.97799 = 100 (approximately).

II) Average Kerosene product properties

Psuedo-component No.	Vol % distilled [A]	Mid pt API	S.G. [B]
6	5.5	53	0.766938
7	10	52	0.771117
8	11	52	0.788301
9	13	46	0.797183
10	14.5	45	0.8017
11	18	42	0.815562
12	13	41	0.82029
13	7	40	0.825073

14	4	39	0.829912
15	4	37	0.839763

Average specific gravity of the kerosene product $= \frac{\sum_{i=1}^{N_{PC}}[A_i][B_i]}{\sum_{i=1}^{N_{PC}}[A_i]} = 0.803868$

ii) Characterization factor (K)

Slope of the kerosene product TBP = $(T_{70} - T_{10})/60$ = (387-297)/60 = 1.5

The volume average boiling point $T_{VABP} = (T_0 + 4T_{50} + T_{100})/6$ = (249+4x365+478)/6 = 364.5 ^oF

From Maxwell's correlation, MEABP = 364.5 – 0 = 364.5 ^oF

Kerosene product characterization factor which is a function of the MEABP and API is obtained from Maxwell's correlation as 11.85.

iii) Molecular weight

Psuedocomponent No.	Vol % on crude	S.G	Weight Factor	Molecular Weight	Mole Factor
6	3	0.82029	2.46087	110	0.022372
7	2.5	0.825073	2.062682	115	0.017936
8	2	0.825073	1.650146	118	0.013984
9	2	0.797183	1.594366	141	0.011308
10	2	0.8017	1.603399	144	0.011135
11	2	0.815562	1.631124	151	0.010802
12	2	0.82029	1.64058	159	0.010318
13	2	0.825073	1.650146	165	0.010001
14	2	0.829912	1.659824	175	0.009485
15	2	0.839763	1.679525	191	0.008793
Total	100		17.63266		0.126134

Average molecular weight = 17.632/0.126134 = 139.79

III) Average LGO product properties

i) Average specific gravity

Psuedo-component No.	Vol % distilled [A]	Mid pt API	S.G. [B]
13	3	40	0.825073
14	5	39	0.829912
15	24	37	0.839763
16	17	37	0.839763
17	17	36	0.844776
18	13	34	0.854985
19	10	33	0.860182
20	10	32	0.865443

Average specific gravity = 0.846336.

ii) Characterization factor

Slope of the LGO product TBP = $(T_{70} - T_{10})/60 = (524 - 451)/60 = 1.2167$

The volume average boiling point $T_{VABP} = (T_0 + 4T_{50} + T_{100})/6 = (407+4x500+592)/6 = 499.83\ ^oF$

From Maxwell's correlation, MEABP = $499.83 + 7 = 506.83\ ^oF$

LGO product characterization factor which is a function of the MEABP and API is obtained from Maxwell's correlation as 11.6.

iii) Molecular weight

Psuedocomponent No.	Vol % on crude	S.G	Weight Factor	Molecular Weight	Mole Factor
13	2	1.076046	2.152091	165	0.013043
14	2	1.076046	2.152091	175	0.012298
15	2	0.825073	1.650146	191	0.00864
16	2	0.839763	1.679525	197	0.008526
17	2	0.844776	1.689552	203	0.008323
18	2	0.854985	1.70997	218	0.007844
19	2	0.860182	1.720365	230	0.00748
20	5	0.865443	4.327217	254	0.017036

Average molecular weight = 17.08/0.08319 = 205.328

IV) Average HGO product properties

i) Average Specific gravity

Psuedo-component No.	Vol % distilled [A]	Mid pt API	S.G. [B]
16	5	37	0.839763
17	4.8	36	0.844776
18	8.3	34	0.854985
19	14.4	33	0.860182
20	40.5	32	0.865443
21	12	31	0.870769
22	7.1	30	0.876161
23	3.9	29	0.88162
24	4	27.5	0.88994

Average specific gravity = 0.864055

ii) Characterization factor

Slope of the HGO product TBP = $(T_{70} – T_{10})/60$ = 1.38

The volume average boiling point $T_{VABP} = (T_0 + 4T_{50} + T_{100})/6$ = 570.83 ^{o}F

From Maxwell's correlation, MEABP = 574.83 ^{o}F

HGO product characterization factor which is a function of the MEABP and API is obtained from Maxwell's correlation as 11.55.

iii) Molecular weight

Psuedo component No.	Vol % on crude	S.G	Weight Factor	Molecular Weight	Mole Factor
16	2	1.076046	2.152091	197	0.010924
17	2	0.611231	1.222462	203	0.006022
18	2	1.076046	2.152091	218	0.009872
19	2	1.076046	2.152091	230	0.009357
20	5	1.076046	5.380228	254	0.021182
21	2	0.870769	1.741538	270	0.00645
22	2	0.876161	1.752322	283	0.006192
23	2	0.88162	1.76324	291	0.006059

24	1	0.887147	0.887147	300	0.002957

Average molecular weight = 20.985/0.084841 = 247.3545

V) Average residue properties

i) Residue specific gravity

The residue specific gravity is determined from mass balance. Crude SG is 0.873996 or 0.874.

Vol % [A]	S.G [B]	[A] X [B]
17.5	0.703	12.3025
16	0.8039	12.8624
13	0.8463	11.0019
7	0.8641	6.0487
46.5	x	46.5x

From volumetric balance, (46.5x+42.2155)/100 = 0.874. From this expression, specific gravity of the residue x = 0.97171 or 0.972 approximately.

ii) Residue characterization factor

Assume MEABP of the residue as 910 ^oF. From residue API of 0.972 and Maxwell's correlation, residue characterization factor K = 14.1.

iii) Residue molecular weight

First determine molecular weight of the crude and from mass balance determine the molecular weight of the residue. This is also due to the reason that residue TBP is not known.

The characterization factor of the crude is determined as follows
Slope of TBP =-(T_{70} - T_{10})/60-=(870-220)/60 = 10.833

VABP = (T_{20} + T_{50}+ T_{80})/3 = 686.6667 ^oF
From Maxwell's correlation, MEABP of the crude = 686.6 – 68 = 618.6 ^oF
Crude SG = 0.874
From Maxwell's characterization factor correlation, crude characterization factor = 11.7.
From Maxwell's molecular weight correlation, crude molecular weight = 252.

Product	Vol % (A)	S.G (B)	Weight Factor	Molecular Weight	Mole Factor
Naphtha	17.5	0.703	12.3025	99.978	0.123052
Kerosene	16	0.8039	12.8624	139.794	0.09201
Light Gas Oil	13	0.8463	11.0019	205.328	0.053582
Heavy Gas Oil	7	0.86455	6.051867	247.35	0.024466
Residue	46.5	0.972	45.198	z	45.198/z

Average molecular weight of the crude = 87.4135/(45.198/z+0.2933). This is equal to 252.

Therefore, equating the above expression to 252, we get z = 843.577 which is the residue molecular weight.

In summary, the evaluated properties of the crude and products are as follows:

Stream	SG	K	Molecular weight
Crude	0.874	11.4	**252**
Naphtha	0.703	12.5	99.978
Kerosene	0.8039	11.85	139.794
Light Gas Oil	0.8463	11.6	205.328
Heavy Gas Oil	0.86455	11.55	247.35
Residue	0.972	14.1	843.5

4.4.1 CDU mass balance table

The overall mass balance across the CDU is carried out using the following procedure:

a) Fix a basis of the total crude flow rate. Typically it is about 30,000 to 50,000 barrels per day of crude oil processing in the CDU.
b) For the chosen crude flow rate, establish volumetric and mass balances with the evaluated specific gravity values of the products
c) Using molecular weights, establish the mole balances across the CDU.

4.4.2 Flash zone mass balance table

The crude stream fed to the flash zone in a partially vaporized state (3 -5 % of overflashing is usually done). The purpose of the over flashing is to enable the arrangement of the internal reflux for the control of the product quality. The following procedure is adopted for the flash zone mass balance.

a) Assume 3% overflash. For the 3 % overflash stream, estimate cumulative volume %, barrels/day, molecular weight, specific gravity etc. from crude assay and first principles of refinery property estimation.
b) Establish the total vapor variables such as flow rates, specific gravity etc using the known variables associated to the overflash and other products (Naphtha to HGO)
c) Estimate the residue variables such as flow rates, specific gravity without the overflash variables from total mass balance. These variables refer to the residue liquid stream that is bereft of condensable part (of the overflash).

We next present an illustrative example for the mass balances across the CDU and overflash zone.

Q 4.3: For the Ecudaor crude stream whose average product properties are estimated, conduct the mass balances across the CDU and the overflash zone. Consider a basis of 50,000 barrels/day of crude oil processing in the CDU.

Solution:

From average product properties, the CDU mass balances are summarized in the following table.

Stream	Range (°F)	Vol %	BPSD	GPH	S.G	Density (lbs/gal)	lbs/hr	Wt %	Mol. Wt	Mol/hr
Crude	-	100	50000	87500	0.874	7.2891	637798.6	100	252	2530.947
CDU Products										
Naphtha	IBP – 310	17.5	8750	15312.5	0.703	5.8629	89776.86	14.07	99.97	898.038
Kero	310 – 475	16	8000	14000	0.804	6.70426	93859.74	14.71	139.9	670.906
LGO	475 – 585	13	6500	11375	0.8463	7.05845	80289.88	12.58	205.3	391.0856
HGO	585 – 680	7	3500	6125	0.8645	7.21034	44163.4	6.92	247.35	178.55
Residue	680 +	46.5	23250	40687.5	0.9716	8.10345	329710.9	51.69	843.577	390.85

For the overflash stream,

% Overflash = 3%

This means that the cumulative volume% = 53.5 to 56.5
TBP range on the crude: 680 – 700 °F
Mid volume % = 55

Mid boiling point = 695 °F
API = 26. SG = 0.898413
Molecular weight = 316

Mass balances across the flash zone are presented as follows

Stream	Range (°F)	Vol %	BPSD	GPH	S.G	Density (lbs/gal)	lbs/hr	Wt %	Mol. Wt	Mol/hr
Crude	-	100	50000	87500	0.874	7.2891	637798.6	100	252	2530.947
Flash zone products										
Overflash	680 - 700	3	1500	2625	0.898413	7.492764	19668.51	3.083812	316	62.24
Products	IBP – 680	53.5	26750	46812.5	0.789067	6.580818	308064.6	48.30123	143.84	2141.66
Total Vapour	IBP - 700	56.5	28250	49437.5	0.794873	6.62924	327733.1	51.38504	148.71	2203.90
Residue	700 +	43.5	21750	38062.5	0.976765	8.14622	310065.5	48.61496	948.08	327.04
Total		100	50000	87500	0.873996	7.289127	637798.6	100	252	2530.947

4.5 Estimation of flash zone temperature

Flash zone temperature estimation is an important task in the design procedure of the CDU. This is due to the fact that flash zone temperature will enable heat balance across the flash zone to yield the residue product temperature. The Crude EFV curve is the starting point for the flash zone temperature. Since live steam is present in the main column (along with hydrocarbon vapors), the hydrocarbon partial pressure needs to be estimated and using the same, the flash zone temperature shall be estimated at the desired hydrocarbon partial pressure. The procedure is elaborated as follows:

a) Draw the equilibrium flash vaporization curve of the crude oil at 1 atm using Maxwell's correlations summarized in refinery property estimation chapter.
b) Assume 1.2 lbs of steam used to produce one gallon of residue. With this assumption, estimate the steam flow rate in lbmol/hr.
c) Assume flash zone pressure = 40 psia.
d) Estimate the partial pressure (PP_{HC}) of the hydrocarbon using the expression
PP_{HC}= moles HC vapor/(moles HC vapor + moles steam) x flash zone pressure
e) Adjust the EFV curve to the desired partial pressure. In other words, we need to draw a new EFV curve. The new EFV curve FRL will be the same as that exists at 1 atm. Therefore, all we require is a point (as the slope is known) to draw the new EFV line. The point is estimated using vapor pressure curve data presented in Table 2.18 (follow solved example 2.9 for further delibeations on calculations). The existing EFV data point is taken at 50 % and 14.7 psia. The new vapor pressure is determined from Table 2.18 at the hydrocarbon partial pressure for 50 % point.

f) For the new EFV curve draw the line with known co-ordinates and slope of the FRL
g) On the new EFV curve, find the flash zone temperature as that temperature that corresponds to the overflash cumulative volume %.

