RECOVERY OF PROPANE FROM REFINERY GASES

Submitted in partial fulfillment of the requirement for award of degree in

M. Tech (Gas Engineering)

2008 - 2010

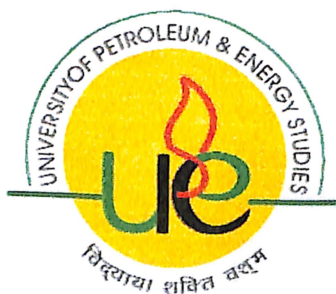By

SANIPINA KASI VISWANADHAM

R030308008

Under the guidance of

Mr. CH. VARA PRASAD
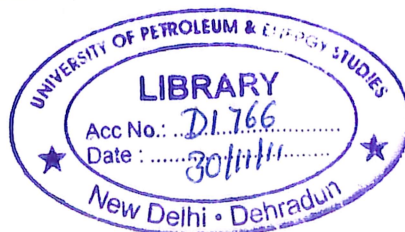
Assistant Professor

COLLEGE OF ENGINEERING STUDIES

UNIVERSITY OF PETROLEUM AND ENERGY STUDIES

DEHRADUN

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I am whole hearted to thank everybody who has encouraged me in completion of this project. The useful experience I gained from this project duration was very much helpful for my future career.

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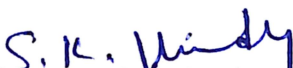
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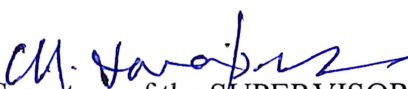
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SANIPINA KASI VISWANADHAM

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CANDIDATE'S DECLARATION

I hereby declare that the work which is being presented in the thesis entitled "RECOVERY OF PROPANE FROM REFINERY GASES" by "KASI VISWANADHAM SANIPINA" in partial fulfillment of requirements for the award of degree of M. Tech. (Gas Engineering) submitted in the Department of Chemical Engineering at UNIVERSITY OF PETROLEUM & ENERGY STUDIES, DEHRADUN is an authentic record of my own work carried out under the supervision of Asst Prof. CHINTA VARA PRASAD. The matter presented in this thesis has not been submitted by me in any other University / Institute for the award of M. Tech Degree. *Due to the confidentiality of data, the original name, location and identity of the data has been changed.*

Signature of the Student

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

Signature of the SUPERVISOR (S)

The M. Tech Viva –Voce Examination of SANIPINA KASI VISWANADHAM has been held on 08/05/2010 and accepted.

Signature of External Examiner
08/05/2010

Signature of Head of Department.

CERTIFICATE

This is to certify that the project work entitled "**RECOVERY OF PROPANE FROM REFINERY GASES**" being submitted by **Mr. SANIPINA KASI VISWANADHAM** (R030308008), in partial fulfillment of the requirement for the award of the degree of **Master of Technology [Gas Engineering]** in University of Petroleum and Energy Studies-Dehradun, is a bonafide project work carried out by him under my guidance.

Dehradun

Date:

Mr. CH.VARA PRASAD

Assistant Professor
UPES - Dehradun

ABSTRACT

The main objective of the project is to design absorber stripper system for recovery of propane from refinery gases. Recovery of propane from petroleum refinery in more quantities leads to meet the demand of the petrochemical industries and other purposes mainly LPG. The uncondensed gases from various refinery units like catalytic cracking, catalytic reforming and coking are processed to separate the propane from other gases.

There are several processes developed to recover the propane from the refinery gases. But the absorption process is the most widely used one because of its low operating cost and flexibility of tailoring solvent composition to suit gas compositions make this process one of most commonly selected.

In this project the absorber stripper system will be designed and simulated by using simulation software like Hysis.

And also the mathematical modelling and simulation of absorber using Kremser Brown Sherwood Method and stripper using FUG method will be carried out. These analytical results will be compared with the simulation software results

Nomenclature

Absorption

A_i	Absorption factor
C_p	Specific heat of water, kcals/kg
D_e	Diameter of the tower, m
G	Total gas flow rate for tower, kmols/hr
H	Enthalpy of vapor, Btu/lb
h	Enthalpy of liquid, Btu/lb
K	Equilibrium condition
L	Molar flow rate of liquid, kmols/hr
L_s	Flow rate of absorbent, kmols/hr
L_b	Flow rate of rich oil, kmols/hr
L_{sa}	Lost solvent in absorbent, kmols/hr
Mw	Molecular weight
N	Total number of moles in the feed, kmols/hr
N_i	Total number of moles in feed of component i, kmols/hr
N_{ai}	Number of absorbed moles of component i, kmols/hr
N_{si}	Number of unabsorbed moles of component i, kmols/hr
N_p	Number of equilibrium trays
P	Total pressure, atm
p	Partial pressure, atm

Q	Heat content, kcals/hr
S	Stripping factor
S_{col}	Surface area of the column, m^2
S_d	Surface area of the down comer, m^2
S_{st}	Amount of steam, kg/hr
t	Temperature, $^{\circ}C$
V	Molar flow rate of rich oil, m^3/sec
v	Velocity in the down spout, m/sec
W	Water required, kg/hr
x_i	Mole fraction of liquid phase components
y_i	Mole fraction of gas phase components
Y_{n+1}	Flow rate of propane in the entering rich gas, kg/hr
Y_n	Flow rate of propane in dry gas, kg/hr
ρ_L	Density of liquid, kg/m^3
ρ_v	Density of gas, kg/m^3
ψ_i	Degree of absorption
λ	Latent heat of vaporization, kcal/kg

Stripping

B	Molar flow rate of bottoms, kmols/hr
B_i	Bottom molal flow of component i
C_p	Specific heat of water, kcals/kg
D	Molar flow rate of distillate, kmols/hr
D_i	Distillate molal flow of component i
D_e	Diameter of the tower, m
F	Mass flow rate of feed, kmols/hr
F_i	Molal flow of component i in feed, kmols/hr
H	Enthalpy of vapor, Btu/lb
h	Enthalpy of liquid, Btu/lb
L	Molar flow rate of liquid, kmols/hr
K	Equilibrium condition
Mw	Molecular weight
m	Constant
N	Number of trays
N_m	Number of stages
N_R	Number of trays in rectifying section
N_S	Number of trays in stripping section
N_p	Number of equilibrium trays
n	Constant

P	Total Pressure, atm
p	Partial pressure, atm
Q	Heat content, kcals/hr
Q_C	Heat content of Condensate, kcals/hr
Q_B	Heat content of bottom, kcals/hr
q	Quality feed of feed
R_{min}	Minimum reflux ratio
R_{actual}	Actual reflux ratio
S	Stripping factor
S_{col}	Surface area of the column, m^2
S_d	Surface area of the down comer, m^2
t	Temperature, °C
V	Molar flow rate of gas, kmol/hr
v	Velocity in the down spout, m/sec
W	Water required, kg/hr
x_i	Mole fraction of liquid phase components
x_{Bi}	Bottom molal flow of component i
x_{Di}	Distillate molal flow of component i
x_F	Mole fraction of liquid phase components in feed
y_i	Mole fraction of gas phase components
Y_{n+1}	Flow rate of propane in the entering rich gas, kg/hr

Y_n Flow rate of propane in dry gas, kg/hr

ρ_L Density of liquid, kg/m^3

ρ_v Density of gas, kg/m^3

λ Latent heat of vaporization, kcal/kg

α Relative volatality

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1.1 INTRODUCTION

Propane is one of the most useful gases obtained in the petrochemical industry. It has variety of commercial, industrial, house hold and agricultural applications. Large percent of propane is recovered from natural gas processing plants and fewer quantities from the petroleum refineries.

The market trend changed towards propane to meet the raw material demand of petrochemical industries. Propane from natural gas and refineries are larger in amount. The recovery of ethane is costly when compared with the recovery of propane is more economical.

The next few years will be a time of rapid change in energy markets. What happens during this period will have a impact on the future of the propane industry. The propane industry should expect major challenges in existing propane markets from electricity, natural gas, and renewable fuels. The complete Propane Market Outlook provides a more detailed discussion of the key issues and trends likely to drive propane markets in the next few years. Maintaining and growing share in these markets will help position propane to capitalize on an eventual rebound in new construction.

To be competitive with liquid based crackers, propane must be extracted at the lowest marginal cost that means
- Large capacities to offset the infrastructure costs and benefit from economies of scale.
- Lowest possible energy consumption.
- Minimise losses of valuable propane during shutdown of the steam cracker when ethane must be rejected to the demethaniser overhead.

Housing starts are unlikely to rebound before 2010, and are unlikely to reach recent housing boom levels in the foreseeable future.

Propane is an energy rich gas. It is usually found mixed with deposits of natural gas and petroleum crude. Under normal conditions propane is gas and under high pressure it becomes liquid. Propane is stored as a liquid in pressurized tanks because it occupies much less space in that form. Gaseous form of propane takes 270 times more space than liquid propane. Raw natural gas is about 90 percent methane, five percent propane and five percent other gases. The propane is separated from the other gases at a natural gas processing plant. The other half of propane comes from petroleum refineries. Many gases are separated from petroleum at refineries and propane is the important one. At atmospheric pressure and temperatures above –44 F, it is a non-toxic, colorless and odorless gas.
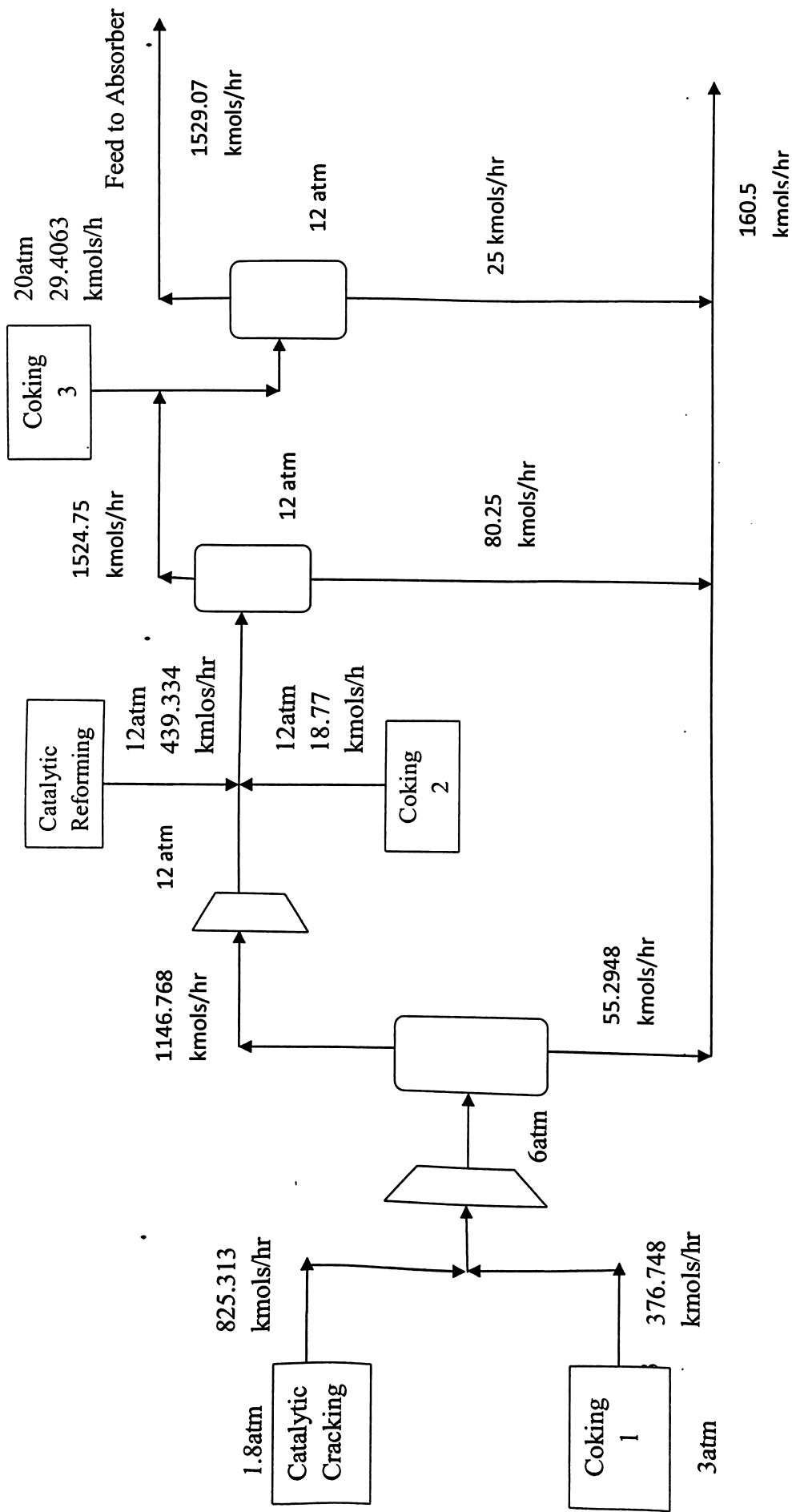
FIGURE: 1 Process Flow Diagram

