

Designing and Process Simulation of A Single Column Crude Distillation Unit for Refineries

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Abstract: *There is quick development in the utilization and interest of unrefined petroleum in different modern fields. Thus, the cost of the petrol is ascending because of the stronger-than-anticipated interest for petroleum items. With the accessible information, most of the distillation models used today is only steady state models. This work aimed at a newly designed single column distillation unit which works on both atmospheric and vacuum condition. The procedure for this work was firstly forming a McCabe graph for the theoretical stage of the column. Then based on the partial Douglas equation, the relationship between data variable and yield variable was contemplated. After that, physical and geometrical dimensions of CDU were calculated by Douglas 1988 method and finally a design of a single column was made. Next, a dynamic simulation of CDU in Chemsep software was run and the element conduct of CDU model was mulled over taking into account the outcomes acquired. From the theoretical calculations, it was proven that the single distillation chamber works on both atmospheric and vacuum pressure conditions. As the conclusion says, the dynamic simulation results describe the whole CDU working and a good control strategy. Thus, this work was extremely useful in comprehension dynamic conduct of CDU and used to expand the productivity of refining methodology which additionally builds the yield of the item and decreases the plant setup support cost.*

KEYWORDS: *single column distillation; Douglas method; McCabe; simulation; Distillation designing.*

1. Introduction

Crude oil distillation is the separation of the hydrocarbons in unrefined petroleum into parts in view of their boiling points. It is changed over to petrol, diesel, kerosene, aviation turbine fuel, bitumen, refinery gas and sulphur^[1]. These portions are mixtures containing hydrocarbon compounds whose boiling points exist in a predefined extent.

Subsequently, refining is the starting stage in refining foul petroleum. The fractionation is done in a vast tower that is worked at atmospheric pressure. The tower contains various plates where hydrocarbon gasses and fluids cooperate. The liquids flow down the tower and the gases up. The parts that ascent most elevated in the segment before gathering are called light portions, and those that consolidate on the least plate are called heavy portion. The extremely lightest division is refinery gas, which is utilized as a fuel as a part of the refinery heaters. Next in place of instability come fuel (utilized for making petrol), kerosene, and light and overwhelming gas oils lastly long deposit^[2]. A simulation test system for refining is done by Chemsep Software^[3]. It consolidates the excellent harmony stage section model with non-equilibrium column shows in one simple and instinctive interface. They can be connected in the configuration of unrefined petroleum refining segment in view of the data got from a working unrefined petroleum refining section of a refinery.

The crude separation process involves numerous complex phenomena which must be controlled in its best position. The input variables of rough refining section are typically vitality supply inputs, product flow rates and reflux ratio, while the yield variables are the oil thing qualities, system working execution or the plant advantage^[4]. If specifications of oil products cannot be reached, the oil supply can bring about a few issues in plant administration. Controlling refining section begins by recognizing controlled, load variables and manipulation. Controlled variables are those variables that must be kept up at a definite quality to satisfy section destinations. These variables for unrefined petroleum fractionators ordinarily incorporate composition of product, column temperatures, column pressure and accumulator levels. The objective of this study is to design and simulate single column distillation unit that will works for the distillation of crude at both atmospheric and vacuum pressure to obtain the petroleum product shown in figure 1^[5].