An example is presented next to elaborate the above steps.

Q 4.4: For the Ecuador crude stream, determine the flash zone temperature in the main column.

Solution: We first determine the slope of the EFV curv

Slope of TBP = S_{TBP} = $(T_{TBP,70} - T_{TBP,10}) / 60 = (890 - 220) / 60 = 11.167$

Equation of DRL
$y - y_{(at\ x=10)} = S_{TBP}(x - x_{(at\ x=10)})$
$y - 220 = 11.167(x - 10)$,
$y = 11.167x + 108.33$

The slope of the FRL, S_{FRL} is found from the Maxwell's correlation as $S_{FRL} = 8.1$

From another Maxwell's correlation, we obtain the difference between the DRL and at 50 % volume distilled, and this is,

$\Delta t_{50(DRL\ -\ FRL)} = 40\ ^0F$

$T_{DRL,50} = 666.7\ ^0F$

Then,
$T_{FRL,50} = 666.7 - 40 = 626.7\ ^0F$

Equation of FRL
$y - y_{(at\ x=50)} = S_{FRL}(x - x_{(at\ x=50)})$
$y - 627 = 8.1(x - 50)$
$y = 8.1x + 222$

The table for Crude TBP, DRL and FRL Temperature Data is presented as follows:

Vol %	T_{TBP} (^{0}F)	T_{DRL} (^{0}F)	T_{FRL} (^{0}F)
0	-30	108	222
10	220	220	303
30	450	443	465
50	650	667	627
70	890	890	789
90	1250	1113	951
100	1430	1225	1032

From Maxwell's third correlation for the EFV, data is summarized as follows

Calculation of EFV:

At Vol % = 30
T_{DRL} = 443 ^{0}F
T_{FRL} = 465 ^{o}F
$\Delta t_{TBP - DRL} = T_{TBP,30} - T_{DRL,30}$
= 450 - 443 = 7 ^{0}F
$\Delta t_{EFV - FRL}/\Delta t_{TBP-DRL}$ = 0.34

from which, $T_{EFV,30}$ = 467 ^{0}F

The EFV data is summarized as follows

Vol %	T_{EFV} (^{0}F)
0	185
10	303
30	467
50	629
70	789
90	998
100	1102

Calculation of flash zone temperature

Assume 1.2 lbs of steam/gal residue

Steam flow rate (lbs/hr) = 48825
Steam flow rate (lbmol/hr) = 2712.5

Flash zone pressure = 40 psia
Partial pressure of hydrocarbon = 17.93 psia

EFV 50 % of the crude = 629 ^{o}F
From Table 2.18, first locate the first point corresponding to 50 % EFV on 14.7 psia line and then extend this line parallel to the existing graph till 17.93 psia to obtain the new EFV value.

From graph, EFV 50 % at 17.93 psia = 665 ^{o}F

Slope of the FRL = 8.1

EFV at 17.93 psia is defined using the line y = 665 + 8.1 (x-50) where x refers to the cumulative volume %. For a value of 56.5 % of x (which is the overflash value), y = 717.65 ^{o}F which is the flash zone temperature.

0% point of cut on atmospheric TBP curve of product ^oF	**Theoretical side draw temperature – Actual side draw temperature**
197.75	9.29
301.27	24.43
300.12	24.43
500.32	66.37
599.31	95.31
686.79	122.62
743.21	142.95
798.48	162.21
851.45	183.06
897.52	201.23

Table 4.1: Packie's correlation data to estimate the draw off temperature.

4.6 Estimation of draw off stream temperatures

The following procedure is adopted for the estimation of draw off temperatures which are useful for the energy balance calculations to yield the side stripper products (kerosene, LGO and HGO) temperatures:

a) Estimate the steam requirements for different products. Use the following assumptions on a mass ratio basis: steam to HGO ratio = 0.5; steam to LGO ratio = 0.5; steam to kerosene ratio = 0.65.

b) Estimate the FRL of the draw off TBP cut. This is done using the TBP cuts of the crude. Establish its IBP at atmospheric pressure. This is regarded as 0% vol temperature on the FRL. This value is required for using packie's correlation to estimate the actual draw off temperature.

c) Predict overflow from the draw-off tray as a liquid reflux to the tray below using the following rules of thumb
Molar ratio of overflow liquid to HGO product = 2.9
Molar ratio of overflow liquid to LGO product = 1.2
Molar ratio of overflow liquid to kero product = 0.9 to 1.0

d) Calculate hydrocarbon partial pressure using the expression:
Partial pressure = (Product vapor moles + Moles overflow)/(product vapor moles + moles steam) x tray pressure
Here product vapor corresponds to total products that are vaporized and leaving the tray. These include HGO + LGO + Kerosene + naphtha for the HGO draw off tray, LGO + Kerosene + naphtha for the LGO draw off tray, Kerosene + naphtha for the kerosene draw off tray. Also in the above expression, steam moles corresponds to steam entering at tray 1 for HGO draw off tray, steam entering at residue and HGO for LGO draw off tray, steam entering at residue, HGO and LGO for kerosene draw off tray. Estimate the tray pressure from assumed pressure drops.

e) Using vapor pressure curves, determine the IBP of the cut at the evaluated partial pressure.

f) Use Table 4.1 (Packie's correlation data) to determine the differential temperature. **For this curve, the x-axis data is to be taken with respect to atmospheric TBP cut point and not the partial pressure based cut point.**

g) Use differential temperature and 0 % point on the FRL (adjusted to the partial pressure) to estimate the draw off temperature.

Q 4.5: For the Ecudaor CDU design problem, determine the draw off streams temperatures associated to HGO, LGO and Kerosene.

Solution:

For all three draw-off stream trays, the tray pressure is evaluated first.

HGO draw off tray = 10
Flash zone pressure = 40 psia
HGO draw off tray is located 6 trays above the flash zone.
Average pressure drop per tray = 0.32 psia per tray

HGO draw off tray pressure = 40 – 6 x 0.32 = 39.1 psia.

Similarly, LGO draw off tray pressure = 40 – (6 + 10) x 0.32 = 34.88 psia

Kerosene draw off tray pressure = 40 – (6 + 10 + 12) x 0.32 = 31.04 psia

Next, we evaluate the steam requirements in the side stream strippers.
Residue zone fresh steam flow rate (from flash zone calculations) = 2712.5 lbmol/hr

HGO zone fresh steam flow rate = 6125 x 0.5/18 = 170.13 lbmol/hr

LGO zone fresh steam flow rate = 11375 x 0.5/18 = 5687.5 lbmol/hr

Kerosene zone fresh steam flow rate = 14000 x 0.65/18 = 505.55 lbmol/hr

From mass balance table summarized previously,

Naphtha vapor flow rate = 898.03 lbmol/hr

Kerosene vapor flow rate = 670.9 lbmol/hr

LGO vapor flow rate = 391.1 lbmol/hr

HGO vapor flow rate = 181.63 lbmol/hr

I) HGO draw off tray temperature calculation

Moles overflow = 2.9 x 181.63 = 526.72 lbmol/hr

Hydrocarbon vapor flow rate = 898.03 + 670.9 + 391.1 + 181.63 = 2140 lbmol/hr

Steam flow rate = 2712.5 lbmol/hr (Only that steam that is reaching the HGO draw off tray is the steam that enters at the bottom of the main column).

Therefore, partial pressure of hydrocarbons at HGO draw off tray = (2140 + 526.72)/(2140 + 526.72 + 2712.5) x 38.1 = 18.89 psia.

Now, the EFV of the Heavy gas oil is determined using Maxwell's correlation. The procedure is not shown here. The obtained IBP from EFV curve is 561 ^{o}F.

From vapor pressure curves, estimated theoretical HGO draw off temperature at a partial pressure of 18.89 psia (instead of 14.7 psia) is 620 ^{o}F.

From Packie's correlation, for an x-axis data point of 561 ^{o}F, the y-axis point (Table 4.1) is 83 ^{o}F.

Therefore, actual HGO draw off temperature = 620 – 83 = 537 ^{o}F.

II) LGO draw off tray temperature calculation

Moles overflow = 1.2 x 391.08 = 469.3 lbmol/hr

Hydrocarbon vapor flow rate = 898.03 + 670.9 + 391.1 = 1960.03 lbmol/hr (All HC vapors other than the HGO and residue products).

Steam flow rate = 2712.5 + 170.13 = 2882.64 lbmol/hr (This is the steam that enters at the residue zone and also in the HGO side stripper).

Therefore, partial pressure of hydrocarbons at HGO draw off tray = (1960.03 + 391.08)/(1960.03+391.08 + 288.26) x 34.2 = 15.66 psia.

Now, the EFV of the LGO is determined using Maxwell's correlation. The procedure is not shown here. The obtained IBP from EFV curve is 485 ^{o}F.

From vapor pressure curves, estimated theoretical LGO draw off temperature at a partial pressure of 15.66 psia (instead of 14.7 psia) is 510 ^{o}F.

From Packie's correlation, for an x-axis data point of 510 ^{o}F, the y-axis point (Table 4.1) is 60 ^{o}F. Therefore, actual HGO draw off temperature = 510 – 60 = 450 ^{o}F.

III) Kerosene draw off tray temperature calculation

Moles overflow = 0.9 x 670.9 = 603.81 lbmol/hr.

Hydrocarbon vapor flow rate = 898.03 + 670.9 = 1568.03 lbmol/hr (All HC vapors other than the LGO, HGO and residue products).

Steam flow rate = 3198.611 lbmol/hr

Therefore, partial pressure of hydrocarbons at HGO draw off tray = (1568 + 603.81)/(1568 + 603.81 + 3198.6) x 30.2 = 12.5 psia.

From EFV, vapor pressure curves and packie's correlation, actual draw off-temperature = 320 °F.

4.7 Estimation of tower top temperature

The tower top temperature is very important to determine, as the know-how of the temperature will lead to further delineating the condenser duty. The tower top temperature is determined from a dew point calculation. The procedure for the same is similar to that provided in the gasoline blending problem illustrated in the refinery mass balances chapter. The procedure for the same is outlined below:

a) Set the reflux drum temperature and pressure as 100 °F and 10 psig. Assume 5 psia pressure drop and hence, tower top pressure = 15 psig or 29.7 psia.
b) Assume external reflux as 0.8 times the total moles overhead product. Determine external reflux flow rate
c) Determine partial pressure of hydrocarbons in the tower top section.
d) Assume a tower top temperature of 250 °F. This value is required to determine the equilibrium constants (K) for different pseudo-components.
e) As reflux will have a composition similar to the gas + naphtha fraction whose pseudo-component distribution is previously known, determine the vapor pressures of each pseudo-component as a function of its mid boiling point and API.
f) Determine mole fraction of each pseudo-component.
g) Calculate equilibrium constant for each pseudo-component as the ratio between its vapor pressure (determined from Maxwell's vapor pressure correlations) and the partial pressure of the hydrocarbons (determined in step c)
h) Assume the determined mole fractions to correspond to the vapor stream. Eventually, determine liquid stream mole fraction using the expression $x = y/K$ where K is the equilibrium constant.
i) Eventually, evaluate summation of all x values
j) For the last pseudo-component, determine its new equilibrium constant using the expression $K_2 = K_1$ (sum of all x values).
k) From the K_2 values, determine the new tower top temperature.
l) If significant differences exist between the new tower top temperature and assumed value (250 °F), then iterate the procedure until a converging tower top temperature is determined.

Q 4.6: For the Ecudaor CDU design problem, determine the tower top temperature.