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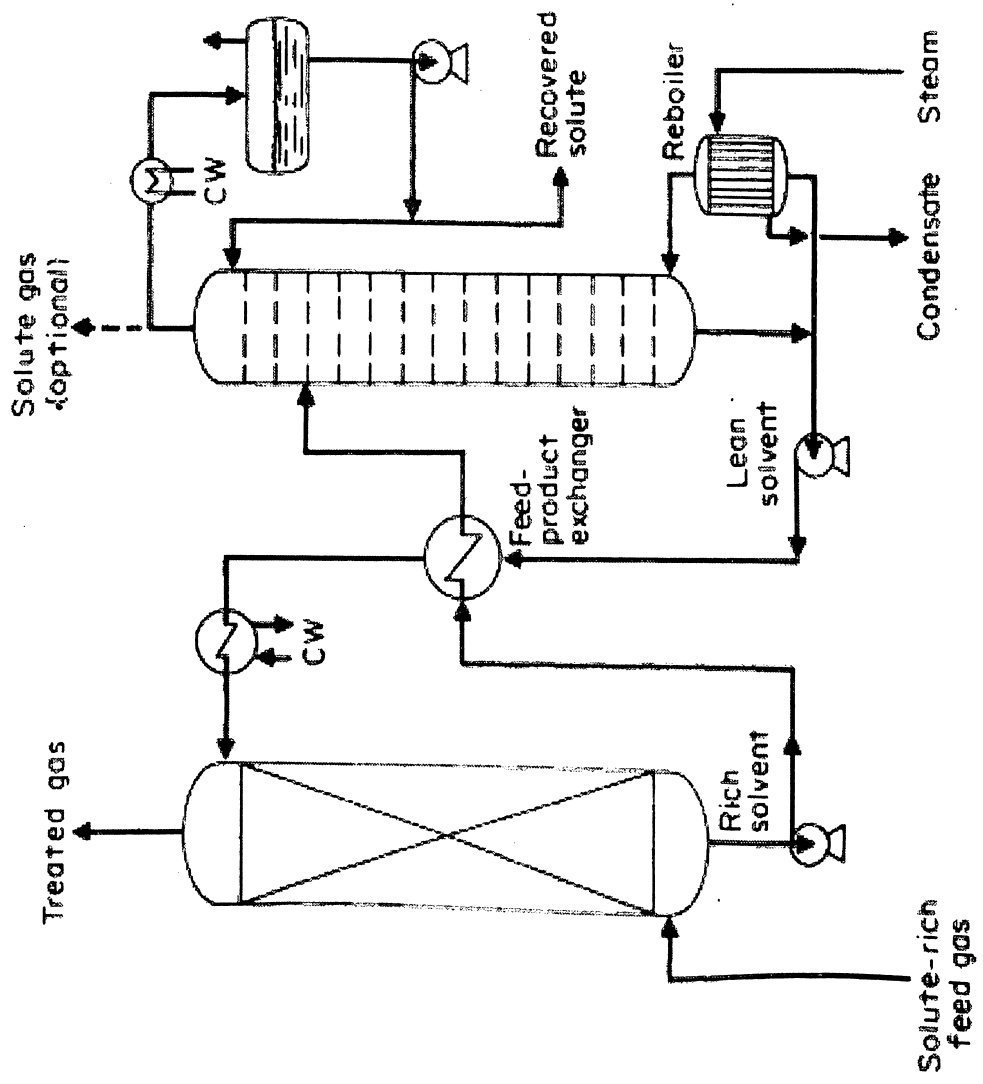
FIGURE: 2 Absorber Stripper System

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1.2 Process Description

The uncondensed gases from various refinery units are merged into single stream which is input to the absorber. In absorber the rich gas is contacted with the lean solvent i.e., gasoline at 40°C temperature and 10atm pressure. As the absorber is of tray type, the rich gas passes through the plates of the column and contacts with the liquid solvent. The plates provide the adequate surface area for mass transfer. The propane from the rich gas is separated and comes out at the top of the tower and the other components are absorbed into the absorbate. The rich solvent is then pumped to the stripping section.

The rich solvent from the bottom of the absorber is pumped into the stripper via feed product exchanger. In the exchanger the heat transfer takes from lean oil to rich oil i.e., the temperature of the rich oil is increased while the temperature of the lean oil is lowered. Then the rich oil enters into the stripper where the feed is contacted with the steam from the reboiler to strip out the more volatile components. The solute gas is taken out from the top of the tower for further processing which is optional.

The reboiler operates at 290°C and 40atm pressure for continuous steam generation and the stripper operates 165°C and 40atm pressure. The condenser operates at 40°C and 40atm pressure in which reflux is maintained to get the desired composition of solvent. The recovered solute is taken to the further processing from the stripping section. The lean oil which is free from absorbed gases pumped back to the absorber via heat exchanger to the absorber. A water cooler is provided in between the heat exchanger and absorber to lower the temperature of the lean oil. The above cycle is repeated continuously to get the desired absorption of propane.

LITERATURE

Absorption

Gas absorption is a unit operation in which a gas mixture is contacted with a liquid for the purpose of preferentially dissolving one or more components of the gas and to provide a solution to them in the liquid. This operation can be carried out in two ways.

1. Physical absorption
2. Reactive absorption

Stripping

Stripping is a unit operation in which volatile components from a liquid are transferred into the gas i.e., mass transfer carried out from liquid to gas.

2.1 General Design Procedure

The designer generally required to determine

1. The best solvent.
2. The best gas velocity through the absorber i.e., vessel diameter.
3. The height of the vessel and its internal components i.e., number of contacting trays.
4. The optimum solvent circulation through the absorber and stripper.
5. The temperatures of streams entering and leaving the absorber and the quality of heat to be re moved to account for heat of solution and other thermal effects.
6. The pressures at which the absorber and stripper will operate and
7. The mechanical design of the absorption and stripping vessels.

2.2 Selection of Solvent

When choice is possible, preference is given to liquids with high solubilities for the solute, a high solubility reduces the amount of solvent to be circulated. The solvent should be relatively nonvolatile, inexpensive, noncorrosive, stable, nonviscous, nonfoaming, and preferably nonflammable. Since the exit gas normally leaves saturated with solvent, solvent loss can be costly and may present environmental contamination problems. Thus, low-cost solvents may be chosen over more expensive ones of higher solubility or lower volatility.

Water generally is used for gases fairly soluble in water, oils for light hydrocarbons, and special chemical solvents for acid gases such as CO2, SO2, and H2S. Sometimes a reversible chemical reaction will result in a very high solubility and a minimum solvent rate.

2.3 Selection of Solubility Data

Solubility values determine the liquid rate necessary for complete or economic solute recovery and so are essential to design. Equilibrium data generally will be found in one of three forms:

1. Solubility data expressed either as solubility in weight or mole percent or as Henry's-law coefficients.
2. Pure component vapor pressures or
3. Equilibrium distribution coefficients.

In order to define completely the solubility of a gas in a liquid, it is generally necessary to state the temperature, the equilibrium partial pressure of the solute gas in the gas phase, and the concentration of the solute gas in the liquid phase.

2.4 Calculation of Liquid to Gas Ratio

The minimum possible liquid rate is readily calculated from the composition of the entering gas and the solubility of the solute in the exit liquor, saturation being assumed. It may be necessary to estimate the temperature of the exit liquid based on the heat of solution of the solute gas. The actual liquid-to-gas ratio (solvent-circulation rate) normally will be greater than the minimum by as much as 25 to 100 percent and may be arrived at by economic considerations as well as by judgment and experience.

When the solute concentration in the inlet gas is low and when nearly all the solute is being absorbed (this is the usual case), the approximation

$$y_1 G_s = x_1 L_s = (y^o_1 / m) L_s \qquad \dots\dots 2.1$$

Leads to the conclusion that the ratio $m G_s / L_s$ represents the fractional approach of the exit liquid to saturation with the inlet gas, i.e.,

$$m G_s / L_s = y^o 1 / y_1 \qquad \dots\dots 2.2$$

Optimization of the liquid-to-gas ratio in terms of total annual costs often suggests that the molar liquid-to-gas ratio L_M / G_M should be about 1.2 to 1.5 times the theoretical minimum corresponding to equilibrium at the rich end of the tower (infinite height), provided flooding is not a problem. This would be an alternative to assuming that $Ls/Gs = m/0.7$.

2.5 Selection of Equipment

Packed columns usually are chosen for very corrosive materials, for liquids that foam badly, for either small or large-diameter towers involving very low allowable pressure drops, and for small-scale operations requiring diameters of less than 0.6 m (2 ft). The type of packing is selected on the basis of resistance to corrosion, mechanical strength and capacity for handling the required flows, mass-transfer efficiency, and cost.

Plate columns may be economically preferable for large-scale operations and are needed when liquid rates are so low that packing would be inadequately wetted, when the gas velocity is so low (owing to a very high L/G) that axial dispersion or "pumping" of the gas back down the (packed) column can occur, or when intermediate cooling is desired. Also, plate towers may have a better turndown ratio and are less subject to fouling by solids than are packed towers.

2.6 Column Diameter and Pressure Drop

Flooding determines the minimum possible diameter of the absorber column, and the usual design is for 60 to 80 percent of the flooding velocity. Maximum allowable pressure drop may be determined by the cost of energy for compression of the feed gas. For systems having a significant tendency to foam, the maximum allowable velocity will be lower than estimated

flooding velocity, especially for plate towers. The safe range of operating velocities should include the velocity one would derive from economic considerations.

2.7 Computation of Tower Height

. The required height of a gas absorption or stripping tower depends on

1. The phase equilibria involved
2. The specified degree of removal of the solute from the gas and
3. The mass-transfer efficiency of the apparatus

These same considerations apply both to plate towers and to packed towers. Items 1 and 2 dictate the required number of theoretical stages (plate tower) or transfer units (packed tower). Item 3 is derived from the tray efficiency and spacing (plate tower) or from the height of one transfer unit (packed tower). Solute-removal specifications normally are derived from economic considerations.

2.8 Selection of Stripper-Operating Conditions

Stripping involves the removal of one or more volatile components from a liquid by contacting it with a gas such as steam, nitrogen, or air. The operating conditions chosen for stripping normally result in a low solubility of the solute (i.e., a high value of m), so that the ratio mG_s/Ls will be larger than unity. A value of 1.4 may be used for rule-of-thumb calculations involving pure physical desorption. For plate-tower calculations the stripping factor $S = KG_s/L_s$,

Where $K = y°/x$, usually is specified for each tray.

More often than not the rate at which residual absorbed gas can be driven from the liquid in a stripping tower is limited by the rate of a chemical reaction, in which case the liquid-phase residence time (and hence, the tower liquid holdup) becomes the most important design factor. Thus, many stripper-regenerators are designed on the basis of liquid holdup rather than on the basis of mass transfer rate.