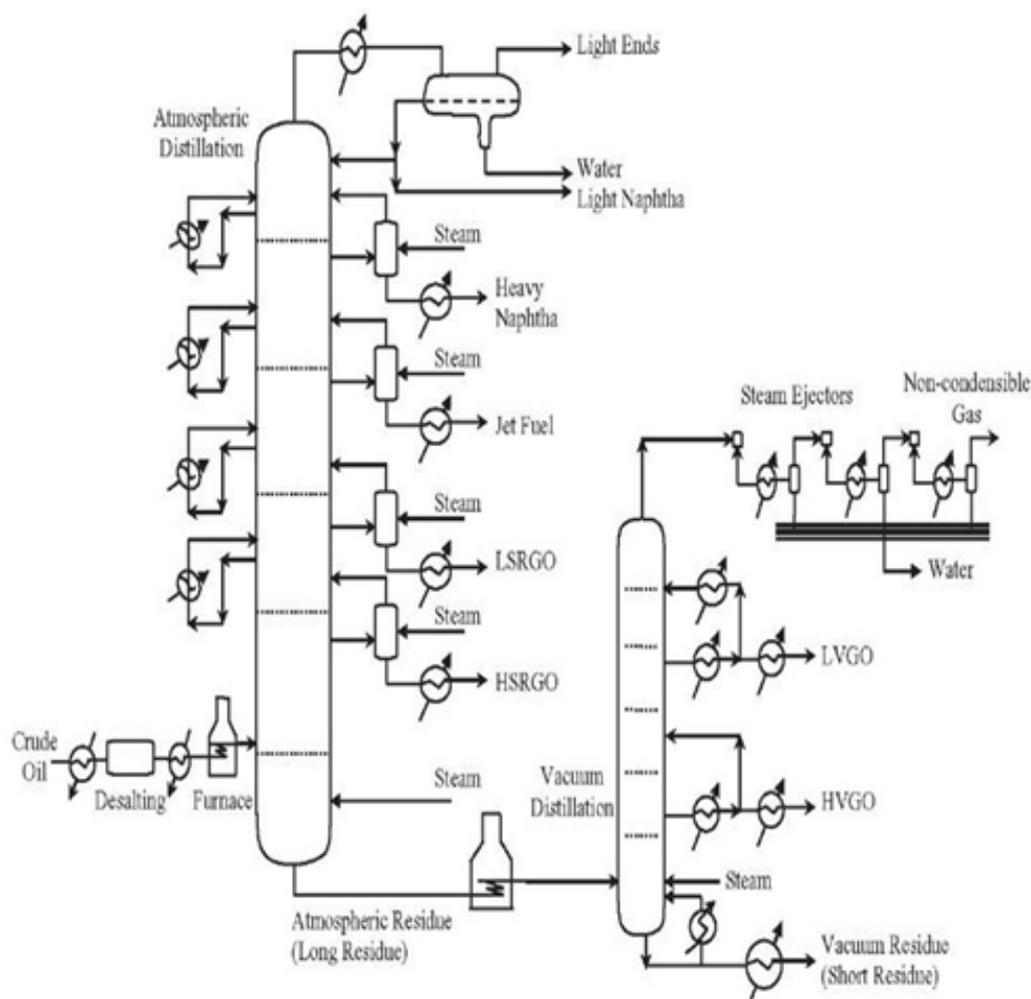


Figure 1. Schematic Diagram of Crude Distillation Unit.

In recent years, the exploration of rough refining procedure was centred around the subject of methodology control and optimization on proposing adjustments to the refining section and the heat exchanger system with a specific end goal to lessen the consumption of energy. Sittig^[6] recommended that with a specific end goal to enhance the proficiency of the refining framework, new internals with higher efficiency should be installed and recommended the use of intermediate reboilers. Bannon and Marple^[7] proposed other column modifications, such as the installation of pump-arounds. Adding pre-flash drums and pre-fractionator column can save energy Harbert^[8] as can adding pump-arounds and reducing operating pressure Fraser and sloley^[9]. Pinch analysis principles guided many researchers Linnloff, *et al*, Dhole and Buckingham^[10] to identify modifications to distillation columns for reducing energy consumption and improving the performance of the system. Liebmann^[11] proposed a two-step approach for improving the execution of refinery refining segments, in

perspective of bits of learning got from crush examination. Another methodology for advancing existing refinery refining sections on pinch analysis principles based and thorough model based simulation was given by Bagajewicz^[12]. An integrated approach regarding designing of the crude oil distillation system was developed Suphanit^[13]. In this study, the shortcut distillation model was used based on pinch analysis and a streamlining system, to create vitality productive grassroots section plan. During the past two decades, there has been a growing awareness among the educated community and mechanical specialists with respect to the control of unrefined petroleum refining segment. After thorough investigation of above researchers, it could be drawn that various approaches used by various scientist were not considering the existing distillation column with details with its huge plant size which must use a capital investment without efficient material use.

The extent of this study is to enhance the expense adequacy and to rearrange the refining procedure

with new CDU. A new design of CDU implies the new process of producing various petroleum products at reduced production cost. A part from that, the scope of this proposed study also include the study on the process and descriptions with this new CDU design which contains a single column that works for both atmospheric and vacuum distillation. Thus, another conservative model of CDU will be produced taking into account the consistent states CDU model. This newly designed model of CDU may reduce the plant size, and though reduces maintenance cost, and which saves time and money on production.

2. Methodology

Actually this design includes a compact model of distillation unit, which contains both atmospheric and vacuum distillation process in a single column. So we need a newly designed distillation column which works for both alternatively. And though, to design a column following parameters we need to calculate: (i) The composition of distillates and feed rate; (ii) The Reflux ratio; (iii) Number of Trays; (iv) Column Height and Diameter.

Determining these according to Douglas method, the column height needs the actual number of trays and which is determined through the McCabe graph by known reflux ratio and the feed rate are summarized below.

2.1. Determination of the Composition of Distillates by Assuming Feed Rate

Let us assume the feed rate with known composition, purity of distillate and bottoms, and the quality of the feed. Performing overall material and component balances to determine the compositions of the distillate and bottoms.^[14]

$$F \cdot Z_F = X_D \cdot D + X_B \cdot B$$

$$F = D + B$$

Where, F = Feed rate of input stream; ZF = Composition of light component in feed; XD = Mole Fraction of light in distillate; XB = Mole Fraction of light in Bottom; D = Total distillate amount; B = Total bottom amount. This number of composition signifies the number of light and heavy distillates to be obtained from the feed, and as when the composition is known, and then the reflux ratio can easily be calculated

2.2. Determination of the Minimum Reflux Ratio

The following assumptions are implied when using the method^[15]: (i) Constant molar overflow,

(ii) Heat effects are negligible, and (iii) For every mole of vapour condensed, another mole of liquid is vaporized. The liquid and vapour leaving the tray is in equilibrium with the vapour and liquid entering the tray. If an equilibrium curve^[16] is not given, draw a Y-X diagram (Y representing the vapour phase and X the liquid). The equilibrium curve can be obtained by relating the relative volatility to the composition of the liquid:

$$Y = \frac{A \cdot X}{1 + X(A-1)}$$

This shows the bubble-point and dew point of a binary mixture at constant pressure. An equilibrium line describes the compositions of the liquid and vapour in equilibrium at a fixed pressure (Figure 2).