Solution:

Tower top pressure = 29.7 psia

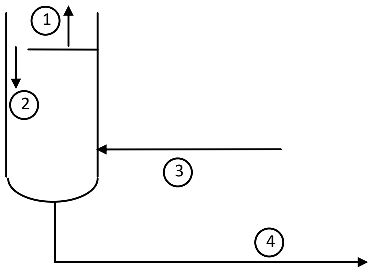
Vapor flow rate = (1+0.8) x 898.03 = 1616.46 lbmol/hr

Steam flow rate in the tower top section = 3704.1 lbmol/hr

Partial pressure of hydrocarbons in the tower top section = 1616.4/(1616.4 + 3704.1) x 29.7 = 9.02 psia.

For determining the tower top temperature, we first assume a tower top temperature of 250 °F. At this temperature, using Maxwell's vapor pressure correlation, we determine the corresponding vapor pressures. All evaluations are presented below in the table.

Component	Mid BP	Vol % on crude	S.G	Wt Factor	Mole Fraction (y)	P^S from Maxwell	K	X=y/K
1	75	1.85567	0.638826	1.185451	0.01631	12.4	20.20128	0.000807
2	125	7.42268	0.665882	4.942632	0.068001	6	9.774811	0.006957
3	160	20.61856	0.698765	14.40753	0.198221	3.8	6.190714	0.032019
4	195	28.86598	0.723785	20.89277	0.287446	2.1	3.421184	0.084019
5	235	19.58763	0.746702	14.62612	0.201228	1.36	2.215624	0.090822
6	265	15.46392	0.766938	11.85986	0.16317	0.7	1.140395	0.143082
7	295	3.092784	0.771117	2.384898	0.032812	0.56	0.912316	0.035965
8	320	3.092784	0.771117	2.384898	0.032812	0.35	0.570197	0.057545
Total				72.68416	1			0.451217



①	Product vapor (Naphtha + Kerosene + LGO + HGO) + Steam (at partial pressure of top stripping tray) + Vapor reflux.
②	Liquid Reflux at top stripping tray.
③	Fresh steam for the flash zone
④	Residue product (At unknown temperature)

Figure 4.3: Envelope for the enthalpy balance to yield residue product temperature.

Psia (psi)	**Saturation Temp. (°F)**	**Enthalpy (Btu/lb) at temperature**										
		350 °F	**400 °F**	**500 °F**	**600 °F**	**700 °F**	**800 °F**	**900 °F**	**1000 °F**	**1100 °F**	**1200 °F**	**1300 °F**
15	213.03	1216.2	1239.9	1287.3	1335.2	1383.8	1433.2	1483.4	1534.5	1586.5	1693.2	1803.4
20	227.96	1215.4	1239.2	1286.9	1334.9	1383.5	1432.9	1483.2	1534.3	1586.3	1693.1	1803.3
30	250.34	1213.6	1237.8	1286.0	1334.2	1383.0	1432.5	1482.8	1534	1586.1	1692.9	1803.2
40	267.25	1211.7	1236.4	1285.0	1333.6	1382.5	1432.1	1482.5	1533.7	1585.8	1692.7	1803.0
50	281.02	1209.9	1234.9	1284.1	1332.9	1382.0	1431.7	1482.2	1533.4	1585.6	1692.5	1802.9
60	292.71	1208.0	1233.5	1283.2	1332.3	1381.5	1431.3	1481.8	1533.2	1585.3	1692.4	1802.8
70	302.93	1206.0	1232.0	1282.2	1331.6	1381.0	1430.9	1481.5	1532.9	1585.1	1692.2	1802.6
80	312.04	1204	1230.5	1281.3	1330.9	1380.5	1430.5	1481.1	1532.6	1584.9	1692	1802.5
90	320.28	1202	1228.9	1280.3	1330.2	1380	1430.1	1480.8	1532.3	1584.6	1691.8	1802.4
100	327.82	1199.9	1227.4	1279.3	1329.6	1379.5	1429.7	1480.4	1532	1584.4	1691.6	1802

Table 4.2: Steam table data relevant for CDU design problem.

$K_2 = 0.451217 \times 0.35 = 0.257283$.

From Maxwell's vapor pressure curves, for pseudo-component 8, this vapor pressure corresponds to 220 + 0.8 x 20 = 236 ^{o}F. We take this as the tower temperature for future calculations.

4.8 Estimation of residue product stream temperature

The residue product stream temperature is determined by consider the envelope shown in Figure 4.3 for the flash zone. For the chosen envelope, the incoming streams are residue liquid (at flash zone temperature), strip out liquid (at flash zone temperature) and fresh steam (entering tray 1). The fresh steam is usually taken as superheated steam at 450 ^{o}F and 50 psig. The outgoing streams are residue liquid at unknown temperature t, strip out vapor at flash zone temperature and steam at flash zone temperature. For stripout vapor + hydrocarbon vapor leaving the flash zone, assume that their temperature is 5 ^{o}F lower than the flash zone temperature. For steam at flash zone temperature, its enthalpy is determined at its partial pressure that exists in the flash zone and using steam tables. Relevant data from steam tables is summarized in Table 4.2. From the enthalpy balance of both consolidated incoming and outgoing streams, the unknown enthalpy of the residue product is determined.

For all calculations, the following important information needs to be remembered always:

a) Hydrocarbon stream's liquid or vapor enthalpy is determined from Maxwell's correlation. The Maxwell's correlations need to be interpolated for both average MEABP of the stream as well as its K value. Therefore, for any stream, it is mandatory to determine first the MEABP and K value. To determine K value, we have to know its API.

b) For all product streams, their MEABP is assumed to be known. If not known, it is approximated to be same as the its VABP.

c) For stripout streams, their MEABP is determined as an average of the MEABP values corresponding to the adjacent cuts. Same is the case for the K value as well as API.

d) For fresh steam, superheated steam data at 450 °F is taken from steam tables. For steam at any other location in the column, its saturated vapor enthalpy at the prevalent partial pressure of the steam is determined.

An illustrative example is presented below to highlight these issues.

Q 4.7: For the Ecudaor stream determine the residue product stream temperature.

Solution:

The enthalpy balance for the envelope shown in Figure 4.3 provides the following information for the incoming streams to the flash zone.

Stream	V/L	API	MEABP	K	°F	Lb/hr	Btu/lb	mmBtu/hr
Residue	L	14.1	910 °F	11.45	717.65	329710.8	389	128.25
Stripout	L	23.1	780	11.75	717.65	18309.4	401	7.34
Fresh Steam	V	-			450	48825	1262.6	61.64
Total						396845.3		197.2

For the outgoing streams, the consolidated stream enthalpy data is presented as follows in the flash zone:

Stream	V/L	API	MEABP	K	°F	Lb/hr	Btu/lb	mmBtu/hr
Residue	L	14.1	910 °F	11.45	t	329710.8	x	0.329710x
Stripout	V	23.1	780	11.75	712	18309.4	487	8.92
Steam	V (PP = 22.07 psia)				712	48825	1389.7	67.85
Total						396845.3		76.77 + 0.329710x

Equating the incoming and outgoing heat balance total, we get the enthalpy of the residue liquid product as x = 365.4 Btu/lb.

From Maxwell's correlation upon interpolation,we get T = 686.5 °F which is the residue product temperature.

4.9 Estimation of side stripper products temperature

Consider the side stripper itself as an envelope for the heat balance. For the side stripper the incoming streams are:

a) Product draw off stream at draw off temperature that was evaluated using packie's correlation. The product draw off stream (liquid) consists of two portions namely the hydrocarbon vapor equivalent to the final product flow rate and strip out stream that enters the side stripper as liquid and leaves the side stripper as vapor.
b) Fresh steam at 450 ^{o}F and 50 psig.

Similarly, the outgoing streams are:

a) Side stripper product stream at unknown temperature.
b) Stripped hydrocarbon vapor whose temperature is assumed to be 5 ^{o}F lower than the draw off temperature.
c) Steam at its partial pressure and vapor temperature.

In these enthalpy balances, the stripout vapor/liquid flowrate is estimated to be 5 % for HGO side stripper unit and 8 % for LGO and Kerosene side stripper units. Subsequently, API is determined to be the average of the corresponding API values of the adjacent cuts. Since in the above consolidated enthalpy balance stream only side striper product stream enthalpy is unknown it can be determined. From the enthalpy values using Maxwell's correlation, one can determine the side stripper product's temperature. The same procedure is applicable for Kerosene, LGO and HGO side strippers.

Q 4.8: For the Ecudaor CDU, determine the side stripper product temperatures.

Solution: We present the enthalpy balance table for the HGO side stripper first.

Incoming streams

Stream	V/L	API	MEABP	K	^{o}F	Lb/hr	Btu/lb	mmBtu/hr
Feed ex strip out	L	32.2	575	11.7	537	44163.4	290	12.81
Stripout	L	33.9	545	11.7	537	2205	292	0.64
Fresh Steam	V				450	3062.5	1262.6	3.86
Total						49430.9		17.31

Outgoing streams

Stream	V/L	API	MEABP	K	^{o}F	Lb/hr	Btu/lb	mmBtu/hr
HGO	L	32.2	575	11.7	t	44163.8	x	0.44163x
Stripout	V	33.9	545	11.7	532	2205	393	0.8665
Steam	V (PP = 19.21 psia)					3062.5	1302.5	3.9889
Total								4.8554 + 0.44163x

Solving for x we get x = 282.09 Btu/lb

From Maxwell's correlations and interpolation, we get HGO product stream temperature = 526.8 ^oF.

Next, we present the LGO stripper enthalpy balance

Incoming streams

Stream	V/L	API	MEABP	K	^oF	Lb/hr	Btu/lb	mmBtu/hr
Feed ex strip out	L	35.7	500	11.65	450	80289.8	234	18.78
Stripout	L	40.1	421	11.6	450	6215.3	241	1.497
Fresh Steam	V				450	5687.5	1262.6	7.181
Total								27.466

Outgoing streams

Stream	V/L	API	MEABP	K	^oF	Lb/hr	Btu/lb	mmBtu/hr
LGO	L	35.7	500	11.65	t	80289.8	x	0.0802x
Stripout	V	40.1	421	11.6	445	6215.3	353	2.194
Steam	V (PP = 17.43 psia)				445	5687.5	1260.6	7.169
Total								0.0802x + 9.36363

From heat balance, x = 225.5 Btu/lb.

From Maxwell's correlation and interpolation, LGO product stream temperature = 435.4 ^oF. Finally, we present Kerosene stripper heat balance

Incoming streams

Stream	V/L	API	MEABP	K	^oF	Lb/hr	Btu/lb	mmBtu/hr
Feed ex strip out	L	44.5	364	11.65	320	93859.7	163	15.299
Stripout	L	57.1	285	12.1	320	7000	178	1.246
Fresh Steam	V				450	9100	1262.6	11.489
Total								28.034

Outgoing streams

Stream	V/L	API	MEABP	K	^oF	Lb/hr	Btu/lb	mmBtu/hr
LGO	L	44.5	364	11.65	t	93859.7	x	0.09386x
Stripout	V	57.1	285	12.1	315	7000	297	2.079
Steam	V (PP = 15.01 psia)				315	9100	1199.6	10.91
Total								12.995

From heat balance, x = 160.3 Btu/lb.

From Maxwell's correlation and interpolation, Kerosene product stream temperature = 315 ^oF.

4.10 Total Tower energy balance and total condenser duty estimation

The total tower energy balance is carried out to consolidate the following streams:

Incoming streams:

a) Crude (Vapor + Liquid): Its flow rate is determined from mass balance for both liquid and vapor streams separately. Eventually, using first principles, total enthalpy (mmBtu/hr) is evaluated.
b) Steam: Fresh steam with its total flow rate is estimated and its total enthalpy (mmBTu/hr) entering the CDU is estimated.

Outgoing streams:

a) All products at their respective stream temperatures that were determined previously.