2.9 Design of Absorber-Stripper Systems

The solute-rich liquor leaving a gas absorber normally is distilled or stripped to regenerate the solvent for recirculation back to the absorber. It is apparent that the conditions selected for the absorption step (e.g., temperature, pressure, L_s/G_s) will affect the design of the stripping tower, and, conversely, a selection of stripping conditions will affect the absorber design. The choice of optimum operating conditions for an absorber-stripper system therefore involves a combination of economic factors and practical judgments as to the operability of the system within the context of the overall process flow sheet.

An approximate procedure for executing the design of an absorber and stripper system is to set up a carefully selected series of design cases and then evaluate the investment costs, the operating costs and the operability of each case.

2.10 Importance of Design Diagrams

One of the first things a designer should try to do is lay out a carefully constructed equilibrium curve, $y^* = f(x)$, on an XY diagram, as shown in Fig. A horizontal line corresponding to the inlet-gas composition Y_1 is then the locus of feasible outlet-liquor compositions, and a vertical line corresponding to the inlet-solvent-liquor composition X_2 is the locus of feasible outlet-gas compositions. These lines are indicated as $Y = Y_1$ and $X = X_2$ respectively on Fig.

For gas absorption, the region of feasible operating lines lies above the equilibrium curve; for stripping, the feasible region for operating lines lies below the equilibrium curve. These feasible regions are bounded by the equilibrium curve and by the lines $X = X_2$ and $Y = Y_1$. By inspection, one should be able to visualize those operating lines that are feasible and those that would lead to "pinch points" within the tower. Also, it is possible to determine if a particular proposed design for solute recovery falls within the feasible envelope.

Once the design recovery for an absorber has been established, the operating curve can be constructed by first locating the point X_2, Y_2 on the diagram. The intersection of the horizontal line corresponding to the inlet gas composition Y_1 with the equilibrium curve $y^* = f(x)$ defines the theoretical minimum liquid-to-gas ratio for systems in which there are no intermediate pinch points. The operating line which connects this point with the point X_2, Y_2 corresponds to the minimum value of L_s/G_s. The actual design value of L_s/G_s normally should be around 1.2 to 1.5

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times this minimum. Thus, the actual design operating line for a gas absorber will pass through the point X_2, Y_2 and will intersect the line $Y = Y_1$ to the left of the equilibrium curve.

For stripping one begins by using the design specification to locate the point X_1, Y_1. Then the intersection of the vertical line $X = X_2$ with the equilibrium curve $y* = f(x)$ defines the theoretical minimum gas-to liquid ratio. The actual value of G_s/L_s is chosen to be about 20 to 50 percent higher than this minimum, so the actual design operating line will intersect the line $X = X_2$ at a point somewhat below the equilibrium curve.

(a)

(b)
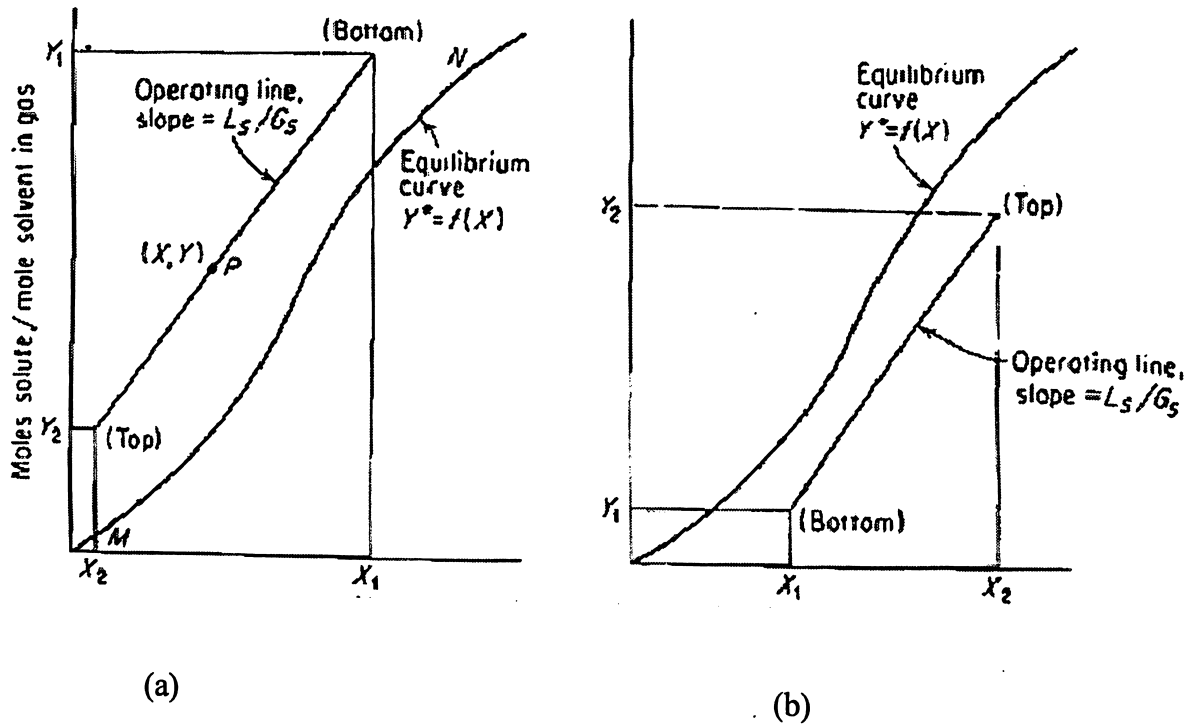
FIGURE: 3 Design diagrams for (a) Absorption and (b) stripping

(Source: Mass Transfer Operations by Robert E. Treybal, Third Edition)

2.11 Choice of Solvent for Absorption

The principle purpose of the absorption is to produce a specific solution from a gas-liquid mixture. The following properties are important considerations in choosing the solvent for absorption

1. Gas Solubility: The gas solubility should be high, thus increasing the rate of absorption and decreasing the quantity of solvent required. Generally solvents of a chemical nature similar to that of the solute to be absorbed will provide good solubility. Thus hydrocarbon oils are used to remove propane from the coke oven gas. For cases where solutions formed are ideal, the solubility of the gas is the same in terms of mole fractions for all solvents, but it is greater in terms of weight fractions for solvents of low molecular weight.

2. Volatility: The solvent should have a low vapor pressure since the gas leaving an absorption operation is ordinarily saturated with the solvent and much may thereby be lost. If necessary a second less volatile liquid can be used to recover the evaporated portion of the first.

3. Corrosiveness: The materials of construction required for the equipment should not be unusual or expensive.

4. Cost: The solvent should be inexpensive, so that losses are not costly, and should be readily available.

5. Viscosity: Low viscosity is preferred for reasons of rapid absorption rates, improved flooding characteristics in absorption towers, low pressure drops on pumping and good heat transfer characteristics.

6. Miscellaneous: The solvent if possible should be nontoxic, nonflammable, and chemically stable and should have a low freezing point.

2.12 Material Balances

The basic material balances and graphical interpretations of absorption and stripping are as follows
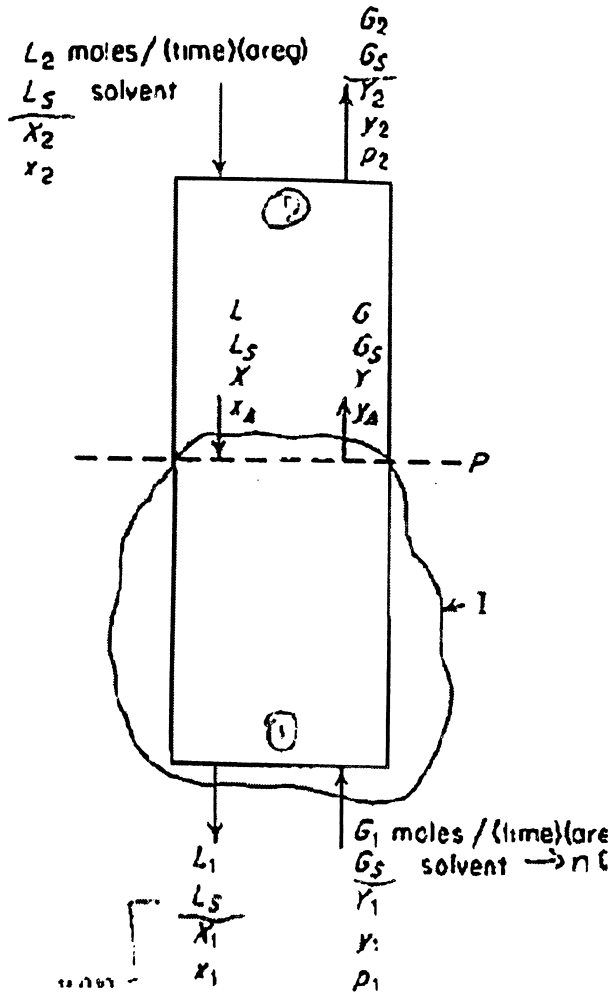
FIGURE: 4 Flow quantities of absorber or stripper

(Source: Mass Transfer Operations by Robert E. Treybal, Third Edition)

For Countercurrent Flow

The above figure represents countercurrent tower, tray tower filled with bubble cap trays to bring about liquid-gas contact. The gas stream at any point in the tower consists of G total mol/(area of cross section) (time), made up of diffusing solute A of mole fraction y, partial pressure p, or mole ratio Y, and non diffusing, essentially insoluble gas G_s mol/(area) (time).

The relationship between these is

$$Y = \frac{y}{1-y} = \frac{p*}{p_t - p*} \qquad \ldots\ldots 2.3$$

$$Gs = G(1 - y) = \frac{G}{1+Y} \qquad \ldots\ldots 2.4$$

Similarly the liquid stream consists of L total mol/(area) (time), containing x mole fraction soluble gas , or mole ratio X, and essentially non volatile solvent Ls mol/(area)(time).

$$X = \frac{x}{1-x}$$

$$L_s = L(1 - x) = \frac{L}{1+X} \qquad \ldots\ldots 2.5$$

Since the solvent gas and solvent liquid are essentially unchanged in quality as they pass through the tower, it is convenient to express the material balance in terms of these. A solute balance about the lower part of the tower is

For packed tower

$$G_s(Y_1 - Y) = L_s(X_1 - X) \qquad \ldots\ldots 2.6$$

For tray tower

$$Gs\,(Y_{N_{p+1}} - Y_1) = Ls(X_{N_p} - Xo) \qquad \ldots\ldots 2.7$$

The above is the equation of a straight line (operating line) on X,Y coordinates, of slope Ls/Gs , which passes through (X1, Y1). Substitution of X2 and Y2 for X and Y shows the line

pass through (X2,Y2) for an absorber. This line indicates the relationship between the liquid and gas concentration at any level in the tower.

2.13 Absorption Factor A

The absorption factor $A = L/mG$ is the ratio of the slope of the operating line to that of the equilibrium curve. For values of A less than unity, corresponding to convergence of the operating line and equilibrium curve for the lower end of the absorber, the fractional absorption of solute is limited, eve for infinite theoretical trays. On the other hand, for values A greater than unity, any degree of absorption is possible if sufficient trays are provided. For a fixed degree of absorption from a fixed amount of gas, as A increases beyond unity, the absorbed solute dissolved more in amount and becomes less valuable, number of trays decreases and therefore operating cost decreases. As a thumb rule, the most economical A will be ranging from1.25 to2.0.