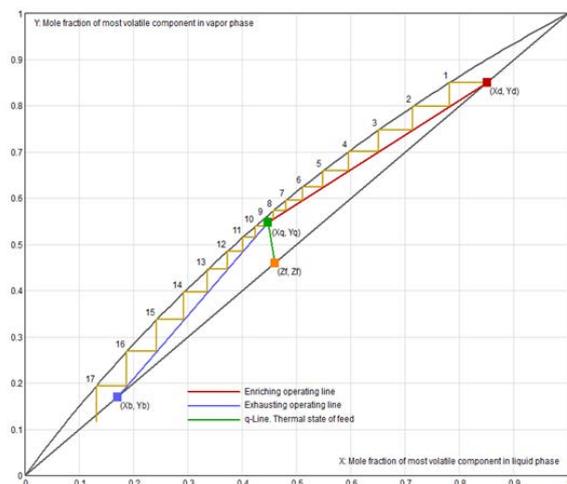


Figure 2. McCabe Equilibrium Curve

The description of the figure 2 is as follows:

The equilibrium line crossing the forty-five degree line is an indication of an azeotropic mixture.

Forty Five Degree Line: Drawing the diagonal line connecting the points (0,0,0,0) to (1,0,1,0). This is your forty-five degree line.

Feed Line (q-Line): The feed line can be constructed by locating the point on the forty-five degree line that corresponds to the feed composition. This point can be extended with a slope of q/(q-1) where q is the feed quality. The feed line can be directly plotted through the following equation:

$$Y = \frac{q}{(q-1)}X - \frac{Z_F}{(q-1)}$$

Upper Operating Line: Draw the operating line for the enriching section. First find the desired top product composition (on the x-axis) and locate the corresponding point on the forty-five degree line. Connect this point to the point where the equilibrium curve and the feed line intersect. This is the upper operating line. The y intercept of this line

is equal to $XD/(R+1)$. The following equation can be used to determine the minimum reflux:

$$R_{\min} = (XD / y \text{ intercept}) - 1$$

Lower Operating Line: Let draw an operating line for the stripping section. First find the desired bottom product composition (on the x-axis) and locate the corresponding point on the forty-five degree line. And drawing a line from the point to the intersection of the equilibrium curve and the feed line. This is our lower operating line.

$$\text{The slope of this line} = (Vb_{\min+1}) / Vb_{\min}$$

Where, Vb is the boil-up ratio.

2.3. Determination of the Actual Reflux Ratio

The actual reflux ratio can be determined by the Douglas method. As the reflux ratio increases, the number of trays and thus the capital cost of the column decreases. However, as a trade-off, an increase in reflux ratio will also increase the vapour rate within the tower, thus increasing expenses such as condensers and reboilers. Most columns are designed to operate between 1.2 and 1.5 times the minimum reflux ratio (as determined above) because this is approximately the region of minimum operating cost. Therefore, based on first estimates, the operating reflux ratio is equated so that:

$$R_{\text{actual}} = R_{\min} * 1.2$$

$$R = L / D$$

2.4. Determination of the Number of Trays

The number of trays can be determined by using the upper and lower operating line and the actual reflux ratio.

Plotting the point $XD/(R+1)$ and drawing a line to the XD . The equation of the upper operating line is:

$$Y = (R/R+1)*X + XD/(R+1)$$

The equation for the lower operating line can be drawn by connecting the XB to the intersection of the feed line and the upper operating line. The equation of the lower operating line is:

$$Y = ((VB+1)/VB)*X + XB/VB$$

Starting from composition of the distillate, a horizontal line is drawn to the equilibrium curve.

This line demonstrates the first tray. From the previous intersection, drop vertically until the upper operating line is obtained. Follow step two to determine next tray. Continue stepping until the liquid composition ends equals the desired bottom composition. The total number of steps is equal to the theoretical number of trays. Then the actual number of trays is determined by taking the quotient of the number of theoretical trays to the tray efficiency. Typical values for tray efficiency range from 0.7 to 0.9^[4]. These values depend on the type of trays being used, as well as the internal liquid and vapour flow rates.

$$N_{\text{actual}} = N_{\text{theory}} / e$$

2.5. Determination of Principal Dimensions of the Column (Diameter/Height)

The tower height can be related to the number of trays in the column. The following formula assumes that a spacing of two feet between trays will be sufficient including additional five to ten feet at both ends of the tower. This includes a 15% excess allowance of space^[4].

$$H_{\text{tower}} = 2.3 N_{\text{actual}} [\text{ft.}]$$

Before we can determine the tower diameter, we need to determine the vapour velocity. The vapour velocity can be derived from the flooding velocity. To limit our column from flooding, we choose a velocity 50-80% of flooding velocity as stated by Douglas.

The diameter of a tower is relatively insensitive to changes in operating temperature or pressure. The main determinant of the diameter is the vapour velocity. The desired vapour velocity is dependent on the limitations of undesired column flooding. This equation allows for a twelve % surplus in area (Douglas, 1998).

$$DT = .0164 (V) * (MG / \rho m)^{0.25}$$

Where, V = Vapour Velocity; MG = molar mass; ρm = Molar density

2.6. Description of detailed process involved in newly designed distillation Unit

The newly designed distillation unit requires a column, furnace, reservoir, pump, stripper and heat exchanger and also a change in refining process; because of a single distillation column.