The balance energy of the incoming and outgoing streams is determined as the energy removed from condenser, bottom and top pump around ($Q_c + Q_{BPA} + Q_{TPA}$). An illustrative example is presented next for the Ecuador crude CDU.

Q 4.9: For the Ecuador crude stream, determine the total condenser and pump around duties with the help of the total tower balance.

Solution:

Crude enthalpy data evaluated from first principles is presented as follows:

Stream	V/L	API	MEABP	K	^oF	lb/hr	Btu/lb	mmBtu/hr
Crude Vapour	V	46.5	360	11.7	717.65	327733.1	526	172.3876
Crude Liquid	L	13.4	910	11.4	390	310065.5	390	120.9255
Total						637798.6		293.3132

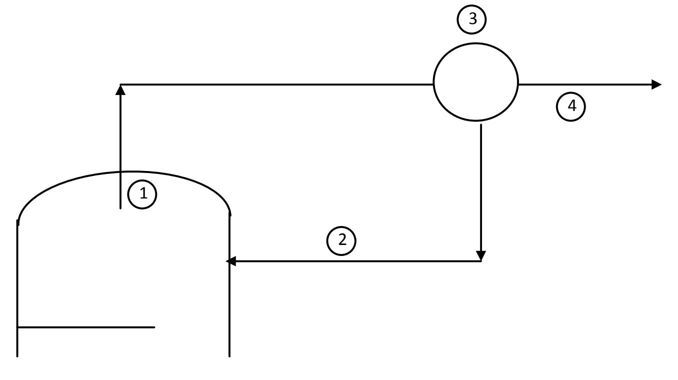
Overall enthalpy balance is summarized in the following table from where the total condenser and pump around duties can be estimated:

Stream	**V/L**	**K**	**T (^{o}F)**	**lb/hr**	**Btu/lb**	**mmBtu/hr**
IN						
Crude	V+L	11.7	717.65	637798.6		293.3132
Steam	V			66675	1263	84.1897
Total				704473.6		377.5028
OUT						
Residue	L	11.5	686.6	329710.9	365.4	120.4773
HGO	L	11.55	526.9	44163.38	282.1	12.4582
LGO	L	11.6	435.4	80289.88	225.5	18.1059
Kero	L	11.9	315.1	93859.74	160.3	15.0467
Naphtha	L	12.75	100	89776.86	53	4.7582
Refluxes						206.6565
Total						377.5028

From the above table, $Q_C + Q_{BPA} + Q_{TPA} = 206.65$ mmBtu/hr.

4.11 Estimation of condenser duty

The condenser duty is estimated using the heat balance envelope presented in Figure 4.4. For the chosen envelope, the incoming streams are naphtha vapor (from tray 45), reflux vapor (from tray 45), steam (at its partial pressure). The temperature of these streams is the tower top temperature that was estimated as a dew point earlier. The outgoing streams for the envelope are naphtha liquid product, water liquid product and cold naphtha reflux. All these streams are assumed to be at 100 ^{o}F. The only term that is missing in the outgoing enthalpy balance is the condenser heat duty. This is therefore estimated from the enthalpy balance. We next present an illustrative example for the determination of condenser duty. Eventually, total pump around duties is determined from the previously known total heat removed minus the condenser duty.



① Product (Naphtha) Vapor + Steam (at PP) + Vapor Reflux.

② Naphtha cold reflux at 100 °F

③ Condenser duty (Unknown)

④ Naphtha product + Condensed water at 100 °F

Figure 4.4: Heat balance Envelope for condenser duty estimation.

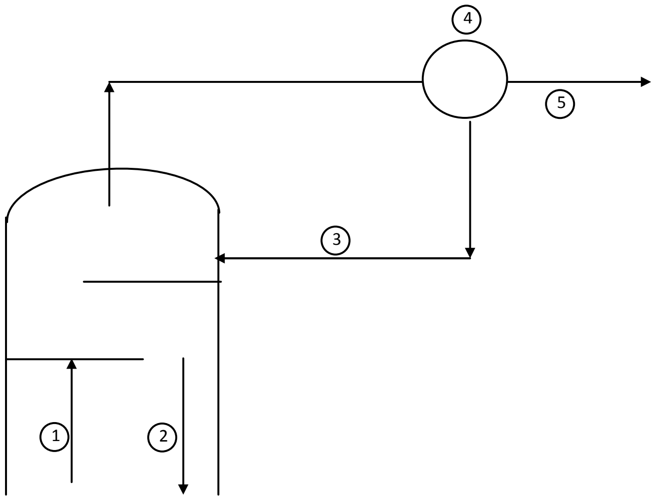
Q 4.10: Determine the condenser duty for the Ecuador crude oil CDU problem.

Solution:

The enthalpy balance table for the evaluation of condenser duty is summarized as follows:

Stream	V/L	K	T	lb/hr	Btu/lb	mmBtu/hr
IN						
Naphtha	V	12.75	236	89776.86	258.33	23.1921
Reflux	V	12.75	236	71821.49	258.33	18.5536
Steam	V		236	66675	1161.8	77.4630
Total				228273.3		119.2087
OUT						
Naphtha	L	12.75	100	89776.86	53	4.4888
Refluxes	L	12.75	100	71821.49	53	3.5911
Water	L		100	66675	67.3	6.6675
Cond duty						104.4613
Total				228273.3	100	119.2087

From the enthalpy balance table, condenser duty = 104.46 mmBtu/hr. Therefore, bottom + top pump around duties = 206.65 – 104.46 = 102.19 mmBtu/hr.



①	Product (Naphtha) Vapor + Steam (at its partial pressure) + Vapor reflux approaching the second tray from top.
②	Liquid Reflux leaving the first tray from the top
③	Naphtha cold reflux
④	Condenser duty (Also known)
⑤	Naphtha and Water products

Figure 4.5: Envelope for the determination of tower top tray overflow.

4.12 Estimation of Overflow from Top tray

The top most tray overflow is important to determine, as later it will be checked to meet the fractionation efficiency. The tower top most tray overflow is determined also from the enthalpy balance by considering an envelope shown in Figure 4.5. For the chosen envelope, the incoming streams are naphtha vapor (from tray 44), overflow vapor (from tray 44) and steam (from tray 44). Eventually heat is lost in the condenser duty and overflow liquid, naphtha liquid product (at 100 ^{o}F), water (at 100 ^{o}F). In these terms, the overflow enthalpy is estimated from where the overflow rate (lbs/hr) is evaluated. However, for these calculations, it is important to assume the overflow liquid and vapor temperatures. It is fair to assume that the overflow liquid is 5 ^{o}F higher than the tower top temperature and the overflow vapor is 5 – 6 ^{o}F higher than the over-flow liquid. **This rule of thumb will be applicable in future calculations as well that the approaching vapor and liquid are at a temperature difference of 5^{o}F.** We next present an illustrative example for the estimation of overflow from the top tray.

Q 4.11: For the Ecuador crude CDU, determine the top tray overflow rate in lbs/hr.

Solution:

The energy balance table for the envelope presented in Figure 4.5 is presented below:

Stream	**V/L**	**API**	**MEABP**	**K**	**T (^oF)**	**lb/hr**	**Btu/lb**	**mmBtu/hr**
IN								
Naphtha	V	69.8	266	12.75	252	89776.86	268.03	24.06289
O/Flow	V	57.1	315	12.25	252	x	264.675	274.5x
Steam	V				252	66675	1169.6	77.98308
Total						156451.9+x		103.43+274.5x
						156451.9		102.046
OUT								
Naphtha	L	69.8	266	12.75	100	89776.86	53	4.758174
O/Flow	L	57.1	315	12.25	246	x	129.975	141.8x
Water	L				100	66675	67.3	4.487228
Cond duty								104.4613
Total						156451.9+x		116.011+141.8x

From the enthalpy balance, we get x = 86568.14 lbs/hr. API of the overflow = 57.1 which corresponds to 6.2497 lbs/gal. Therefore, the top tray overflow in gallons per hour = 13851.55 GPH. This is regarded as cold GPH. In fractionation efficiency calculations, we need to determine the hot GPH. Therefore, it is important to note this concept in the design calculations.

4.13 Verification of fractionation criteria

In a distillation column, the generation of adequate amount of vapor requires adequate amount of liquid. Therefore, fractionation criteria is usually defined as the ratio of vapor and liquid (reflux) flow rates. However, in the case of CDU, the fractionation criteria is defined using the product vapor flow rates and corresponding reflux rates.

Fractionation criteria is defined using the following terms for the performance of the CDU:

A) Factor F: Defined as the ratio of the hot gallons per hour (GPH) of the lighest product from draw off tray to the total vapor product (cold GPH) leaving the lightest product draw off tray.
For different combinations of product streams, the factor F is defined as

i) Naphtha-Kerosene

$$F = \frac{Hot\ GPH\ leaving\ top\ tray}{Naphtha\ product\ flow\ rate\ (GPH)\ in\ the\ vapor}$$

ii) Kerosene-LGO

$$F = \frac{Hot\ GPH\ leaving\ Kerosene\ Draw\ off\ tray}{Naphtha + Kerosene\ product\ flow\ rate\ (GPH)\ in\ the\ vapor}$$

iii) LGO-HGO

$$F = \frac{Hot\ GPH\ leaving\ LGO\ Draw\ off\ tray}{Naphtha + Kerosene + LGO\ product\ flow\ rate\ (GPH) in\ the\ vapor}$$

iii) HGO-LGO

$$F = \frac{Hot\ GPH\ leaving\ HGO\ Draw\ off\ tray}{Naphtha + Kerosene + LGO + HGO\ product\ flow\ rate\ (GPH) in\ the\ vapor}$$

A convenient correlation between the cold stream specific gravity (at 60 °F) and the hot stream specific gravity (at desired temperature) is presented in Table 4.3. Using these two specific gravities, the following expression is used to convert the cold GPH values (which are usually known from CDU mass balances):

$$Hot\ GPH = cold\ GPH \frac{cold\ SG}{Hot\ SG}$$

Since usually hot stream SG is lower than the cold stream SG, the hot GPH is usually higher than the cold GPH.

After applying the fractionation criteria, one would be able to carry out mass and energy balances as follows:

a) Top tray: Provides appropriate liquid flow rate (reflux) for sections below the top tray.
b) Kerosene draw off tray: Provides appropriate liquid flow rate (reflux) to sections below the kerosene draw off tray
c) LGO draw off tray: Provides appropriate liquid flow rate (reflux) to sections below the LGO draw off tray.
d) HGO draw off tray: Provides appropriate liquid flow rate (reflux) to sections below the HGO draw off tray.

B) No of trays in various sections (N):

The number of trays in various sections is taken as follows for various combinations:

i) Naphtha-Kerosene: N = 12 (No pump arounds, therefore, all trays included in the evaluation)

ii) Kerosene-LGO: N = 11 (2 Pump around trays get a credit of only 1 tray).
iii) LGO-HGO: N = 11 (2 pump around trays get a credit of only 1 tray).
iv) HGO-residue: Criteria not required. This is due to the fact that usually ASTM gaps are defined for Naphtha-kerosene, Kerosene-LGO, LGO-HGO combinations and the final TBP temperature of the residue is usually specified.

C) ASTM Gaps (G):

ASTM Gaps are usually defined as per the following product specifications for the CDU

i) Naphtha-Kerosene: G = 25 ^{o}F
ii) Kerosene-LGO: G = -10 ^{o}F
iii) LGO-HGO: G = -25 ^{o}F

The fractionation criteria is defined as a correlation between Reflux Ratio, N, G and $\Delta T_{50\%}$ TBP between the adjacent cuts. Typically, fractionation criteria is correlated between (Reflux ratio x N) and G for various values of $\Delta T_{50\%}$ TBP. The fractionation criteria correlation data are presented in Table 4.4 – 4.5 for steam stripping conductions in the CDU.