In terms of mole fractions

Absorption:

$$\frac{y_{N_p+1} - y_1}{y_{N_p+1} - mx_0} = \frac{A^{N_p+1} - A}{A^{N_p+1} - 1} \qquad .....2.8$$

$$N_p = \frac{log\left[\frac{y_{N_p+1} - mx_0}{y_1 - mx_0}\left(1 - \frac{1}{A}\right) + \frac{1}{A}\right]}{log\,A} \qquad ......2.9$$

Stripping:

$$\frac{x_0 - x_{N_p}}{x_0 - y_{N_p+1}/m} = \frac{S^{N_p+1} - S}{S^{N_p+1} - 1} \qquad .....2.10$$

$$N_p = \frac{log\left[\frac{x_0 - y_{N_p+1}/m}{x_{N_p} - y_{N_p+1}/m}\left(1 - \frac{1}{S}\right) + \frac{1}{S}\right]}{log\,S} \qquad ......2.11$$

Where $A = L/mG$, and $S = mG/L$

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2.14 Non Isothermal Operation

Absorbers and strippers deal with dilute gas mixtures and liquids, and it is frequently satisfactory in these cases to assume that the operation is isothermal. But actually absorption operations are usually exothermic, and when large quantities of solute gas absorbed to form concentrated solutions, the temperature effects cannot be ignored. If by absorption the temperature of the liquid is raised to considerable extent, the equilibrium solubility of the solute will be appreciably reduced and the capacity of the absorber decreased. If the heat evolved is excessive, cooling coils can be installed or liquid can be removed at intervals, cooled, and returned to the absorber. For stripping, an endothermic action, the temperature tends to fail.

If Q_T is the heat removed per unit time from entire tower, an enthalpy balance for the entire tower is

$$L_0 H_{L0} + G_{N_p+1} H_{G,N_p+1} = L_{N_p} H_{L,N_p} + G_1 H_{G1} + Q_T \quad \ldots\ldots 2.12$$

Where H represents in each case the molal enthalpy of the stream at its particular concentration and condition. It is convenient to to refer all enthalpies to the condition of pure liquid solvent, pure diluents gas, and pure solute at some base temperature t_0, with each substance assigned zero enthalpy for its normal state of aggregation at t_0 and 1 atm pressure. Thus the molal enthalpy of a liquid solution, temperature t_L, composition x mole fraction solute, available from the following equation

$$H_L = C_L (t_L - t_0) + \Delta H_S \qquad \ldots\ldots .2.13$$

The first term on the right represents the sensible heat and the second the enthalpy of mixing, or integral of solution at the prevailing condition and at the base temperature t_0, per mole of solution. If heat evolves ΔH_S is negative quantity.

For adiabatic operation, Q_T is zero and the temperature of the streams leaving an absorber will generally be higher than the entering temperature owing to the heat of solution. The total and solute balances up to tray n, given by the following equations

$$L_n + G_{N_p + 1} = L_{N_p} + G_{n + 1} \qquad \ldots\ldots 2.14$$

$$L_n \, x_n + G_{N_p + 1} \, y_{N_p+1} = L_{N_p} \, x_{N_p} + G_{n+1} \, y_{n+1} \qquad \ldots 2.15$$

From which L_n and x_n are computed. An enthalpy balance is

$$L_n H_{L,n} + G_{N_{p+1}} H_{G,N_{p+1}} = L_{N_p} H_{L,N_p} + G_{n+1} H_{G,n+1} \qquad \ldots 2.16$$

From which the temperature of stream L_n can be obtained.

DESIGN PROCEDURES

3.1 Absorption:

Kremser Brown - Sherwood Method

1. Calculate the total moles of gas inlet to the absorber indentifying the quantities of individual components.

2. Assuming the tower pressure as set and an average of top and bottom temperatures can be selected, read equilibrium K_i values from charts for each component in gas.

3. Assuming the fix lean oil rate.

4. Calculate

$$\frac{L}{V} = \frac{Mols/hr\ lean\ oil}{Mols/hr\ rich\ gas} \qquad \ldots\ldots 3.1$$

Assume this value constant for tower design

5. Calculate absorption factor

$$A_i = L/(V)\ (K_i) = \frac{y_{i,n+1} - y_{i,n}}{y_{i,n} - y_{i,n-1}} \qquad \ldots\ldots 3.2$$

6. Calculate fraction absorbed for each component, assuming a fixed overall tray efficiency for and assume number of trays or existing column with trays.

Theoretical trays, N = (tray efficiency E_o) (no of actual trays)

7. Calculate Degree of absorption

$$\psi_i = \frac{N_{ai}}{N_i} = \frac{A_i^{n+1} - A_i}{A_i^{n+1} - 1} \qquad \ldots\ldots 3.3$$

8. Moles of each component absorbed /hr

$$= (V * y_{(n+1)i}) \, (\psi_i) \qquad \dots\dots 3.4$$

9. Quantity of each component absorbed is

$$N_{ai} = \psi_i \, N_i \qquad \dots\dots 3.5$$

10. The quantity of non absorbed material is

$$N_{si} = (1 - \psi) \, N_i \quad \text{or} \qquad \dots\dots\dots 3.6$$

$$N_{si} = N_i - N_{ai}$$

11. Flow rate of absorbent at top condition is given by

$$L_s = A_i \, K_i \, N_s \qquad \dots\dots 3.7$$

12. The amount of rich oil

$$L_b = N_a + L_s \qquad \dots\dots 3.8$$

Where $L_b = A_i \, K_i \, N$

13. Flow rate of absorbent at bottom condition

$$L_s = L_b - N_a \qquad \dots\dots\dots 3.9$$

14. Taking the average of absorbent flow rate at top and bottom conditions.

$$L_s = \frac{L_{top} + L_{bottom}}{2} \qquad \dots\dots 3.10$$

15. Tower diameter

$$D_e = \sqrt{\frac{S_{col} * 4}{\pi}} \qquad \dots\dots 3.11$$

We have, $\dfrac{S_d}{S_{col}} = 0.1$

$$S_d = \frac{V}{v} \quad \text{(Assuming the velocity of the down spout as v m/sec)}$$

16. Steam requirement

$$\frac{L_s}{steam} = \frac{L}{V} = \frac{S_{ni}}{K_{ni}} \qquad \text{......3.12}$$

17. Solvent loss by entertainment for steam desorption

$$L_{sa}(L_s + N_a) = K_{Ls} L_s L_{sa} + K_{Ls} L_s (N_d + S_{st}) \qquad \text{.......3.13}$$

Where, $K_{Ls} = \dfrac{p_i}{P}$

18. Water requirement

$$W = \frac{Q}{C_p(t_2 - t_1)} \qquad \text{......3.14}$$

3.2 Stripping

Fenske Underwood Gilliland Method

1. The rich oil flow rate and absorbate composition is known. From the available temperature of condensing fluids, column operating pressure can be determined.

2. From K charts, determining K_i values for each component at the column temperature and pressure.

3. From a fixed percentage for key component, moles component stripped/hr

$$G_{mi} = (L_{m+1}) (X_{m+1}) (E_{si}) \qquad \text{.......3.15}$$

4. Calculating minimum reflux by Underwood equation

$$\sum \frac{x_{Fi} \, \alpha_{l/h}}{\alpha_{l/h} - \theta} = 1 - q \qquad \text{......3.16}$$

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5. Using the value of minimum reflux ratio R_{min} is calculated by

$$\sum \frac{x_{Di} \, {}^{\alpha_{i/h}}}{\alpha_{i/h} - \theta} = 1 + R_{min} \qquad , \qquad \text{.......3.17}$$

$\dfrac{N - N_m}{N+1}$ value is calculated from the Gilliland graph at $\dfrac{R - R_m}{R+1}$
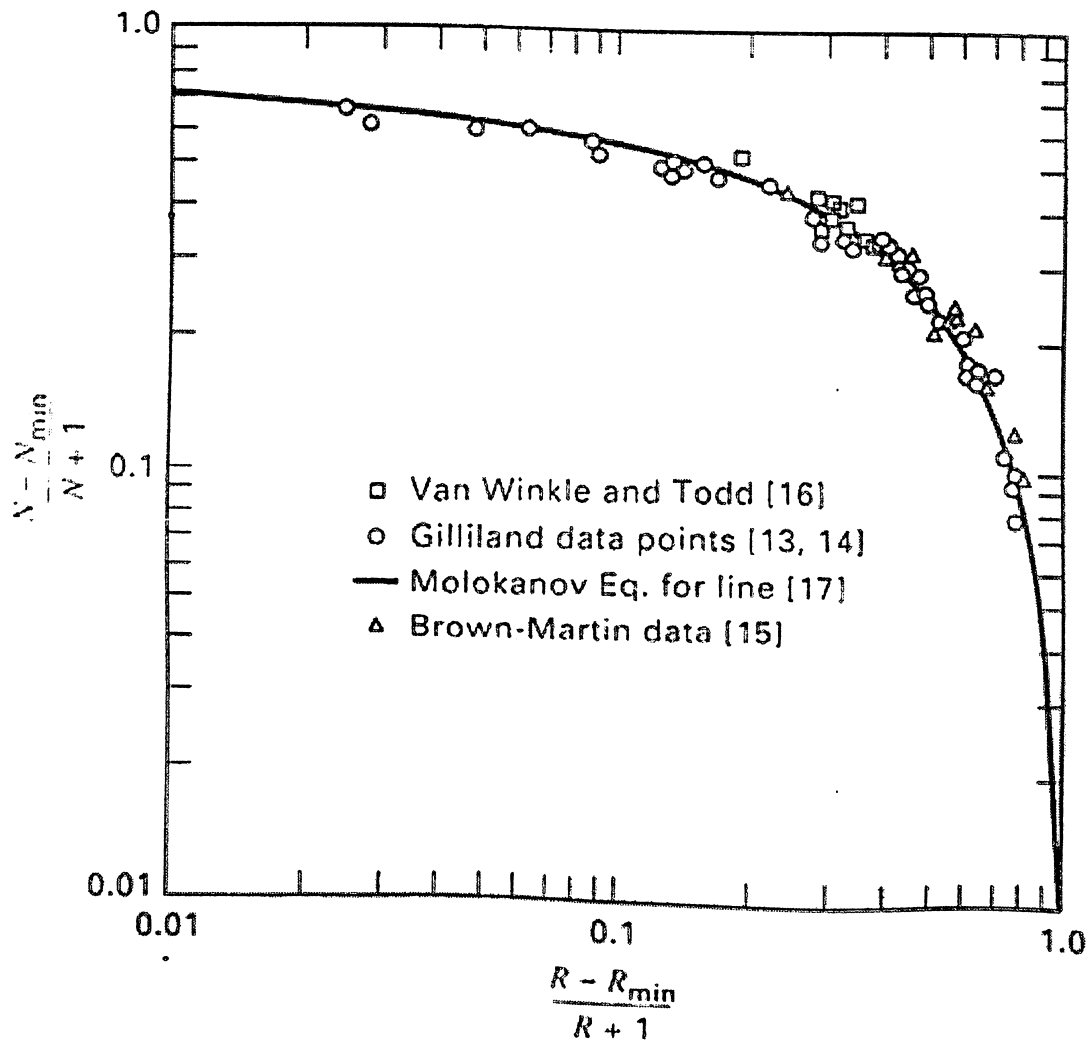
FIGURE: 5 Gilliland Graph

(Source: Mass Transfer Operations by Robert E. Treybal, Third Edition)

6. Now calculating number of stages by Fenske equation

$$N_m = \frac{\log(\frac{x_{DL}}{x_{Dh}} \frac{x_{Bh}}{x_{Bl}})}{\log \alpha/_h} \qquad \ldots\ldots 3.18$$

7. Calculating number of trays in stripping section and rectifying section by Hengestebeck - Geddes equation.

$$\frac{N_R}{N_S} = \frac{\log\left[\left(x_l/_{x_h}\right)_D \left(x_h/_{x_l}\right)_F\right]}{\log\left[\left(x_l/_{x_h}\right)_F \left(x_h/_{x_l}\right)_B\right]} \qquad \ldots\ldots 3.19$$

8. Material Balance to find out the flow rate of liquid

$$V = L + D \qquad \ldots\ldots 3.20$$

Where V - flow rate of vapor kmols/hr

L - flow rate of liquid kmlos/hr

D - flow rate of distillate kmols/hr

9. Calculating diameter of the column

$$D_e = \sqrt{\frac{S_{col} *4}{\pi}} \qquad \ldots\ldots 3.21$$

We have, $\dfrac{S_d}{S_{col}} = 0.1$

$$S_d = \frac{V}{v}$$

Assuming the velocity of the down spout as v m/sec

10. Heat Balance

$$Q_c = V\, H_V - [D\, H_D - L\, H_L] \qquad \ldots\ldots 3.22$$

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Where Q_e - heat content of condensate kcals/hr

$$Q_B = V\,H_V + B\,h_B - [F\,h_F - L\,h_L] \quad \ldots\ldots 3.23$$

Where Q_B - heat content of bottoms kcals/hr

11. Water requirement

$$W = \frac{Q_c}{C_p(t_2 - t1)} \qquad \ldots\ldots 3.24$$

12. Steam requirement

$$\text{Steam} = \frac{Q_B}{\lambda} \qquad \ldots\ldots 3.25$$

13. Computation of the column height approximately basing on the column diameter.

DESIGN CALCULATIONS

PROBLEM

The uncondensed gases from the refinery units like Catalytic Cracking, Catalytic Reforming, Coking-1, Coking-2 and Coking-3 are absorbed in hydrocarbon oil in an absorber. The absorber is a bubble cap tower. It operates at a pressure of 10atm and an average temperature of 40°C. The absorption of components is calculated for a fixed number of trays of 10. The absorption of Propane is 96%.