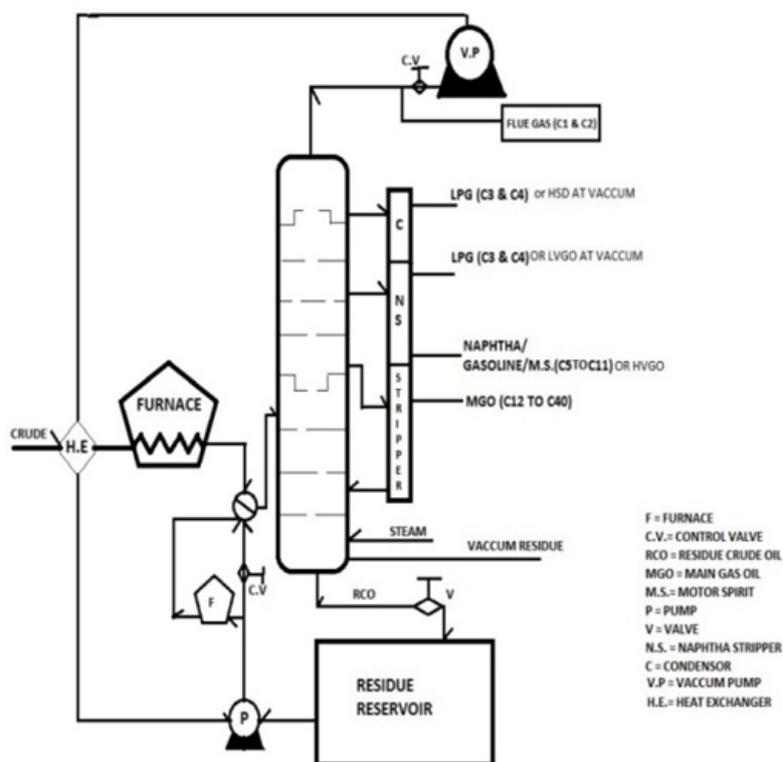


Figure 3. Schematic Diagram of Newly designed Distillation unit

The figure 4 shows about an internal design of the whole CDU column and which contains a major side of sieves trays.

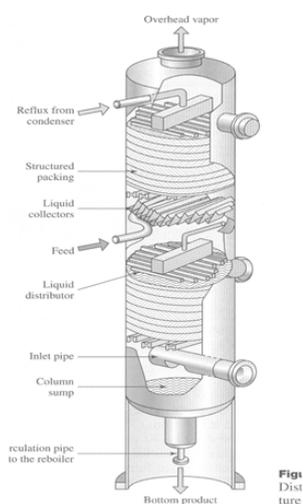


Figure 4. The Column Internal Design and Trays

The sieve trays design is shown in figure 5. The sieve trays are simply metal plates with holes in them. Vapor passes straight upward through the liquid on the plate. The arrangement, number and size of the holes are design parameters. Because of their efficiency, wide operating range, ease of maintenance and cost factors, sieve^[18] and valve trays have replaced the once highly thought of bubble cap trays in many applications.

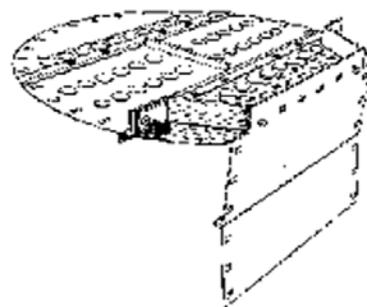


Figure 5. The Design of Sieves Tray

The figure 6 is showing detailed process for distillation through newly designed column.

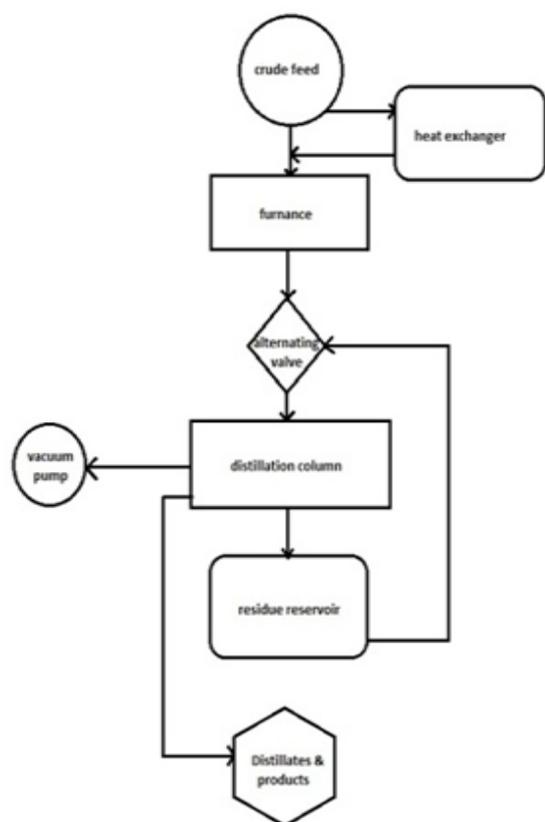


Figure 6. Process Flow Chart

After de-saltation process, the feed is allowed to be heated in the heat exchanger (H.E.) with the overhead distillates. Then, feed is send to the furnace for further heating up to 360 °C. And, all the light distillates are moved to the overhead and the heavy distillates are allowed to move to the lower end of the distillation column.

On heating with steam at atmospheric pressure (760mm of Hg.) all the major distillates are obtained according to their boiling ranges as given in table 1.

Table 1. The boiling point and carbon number ranges of atmospheric distillates

Distillates	Numbers of Carbon	Boiling Range (°C)
Gases	1-4	<30
Gasoline	5-12	30 – 210
Naphtha	8-12	100 – 200
Kerosene	11-13	150 – 250
Diesel, fuel oil	13-17	160 – 380
Atmospheric residue	>20	>380

Then, the atmospheric residues are forced to the residue reservoir, and after filling up the reservoir,

and then the pump automatically sends the residue to the column through an alternating valve.