The fractionation criteria can be applied in two ways which are summarized as follows:

a) Assume ASTM Gap as desired for the combination of products and determine the hot GPH liquid reflux flow rates. Eventually, conduct energy balance to determine unknown parameters such as pump around duties etc.
b) Assume the value of the unknown parameter such as pump around duty and determine the hot GPH as an unknown parameter from energy balances of respective sections of the CDU. Eventually, with the known values of the hot GPH, verify whether desired ASTM gap has been met or not. If desired ASTM gap values are not met, then unknown parameters need to be adjusted.

A careful dissection of the CDU design calculations conducted so far indicates the following information:

a) Reflux ratio of 0.8 (on a molar basis) has been assumed previously. The reflux ratio enabled the determination of the tower top temperature. The tower top temperature enabled the evaluation of liquid and vapor enthalpies at the top section for the energy balance.
b) From energy balance of the tower top section, we have been able to determine the condenser duty. From this the total pump around duties can be estimated.

SG = 0.5		SG = 0.52		SG = 0.54		SG = 0.58		SG = 0.6		SG = 0.62		SG = 0.64		SG = 0.66		SG = 0.68		SG = 0.7	
T (°F)	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**
80.07	0.48	81.38	0.50	79.02	0.52	80.39	0.56	87.81	0.58	86.67	0.60	90.42	0.62	85.62	0.64	101.59	0.65	98.01	0.68
100.84	0.46	102.17	0.48	101.03	0.51	99.96	0.55	115.94	0.56	117.26	0.58	122.24	0.60	110.09	0.63	139.54	0.64	139.62	0.66
119.16	0.44	121.73	0.47	120.59	0.49	121.98	0.53	137.96	0.54	146.61	0.56	151.60	0.58	138.24	0.62	171.36	0.62	181.23	0.63
139.92	0.42	141.28	0.45	140.15	0.48	141.55	0.52	164.86	0.52	177.18	0.54	182.18	0.56	168.82	0.60	201.95	0.60	217.94	0.61
160.66	0.39	160.82	0.43	160.93	0.46	163.55	0.50	197.88	0.50	204.10	0.52	217.67	0.54	200.63	0.58	233.77	0.58	257.10	0.59
178.94	0.36	184.00	0.40	180.47	0.44	183.12	0.49	227.23	0.48	228.56	0.51	251.93	0.52	230.00	0.56	260.68	0.57	287.69	0.58
182.58	0.35	198.64	0.38	213.42	0.39	200.23	0.47	262.66	0.45	267.70	0.48	293.50	0.49	259.37	0.55	288.83	0.55	339.09	0.55
		212.02	0.35	232.93	0.37	216.12	0.46	289.53	0.42	301.92	0.45	330.20	0.47	285.06	0.53	335.32	0.53	377.03	0.53
				239.00	0.35	249.10	0.43	309.05	0.39	331.24	0.42	368.10	0.44	322.99	0.51	367.14	0.51	411.29	0.51
						267.40	0.40	323.68	0.37	367.87	0.38	397.43	0.41	351.12	0.49	395.26	0.49	445.55	0.49
						289.37	0.37			386.16	0.35	427.95	0.38	373.13	0.47	430.74	0.47	487.14	0.46
						301.54	0.35					446.23	0.35	403.72	0.45	463.76	0.44	520.17	0.44
														428.16	0.43	491.89	0.42	549.52	0.42
														453.82	0.41	523.66	0.39	580.08	0.39
														483.13	0.38	557.85	0.36	604.51	0.37
														501.44	0.35			616.69	0.35

(a)

SG = 0.72		**SG = 0.74**		**SG = 0.76**		**SG = 0.78**		**SG = 0.8**		**SG = 0.82**		**SG = 0.84**		**SG = 0.88**		**SG = 0.94**		**SG = 0.98**	
T (°F)	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**	**T (°F)**	**SG**
100.53	0.70	114.07	0.71	101.91	0.74	100.78	0.76	102.08	0.78	118.07	0.79	101.01	0.82	119.54	0.86	111.21	0.92	104.01	0.96
140.91	0.67	165.48	0.68	139.85	0.72	142.41	0.74	160.85	0.76	175.62	0.77	157.35	0.80	185.67	0.83	178.55	0.90	154.23	0.95
181.30	0.65	225.45	0.66	182.70	0.70	188.92	0.72	223.29	0.73	230.73	0.75	208.78	0.78	248.12	0.81	254.48	0.87	195.86	0.93
221.69	0.63	285.43	0.63	224.33	0.68	237.88	0.70	283.28	0.71	291.96	0.73	263.89	0.77	311.80	0.79	310.81	0.85	244.85	0.92
263.30	0.61	341.73	0.60	263.49	0.66	286.87	0.68	344.51	0.68	343.38	0.71	311.63	0.74	375.46	0.76	385.51	0.82	302.41	0.90
293.90	0.60	401.70	0.57	299.00	0.65	323.58	0.66	404.50	0.66	398.48	0.69	353.27	0.73	451.38	0.73	440.62	0.80	357.52	0.88
336.73	0.57	462.88	0.53	336.94	0.63	361.54	0.65	458.35	0.63	462.13	0.66	404.69	0.71	515.04	0.70	485.92	0.79	413.85	0.86
383.24	0.55	522.83	0.50	381.00	0.61	404.38	0.63	504.86	0.61	541.70	0.62	447.54	0.69	581.14	0.67	544.70	0.76	477.53	0.84
424.83	0.52	569.31	0.47	422.62	0.59	434.97	0.61	550.15	0.58	593.09	0.59	491.60	0.67	655.81	0.64	588.77	0.74	524.07	0.82
477.45	0.50	612.12	0.44	465.46	0.56	475.36	0.59	588.09	0.57	635.92	0.57	530.78	0.65	721.89	0.60	634.06	0.72	576.72	0.80
520.27	0.47	659.79	0.40	503.38	0.54	524.32	0.56	635.80	0.54	675.08	0.55	566.26	0.63	773.28	0.57	690.39	0.70	624.47	0.78
565.53	0.44			549.88	0.51	564.69	0.54	671.28	0.52	721.55	0.51	604.22	0.61	816.10	0.54	752.83	0.68	668.54	0.76
610.77	0.40			593.90	0.48	603.84	0.52	710.43	0.49	772.92	0.48	645.83	0.59	879.71	0.51	810.37	0.65	717.53	0.74
				639.17	0.46	644.21	0.49	744.66	0.46	814.50	0.45	686.20	0.57			865.44	0.62	766.51	0.72
				673.41	0.43	678.45	0.47	776.44	0.44	842.62	0.43	722.89	0.54			903.38	0.60	815.47	0.70
				702.73	0.40	712.68	0.44	811.86	0.40			754.71	0.52					863.23	0.68
						756.68	0.40					785.30	0.51					904.86	0.67
												819.54	0.48						
												850.10	0.46						

(b)

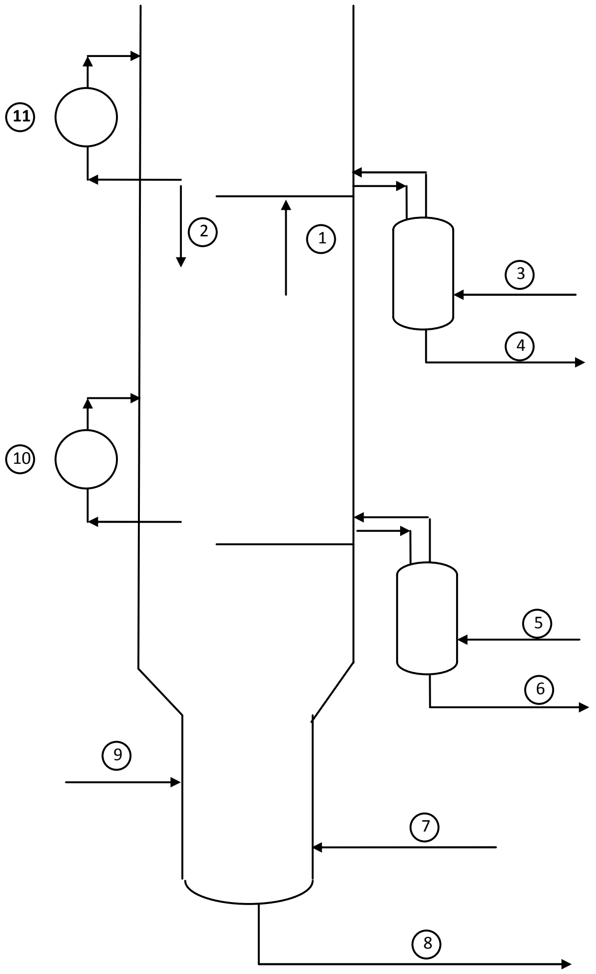
Table 4.3: Variation of specific gravity with temperature (a) Data range: SG = 0.5 to 0.7 at 60 °F (b) Data range: SG = 0.72 to 0.98 at 60 °F.

$\Delta TBP_{50\%}$ = 100 °F		$\Delta TBP_{50\%}$ = 150 °F		$\Delta TBP_{50\%}$ = 200 °F		$\Delta TBP_{50\%}$ = 250 °F	
ASTM Gap (°F)	**F = RR x N**	**ASTM Gap (°F)**	**F = RR x N**	**ASTM Gap (°F)**	**F = RR x N**	**ASTM Gap (°F)**	**F = RR x N**
23.30	100.63	45.40	101.75	65.18	101.56	81.08	102.75
22.13	69.63	43.85	70.40	64.79	81.21	79.92	49.84
19.81	50.12	41.52	50.02	63.63	60.80	77.59	40.39
15.54	34.69	38.03	34.62	62.85	49.92	76.81	34.04
12.83	29.24	33.38	24.60	61.30	39.92	75.26	29.85
9.72	24.65	29.50	19.93	59.75	34.54	73.32	24.50
6.62	21.62	22.13	14.93	56.26	24.54	69.83	19.60
4.29	19.46	14.76	11.64	53.16	19.37	64.79	14.68
-0.36	16.41	9.72	10.08	46.95	14.90	59.75	12.06
-3.85	14.78	5.46	8.84	39.97	11.02	54.32	9.90
-7.73	13.48	0.80	7.96	36.09	9.92	50.83	8.80
-11.22	12.46	-3.85	7.07	29.50	8.04	46.95	7.92
-15.87	10.93					45.01	7.32
-19.36	9.84						
-24.79	8.98						
-28.67	8.19						

Table 4.4: Fractionation criteria correlation data for naphtha-kerosene products.

$\Delta TBP_{50\%}$ = 100 °F		$\Delta TBP_{50\%}$ = 150 °F		$\Delta TBP_{50\%}$ = 200 °F		$\Delta TBP_{50\%}$ = 250 °F		$\Delta TBP_{50\%}$ = 300 °F	
ASTM Gap (°F)	**F = RR x N**	**ASTM Gap (°F)**	**F = RR x N**	**ASTM Gap (°F)**	**F = RR x N**	**ASTM Gap (°F)**	**F = RR x N**	**ASTM Gap (°F)**	**F = RR x N**
-24.82	5.72	-16.37	4.75	-2.75	5.26	1.47	4.51	12.65	4.33
-22.69	6.73	-9.70	6.13	1.19	6.38	8.44	5.88	16.29	5.15
-19.36	7.68	-2.11	8.41	6.65	8.07	16.02	7.82	20.54	6.31
-16.02	9.14	4.86	11.42	12.41	9.60	24.51	9.99	26.60	7.58
-11.77	10.76	10.33	14.44	16.96	11.18	30.27	12.00	33.27	9.20
-7.22	12.79	17.61	21.28	24.23	14.43	37.85	15.64	38.43	11.17
-3.58	15.06	21.26	27.47	28.18	18.06	46.34	22.81	43.89	13.69
0.36	18.10	24.61	38.08	32.43	22.84	51.21	32.27	51.16	18.21
4.31	21.52			36.69	29.47	51.82	36.48	54.50	21.44
7.35	26.94			37.91	36.89			59.36	28.24
10.09	32.37							62.11	37.97
12.21	37.72								

Table 4.5: Fractionation criteria correlation data for side stream-side stream products.