The absorbent is Gasoline of the following composition

Component	x_i
Hexane	0.02656
Benzene	0.02604
2-3 Dimethyl Pentene	0.04016
Heptane	0.09955
Tolune	0.57656
Octane	0.04269
Xylene	0.18841

Input Data

Pressure   : 10 atm

Temperature   : 40°C

Flow rate of Gas   : 1529.07 kmoles/hr

Number of trays   : 10

Density of the liquid (ρ_L) : 660 kg/m³

Absorber Calculations:

The absorption factor for the tower with 10 plates is read from the graph

For propane at 96% recovery $\frac{1}{A} = 0.865$

$$A_{C_3} = \frac{1}{0.865} = 1.156$$

A_{C_3}- Absorption factor of Propane

Now,

$$\frac{L}{V} = A_{C_3}\, K_{C_3}$$

$$\left(\begin{array}{l} Value\ of\ K_{C_3}\ taken\ from\ the\ graph \\ Source : GPSA\ Data\ Book, Section\ 25 \end{array} \right)$$

$$= 1.156 * 1.25$$

$$= 1.445$$

Degree of absorption

$$\psi_i = \frac{N_{ai}}{N_i} = \frac{A_i^{n+1} - A_i}{A_i^{n+1} - 1}$$

For H_2o,

$$A_{H_2O}\, K_{H_2O} = A_{C_3}\, K_{C_3}$$

$$A_{H_2o} = \frac{A_{C_3}\, K_{C_3}}{K_{H_2o}} = \frac{1.445}{11} = .13136$$

$$\psi_{H_2o} = \frac{AC_{H_2o}^{11} - A_{C_{H_2O}}}{AC_{H_2o}^{11} - 1} = \frac{0.13136^{11} - 0.13136}{0.13136^{11} - 1} \approx 0.13136$$

Similarly for other components the ψ_i values are provided in the table

Component	y_i	K_i (40°C,10atm)	A_i	ψ_i
H_2O	0.0206508	11	0.131363636	0.131363636
CO_2	0.0147732	9.6	0.150520833	0.150520833
H_2S	0.0170543	3.6	0.401388889	0.401362804
H_2	0.3254882	78	0.018525641	0.018525641
CH_4	0.173918	18	0.080277778	0.080274711
C_2H_4	0.0367611	5.6	0.258035714	0.258035464
C_2H_6	0.0804506	4.0	0.36125	0.361241266
C_3H_6	0.09929	1.4	1.032142857	0.922778714
C_3H_8	0.0575329	1.25	1.156	0.941933032
C_4H_8	0.070835	0.48	3.010416667	0.999989076
IC_4H_{10}	0.0402608	0.6	2.408333333	0.999910906
nC_4H_{10}	0.0228721	0.45	3.211111111	0.999994093
C_5H_{12}	0.0341359	0.15	9.633333333	1
C_6+	0.006036	0.05	28.9	1
Total	1.0000589			

TABLE 4.1

The quantity of each component absorbed is given by

$$N_{ai} = \psi_i \, N_i$$

Sample calculation of H_2o,

N_i = 31.57651876 kmols/hr

ψ_i = 0.1313636 .

N_{ai} = 31.57651876 * 0.1313636= 4.148006322 kmols/hr

Similarly for other components, the values of N_{ai} are listed in the table

Component	y_i	N_i (kmol/hr)	Ψ_i	$N_{ai}=\psi_i N_i$ (kmols/hr)
H_2O	0.0206508	31.57651876	0.131363636	4.148006322
Co_2	0.0147732	22.58925692	0.150520833	3.400153759
H_2S	0.0170543	26.0772185	0.401362804	10.46642555
H_2	0.3254882	497.694242	0.018525641	9.220104867
CH_4	0.173918	265.9327963	0.080274711	21.34767845
C_2H_4	0.0367611	56.21029518	0.258035464	14.50424958
C_2H_6	0.0804506	123.0145989	0.361241266	44.43794943
C_3H_6	0.09929	151.8213603	0.922778714	140.0975197
C_3H_8	0.0575329	87.9718314	0.941933032	82.86357386
C_4H_8	0.070835	108.3116735	0.999989076	108.3104902
IC_4H_{10}	0.0402608	61.56158146	0.999910906	61.55609666
nC_4H_{10}	0.0228721	34.97304195	0.999994093	34.97283534
C_5H_{12}	0.0341359	52.19618061	1	52.19618061
C_6+	0.006036	9.22946652	1	9.22946652
Total	1.0000589	1529.160062		596.7507309

TABLE 4.2

The quantity of non absorbed components is given by

$$N_{si} = N_i - N_{ai}$$

Sample calculation of H_2o,

N_i = 31.57651876 kmols/hr

N_{ai} = 4.148006322 kmols/hr

$$N_s = 31.57651876 * 4.148006322 = 27.42851243 \text{ kmols/hr}$$

Similarly for other components the N_s values are tabulated

Component	y_i	N_i (kmol/hr)	$N_{ai}=\psi_i N_i$ (kmols/hr)	$N_s=N_i-N_{ai}$ (kmols/hr)
H_2O	0.0206508	31.57651876	4.085688351	27.49083041
CO_2	0.0147732	22.58925692	3.334489658	19.25476727
H_2S	0.0170543	26.0772185	9.387819265	16.68939924
H_2	0.3254882	497.694242	9.216999135	488.4772428
CH_4	0.173918	265.9327963	21.34767845	244.5851178
C_2H_4	0.0367611	56.21029518	14.45650029	41.75379489
C_2H_6	0.0804506	123.0145989	44.37588579	78.63871316
C_3H_6	0.09929	151.8213603	132.135863	19.68549733
C_3H_8	0.0575329	87.9718314	82.86357386	5.108257542
C_4H_8	0.070835	108.3116735	108.3009498	0.010723665
IC_4H_{10}	0.0402608	61.56158146	61.55609666	0.005484792
nC_4H_{10}	0.0228721	34.97304195	34.97283534	0.000206603
C_5H_{12}	0.0341359	52.19618061	52.19618061	7.32427E-11
C_6+	0.006036	9.22946652	9.229466505	1.52931E-08
Total	1.0000589	1529.160062		941.7000355

TABLE 4.3

Now the flow rate of the absorbent L_s is found by

$$L_s = A_{C_3H_8} \ K_{C_3H_8} \ N_s$$

Where N_s - number of moles unabsorbed

From the table the value of N_s = 932.4093314 kmols/hr

$L_s = 1.156 * 1.25 * 932.4093314 = 1347.331484$ kmols/hr

L_s at top condition

The amount of rich oil

$$L_b = N_a + L_s$$

$$\text{Where } L_b = A_{C_3H_8} \, K_{C_3H_8} \; N$$

N = total number of moles of feed

$L_b = 1.156 * 1.25 * 1529.07$

    = 2209.50615 kmols/hr

$L_s = L_b - N_a$

    = 2209.50615 - 596.75

    = 1612.75615 kmols/hr

L_s at bottom condition

Taking the average of L_s at top and bottom condition

$$L_s = \frac{1347.331484 + 1612.75615}{2} = 1480.043817 \text{ kmols/hr}$$

Calculation of the column diameter

Flow rate   = 2209.50615 kmols/hr

The tower diameter is to be based on this liquid load.

Density of the liquid $= \rho_L^{40^\circ C} = 660 \text{ kg/m}^3$

Molecular Weight of absorbent is calculated from the table

Component	x_i	M_{wi}	$x_i * M_{wi}$
Hexane	0.02656	86	2.28416
Benzene	0.02604	78	2.03112
2-3 Dimethyl Pentene	0.04016	100	4.016
Pentane	0.09955	100	9.955
Tolune	0.57656	92	53.04352
Octane	0.04269	114	4.86666
Xylene	0.18841	106	19.97146
Total		**676**	**96.16792**

Molecular Weight of absorbent = 96.16792

$$\text{Flow rate} = \frac{2209.50615 * 96.16792}{660*3600} = 0.08421764 \text{ m}^3/\text{sec}$$

Taking v the velocity in the down spout as 0.12 m/sec

$$S_d = \frac{V}{0.12} = \frac{0.08421764}{0.12} = 0.745181367 \text{ m}^2$$

But $\dfrac{S_d}{S_{col}} = 0.1$

$$S_{col} = \frac{S_d}{0.1} = \frac{0.745181367}{0.1} = 7.451813671 \text{ m}^2$$

$$\frac{\pi}{4} D_e^2 = 7.451813671$$

$$D_e = \sqrt{\frac{7.451813671*4}{\pi}}$$

$$= 3.081031689 \text{ m}$$

So the area taken up by the down comer is 20%

Area available for bubble cap trays is 80%

Computation of tower height

Plate Spacing = 0.4m

Number of plates = 10

Space above the top plate	2*0.4	0.8
Two sets of three plates each	2*3*0.4	2.4
Two sets of two plates each	2*2*0.4	1.6
Three manhole spaces	3*0.7	2.1
Free space at bottom	1*0.4	0.4
Total		**7.3m**

The total height of the tower is 7.3m

Steam requirement if desorption is at 40atm 99% of C5H12 recovered

$$0.99 = \frac{S^{11}_{nC_5H_{12}} - S_{nC_5H_{12}}}{S^{11}_{nC_5H_{12}} - 1}$$

$$For\ pentane\ \frac{1}{S} = 0.71 \quad (from\ graph)$$

$$\frac{L_S}{Steam} = \frac{S_{nC_5H_{12}}}{K_{nC_5H_{12}}} = \frac{1.4085}{0.725} = 1.9426 \text{ kmols/hr}$$

$$Steam = \frac{L_S}{1.9426} = \frac{1480.043817}{1.9426} = 761.8529357 \text{ kmols/hr}$$

$$Steam = 761.8529357 * 18 = 13713.35284 \text{ kg/hr}$$

(Molecular weight of H_2o is 18)

