DESIGNING OF DEPROPANIZER TRAY COLUMN USING ASPEN HYSYS AS A SIMULATOR

By
JAI DUBEY
&
MAYANK KANODIA

विद्यया अमृतम् श्नुते

COLLEGE OF ENGINEERING
UNIVERSITY OF PETROLEUM & ENERGY STUDIES
DEHRADUN
MAY, 2009

“DESIGNING OF DEPROPANIZER TRAY COLUMN USING ASPEN HYSYS AS A SIMULATOR”

A dissertation submitted in partial fulfillment of the requirements for the Degree of
Bachelor of Technology
(Applied Petroleum Engineering)

By
JAI DUBEY
&
MAYANK KANODIA

Under the guidance of
Mrs. PULLA ROSE HAVILAH
Asst. Professor, COE, UPES

Approved By
Dr. B.P. Pandey
Dean Emeritus, COE, UPES

College of Engineering
University of Petroleum & Energy Studies
Dehradun
May, 2009

ACKNOWLEDGEMENT

We express our heartiest thanks to **Dr. B.P. PANDEY** (Dean Emeritus, College Of Engineering, UPES) for his constant endeavor and full cooperation in completion of the project.

We acknowledge indebtness to our mentor, **Mrs. PULLA ROSE HAVILAH** (Asst. Professor, College Of Engineering, UPES) whose supervision, constant inspiration and guidance helped us throughout our work presented in this report.

We also express our deep sense of gratitude to **Mr. RAJESHWAR MAHAJAN** and **Mr. VINAY AVASTHI** (Course Coordinators, College Of Engineering, UPES) for giving us valuable information regarding the curriculum during the entire academic year.

At last we would like to thank **Mr. KARUN GARG** (Senior Engineer, Engineers India Limited) and our family and friends who have been with us throughout all the thick and thin of our life.

Jai
Mayank

Date: 08th MAY 2009

Jai Dubey

&

Mayank Kanodia

CERTIFICATE

This is to certify that the work contained in this thesis titled "DESIGNING OF A DEPROPANIZER TRAY COLUMN USING ASPEN HYSYS AS A SIMULATOR" has been carried out by JAI DUBEY & MAYANK KANODIA under my supervision and has not been submitted elsewhere for a degree.

P. Rose Havilah

Mrs. Pulla Rose Havilah
Asst. Professor
College Of Engineering
University of Petroleum & Energy Studies
Dehradun

Date: 8/5/09

ABSTRACT

Distillation utilizes vapor and liquid phases at essentially the same temperature and pressure for the coexisting zones. Plates or trays are used to bring the two phases into intimate contact. Trays are stacked one above the other and enclosed in a cylindrical shell to form a column. The overall flow pattern in a distillation column provides countercurrent contacting of vapor and liquid streams on all the trays through the column. These phases on a given tray approach thermal, pressure, and composition equilibriums to an extent dependent upon the efficiency of the contacting tray. Column process design specifies the separation, and sets column and utility pressure, reflux, stages, and feed point. These in turn yield internal flows and reboiler and condenser duties.

Designing of Depropanizer column deals in determining various parameters like number of equilibrium stages, column internals, thermodynamic properties of fluid, etc. Initial data for designing column is taken form a refinery. Column is designed using Fenske Underwood Gilliland theoretical method. Consecutively, calculation of column internals is also done. A simulation for depropanizer column is done using Aspen Hysys as simulating software. Thus, a correlating study is done between the theoretical and simulation method. A complete understanding of software is done under the able guidance of mentor before using it. A layout model of column is also prepared based upon the data obtained. This project provides the designing of the depropanizer column and use of simulation software to validate the calculated data.