This technique (alternating valve) cancels the flow of pure feed crude and allows flowing residue to the column at vacuum pressure (684 mm of Hg) i.e. lower than the atmospheric pressure. This process is necessary to avoid cracking and produce distillates as depicted in table 2:

Table 2. The boiling point and carbon number ranges of vacuum distillates

Distillates	Numbers of Carbon	Boiling Range °c
LVGO (light vacuum gas oil)	20-34	>370°c
HVGO (heavy vacuum gas oil)	34-54	370-565°c
Vacuum Residue	>54	503-671°c

Then the residues (VR) are sending to the Coker for formation of petroleum coke through thermal cracking.

2.7. Simulation of the newly Designed Distillation Chamber

Simulation is done by Free licenced Chemsep software v6.98^[14]. Chemsep™ Modelling Separation Processes version 6 software was made by Harry Kooijman and Ross Taylor for making it easy to simulate a newly designed distillation chamber. For simulation of the distillation chamber following parameters should be taken: (i) feed Components, Operation; (ii) Properties of Column (thermodynamic, physical property and reaction); (iii) Feeds (Feed rate and Quantity); (iv) Column Specifications

After, these all parameters are known, the Chemsep software give its result in form of a flow profile table, Mc-Cabe graph, Liquid and vapour component graph and the FUG (Fenske-Underwood- Gilliland Analysis)

3. Results and Discussions

This graphical data are used to plot a McCabe equilibrium graph with X-coordinates taken as mole fraction of most volatile component in liquid phase and Y-coordinates taken as mole fraction of most volatile component in vapour phase.^[17]

Table 3: The graphical data to plot a McCabe graph obtained from Chemsep soft

STAGE	KL	KH	RV	dy*/dx	xL	xH	E- O'Connel	DV (m ² /sec)	DL (m ² /sec)
2	1.719985	0.219832	7.824080	0.128386	0.997429	0.002571	0.444537	0.0000029640	0.0000000041
3	2.227788	0.260619	8.548054	0.119642	0.987357	0.012643	0.436385	0.0000031088	0.0000000045
4	2.629494	0.293832	8.948969	0.121875	0.952198	0.047802	0.432982	0.0000032026	0.0000000047
5	2.858425	0.312317	9.152332	0.140557	0.867158	0.132842	0.431254	0.0000032495	0.0000000049
6	2.981950	0.323530	9.216932	0.194066	0.717005	0.282995	0.430339	0.0000032775	0.0000000049
7	3.091819	0.336885	9.177665	0.332536	0.520133	0.479867	0.429574	0.0000033116	0.0000000050
8	3.315464	0.365706	9.065916	0.664439	0.333979	0.666021	0.427714	0.0000033834	0.0000000053
9	4.116836	0.456785	9.012643	1.271882	0.207418	0.792582	0.420173	0.0000036078	0.0000000059
10	5.434897	0.656308	8.281022	1.513446	0.183924	0.816076	0.423607	0.0000038711	0.0000000068
11	5.434962	0.656494	8.278765	1.515076	0.183764	0.816236	0.423638	0.0000038713	0.0000000068
12	5.435075	0.656826	8.274759	1.517971	0.183481	0.816519	0.423694	0.0000038716	0.0000000068
13	5.435274	0.657415	8.267647	1.523120	0.182979	0.817021	0.423793	0.0000038721	0.0000000068
14	5.435624	0.658465	8.254987	1.532296	0.182090	0.817910	0.423971	0.0000038730	0.0000000068
15	5.436239	0.660352	8.232336	1.548725	0.180516	0.819484	0.424287	0.0000038746	0.0000000068
16	5.437346	0.663789	8.191379	1.578392	0.177725	0.822275	0.424861	0.0000038774	0.0000000068
17	5.439477	0.670223	8.115919	1.632790	0.172779	0.827221	0.425918	0.0000038825	0.0000000068
18	5.444309	0.682827	7.973193	1.735212	0.163997	0.836003	0.427927	0.0000038919	0.0000000069
19	5.458328	0.709014	7.698473	1.935908	0.148416	0.851584	0.431861	0.0000039108	0.0000000069
20	5.507844	0.766125	7.189222	2.346023	0.121267	0.878733	0.439543	0.0000039537	0.0000000071
21	5.667728	0.892009	6.353891	3.165080	0.077862	0.922138	0.453815	0.0000040730	0.0000000075

The above graphical data are plotted and which is shown in figure 7 and it signifies the Forty Five

Degree Line, Feed Line (q-Line), Upper Operating Line, and Lower Operating Line^[16].