①	Product (Naphtha+Kerosene +LGO) Vapor + Steam (at PP) + Vapor Reflux approaching LGO Draw off tray.
②	Liquid reflux leaving LGO draw off tray.
③	Fresh steam entering LGO SS
④	LGO product
⑤	Fresh steam entering HGO SS
⑥	HGO product
⑦	Fresh stream entering the flash zone.
⑧	Residue product
⑨	Crude (V +L)
⑩	Bottom Pump around duty
⑪	Top pump around duty

Figure 4.6: Energy balance envelope for the estimation of reflux flow rate below the LGO draw off tray.

Since the design calculations of the CDU follow in a sequential manner, it is inevitable that **the selection of the reflux ratio (on a molar basis) at the tower top section is very important. This unique variable has the key for all design calculations and satisfaction of the fractionation criterion.**

a) A wrong choice of reflux ratio may provide negative flow rates. This is unacceptable.

b) **Even if one chooses an appropriately high value of reflux ratio, positive reflux flow rate values may be obtained. However, fractionation criteria needs to be satisfied.**
c) **Very high values of reflux ratio enhances the column diameters and contributes to the cost.**

Therefore, in this lecture notes we uniquely wish to provide elabotations towards this important variable. We consider different values of naphtha reflux ratio in the tower top section and we wish to elaborate upon this

Coming back to the fractionation criteria, the immediate issue that we can consider in the CDU calculations is to execute the following calculations:

a) Apply fractionation criteria for the naphtha-kerosene section and verify whether fractionation crtieria is satisfied or not.
b) Conduct energy balances for the section below the kero draw-off tray whose envelope is shown in Figure 4.6. The heat balance envelope consists of the following consolidated stream data

Heat in	Heat out
Crude Vapor + Liquid (Obtain data from flash zone mass and energy balance calculations)	Overflow vapor (Mass will be same as unknown reflux rate. Temperature is known from tower temperature profile).
Total fresh superheated steam at 450 ^oF	Vapor products to draw off tray (known from mass balance tables).
LGO draw off tray liquid reflux rate (Flow rate is unknown but temperature is known from assumed tower temperature profile)	Steam to tray 34 (all steam other than kerosene fresh steam)
	LGO product
	HGO product
	Total Pump around duties

From the above table, for the heat balance envelope, one requires the knowledge of the total pump around duties, which is obtained from the total refluxes and condenser duties. With this knowledge, determine the liquid reflux flow rates (GPH) and apply fractionation criteria. **If we get negative flow rates then we enhance the reflux ratio systematically until we get both applicable criteria satisfied i.e., positive reflux flow rates (hot GPH) and fractionation criteria.**
We next present an illustrative example that elaborates upon the choice of the reflux ratio and subsequent energy balance calculations:

Q 4.12: Verify the fractionation criteria for the Ecudaor crude oil case and determine the optimal reflux ratio that provides all criteria (positive flow rates and ASTM gap based fractionation criteria) satisfied.

Solution:

From previous solutions, we obtain the following information:

Condenser duty = 104.46 mmBtu/hr (from previous solution).
Condenser duty + TPA duty + BPA duty = 206.65 mmBtu/hr (from previous solution).

Therefore, pump around duty = 102.19 mmBtu/hr.

First, we apply the fractionation criteria for naphtha-kerosene combination:

From the solution of Q9, we obtain the cold GPH of the liquid reflux as 13851.5 GPH.

Specific gravity of the stream at tower top temperature (246 °F) = 0.587393
From mass balance table, SG of Naphtha = 0.705736 (at 60 °F)
Hot GPH = 13851.55*0.705736/0.587393. = 17692.29 GPH
Cold Naphtha product GPH (from mass balance table) = 15271.35 GPH

Factor F = Hot GPH/Naphtha product rate = 17692.29/15271.35 = 1.16

N = 12

Ordinate on fractionation criteria correlation = 1.16 x 12 = 13.9

From product TBP curves, $\Delta T_{50\%}$ = 157.

From fractionation criteria correlation data (Table 4.4), ASTM gap = 27 °F
Desired minimum ASTM gap = 25 °F

Since obtained ASTM gap is greater than desired ASTM gap, the fractionation criteria is concluded to be satisfied.

We next carry out the energy balances for the energy balance envelope presented in Figure 4.6. A summary of the energy balance table is presented below:

Stream			API	MeABP	K	T (°F)	lb/hr	Btu/lb	mmBtu/hr
IN									
Crude		V+L					637798.6		293.3132
Steam		V				450	75425.0	1262.6	95.2316
Tray 34 overflow		L	45.5	408	11.9	321.83	X	162.3	
Total							713223.6		388.5448+162.3X
OUT									
O/F to tray 34		V	45.5	408	11.9	325.8	X	293.0	293X
Vapor to tray 34		V	57.2	315	12.25	325.8	183636.6	302.4	55.5317
Steam to tray 34		V		PP=	22.1	325.8	75425.0	1203.4	90.7694
LGO		L	35.7		11.6	435.4	80289.9	225.5	18.1059
HGO		L	32.2		11.55	526.9	44163.4	282.1	12.4582
Residue		L	14.1		11.5	686.6	329710.9	365.4	120.4773
PA									102.1952
Total							713225.8		399.5378+293X

Solving for X, we get X = -84.106.2

Since we obtained a negative flow rate, we conclude that the naphtha reflux ratio is not enough to meet even the column mass balance requirements.

We next increase column reflux ratio to a value of 1.5 on a molar basis and repeat all relevant calculations. For First we determine the tower top temperature. From dew point calculations, tower top temperature = 248 ^{o}F. Eventually, condenser duty and total pump around duty are evaluated as follows:

Stream		K	T (^{o}F)	lb/hr	Btu/lb	mmBtu/hr
IN						
Naphtha	V	12.75	248	89776.86	268.1	24.0692
Reflux	V	12.75	248	134665.3	268.1	36.1038
Steam	V		248	66675	1167.9	77.8697
Total				291117.2		138.0427
OUT						
Naphtha	L	12.75	100	89776.86	53	4.7582
Refluxes	L	12.75	100	134665.3	53	7.1373
Water	L		100	66675	67.3	4.4872
Cond duty					165	121.6600
Total				291117.2	100	138.0427
Total pump around duty=			**84.9965**	mmBtu/hr		

Next we carry out the energy balance for the envelope presented in Figure 4.6 to determine the tower top tray reflux stream flow rate. Relevant energy balances are summarized in the following table.

Stream		K	T	lb/hr	Btu/lb	mmBtu/hr	
IN							
Naphtha	V	12.75	264	89776.86	274.6	24.65273	
O/Flow	V	12.25	264	x	274.1	274.5x	
Steam	V		264	66675	1175.7	78.3898	
Total				156451.9+x		103.43+274.5x	
OUT							
Naphtha	L	12.75	100	89776.86	53	4.758174	
O/Flow	L	12.25	258	x	143.9	141.8x	
Water	L		100	66675	67.3	4.487228	
Cond duty						121.6600	
Total				156451.9+x		116.011+141.8x	
x=	214000.7	lb/hr	34241.71	GPH			

Using information from the above table, we next apply the fractionation criteria for naphtha-kerosene combination.

Cold GPH = 34241.7
SG at 258 ^{o}F = 0.5455
SG at 60 ^{o}F = 0.7057
Hot GPH = 47090
Naphtha GPH = 15721.3
F = 3.08
N = 12
Factor = 37

From fractionation criteria correlation ASTM gap = 42 ^{o}F. Desired ASTM gap = 25 ^{o}F. Therefore, criteria is satisfied.

Next we present the summary of the energy balances carried out for the heat balance envelope presented in Figure 4.6.

Stream			API	MeABP	K	T	lb/hr	Btu/lb	mmBtu/hr
IN									
Crude		V+L					637798.6		293.3132
Steam		V				450	75425.0	1262.6	95.2316
Tray 34 overflow		L	45.5	408	11.9	321.83	X	162.3	162.3X
Total							713223.6		388.5448+162.3X
OUT									
O/F to tray 34		V	45.5	408	11.9	325.8	X	293.0	293X
Vapor to tray 34		V	57.2	315	12.25	325.8	183636.6	302.4	55.5317
Steam to tray 34		V		PP=	22.1	325.8	75425.0	1203.4	90.7694
LGO		L	35.7		11.6	435.4	80289.9	225.5	18.1059
HGO		L	32.2		11.55	526.9	44163.4	282.1	12.4582
Residue		L	14.1		11.5	686.6	329710.9	365.4	120.4773
PA									84.9965
Total							713225.8		382.3391+293X

Solving for X we get X = 47479.03 lb/hr which corresponds to 7129.74 cold GPH. **Since we got positive flow rates, we proceed towards applying fractionation criteria based on correlation data presented in Table 4.5.**

SG at 326 = 0.64
Hot GPH = 8905.8
Kero + Naphtha product flow rate = 29312.5
F = 0.3
N = 11
Ordinate = 3.3

From product TBP data obtained earlier, $\Delta T_{50\%}$ TBP = 135 ^{o}F.
ASTM gap from Table 4.5 = -28 ^{o}F
Desired ASTM gap = -10 ^{o}F

This time we failed only in obtaining the desired ASTM gap. We next consider a reflux ratio of 2 and repeat all calculations. Hopefully, this time we should satisfy all criteria desired. Therefore, we attempt to be more careful to all calculations desired.

Tower top temperature calculations:

Moles of overhead product = 898.038 moles/hr
Total moles HC in the overhead vapour = 2694.114

Total moles of steam in the overhead vapour = 3704.167

PP_{HC} in overhead vapour = 12.50573 psia 0.85073 atm

To find the tower top temperature, assume initially a value of 250^{o}F, and then conduct dew point calculations as given under.

Table: Calculation of Tower Top Temperature

Component	Mid BP	Vol % on crude	S.G	Wt Factor	Mole Fraction	P^S	K	X=y/K
1	75	0.5	0.638826	0.319413	0.024833	12.4	14.57572	0.001704
2	125	1.5	0.665882	0.998824	0.077653	6	7.052766	0.01101
3	160	2.5	0.698765	1.746914	0.135813	3.8	4.466752	0.030405
4	195	3.5	0.723785	2.533248	0.196946	2.1	2.468468	0.079785
5	235	2	0.746702	1.493404	0.116104	1.36	1.598627	0.072627
6	265	3	0.766938	2.300813	0.178876	0.7	0.822823	0.217393
7	295	2.5	0.771117	1.927793	0.149875	0.56	0.658258	0.227685
8	320	2	0.771117	1.542234	0.1199	0.35	0.411411	0.291436
Total				12.86264	1			0.932045

$K_2 = K_1 * \Sigma X =$ 0.383454

Vapour Pressure corresponding to K2 = 0.326216 Atm

From Maxwell's correlation, we get tower top temperature = 250 ^{o}F. Next, we proceed towards condenser duty calculations and total pump around duties. A summary of the calculations is presented below:

Stream		K	T	lb/hr	Btu/lb	mmBtu/hr
IN						
Naphtha	V	12.75	250	89776.86	269	24.1500
Reflux	V	12.75	250	179553.7	269	48.3000
Steam	V		250	66675	1167.9	77.8697
Total				336005.6		150.3197
OUT						
Naphtha	L	12.75	100	89776.86	53	4.7582
Refluxes	L	12.75	100	179553.7	53	9.5163
Water	L		100	66675	67.3	4.4872
Cond duty					165	131.5579
Total				336005.6	100	150.3197
Total pump around duty=			**75.0986**	mmBtu/hr		

Next, we determine the reflux overflow rate from the tower top tray. Calculations are summarized below:

Stream		K	T	lb/hr	Btu/lb	mmBtu/hr	
IN							
Naphtha	V	12.75	266	89776.86	275.5	24.73352	
O/Flow	V	12.25	266	x	275	274.5x	
Steam	V		266	66675	1175.7	78.3898	
Total				156451.9+x		103.43+274.5x	
				156451.9		103.1233	
OUT							
Naphtha	L	12.75	100	89776.86	53	4.758174	
O/Flow	L	12.25	260	x	145	141.8x	
Water	L		100	66675	67.3	4.487228	
Cond duty						131.5579	
Total				156451.9+x		116.011+141.8x	
						140.8033	
x=	289846.1	lb/hr					
	46377.53	GPH					

We next apply the fractionation criteria for naphtha-kerosene products

Cold GPH = 46377.53
Hot GPH = 64606.02
Naphtha product GPH = 15721.35
F = 4.23
Ordinate = 50.8
$\Delta T_{50\%}$ TBP = 157 ^{o}F
From fractionation correlation, ASTM gap = 45 ^{o}F. ASTM gap desired = 25 ^{o}F. Therefore, the criteria is satisfied.