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

ABBREVIATIONS

EOS	Equation of state
RK	Redlich Kwang
VLE	Vapor Liquid Equilibrium
PR	Peng Robinson
PFD	Process flow diagram
SB	Smith Brinkley
FUG	Fenske Underwood Gilliland

NOMENCLATURE

$[x_i/ x_r]$	The ratio of the concentration of any component i to the concentration of a reference component r
x_{LK}	Light key concentrations
x_{HK}	Heavy key concentrations
$X_{f, HK}$	Concentration of the heavy key in the feed
$X_{f, LK}$	Concentration of the light key in the feed
$X_{d, HK}$	Concentration of the heavy key in the top product
$X_{b, LK}$	Concentration of the light key if in the bottom product
N_m	Minimum number of stages at total reflux, including the reboiler
N_r	Number of stages above the feed, including any partial condenser
N_s	Number of stages below the feed, including the reboiler
N_{min}	Minimum number of stages
α_i	Average relative volatility of the component / with respect to the reference component
α_{LK}	Average relative volatility of the light key with respect to the heavy key
q	Thermal condition of the feed
R_m	Minimum reflux
B	Molar flow bottom product
D	Molar flow top product
u_v	Maximum allowable vapor velocity, based on the gross (total) column cross – sectional area, m/s
l_t	Plate spacing, m
D_c	Column diameter, m
V_w	Maximum vapor rate, kg/s
A_c	Total column cross-sectional area, m^2
A_d	Cross-sectional area of downcomer, m^2
A_n	Net area available for vapor-liquid disengagement, m^2
A_a	Active or bubbling, area, m^2
A_h	Hole area, the total area of all the active holes, m^2

A_p	Perforated area (including blanked areas), m^2
A_{ap}	The clearance area under the downcomer apron, m^2
u_f	Flooding vapor velocity, m/s
F_{LV}	Liquid-vapor flow factor
L_w	Liquid mass flow-rate, kg/s
L_{wd}	Liquid flow rate in downcomer, kg/s
V_w	Vapor mass flow-rate, kg/s
u_h	Minimum vapor velocity through the holes(based on the hole area), m/s
d_h	Hole diameter, mm
ρ_V	Density of vapor, Kg/m^3
ρ_L	Density of liquid, Kg/m^3
h_w	Height of the weir, mm
h_{ow}	Depth of the crest of liquid over the weir, mm
h_d	The dry plate drop, mm liquid
h_r	Residual head loss, mm liquid
h_t	Total plate pressure drop, mm liquid
h_b	Downcomer back-up, measured from plate surface, mm
h_{dc}	Head loss in the downcomer, mm
h_{ap}	Height of the bottom edge of the apron above the plate, mm
h_{bc}	Clear liquid back-up, mm
t_r	Residence time, sec
C_o	Orifice coefficient
ΔP_t	Total plate pressure drop, Pa (N/m^2)
l_w	Weir length, mm
l_p	Hole pitch, mm
T_C	Critical temperature, °C
T_R	Residual temperature, °C
P_C	Critical pressure, psi
H^{IG}	Ideal gas enthalpy calculated at temperature, T
S^{IG}	Ideal gas entropy calculated at temperature, T
C_P^{IG}	Ideal gas heat capacity calculated at temperature, T
Φ_i	Fugacity coefficient

CHAPTER 1

INTRODUCTION

1. INTRODUCTION

1.1 OVERVIEW OF DISTILLATION

Distillation is simply defined as a process in which a liquid or vapor mixture of two or more substances is separated into its component fractions of desired purity, by the application and removal of heat. The process is based on the fact that the vapor of a boiling mixture will be richer in the components that have lower boiling points. Hence, when this vapor is cooled and condensed, the condensate will contain more volatile components. At the same time, the original mixture will contain more of the less volatile material. The primary equipment employed in the process of distillation are distillation columns, which are designed to achieve this separation efficiently.

The best way to reduce operating costs of existing units is to improve their efficiency and operation via process optimization and control. To achievé this improvement, a thorough understanding of distillation principles and how distillation systems are designed is essential. As stated, distillation is the process of heating a liquid until some of its ingredients pass into the vapor phase, and then cooling the vapor to recover it in liquid form by condensation. The main purpose of distillation is to separate a mixture. If the difference in boiling points between two substances is great, complete separation may be easily accomplished by a single-stage distillation. If the boiling points differ only slightly, many redistillations may be required. In the simplest mixture of two mutually soluble liquids with similar chemical structures, the readiness to vaporize of each is undisturbed by the presence of the other. For example, would be halfway between the boiling points of the pure substances, and the degree of separation achieved by a single distillation would depend only on each substance's readiness to vaporize at this temperature. This simple law was first stated by 19th- century by the French chemist Frangois Marie Raoult (known as Raoult's law).