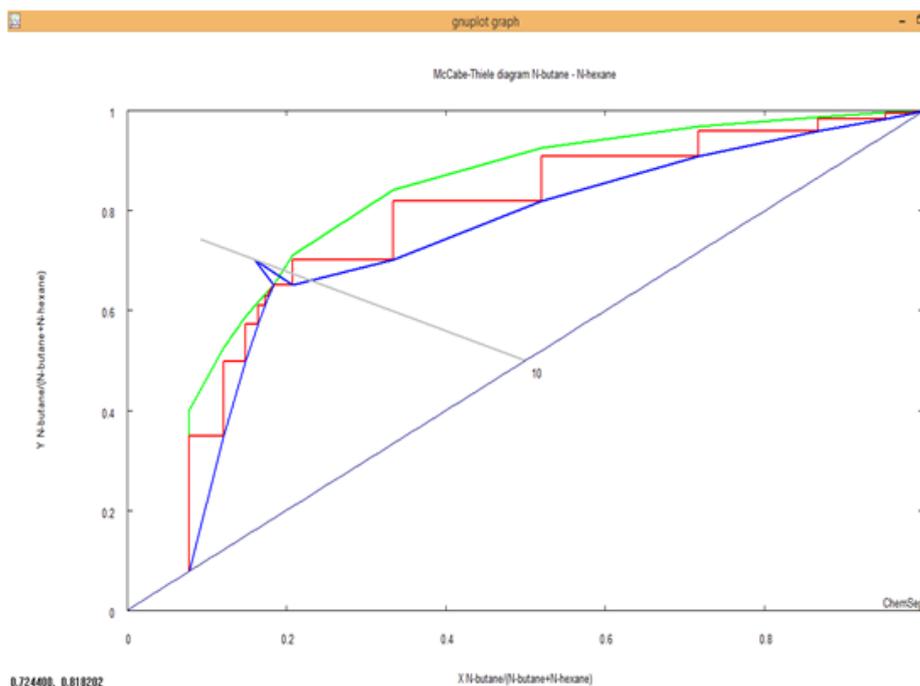


Figure 7. Plotted McCabe Equilibrium Graph by Chemsep Software

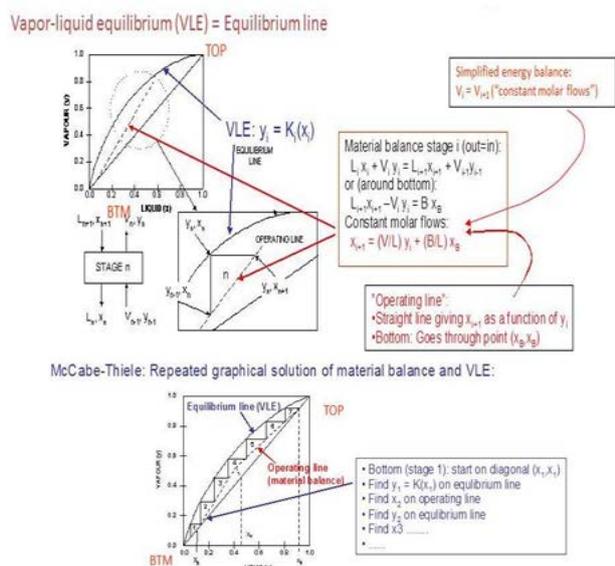


Figure 8. Calibration of McCabe Equilibrium Graph

Now, from the above plotted line all the parameters are determined.

Table 4: The physical parameters of the column

Parameters	Data determined
Mole fraction, X	0.14
Mole fraction, Y	0.2422
Relative volatility, α	1.59
Thermal state, q	0.87
Mole fraction of distillates, Xd	0.85
Mole fraction in Feed, Zf	0.46
Number of feed plate, Nf	10
Overall efficiency, e	0.86
Number of theoretical plates, N_{theory}	19
Mole fraction in bottoms, Xb	0.17
Minimum Reflux Ratio, R_{min}	2.5092

3.1. The Actual Reflux Ratio

The Reflux ratio is the ratio of reflux flow (L) to distillate flow (D), which is also 1.2 times the theoretical reflux ratio.

Also denoted by $R = L/D$

Now, the Actual Reflux ratio, $R = 3.08$

The result determines the amount of reflux flow necessary in ratio with the distillates flow. So that,

the higher molecular weight hydrocarbon refluxes well with the light distillates. And so it decreases the cost shown in figure 7.

3.2. The Actual Number of Trays

The most important thing for designing a distillation column is to determine the number of trays. Number of trays should always be in such a way that the distillates should break into smaller molecules after passing from each tray. And secondly the slots made in the trays should be such that, the distillate flow be in order to the overhead. The actual number of trays can be determined by dividing the theoretical value with the efficiency (e).

Therefore, the actual number of trays:

$$N_{actual} = 22 \text{ trays}$$

Comparing the result of actual number of trays to the number of feed trays shows that feed is actually given from the middle of the fractionating column. And, also the height and diameter are now determined after calculating the actual number of trays.

3.3. Height of the Tower and the Internal Diameter of the Column

As discussed in chapter 3 that the height of the column should not exceed 175 feet's and the diameter should be less than 20 to 30. Taking in account all these limitations the height of the tower is calculated by 2.3 times of the actual number of trays and the internal diameter is in ratio of 5.

Table 5: The geometrical dimensions of the column

Parameters	Data
Height of Tower, H	50 feet
Internal Diameter, D_i	10 feet
Wall Thickness, t	0.3125
Pressure, P	160psi.g.

The above all tables shows all the graphical, geometrical and physical parameters for designing a column that works for both atmospheric and vacuum pressure alternatively.

The 22 trays should be sieves type shown in figure 5 because it increases the efficiency (e) and flow rate.

3.4. Sieve Tray Geometry

Following were the dimensions of various parts of sieve tray:

- (i) Hole diameter: 0.005-0.025 m;

- (ii) Fractional free area: 0.06-0.16 m²;
- (iii) Fractional down comer area: 0.05-0.3 m²;
- (iv) Pitch / Hole diameter ratio: 2.5-4.0;
- (v) Tray spacing: 0.305-0.915 m;
- (vi) Weir height: 0.025-0.075 m.

After designing a whole distillation unit as shown in Figure 2, it was seen that not only it's a compact process but it is also cost effective. Because, the most important parameters increasing the cost of oil refining are the production and maintenance cost; and these two factors were considered to minimise by lowering the steam cost, furnace oil cost and the metallurgical cost to set up this process plant.