Next we carry out the energy balances for the envelope shown in Figure 4.6. The following table summarizes the energy balances and evaluated liquid reflux flow rate from the LGO draw off tray.

Stream			API	MeABP	K	T	lb/hr	Btu/lb	mmBtu/hr
IN									
Crude		V+L					637798.6		293.3132
Steam		V				450	75425.0	1262.6	95.2316
Tray 34 overflow		L	45.5	408	11.9	321.83	X	162.3	162.3X
Total							713223.6		388.5448+162.3X
OUT									
O/F to tray 34		V	45.5	408	11.9	325.8	X	293.0	293X
Vapor to tray 34		V	57.2	315	12.25	325.8	183636.6	302.4	55.5317
Steam to tray 34		V		PP=	22.1	325.8	75425.0	1203.4	90.7694

LGO		L	35.7		11.6	435.4	80289.9	225.5	18.1059
HGO		L	32.2		11.55	526.9	44163.4	282.1	12.4582
Residue		L	14.1		11.5	686.6	329710.9	365.4	120.4773
PA									75.0986
Total							713225.8		372.4412+293X

From energy balance calculations X = 123206.6 lbs/hr which corresponds to 18501.4 GPH.

We next apply the fractionation criteria

LGO cold GPH = 18501.4 GPH
SG at 326 ^{o}F = 0.64
Hot GPH = 23110.49
Kero + Naphtha product GPH = 29312.5

Factor F = 0.79
Ordinate = 8.7.
$\Delta T_{50\%}$ TBP = 135 ^{o}F
ASTM gap from fractionation criteria correlation = -7.5 ^{o}F
ASTM gap desired = -10 ^{o}F
Since the obtained ASTM gap is higher than the desired ASTM gap, we accept this value. **Since both flow rates and fractionation criteria are satisfied, we conclude with this value.**

After observing the sequence of calculations, one may think that its better to force the ASTM gap criteria first and then determine the tower top temperature. One may come out with an alternate method of obtaining the desired criteria satisfied. However, what we presented here is more relevant and suitable as **it enabled a systematic exploration of various associated parameters. The procedure outlined will provide an outlook for the design engineer to explore and understand the intracacies involved in pump around duties. Also, it can be observed that we may have slightly approximated the reflux ratio to be on the higher side. A very accurate evaluation of the reflux ratio will provide a value of about 1.9. For all future calculations, we use the values generated for a reflux ratio of 2 only.**

We next present the discussion for the evaluation of top pump around duties. This calculation is carried out only after the desired criteria are satisfied.

4.14 Estimation of Top and Bottom Pump Around Duty

The estimation of TPA duty is tricky. Very fine estimation of vapor and liquid enthalpies is required for the case. In many text books, TPA duty estimation is not presented including the famous Jones and Pujado (2006) book. In this lecture notes, we present a concise presentation of the procedure for the estimation of top pump around duty.

The top pump around duty is estimated using the heat balance envelope presented in Figure 4.7. The envelope considered is from tower top section to the LGO draw off tray along with the Kerosene and LGO side strippers. For the chosen heat balance envelope, the incoming streams enthalpy terms are itemized as:

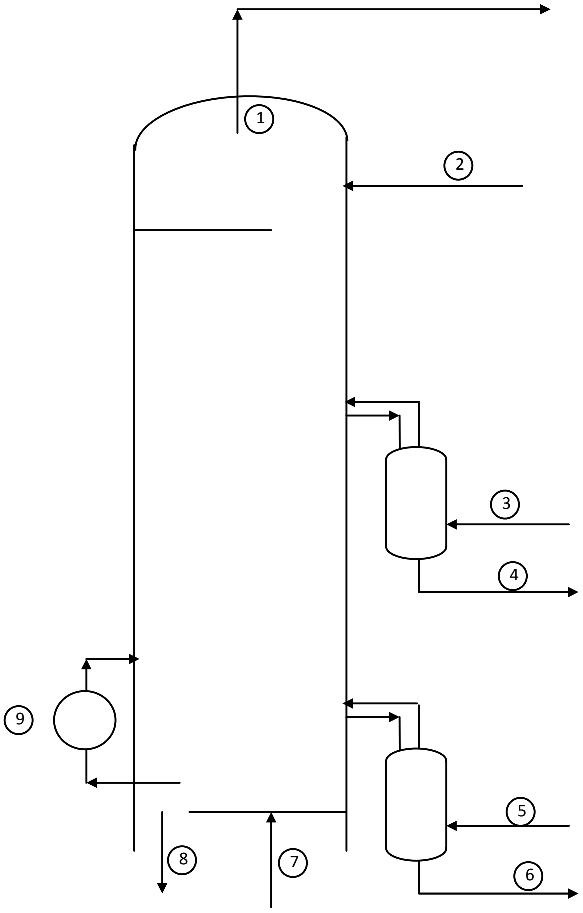
<table>
<tr><td rowspan="9"> </td><td>①
Product (Naphtha) Vapor + Steam (at PP) + Vapor Reflux at Tower Top temperature</td></tr>
<tr><td>②
Naphtha cold reflux at 100 ^{o}F</td></tr>
<tr><td>③
Fresh steam entering Kero SS</td></tr>
<tr><td>④
Kero product</td></tr>
<tr><td>⑤
Fresh steam entering LGO SS</td></tr>
<tr><td>⑥
LGO product</td></tr>
<tr><td>⑦
Vapor approaching LGO Draw off tray. Consists of Product (Naphtha + Kero + LGO) vapor + Steam (Residue steam) + Vapor reflux.</td></tr>
<tr><td>⑧
Liquid reflux leaving LGO draw off tray.</td></tr>
<tr><td>⑨
Top pump around exchanger duty.</td></tr>
</table>

Figure 4.7: Heat balance envelope for the estimation of top pump around duty

a) Hydrocarbon product vapors (Naphtha + Kerosene + LGO) entering tray 34 (LGO draw off tray). By assuming the temperature of these vapors, their enthalpy can be estimated.

b) Steam in vapors entering tray 34 (LGO draw off tray). Once again by assuming the partial pressure of the steam (from Draw off temperature calculations), and assumed temperature of the consolidated vapor stream entering the tray 34, using steam tables, the enthalpy contributed by the steam in vapors entering tray 34.

c) Cold naphtha liquid reflux entering tray 1. The enthalpy of this stream is known from previous calculations.

d) Reflux vapor stream entering tray 34. Since consolidated vapor stream temperature is known (assumed), the enthalpy of the reflux vapor entering tray 34 is known. However, its flow rate is not known. **The vapor reflux flow rate is determined from the fractionation criteria for LGO – HGO product combination which will give hot GPH reflux and then this is eventually converted to lbs/hr.** Therefore, vapor reflux stream enthalpy is known.

The outgoing streams for the chosen heat balance envelope are itemized as:

a) Reflux liquid leaving tray 34. This is at LGO draw off tray temperature. With the temperature known, enthalpy can be estimated and heat lost through this stream can be estimated.
b) Consolidated vapor stream leaving tray 1 at its temperature. The consolidated vapor stream consists of naphtha reflux vapor + naphtha product + steam. The consolidated vapor stream enthalpy data need not be calculated as it is available as a total enthalpy leaving the heat balance envelope chosen to estimate the condenser duty. Therefore, the stream data is known.
c) Kerosene product. Its flow rate and enthalpy are known from total tower energy balance table.
d) LGO product. Its flow rate and enthalpy are known from total tower energy balance table.
e) Top pump around duty. This is unknown and should be estimated from the energy balance.

By equating total energy out and total energy coming in, we can estimate the top pump around duty. **The most tricky issues in these calculations is the vapor and liquid enthalpy values estimation. Since enthalpy estimation is a function of MEABP and K value, it needs to be assumed. For instance, since hydrocarbon vapors consist of naphtha + kerosene + LGO products, a rough estimation of MEABP and K would be based on their cut ranges and average values. Similarly, for the liquid stream, the LGO product MEABP and K values can be assumed. However, the missing data is for the reflux liquid. Since reflux liquid could significantly contribute to the average properties, in reality, one needs to adopt an iterative procedure where by the average enthalpies of the hydrocarbon vapor and liquid are updated in every instance to obtain a converged value of the pump around duty.** For sample calculations, we can ignore the iterative approach and do calculations for one forward step.

We next present an illustrative example for the estimation of TPA duty.

Q 4.13: For the Ecuador crude oil CDU problem assuming data from solutions generated in Q 1 – 10, estimate the top pump around exchanger duty.

Solution:

First we apply fractionation criteria for LGO-HGO combination to estimate the liquid and vapor reflux flow rates (lbs/hr).

From TBP data evaluated for LGO and HGO, $\Delta T_{50\%} = 575 - 505 = 70\ ^oF$

ASTM gap desired = -35 ^oF

From Graph, Ordinate – 5

No. of plates = 11

Hot GPH = 5/11 x (15312.5+14000+11375) = 18494.31 GPH.

Draw off temperature = 537 ^{o}F

Vapor stream temperature = 547 ^{o}F

Liquid stream temperature = 542 ^{o}F

Liquid API = (6500*0.8463+3500*0.8645)/(6500+3500) = 0.85267

Vapor API = (8750*0.702995+6500*0.8463+8000*0.80387)/(8750+6500+8000) = 0.7777684

Steam entering vapor = 48825+3062.5 = 51887.5 lbs/hr

Fresh steam entering the envelope = (315.97+505.55)*18 = 14787.36 lbs/hr

Hydrocarbon vapor flow rate = 89776.85+93859.7+80289.87 = 263926.42 lbs/hr

LGO SG = 0.846337

Hot LGO SG = 0.71

Cold GPH = 18494.31 x 0.71/0.846337 = 15515.056 GPH = 109380.96 lbs/hr

Hydrocarbon vapor enthalpy at T = 547 ^{o}F, MEABP = 340 ^{o}F and K = 11.5 (Assumed values) = 411.3 Btu/lb. Hydrocarbon liquid enthalpy at T = 542 ^{o}F, MEABP = 300 ^{o}F and K = 12 (Assumed values) = 340 Btu/lb.

We next apply the energy balances for the envelope presented in Figure 4.7. For the envelope, entering streams enthalpies are:

a) Hydrocarbon vapors = 263926.2 x 411.3/1000000 = 108.55 mmBtu/hr
b) Steam fresh enthalpy = 14787.36 x 1262.6/1000000 = 18.67 mmBtu/hr
c) Steam enthalpy in vapor at a partial pressure of 1.2 bar and vapor temperature = 1304.2 Btu/lb. Thereby, steam enthalpy = 51887.5 x 1304.2/100000 = 67.7 mmBtu/hr
d) Enthalpy of told cold naphtha reflux stream at 100 ^{o}F from previous heat balance tables = 4.7582 mmBtu/hr
e) Hydrocarbon vapor reflux = 109380.96 x 411.3/100000 = 44.98 mmBtu/hr

Total energy in = 244.648 mmBtu/hr

For the envelope, leaving stream enthalpies are

a) Hydrocarbon liquid reflux = 109380.5 x 340/1000000 = 37.18 mmBtu/hr
b) Vapor enthalpy at tower top temperature (250 ^{o}F) = 150.32 mmBtu/hr
c) LGO product enthalpy from overall tower energy balance table = 18.11 mmBtu/hr
d) Kerosene product enthalpy from overall tower energy balance table = 15.0467 mmBtu/hr

Total energy going out = 220.66 mmBtu/hr. Therefore, Top pump around duty = 244.648 – 200.66 = 23.98 mmBtu/hr. Bottom pump around duty = 75.0986 – 23.98 = 51.11 mmBtu/hr.