Solvent loss by entertainment for steam desorption

$$L_{sa}(L_s + N_a) = K_{Ls} L_s \, L_{sa} + K_{Ls} L_s \, (N_d + S_{st})$$

Where L_{sa} - Solvent loss in absorbent kmols/hr

K_{Ls} - Equilibrium constant for lost solvent

Vapor pressure of Octane at $40^0C = 0.05$ atm

$$K_{Ls} = \frac{pc_8}{10} = \frac{0.05}{10} = 0.005 \text{ atm}$$

L_s = 1480.043817 kmols/hr

$N_a = N_d$ = 596.75 kmols/hr

$$L_{sa} = \frac{.005*1480.043817 *[596.74+13713.35284\,]}{1480.043817 +596.75-0.005*1480.043817}$$

$$= 52.73203347 \text{ kmols/hr}$$

Water requirement

The values of the enthalpies of individual components are provided in the table

Component	$N_{ai}=\psi_i N_i$ (kmols/hr)	x_i	Mw	$x_i * Mw$	H_i (40°C, 10 atm)	h_i (40°C) Btulb	$L_i =(H_i-h_i)$ Btu/lb	$Q = \sum(N_{ai}*L_i)$
H_2O	4.148006322	0.006950987	16	0.111215785	355	275	80	331.8405058
CO_2	3.400153759	0.005697779	44	0.250702274	355	275	80	272.0123007
H_2S	10.46642555	0.017539024	34	0.596326825	355	275	80	837.3140439
H_2	9.220104867	0.015450513	2	0.030901026	355	275	80	737.6083894
CH_4	21.34767845	0.035773192	16.04	0.573801999	355	275	80	1707.814276
C_2H_4	14.50424958	0.024305374	28.05	0.681765735	310	225	85	1232.861215
C_2H_6	44.43794943	0.074466519	30.07	2.239208216	330	240	90	3999.415449
C_3H_6	140.0975197	0.234767236	42.08	9.879005292	310	173	137	19193.3602
C_3H_8	82.86357386	0.138857934	44.1	6.123634909	310	175	135	11186.58247
C_4H_8	108.3104902	0.18150039	56.11	10.18398687	307	162	145	15705.02108
IC_4H_{10}	61.55609666	0.103152109	58.12	5.995200597	290	155	135	8310.07305
nC_4H_{10}	34.97283534	0.058605434	58.12	3.406147802	310	160	150	5245.925302
C_5H_{12}	52.19618061	0.087467309	72.15	6.310766348	310	155	155	8090.407994
C_6+	9.229466652	0.015466201	86	1.330093253	300	150	150	1384.419978
Total	596.7507309			47.71275693				78234.65625

TABLE 4.4

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$Q = \sum (N_{ai} * L_i) = 78234.65625$ kcal/hr

$$= \frac{78234.65625 * 0.252 * 47.71275693}{0.45359}$$

$$= 2073818.573 \text{ kcal/hr}$$

Water requirement (W) $= \dfrac{Q}{C_p(t_2 - t_1)}$

$$= \frac{2073818.573}{1\,(42-25)}$$

$$= 121989.3278 \text{ kg/hr}$$

Stripper Calculations:

The rich oil from the absorber is taken into the stripper to remove the more volatile components. The feed to the distillation tower is provided in the following table.

Component	F_i kmol/hr	x_F
H_2O	4.1479	0.001998005
CO_2	3.4001	0.001637797
H_2S	9.9162	0.004776543
H_2	9.22	0.004441189
CH_4	21.3484	0.010283329
C_2H_4	14.5042	0.00698654
C_2H_6	45.5784	0.021954698
C_3H_6	140.0974	0.067483634
C_3H_8	84.4529	0.040680188
C_4H_8	108.3116	0.052172705
IC_4H_{10}	61.5615	0.02965361
nC_4H_{10}	34.973	0.016846174
C_5H_{12}	52.1961	0.025142383
C_6^+	9.2294	0.004445717
Hexane	39.2386	0.018900874
Benzene	38.2386	0.018419183
2,3 - diethyl pentane	59.3307	0.028579054
Pentane	147.0711	0.070842801
Tolune	851.7866	0.410297799
Octane	63.0684	0.03037947
Xylene	278.3493	0.134078307
Total	**2076.0204**	**1**

TABLE 4.5

ROUGH DISTRIBUTION

DISTILLATE

Considering rough distribution in the stripper, the distillate and bottoms are listed in the following tables basing on the light key and heavy key components.

Component	D_i kmol/hr	x_{Di}	k_i (40 atm, 40°C)	$K_i x_{Di}$ (kmol/hr)
H_2O	4.1479	0.006925435	2.6	0.018006131
CO_2	3.4001	0.00567689	3	0.01703067
H_2S	9.9162	0.01655633	1.5	0.024834494
H_2	9.22	0.015393937	22	0.338666615
CH_4	21.3484	0.03564381	6	0.213862858
C_2H_4	14.5042	0.024216566	1.8	0.043589819
C_2H_6	45.5784	0.076098809	1.3	0.098928452
C_3H_6	140.0974	0.233910038	0.525	0.12280277
C_3H_8	84.4529	0.141004623	0.46	0.064862127
C_4H_8	108.3116	0.180839691	0.19	0.034359541
IC_4H_{10}	61.5615	0.102784583	0.22	0.022612608
nC_4H_{10}	34.973	0.058391774	0.17	0.009926602
C_5H_{12}	52.1961	0.087147882	0.06	0.005228873
C_6^+	9.2294	0.015409631	0.0225	0.000346717
nC_6	0	0	0	0
Benzene	0	0	0	0
2,3 - diethyl pentane	0	0	0	0
nc_7	0	0	0	0
Tolune	0	0	0	0
nC_8	0	0	0	0
Xylene	0	0	0	0
Total	**598.9371**			**1.015058277**

TABLE 4.6

BOTTOM STREAM

The individual bottom components are listed in the table according to the rough distribution of the feed.

Component	B_i (kmol/hr)	x_{Bi}	K_i (40 atm, 290°C)	$K_i x_{Bi}$
H_2O	0	0	0	0
Co_2	0	0	0	0
H_2S	0	0	0	0
H_2	0	0	0	0
CH_4	0	0	0	0
C_2H_4	0	0	0	0
C_2H_6	0	0	0	0
C_3H_6	0	0	0	0
C_3H_8	0	0	0	0
C_4H_8	0	0	0	0
IC_4H_{10}	0	0	0	0
nC_4H_{10}	0	0	0	0
C_5H_{12}	0.52198	0.000353261	3.8	0.001342391
C_6^+		0	0	0
nC_6	39.2386	0.026555536	2	0.053111072
Benzene	38.2386	0.025878765	2	0.05175753
2,3 - diethyl pentane	59.3307	0.040153281	1.05	0.042160945
nc_7	147.0711	0.099533415	1.05	0.104510086
Tolune	851.7866	0.576464237	1.05	0.605287449
nC_8	63.0684	0.042682847	0.6	0.025609708
Xylene	278.3493	0.188378658	0.6	0.113027195
Total	**1477.60528**	**1**		**0.996806376**

TABLE 4.7

The Condenser temperature is 40°C

The Reboiler temperature is 290°C

$$Tavg = \frac{t_{top} + t_{bottom}}{2}$$

$$= \frac{40+290}{2}$$

$$= 165°C$$

Calculation of relative volatality at average temperature of 165°C and 40atm

Component	F_i mol/hr	x_f	K_i (40atm,165°C)	$\alpha = (K_i/K_h)$
H2O	4.1479	0.001997782	5	10.63829787
CO2	3.4001	0.001637614	5.25	11.17021277
H2S	9.9162	0.004776009	3.3	7.021276596
H2	9.22	0.004440694	14.5	30.85106383
CH4	21.3484	0.010282181	70	148.9361702
C2H4	14.5042	0.00698576	27	57.44680851
C2H6	45.5784	0.021952246	18	38.29787234
C3H6	140.0974	0.067476099	8.2	17.44680851
C3H8	84.4529	0.040675646	7.2	15.31914894
C4H8	108.3116	0.05216688	1.1	2.340425532
iC4H10	61.5615	0.0296503	3.6	7.659574468
nC4H10	34.973	0.016844293	2.85	6.063829787
C5H12	52.1961	0.025139576	1.1	2.340425532
nC6	86.9384	0.041872755	0.47	1
nc7	1058.1884	0.509662747	0.19	0.404255319
nC8	341.4177	0.164439416	0.085	0.180851064
Total	**2076.2522**	1		

TABLE 4.8

Considering C_6, C_7 and C_8 streams of the components together, the individual molar flow rate of the components in the feed and their mole fractions are listed in the table.

Component	F_i kmol/hr	x_f
H_2O	4.1479	0.001997782
Co_2	3.4001	0.001637614
H_2S	9.9162	0.004776009
H_2	9.22	0.004440694
CH_4	21.3484	0.010282181
C_2H_4	14.5042	0.00698576
C_2H_6	45.5784	0.021952246
C_3H_6	140.0974	0.067476099
C_3H_8	84.4529	0.040675646
C_4H_8	108.3116	0.05216688
IC_4H_{10}	61.5615	0.0296503
nC_4H_{10}	34.973	0.016844293
C_5H_{12}	52.1961	0.025139576
C_6	86.9384	0.041872755
C7	1058.1884	0.509662747
C8	341.4177	0.164439416
Total	**2076.2522**	**1**

DISTILLATE AND BOTTOMS DISTRIBUTION

The new distribution of the light key and heavy key components are assumed to be

Copmponent	Distillate	Bottoms
Pentane	0.99	0.01
Hexane	0.1061	0.8939

MATERIAL BALANCE

$F_i = D_i + B_i$

For Pentane,

$D_i = F_i * 0.99 = 52.1961 * 0.99 = 51.674139$ kmol/hr

$B_i = F_i * 0.01 = 52.1961 * 0.01 = 0.521961$ kmol/hr

For Hexane,

$D_i = F_i * 0.1061 = 86.9384 * 0.1061 = 9.22416424$ kmol/hr

$B_i = F_i * 0.8939 = 86.9384 * 86.9384 = 77.71423576$ kmol/hr

Copmponent	Distillate	Bottoms	D_i/B_i	$\ln(D_i/B_i)$
Pentane	51.674139	0.521961	99	4.59511985
Hexane	9.22416424	77.71423576	0.118693366	-2.131211866

The new distribution equation can be found by $ln\left(D_i/B_i\right) = m + \alpha n$

$4.5852 = 2.34n + m$

$-2.1312 = 1.00n + m$

Solving above two equations,

We have, $m = 5.0197$

$n = -7.1509$

$\ln(D_i/B_i) = 5.0197\alpha_i - 7.1509$

The new distribution of the components is calculated from the above equation

NEW DISTRIBUTION

The new distribution of the components of distillate and bottoms are calculated from the equation $\ln(D_i/B_i) = 5.0197\alpha_i - 7.1509$ and listed in the following tables.

DISTILLATE

Component	D_i kmol/hr	x_{Di}	K_i (40atm and 40°C)	$K_i * x_{Di}$
H_2O	4.1479	0.006864479	2.6	0.017847644
Co_2	3.4001	0.005626923	3	0.016880769
H_2S	9.9162	0.016410603	1.5	0.024615905
H_2	9.222	0.015261752	22	0.335758543
CH_4	21.3484	0.035330079	6	0.211980472
C_2H_4	14.5042	0.024003416	1.8	0.043206149
C_2H_6	45.5784	0.075429	1.3	0.0980577
C_3H_6	140.0974	0.2318512	0.525	0.12172188
C_3H_8	84.45	0.139758724	0.46	0.064289013
C_4H_8	107.228	0.177454689	0.19	0.033716391
IC_4H_{10}	61.5615	0.10187989	0.22	0.022413576
nC_4H_{10}	34.973	0.057877819	0.17	0.009839229
C_5H_{12}	51.674	0.085516783	0.6	0.05131007
C_6	9.224	0.015265062	0.0225	0.000343464
$C7$	6.2695	0.010375575	0.0095	9.8568E-05
$C8$	0.66106	0.001094007	0.0052	5.68884E-06
Total	604.25566	1		1.052085061

TABLE 4.9

BOTTOMS

Component	B_i	x_{Bi}	K_i	$K_i x_{Bi}$
C_4H_8	1.0831	0.000735812	2.2	0.001618786
C_5H_{12}	0.5219	0.000354557	1.7	0.000602746
C_6	77.7144	0.052795848	1.3	0.068634602
C_7	1051.91	0.714622773	1	0.714622773
C_8	340.75	0.231491011	0.9	0.20834191
Total	1471.9794	1		0.993820817

Calculation of minimum reflux ratio for the stripper by UNDERWOOD METHOD

Relative volatality at average temperature of 165°C and 40atm

Component	F_i kmol/hr	x_f	K_i (40atm,165°C)	$\alpha=(K_i/K_h)$
H2O	4.1479	0.001997782	5	10.63829787
CO2	3.4001	0.001637614	5.25	11.17021277
H2S	9.9162	0.004776009	3.3	7.021276596
H2	9.22	0.004440694	14.5	30.85106383
CH4	21.3484	0.010282181	70	148.9361702
C2H4	14.5042	0.00698576	27	57.44680851
C2H6	45.5784	0.021952246	18	38.29787234
C3H6	140.0974	0.067476099	8.2	17.44680851
C3H8	84.4529	0.040675646	7.2	15.31914894
C4H8	108.3116	0.05216688	1.1	2.340425532
iC4H10	61.5615	0.0296503	3.6	7.659574468
nC4H10	34.973	0.016844293	2.85	6.063829787
C5H12	52.1961	0.025139576	1.1	2.340425532
nC6	86.9384	0.041872755	0.47	1
nc7	1058.1884	0.509662747	0.19	0.404255319
nC8	341.4177	0.164439416	0.085	0.180851064
Total	2076.2522	1		

The following are the equations to calculate the minimum reflux ratio

$$\sum \frac{x_{Fi}\alpha_{l/h}}{\alpha_{l/h} - \theta} = 1 - q$$

Since the feed is at the boiling point q = 1

By iteration it is found that θ has the value of 1.317125.