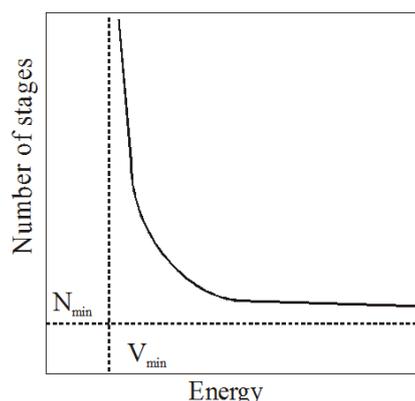


Figure 9. Energy Vs. Number of Trays/Stages

The figure 9 shows that, on increase of number of stages there is a decrease of vapour velocity.

Energy versus number of stages (N)

Trade-off between number of stages and energy

Actual V approaches V_{min} for N approximately 2 x N_{min} or larger, typically:

- 2N_{min} → + 25% V_{min}
- 3N_{min} → + 3 % V_{min}
- 4N_{min} → + 0.3 % V_{min}

Table 6: The Flow profiles of the column simulation

Stage	Temperature (K)	Pressure N/m ²	Flow rates (kmol/s)		Feed	Product	State
			Liquid	Vapour			
1	280.639	101325	65.3643	RR=3.08		21.2222	Vapour
2	287.552	101325	62.3845	86.5865			
3	294.577	101325	61.1772	83.6066			
4	299.053	101325	60.9298	82.3993			
5	301.267	101325	60.7264	82.152			
6	302.582	101325	60.1584	81.9486			
7	304.177	101325	58.7062	81.3806			
8	307.513	101325	54.4912	79.9284			
9	317.743	101325	45.5196	75.7133			
10	329.394	101325	89.586	12.7417	100.0000		
11	329.401	101325	89.5891	10.8082			
12	329.413	101325	89.5946	10.8113			
13	329.435	101325	89.6043	10.8168			
14	329.474	101325	89.6215	10.8265			
15	329.543	101325	89.652	10.8437			
16	329.666	101325	89.7056	10.8741			
17	329.888	101325	89.7996	10.9278			
18	330.299	101325	89.9625	11.0218			
19	331.117	101325	90.236	11.1847			
20	332.975	101325	90.6495	11.4582			
21	338.096	101325	92.1701	11.8716			
22	338.096	101325	BR=0.17	13.3922		78.7778	Liquid

The simulation is checked by Chemsep v6.98 software which also works for McCabe graph. Feed

were assumed as 100 % (n-butane, n-pentane, n-hexane, n-heptane and n- octane) on the 10 trays. And the result is 95.0% profile recovery of n-butane

and n-hexane at 101.35 KN/m², which defines as the crude distillation column works fine for

fractionation. Shown in table 6 and figure 10.