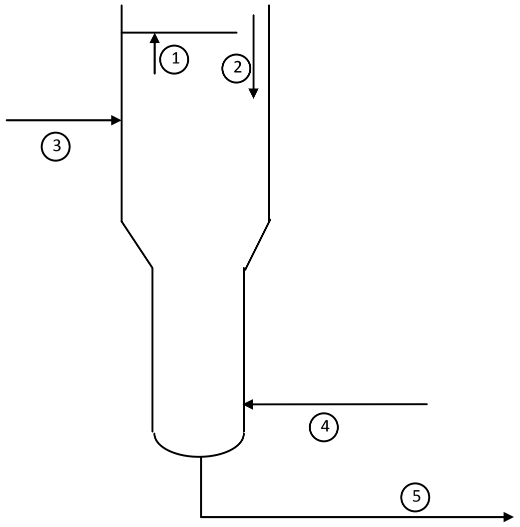
	① Product Vapor + Steam (at Partial pressure) + Vapor reflux
	② Liquid Reflux
	③ Crude (Vapor + Liquid)
	④ Fresh steam
	⑤ Residue product

Figure 4.8: Heat balance envelope for the estimation of flash zone liquid reflux rate.

4.15 Estimation of flash zone liquid reflux rate

The flash zone liquid reflux rate is determined by conducting the energy balance across the heat balance envelope taken from below the HGO draw off tray and for the flash zone (Figure 4.8). For this heat balance envelope, the incoming streams are crude (V +L), fresh steam entering at the bottom most tray and liquid reflux stream. The outgoing streams in the envelope are product vapor stream consisting of Naphtha + Kerosene + LGO + HGO, steam in vapor at its partial pressure of the draw off tray pressure, reflux vapor stream and residue product stream. Of these, the liquid reflux stream temperature is taken as the HGO draw off tray temperature. The vapor stream temperature is taken as 5 ^{o}F plus the liquid reflux temperature. At these temperatures, from Maxwell's correlation, approximate enthalpy values can be obtained. Eventually, one can determine the unknown reflux liquid/vapor mass flow rate. The following illustrative example summarizes the calculations for Ecuador crude oil.

Q 4.14: For the Ecuador crude oil CDU design problem, determine the reflux liquid stream flow rate at the flash zone.

Solution:

The heat balance taken for the heat balance envelope taken as Figure 4.8 is presented below:

Stream			API	MeABP	K	T	lb/hr	Btu/lb	mmBtu/hr
IN									
Crude		V+L					637798.6		293.3132
Steam		V				450	48825.0	1262.6	61.6464
Liquid Reflux		L	32.2		11.55	537	X	282.1	282.1X
Total							686623.6		354.9596 +282.1X
OUT									
Product vapor		V	47.8	426	12.15	542	308089.9	415.0	127.8573
Steam in vapor		V		PP=19.1		542	48825.0	1304.0	63.6678
Reflux vapor		V	32.2	575	11.55	542	X	404.0	404X
Residue		L	14.1		11.5	686.6	329710.9	365.4	120.4773
Total							686625.8		312.0024 +404X

From energy balance table, X = 353277.9 lbs/hr which corresponds to a GPH of 48939 for an assumed HGO SG of 0.8643 for the liquid reflux stream.

Sieve & Valve tray flood line		Sieve and valve tray design		Bubble cap trays (flood)	
Tray spacing (inches)	K_f	Tray spacing (inches)	K_f	Tray spacing (inches)	K_f
10.1	309.1	13.4	505.7	9.9	151.5
10.6	372.4	14.2	580.7	11.3	226.8
11.2	441.5	15.0	635.5	12.4	304.7
12.0	507.8	16.1	713.5	13.2	359.6
13.2	605.9	17.2	774.3	14.7	480.7
14.7	712.7	18.7	835.3	15.6	538.5
16.2	828.1	20.1	902.0	17.2	596.7
17.4	909.0	22.1	980.4	18.4	643.3
18.9	964.3	23.3	1032.7	20.0	712.9
19.9	1010.7	25.0	1082.3	23.4	786.4
21.7	1063.3	26.4	1117.5	25.1	816.1
23.7	1116.0	28.5	1155.9	28.8	884.0
25.0	1156.8	30.1	1185.5	30.3	904.9
27.1	1212.4	32.2	1203.9	32.2	917.5
28.7	1256.3	35.0	1225.5	33.6	926.9
30.1	1294.4				
30.9	1317.7				

Table 4.6: Variation of K_f (Flooding factor) for various tray and sieve specifications.

4.16 Estimation of column diameters

A critical observation of vapor and liquid flow rates estimated at various sections of the CDU indicate that column diameters will different for

a) Tower top section
b) LGO draw off tray
c) HGO draw off tray
d) Section below the Flash zone

Since HGO draw off tray provides maximum vapor flow rates, HGO draw off tray diameter will be maximum. Therefore, for operational convenience, the CDU is usually designed as a column with three different diameters. These refer to diameters of the tower top section, HGO draw off tray, section below the flash zone.

For these sections, the column diameter calculations involve the following steps:

a) Calculate the consolidated liquid flow rate on the tray.
    a. This will be equal to tray 45 liquid reflux rate for the tower top section.
    b. This will be equal to liquid reflux + pump around stream flow rate + LGO product stream flow rate for the LGO draw off tray
    c. This will be equal to liquid reflux + pump around stream flow rate + HGO product stream flow rate for the HGO draw off tray
    d. This will be equal to the liquid reflux for the section below the flash zone.

In these calculations, the pump around flow rate is estimated using TPA and BPA cooling duties that were estimated from energy balance. The pump around cooling fluid outlet is assumed as 300 ^{o}F to evaluate the pump around flow rate for BPA. For LGO, if desired these calculations can be done by assuming the cooling fluid outlet to be about 150 - 200 ^{o}F. Actually, these values are important design variables as the crude oil is heat integrated with these pump around exchangers.

b) Calculate the consolidated vapor flow rate to the tray
    a. For the tower top section, this will be equal to the tray 45 consolidated vapor including naphtha product vapor + steam vapor + reflux vapor
    b. For the HGO draw off tray, this will be equal to the sum of (Naphtha + Kerosene + LGO + HGO product vapor), residue steam, and reflux vapor.
    c. For the section below the flash zone, this will be equal to the sum of residue steam + reflux vapor.
c) Determine the liquid and vapor densities using suitable approximations. For liquid, SG of various products can be assumed. For the vapor, SG can be estimated using ideal gas law.
d) Use flooding correlation data as those presented in Table 4.6 to determine the K_f value. Ideally, it is assumed that K_f = 1100 for a tray spacing of 24 inches.
e) From known value of K_f determine the maximum allowed vapor mass flux using the expression

$$G_f = K_f\sqrt{(\rho_v)(\rho_L - \rho_v)}$$

f) From the known value of G_f, determine the maximum permissible value of G_f during operation at 80 % flooding value.
g) Determine the diameter of the tray from the 80% flood mass flux and total vapor mass flux on the tray. This is termed as A_v

h) For downcomer sizing, determine first the volumetric flow rate of liquid on the tray using total liquid mass flow rate and liquid density. Eventually, the down comer area is determined by assuming that the permissible value is 0.6 cft/s using the expression:

$$A_d = \frac{v_L}{0.6}$$

i) Assume an additional 20 % area as waste area

j) The total area of the tray is evaluated using the expression

$$Total\ area = Vapor\ area + downcomer\ area\ \times 2 + waste\ area$$

k) Calculate the diameter of the tray using the expression

$$D = \sqrt{\frac{4 \times Total\ area}{\pi}}$$

We next demonstrate the calculations for the design of Ecuador crude oil CDU.

Q 14.15: Determine the diameter of the CDU processing Ecuador crude oil at three different sections namely, tower top section, HGO draw off tray and section below the flash zone.

Solution:

Tower top section

Reflux Liquid =	**289846.1**	lb/hr
Total, G =	**446297.9**	lb/hr
So, G/x =	7501.156	
So, x =	59.49722	Mol. Wt.
ρv =	0.193156	lb/ft^3
SG of liquid =	5.855948	lb/gal
	43.80249	lb/cft
Gf =	3192.544	
A =	174.7423	ft2
Downcomer liquid velocity=0.6 ft/s		
cfs of liquid flow	1.838087	ft^3/s
Aw =	34.94845	
Adc =	3.063478	
	6.126957	2 down comers
At =	215.8177	
D =	**16.57673**	**ft**
	5.052586	**m**

HGO Draw off tray

Liquid reflux	375354.8	lbs/hr
HGO K = 11.75 MEABP = 575		
Enthalpy at 300 ^{o}F	142.9	
Enthalpy at 560 ^{o}F	306.5	Btu/lb
BPA duty	51.11	mmBtu/hr
PA flow rate	312423.3841	lbs/hr
Liquid reflux	375354.8	lbs/hr
HGO product	44163.38	lbs/hr
Total liquid flow	**731941.5641**	lbs/hr
Liquid SG	0.864386072	
GPH of liquid	101653.807	
Reflux vapor	375354.8	lbs/hr
Steam flow rate	48825	lbs/hr
Product vapor	308089.84	lbs/hr
Total vapor	**732269.64**	lbs/hr
Total, G=	732269.64	lb/hr
So, G/x =	6368.462308	
So, x=	114.983744	Mol. Wt.
ρv=	0.387837589	lb/ft3
SG of liquid=	7.20033598	lb/gal
	53.85851313	lb/cft
Gf=	5009.28187	
A=	182.7281981	ft^2
Downcomer liquid velocity=0.6 ft/s		
cfs of liquid flow=	3.775022542	ft^3/s
Aw=	36.54563962	
Adc=	6.291704236	
	12.58340847	2 down comers
At=	231.8572462	
D=	**17.18167815**	**ft**
	5.236975499	**m**

Section below the flash zone

Steam flow rate=	48825	lb/hr
Reflux vapor	352377.9	lb/hr
Total, G=	**401202.9**	lb/hr
So, G/x =	6368.49	
So, x=	62.99812	Mol. Wt.
ρv=	0.212491	lb/ft3
SG of liquid=	7.201702	lb/gal
	53.86873	lb/cft
Gf=	3714.269	
A=	135.0208	ft^2
Downcomer liquid velocity=0.6 ft/s		
cfs of liquid flow=	1.817061	ft^3/s
Aw=	27.00416	
Adc=	3.028434	
	6.056869	2 downcomers
At=	168.0819	
D=	**14.62903**	**ft**
	4.45893	m

In summary, we observe that the diameters of the CDU are 4.45 m at the flash zone, 5.23 m at the HGO draw off tray and 5.05 m at the tower top section. In comparison the values reported by Jones and Pujado (2006) for 30,000 barrels/day capacity of the crude are 2.13 m for flash zone, 3.04 m for tower top section and 4.14 m. The obtained diameters are in good agreement with those obtained from Jones and Pujado (2006) with the fact that for our case both reflux ratio and feed flow rate are high and these do contribute towards the variations in the diameter.

4.17 Summary

In this lecture notes, all CDU calculations have been demonstrated. Many assumptions have been taken for the specific gravities, enthalpies of steam and hydrocarbon etc. These need to be further justified in due course of calculations. After obtaining significant experience in the refinery operation, a design engineer will develop maturity for these assumptions. One of the primary objectives of this section of the lecture notes is to demonstrate the procedure for the selection of naphtha reflux ratio and evaluation of pump around duty for each individual pump around. With these procedures, it is opined that the CDU calculations can be conducted by a process engineer with ease.

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