Using the value of θ the minimum reflux ratio can be calculated

$$\sum \frac{x_{Di}\alpha_{l/h}}{\alpha_{l/h} - \theta} = 1 + R_{min}$$

$1 + R_{min} = 1.3223$

$R_{min} = 0.3223$

$R_{actual} = 0.5$ (1.5 times of R_{min})

$$\frac{R - R_m}{R + 1} = \frac{0.5 - 0.\overset{\cdot}{3}223}{1.5}$$

$\dfrac{N - N_m}{N + 1}$ at this value of $\dfrac{R - R_m}{R + 1}$ is 0.51 (from Gilliland Graph)

Calculating number of stages by Fenske equation

$$N_m = \frac{\log\left(\frac{x_{Dl}}{x_{Dh}} \frac{x_{Bh}}{x_{Bl}}\right)}{\log \alpha/_h}$$

$$= \frac{\log\left(\frac{0.085516}{0.015265} * \frac{0.052795}{0.0003545}\right)}{\log(2.34)}$$

$= 7.9122$

Now we have,

$$\frac{N - 7.9122}{N + 1} = 0.51$$

N - 7.9122 = 0.51N + 0.51

0.49N = 8.4222

N = 17.188

The number of trays in the absorber comes around 17.188

Calculating the number of trays in stripping section and rectifying section by Hengestebeck - Geddes equation

$$\frac{N_R}{N_S} = \frac{\log\left[\left(\frac{x_l}{x_h}\right)_D \left(\frac{x_h}{x_l}\right)_F\right]}{\log\left[\left(\frac{x_l}{x_h}\right)_F \left(\frac{x_h}{x_l}\right)_B\right]}$$

$$= \frac{\log\left[\left(\frac{0.08551}{0.01526}\right)\left(\frac{0.04187}{0.03513}\right)\right]}{\log\left[\left(\frac{0.02513}{0.04187}\right)\left(\frac{0.05279}{0.0003545}\right)\right]}$$

$= 0.4972$

$N = N_R + N_S$

$\quad = 0.4972 N_S + N_S$

$$N_S = \frac{N}{1.4972} = \frac{17.1}{1.4972}$$

$\quad = 11.42$

Similarly, $N_R = 17.1 - 11.42 = 5.678$

Taking an efficiency of 70%

$$N_S = \frac{11.42}{0.7} = 16.31 \approx 17$$

$$N_R = \frac{5.678}{0.7} = 8.11 \approx 8$$

So the total number of plates in the column is

$N = N_R + N_S$

$\quad = 17 + 8 = 25$

The total number of trays in the column is 25

Material Balance

$$V = L + D$$

$$= RD + D$$

$$= (R + 1) D$$

$$= 1.5 * 604.2572 = 906.38 \text{ kmol/hr}$$

$$L = V - D = 906.38 - 604.25 = 302.125 \text{ kmol/hr}$$

Diameter of the Column

Flow rate   = 2076 kmols/hr

The tower diameter is to be based on this liquid load.

Density of the liquid = $\rho_L^{190^0C}$ = 480 kg/m^3

Molecular Weight of absorbent is calculated from the table

Component	x_i	M_{wi}	$x_i * M_{wi}$
Hexane	0.02656	86	2.28416
Benzene	0.02604	78	2.03112
2-3 Dimethyl Pentene	0.04016	100	4.016
Pentane	0.09955	100	9.955
Tolune	0.57656	92	53.04352
Octane	0.04269	114	4.86666
Xylene	0.18841	106	19.97146
Total		676	96.16792

Molecular Weight of absorbent = 96.16792

$$\text{Flow rate} = \frac{2076 * 96.16792}{480 * 3600} = 0.11533 \text{ m}^3/\text{sec}$$

Taking v the velocity in the down spout as 0.1 m/sec

$$S_d = \frac{L}{0.12} = \frac{0.115333}{0.1} = 1.1533 \text{ m}^2$$

But $\dfrac{S_d}{S_{col}} = 0.1$

$$S_{col} = \frac{S_d}{0.1} = \frac{1.15333}{0.1} = 11.5333 \text{ m}^2$$

$$\frac{\pi}{4}D_e^2 = 11.533$$

$$D_e = \sqrt{\frac{11.5333*4}{\pi}}$$

$$= 3.8325 \text{ m}$$

The diameter of the tower is 3.8325m

Computation of Tower Height

Plate Spacing = 0.4m

Number of Plates = 25

Free space at the top	(2 * 0.4)	0.8
Three sets of four plates	(3 * 3 * 0.4)	3.6
One set of two plates	(1 * 2 * 0.4)	0.8
One set of two plates	(1 * 0.4)	0.4
Four man holes	(4 * 0.7)	2.8
Two free spaces	(2 * 0.4)	0.8
Two sets of four plates	(2 * 3 * 0.4)	2.4
One man hole	(1 * 0.7)	0.7
Free space at the bottom	(1 * 2.0)	2
Total		**14.3m**

The total height of the column is 14.3m

HEAT BALANCE

H_V = 8138 kcals/hr

h_L = 4582 kcals/hr

Condenser heat duty (Q_c) = $V H_V$ - [$D h_D$ - $L h_L$]

$$= V [H_V - h_L]$$

$$= 906 (8138 - 4582)$$

$$= 3223069.5 \text{ kcals/hr}$$

$B h_B$ = 1471 * 27045 = 39783195 kcals/hr

$L h_L$ = 302 * 4582 = 1383764 kcals/hr

$F h_F$ = 906 * 8138 = 7373028 kcals/hr

Reboiler heat duty (Q_B) = $V H_V$ + $B h_B$ - $(F h_F + L h_L)$

$$= 7373028 + 39783195 - 43351459 - 1383764$$

$$= 3421000 \text{ kcals/hr}$$

Water Requirement

$$= \frac{3221736}{17} = 189513 \text{ kg/hr}$$

The quantity of the water required to generate the steam in the column is 189513 kg/hr

Steam Requirement

$$= \frac{3421000}{7000} \qquad \text{(Latent heat of vaporization } \lambda = 7000 \text{ kcal/kg)}$$

$$= 488.7 \text{ kg/hr}$$

The amount of fuel required in the column to strip the volatile components is 488.7 kg/hr

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SIMULATION RESULTS OF ABSORBER AND STRIPPER BY HYSIS SOFTWARE

The product composition of the individual components from the Absorber Column are provided in the following table

Component	Top Composition	Bottom Composition
H_2O	1.82E-02	1.47E-03
Co_2	1.32E-02	9.35E-04
H_2S	1.11E-02	3.63E-03
H_2	0.321824676	1.27E-03
CH_4	0.166966766	3.79E-03
C_2H_4	0.317060398	3.04E-02
C_2H_6	6.61E-02	8.69E-03
C_3H_6	4.55E-02	3.32E-02
C_3H_8	2.74E-02	1.86E-02
C_4H_8	9.66E-05	4.39E-02
IC_4H_{10}	3.09E-04	2.48E-02
nC_4H_{10}	7.98E-06	1.42E-02
C_5H_{12}	3.24E-10	2.12E-02
nC_6	1.38E-03	2.41E-02
Benzene	8.34E-04	2.03E-02
2,3 - diethyl pentane	9.92E-04	3.14E-02
nc_7	1.97E-03	7.82E-02
Tolune	6.11E-03	0.456196745
nC_8	3.20E-04	3.39E-02
Xylene	5.51E-04	0.149978124

TABLE 4.10

Stage wise temperature and pressure profiles of absorber column

Stages	Pressure(kpa)	Temperature(°C)
0	840.9975	38.27091552
1	843.2491	40.59047579
2	845.5008	41.80254811
3	847.7525	42.70074294
4	850.0041	43.63289181
5	852.2558	44.82949982
6	854.5075	46.45435196
7	856.7591	48.53147117
8	859.0108	50.62294427
9	861.2625	50.26989618

TABLE 4.11

The stage wise compositions of the components in the Absorber Column

Stages	H2o	Co2	H2s	H2	CH4	C2H4	C2H6	C3H6	C3H8	C4H8
0	1.80E-03	1.17E-03	3.65E-03	1.35E-03	4.46E-03	3.63E-02	9.95E-03	2.32E-02	1.33E-02	1.60E-04
1	1.77E-03	1.13E-03	4.51E-03	1.23E-03	4.07E-03	3.63E-02	1.04E-02	3.36E-02	1.89E-02	4.46E-04
2	1.71E-03	1.09E-03	4.68E-03	1.22E-03	3.97E-03	3.52E-02	1.01E-02	3.86E-02	2.15E-02	9.68E-04
3	1.67E-03	1.06E-03	4.64E-03	1.22E-03	3.93E-03	3.44E-02	9.88E-03	4.07E-02	2.25E-02	1.91E-03
4	1.65E-03	1.05E-03	4.54E-03	1.22E-03	3.90E-03	3.39E-02	9.71E-03	4.12E-02	2.27E-02	3.58E-03
5	1.62E-03	1.03E-03	4.41E-03	1.23E-03	3.88E-03	3.33E-02	9.53E-03	4.06E-02	2.23E-02	6.43E-03
6	1.58E-03	1.00E-03	4.23E-03	1.24E-03	3.85E-03	3.26E-02	9.30E-03	3.91E-02	2.15E-02	1.10E-02
7	1.54E-03	9.75E-04	4.01E-03	1.25E-03	3.81E-03	3.16E-02	9.01E-03	3.69E-02	2.03E-02	1.80E-02
8	1.49E-03	9.43E-04	3.77E-03	1.26E-03	3.77E-03	3.06E-02	8.71E-03	3.43E-02	1.90E-02	2.80E-02
9	1.47E-03	9.35E-04	3.63E-03	1.27E-03	3.79E-03	3.04E-02	8.69E-03	3.32E-02	1.86E-02	4.39E-02

(Continued)

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Stages	iC4H10	nC4H10	C5H12	C6	Benzene	2,3 Dimethyk pentane	n heptane	Tolune	n Octane	Xylene
0	4.15E-04	1.55E-05	1.94E-09	2.42E-02	2.36E-02	3.64E-02	9.01E-02	0.52124	3.86E-02	0.170083446
1	1.00E-03	4.79E-05	1.50E-08	2.38E-02	2.32E-02	3.57E-02	8.84E-02	0.511025	3.78E-02	0.166662168
2	1.87E-03	1.17E-04	1.01E-07	2.36E-02	2.30E-02	3.54E-02	8.77E-02	0.506668	3.75E-02	0.165184192
3	3.15E-03	2.64E-04	6.55E-07	2.35E-02	2.29E-02	3.53E-02	8.73E-02	0.50409	3.73E-02	0.164289027
4	4.97E-03	5.66E-04	4.11E-06	2.35E-02	2.28E-02	3.51E-02	8.70E-02	0.501985	3.71E-02	0.16353362
5	7.48E-03	1.17E-03	2.50E-05	2.34E-02	2.28E-02	3.50E-02	8.66E-02	0.49961	3.69E-02	0.162662663
6	1.07E-02	2.31E-03	1.45E-04	2.33E-02	2.27E-02	3.48E-02	8.61E-02	0.496348	3.67E-02	0.161470761
7	1.46E-02	4.36E-03	7.90E-04	2.32E-02	2.25E-02	3.45E-02	8.53E-02	0.49139	3.63E-02	0.159725373
8	1.89E-02	7.81E-03	4.06E-03	2.29E-02	2.20E-02	3.38E-02	8.36E-02	0.482464	3.56E-02	0.156899002
9	2.48E-02	1.42E-02	2.12E-02	2.41E-02	2.03E-02	3.14E-02	7.82E-02	0.456197	3.39E-02	0.149978124