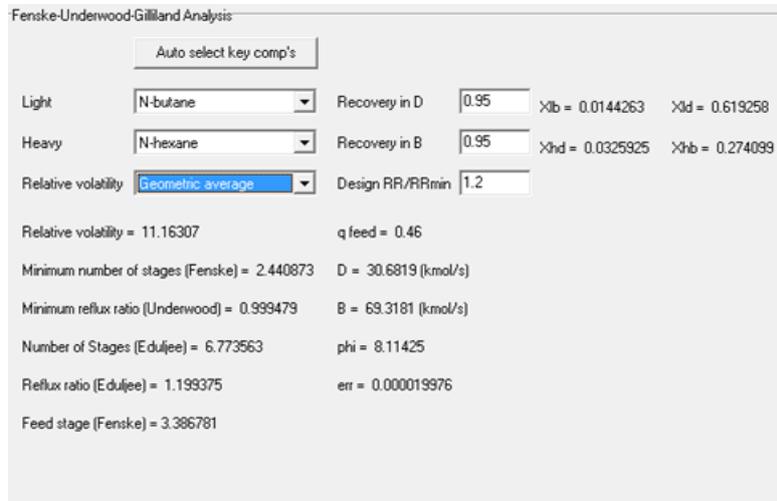


Figure 10. Physical Parameter Determined from Chemsep v6.98

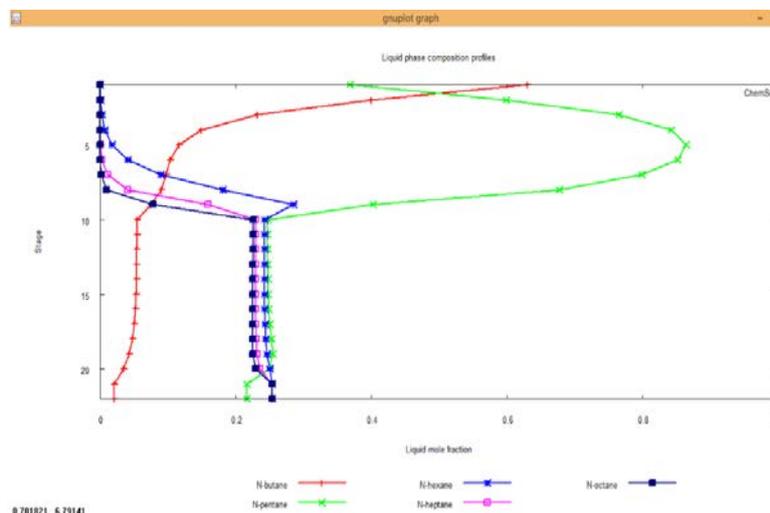


Figure 11. Liquid Phase Composition Profile by Chemsep Software

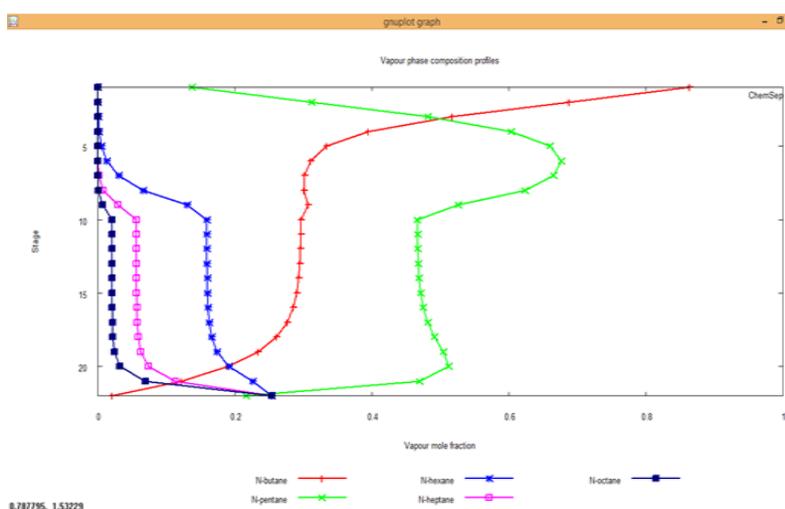


Figure 12. Vapour Phase Composition Profile by Chemsep Software

From the figures 11 and 12 it is determined that n-butane is obtained at vapour phase in the overhead

product and n-hexane is obtained at liquid phase at a bottom distillate product with 95% recovery.

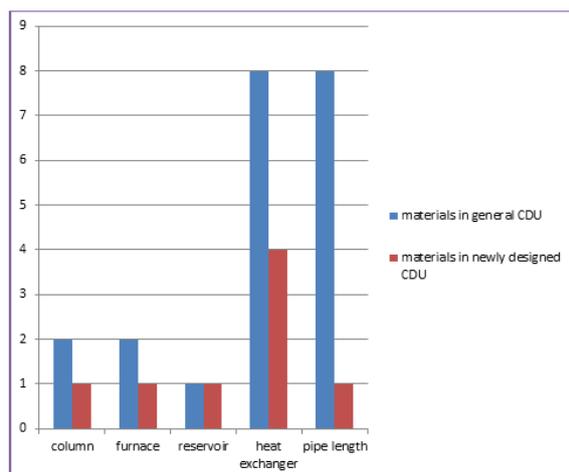


Figure 13. Materials Used to Setup a Plant with Newly Designed CDU in Comparison with the Design used now.

The above graph shows that the difference in use of material to setup this distillation unit in comparison with the unit used now. From this graph we can estimate the cost difference to setup a plant with this newly designed distillation process.

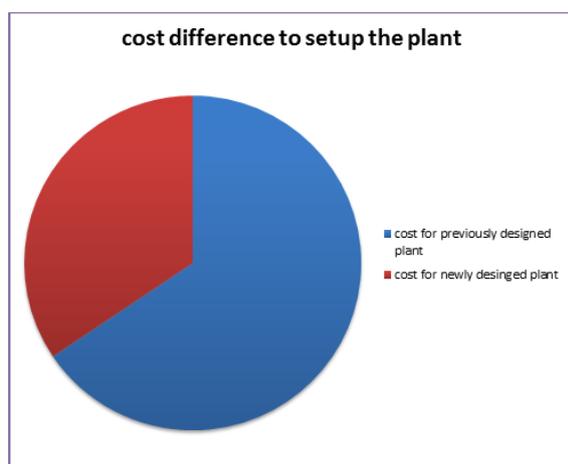


Figure 14. Cost Effectiveness to setup a plant with this new design

From the above graph the difference defines that the production cost per day is also reduced due to the reduction of material cost by 13: 25.

4. Conclusion

This work speaks to the consequences of recently outlined refining unit prepare that will really lessen the maintenance cost, plant setup cost, and the production cost. This in turn can help in refinery planning and scheduling. This work has considered a grass root design of crude distillation column consisting of a, furnace, heat exchanger, a single column (works for both vacuum distillation and atmospheric distillation) with minimal number of stages/plates, pump and column stripper. As

always, the execution of any plant is broke down on the benefits made by the plant, so here the target capacity considered is amplification of benefits made every day. Inferences drawn from above works: (i) The column designed for the distillation process can be used for both atmospheric and vacuum pressure distillation process and which gives 95.0% flow profile. (ii) Material and metals used to setup a plant is less than the current setup in refineries of 13: 25 ratio. (iii) Maintenance cost is also reduced as the plant size reduced by 47 % of the current refinery plant. (iv) A single distillation column is used for both atmospheric and vacuum pressure so furnace oil cost and numbers of heat exchangers are also reduced. So in total it is clear that the overall production cost per day will be reduced per day as the new refining process is simple and compact.

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6. Nomenclature

F	Feed rate of input stream;
ZF	Composition of light component in feed;
XD	Mole Fraction of light in distillate;
XB	Mole Fraction of light in Bottom;
D	Total distillate amount;
B	Total bottom amount.
R_{actual}	Actual Reflux Ratio.
R_{theory}	Theoretical Reflux Ratio.
N_{actual}	Actual Number of Trays.
N_{theory}	Theoretical Number of Trays.
e	Efficiency.
H_{tower}	Height of the Column.
DT	Diameter of the Tower.
V	Vapour Velocity;
MG	Molar mass;
ρ_m	Molar density.
Vb	Boilup ratio.
X	Mole Fraction (Liquid).

Y	Mole Fraction (Vapour).
α	Relative Volatility.
Q	Thermal State.
D_1	Internal Diameter.
T	Wall thickness.
P	Pressure

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