The term "still" is applied only to the vessel in which liquids are boiled during distillation, but the term is sometimes applied to the entire apparatus, including the fractionating column, the condenser, and the receiver in which the distillate is collected. If a water and alcohol distillate is returned from the condenser and made to drip down through a long column a series of plates, and if the vapor, as it rises to the condenser, is made to bubble through this liquid at each plate, the vapor and liquid will interact so that some of the water in the vapor condenses and some of the alcohol in the liquid vaporizes. The interaction at each plate is equivalent to a redistillation. This process is referred to by several names in the industry; namely rectification, fractionation, or fractional distillation.

If two insoluble liquids are heated, each is unaffected by the presence of the other and vaporizes to an extent determined only by its own nature. Such a mixture always boils at a temperature lower than is true for either substance alone. This effect may be applied to substances that would be damaged by overheating if distilled in the usual fashion. Substances can also be distilled at temperatures below their normal boiling points by partially evacuating the still. The greater the vacuum, the lower is the distillation temperature.

1.2 BASIC COMPONENTS OF DISTILLATION COLUMN

There are a variety of configurations for distillation columns, each designed to perform specific types of separations. The two major types are batch and continuous columns. In a batch operation, the feed to the column is introduced batch-wise. That is, the column is charged with 'batch' and then the distillation process is conducted. When the desired separation is achieved, a next batch of feed is introduced. In contrast, continuous columns process a continuous feed stream. They are capable of handling high throughputs and are more common of the two types.

Continuous columns can be further classified according to:

- The nature of the feed that they are processing (binary column - feed contains only two components, and multi-component column - feed contains more than two components);
- The number of product streams they have (multiproduct column - column has more than two product streams);
- Where the extra feed exits when it is used to help with the separation (extractive distillation – where the extra feed appears in the bottom product stream , and azeotropic distillation - where the extra feed appears at the top product stream);
- The type of column internals (tray column - where trays of various designs are used to hold up the liquid to provide better contact between vapor and liquid, and hence achieve better separation, and the packed column - where instead of trays, packings are employed to effect contact between vapor and liquid).

There are several important components in a distillation column, each of which is used either to transfer heat energy or enhance mass transfer. The major components in a typical distillation are:

- **Vertical shell** where the separation of liquid components is carried out,
- **Column internals** such as tray, plates and/or packings which are used to enhance component separations,
- **Reboiler** to provide the necessary vaporization for the distillation process
- **Condenser** to cool and condense the vapor leaving the top of the column
- **Reflux drum** to hold the condensed vapor from the top of the column
- **Liquid (reflux)** is recycled back to the column.

The column internals are housed within a vertical shell, and together with the condenser and reboiler, constitute a distillation column. The liquid mixture that is to be processed is called the feed. The feed introduced usually somewhere near the middle of the column to a tray known as the feed tray. The feed tray divides the column into a top (enriching or rectification) section and a bottom (stripping) section. The feed flows down the column where it is collected at the bottom in the reboiler. Heat is supplied to the reboiler to generate vapor. The source of heat input can be any suitable fluid, although in most chemical plants this is normally steam. In refineries, the heating source may be the output streams of other columns. The vapor raised in the reboiler is re-introduced into the unit at the bottom of the column. The liquid removed from the reboiler is known as the bottoms product or simply, the bottoms.. The vapor travels up the column, and as it exits the top of the unit, it is cooled by a condenser. The condensed liquid is stored in a holding vessel known as the reflux drum. Some of this liquid is recycled back to the top of the column and this is called the reflux. The condensed liquid that is removed from the system is known as the distillate or top product.