TABLE 4.12

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The product composition of the individual components from the Stripper Column are listed in the following table

Component	Product Composition	Solvent Composition
H_2O	3.89E-03	1.14E-20
CO_2	2.47E-03	1.58E-21
H_2S	9.61E-03	6.63E-17
H_2	3.35E-03	1.99E-29
CH_4	1.00E-02	7.05E-24
C_2H_4	8.03E-02	6.40E-19
C_2H_6	2.30E-02	1.07E-18
C_3H_6	8.78E-02	1.07E-14
C_3H_8	4.92E-02	4.61E-15
C_4H_8	0.116077617	3.35E-11
IC_4H_{10}	6.56E-02	5.19E-12
nC_4H_{10}	3.75E-02	3.06E-11
C_5H_{12}	5.60E-02	6.57E-08
nC_6	6.35E-02	7.57E-05
Benzene	5.13E-02	1.37E-03
2,3 - diethyl pentane	6.89E-02	8.65E-03
nc_7	0.136697054	4.26E-02
Tolune	0.133891474	0.652146663
nC_8	7.91E-04	5.40E-02
Xylene	4.35E-07	0.241159119

TABLE 4.13

Stage wise temperature and pressure profiles of Stripper column

Stages	Pressure(kpa)	Temperture(°C)
0	101.325	71.1581412
1	103.325	87.1728133
2	104.1403	89.8324415
3	104.9556	91.1034797
4	105.7708	91.9898238
5	106.5861	92.6712029
6	107.4014	93.2255008
7	108.2167	93.7006252
8	109.0319	94.1481988
9	109.8472	94.6362432
10	110.6625	95.2812418
11	111.4778	96.2962534
12	112.2931	98.03695
13	113.1083	109.301564
14	113.9236	111.216739
15	114.7389	112.307957
16	115.5542	113.287842
17	116.3694	114.306084
18	117.1847	115.494154
19	118	117.177477
20	118	120.063871

TABLE 4.14

The stage wise compositions of the components in the Stripper Column

Stages	H2o	Co2	H2s	H2	CH4	C2H4	C2H6	C3H6	C3H8	C4H8
0	3.68E-05	2.04E-05	2.23E-04	2.65E-06	3.53E-05	9.17E-04	3.46E-04	3.56E-03	2.08E-03	1.31E-02
1	1.67E-05	9.16E-06	9.85E-05	1.39E-06	1.67E-05	4.08E-04	1.50E-04	1.49E-03	8.56E-04	5.49E-03
2	1.59E-05	8.68E-06	9.37E-05	1.32E-06	1.57E-05	3.84E-04	1.39E-04	1.37E-03	7.73E-04	4.80E-03
3	1.57E-05	8.58E-06	9.38E-05	1.29E-06	1.54E-05	3.78E-04	1.36E-04	1.34E-03	7.50E-04	4.67E-03
4	1.57E-05	8.55E-06	9.43E-05	1.27E-06	1.53E-05	3.75E-04	1.35E-04	1.33E-03	7.38E-04	4.61E-03
5	1.57E-05	8.55E-06	9.47E-05	1.26E-06	1.52E-05	3.74E-04	1.34E-04	1.32E-03	7.31E-04	4.57E-03
6	1.58E-05	8.56E-06	9.52E-05	1.26E-06	1.52E-05	3.75E-04	1.34E-04	1.32E-03	7.28E-04	4.55E-03
7	1.58E-05	8.60E-06	9.56E-05	1.27E-06	1.53E-05	3.76E-04	1.34E-04	1.32E-03	7.27E-04	4.55E-03
8	1.59E-05	8.64E-06	9.60E-05	1.28E-06	1.54E-05	3.77E-04	1.34E-04	1.32E-03	7.28E-04	4.55E-03
9	1.60E-05	8.68E-06	9.65E-05	1.29E-06	1.55E-05	3.79E-04	1.35E-04	1.33E-03	7.30E-04	4.55E-03
10	1.61E-05	8.72E-06	9.69E-05	1.30E-06	1.56E-05	3.80E-04	1.35E-04	1.33E-03	7.30E-04	4.54E-03
11	1.61E-05	8.74E-06	9.75E-05	1.31E-06	1.57E-05	3.80E-04	1.35E-04	1.32E-03	7.27E-04	4.50E-03
12	1.61E-05	8.72E-06	9.82E-05	1.32E-06	1.57E-05	3.78E-04	1.34E-04	1.30E-03	7.15E-04	4.41E-03
13	2.16E-07	9.93E-08	3.10E-06	1.93E-09	8.43E-08	5.73E-06	2.49E-06	5.82E-05	3.14E-05	4.71E-04
14	2.79E-09	1.08E-09	9.38E-08	2.73E-12	4.35E-10	8.31E-08	4.42E-08	2.47E-06	1.30E-06	4.75E-05
15	3.57E-11	1.18E-11	2.82E-09	3.86E-15	2.23E-12	1.20E-09	7.76E-10	1.03E-07	5.33E-08	4.71E-06
16	4.57E-13	1.27E-13	8.46E-11	5.45E-18	1.14E-14	1.71E-11	1.35E-11	4.29E-09	2.15E-09	4.61E-07
17	5.82E-15	1.36E-15	2.54E-12	7.65E-21	5.78E-17	2.43E-13	2.32E-13	1.76E-10	8.56E-11	4.46E-08
18	7.39E-17	1.45E-17	7.60E-14	1.06E-23	2.91E-19	3.43E-15	3.96E-15	7.16E-12	3.35E-12	4.24E-09
19	9.29E-19	1.54E-19	2.27E-15	1.47E-26	1.45E-21	4.77E-17	6.62E-17	2.85E-13	1.28E-13	3.92E-10
20	1.14E-20	1.58E-21	6.63E-17	1.99E-29	7.05E-24	6.40E-19	1.07E-18	1.07E-14	4.61E-15	3.35E-11

(Continued)

Stages	iC4H10	nC4H10	C5H12	C6	Benzene	2,3 Dimethyl pentane	n heptane	Tolune	n Octane	Xylene
0	6.40E-03	4.93E-03	1.95E-02	5.72E-02	6.03E-02	0.120169183	0.29605783	0.411124	4.04E-03	4.08E-06
1	2.66E-03	2.07E-03	9.10E-03	3.49E-02	4.29E-02	0.101761749	0.27950474	0.511662	6.83E-03	1.20E-05
2	2.33E-03	1.79E-03	7.23E-03	2.62E-02	3.44E-02	8.50E-02	0.24692571	0.578595	9.86E-03	3.12E-05
3	2.26E-03	1.73E-03	6.81E-03	2.30E-02	3.06E-02	7.35E-02	0.21656961	0.625077	1.30E-02	7.74E-05
4	2.23E-03	1.71E-03	6.64E-03	2.16E-02	2.89E-02	6.61E-02	0.19276477	0.656401	1.62E-02	1.87E-04
5	2.21E-03	1.69E-03	6.55E-03	2.09E-02	2.81E-02	6.14E-02	0.17568349	0.676322	1.94E-02	4.43E-04
6	2.20E-03	1.68E-03	6.49E-03	2.05E-02	2.76E-02	5.86E-02	0.16408336	0.687842	2.26E-02	1.04E-03
7	2.20E-03	1.68E-03	6.46E-03	2.03E-02	2.74E-02	5.69E-02	0.15647585	0.693024	2.60E-02	2.41E-03
8	2.20E-03	1.68E-03	6.45E-03	2.02E-02	2.71E-02	5.58E-02	0.1515034	0.692888	2.94E-02	5.54E-03
9	2.20E-03	1.68E-03	6.43E-03	2.01E-02	2.69E-02	5.49E-02	0.14795622	0.687055	3.29E-02	1.26E-02
10	2.19E-03	1.68E-03	6.39E-03	1.99E-02	2.66E-02	5.40E-02	0.14456694	0.673034	3.62E-02	2.82E-02
11	2.18E-03	1.66E-03	6.30E-03	1.95E-02	2.60E-02	5.24E-02	0.13969016	0.645101	3.87E-02	6.13E-02
12	2.13E-03	1.62E-03	6.11E-03	1.86E-02	2.49E-02	4.95E-02	0.13115717	0.593966	3.91E-02	0.125826
13	1.98E-04	1.95E-04	1.67E-03	1.11E-02	2.07E-02	4.86E-02	0.13764207	0.616238	4.00E-02	0.123024
14	1.73E-05	2.21E-05	4.25E-04	6.18E-03	1.59E-02	4.42E-02	0.13493263	0.633452	4.13E-02	0.123489
15	1.49E-06	2.47E-06	1.06E-04	3.33E-03	1.19E-02	3.88E-02	0.12804652	0.650587	4.29E-02	0.124294
16	1.27E-07	2.71E-07	2.61E-05	1.75E-03	8.70E-03	3.29E-02	0.1174665	0.668664	4.49E-02	0.12558
17	1.06E-08	2.94E-08	6.29E-06	8.94E-04	6.14E-03	2.68E-02	0.10334415	0.68699	4.75E-02	0.128406
18	8.75E-10	3.13E-09	1.48E-06	4.36E-04	4.13E-03	2.06E-02	8.59E-02	0.70188	5.05E-02	0.136494
19	7.00E-11	3.23E-10	3.32E-07	1.97E-04	2.57E-03	1.45E-02	6.56E-02	0.701386	5.36E-02	0.162208
20	5.19E-12	3.06E-11	6.57E-08	7.57E-05	1.37E-03	8.65E-03	4.26E-02	0.652147	5.40E-02	0.241159

TABLE 4.15

RESULTS AND CONCLUSION

The absorber and stripper design calculations are carried out manually and simulated in the hysis software and the percentage of deviation is tabulated.

Absorption Column

Parameter	Calculated Results	Hysis Results
Solvent flow rate kmols/hr	2076.25	1870.45
Number of stages	10	10
Diameter, m	3.08	3.5
Column Height, m	7.43	8

Stripper Column

Parameter	Calculated Value	Designed Value
Column Height, m	14.3	14.8
Diameter, m	3.82	3.9
Number of stages	25	25

From the above results it is observed that the calculated value of solvent flow rate is deviated from the designed value, which might have resulted out of manual errors in reading the graph. There can be still improvement in the results if suitable assumptions are made.

Column diameter is calculated and based on the diameter column height is calculated suitably referring to the normal design practices. Hence, this may vary from one original equipment supplier (OES) to other.

References

1. Engineering Data Book, SI Version , Gas Processors Suppliers Association (GPSA), Eleventh Edition – SI, 1998

2. Ludwig, Ernest E; Applied Process Design for Chemical and Petrochemical Plants, Third Edition,Volume – 2, Gulf Professional Publishers, Houston

3. Mc Cabe, Warren L, Smith, Julian C; Harriot, Peter; Unit Operations of Chemical Engineering, Sixth Edition, McGraw-Hill Publications

4. James O. Maloney, Perry's Chemical Engineers' Handbook, 8th Edition, McGraw-Hill Publications

5. Treybal, Robert E, 'Mass Transfer Operations', Third Edition, 2002 McGraw-Hill Publications.

6. R.K.Sinnot, Coulson & Richardson's Chemical Engineering Design, Third Edition.Vol-6, Published by Butterworth Heinemann, 2003,Oxford.

7. Donald Q. Kern, 'Process Heat Transfer', McGraw-Hill Publications, 1950, Singapore.

8. James G. Speight, Chemical and Process Design Hand Book, McGraw-Hill Publications

9. Jean-Pierre Wauquier, Seperation Processes, t Editions Technip, Paris

10. David S. J. Stan Jones and Peter R. Pujado, Hand Book of Petroleum Processing, Springer Publications