1.3 CONTINUOUS DISTILLATION: PROCESS DESCRIPTION

The separation of liquid mixtures by distillation depends on differences in volatility between the components. The greater the relative volatilities, the easier are the separation. The basic equipment required for continuous distillation is shown in Figure 1. Vapor flows up the column and liquid counter-currently down the column. The vapor and liquid are brought into contact on plates, or packing. Part of the condensate from the condenser is returned to the top of the column to provide liquid flow above the feed point (reflux), and part of the liquid from the base of the column is vaporized in the reboiler and returned to provide the vapor flow. In the section below the feed, the more volatile components are stripped from the liquid and this is known as the stripping section. Above the feed, the concentration of the more volatile components is increased and this is called the enrichment, or more commonly, the rectifying section.

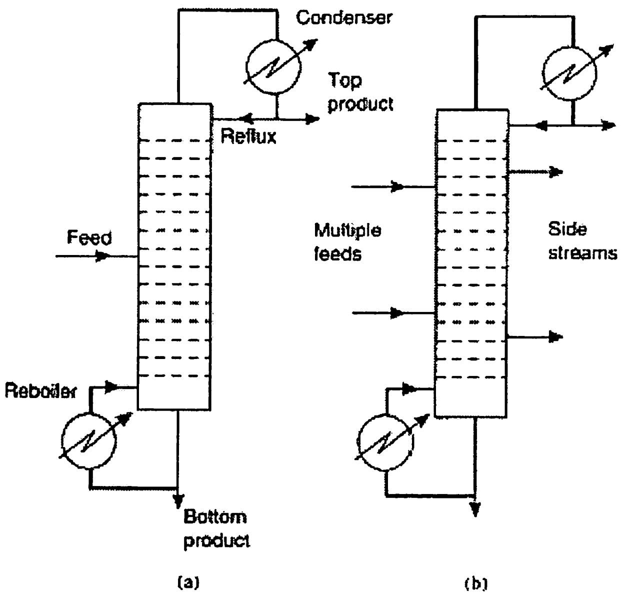


Figure 1 Distillation column (*a*) Basic column (*b*) Multiple feeds and side streams

Figure 1(a) shows a column producing two product streams, referred to as tops and bottoms, from a single feed. Columns are occasionally used with more than one feed and with side streams withdrawn at points up the column, Figure 1(b). This does not alter the basic operation, but complicates the analysis of the process, to some extent. If the process requirement is to strip a volatile component from a relatively non-volatile solvent, the rectifying section may be omitted, and the column would then be called as stripping column. In some operations, where the top product is required as a vapor, only sufficient liquid is condensed to provide the reflux flow to the column, and the condenser is referred to as a partial condenser. When the liquid is totally condensed, the liquid returned to the column will have the same composition as the top product. In a partial condenser the reflux will be in equilibrium with the vapor leaving the condenser. Virtually pure top and bottom products can be obtained in a single column from a binary feed, but where the feed contains more than two components; only a single "pure" product can be produced, either from the top or bottom of the column. Several columns will be needed to separate a multicomponent feed into its constituent parts. Continuous distillation is of many types but here we are considering only multi component distillation.

1.4 MULTICOMPONENT DISTILLATION

A tower comprised of rectifying (above the feed) and stripping (below the feed) sections is capable of making a more or less sharp separation between two products or pure components of the mixture, that is, between the light and heavy key components. Key components are the two components in the feed mixture whose separation is specified. The more volatile of these components is the light key (Propane), and the less volatile is the heavy key (Butane). Other components are termed nonkeys. The key components appear to a significant extent in both overhead and bottom products. Light nonkeys end up almost exclusively in the overhead product, and heavy nonkeys end up almost exclusively in the bottom product in many separations, components are present whose relative volatilities are intermediate between the light key and the heavy key. These components are termed intermediate keys or distributed keys. Intermediate keys are split between the top and bottom products. We are considering a depropanizer column over here.

1.5 DEPROPANIZER COLUMN

The bottoms from the de-ethanizer serves as the principal feed to the depropanizer. A second feed stream to this column consists of the bottoms liquid from the condensate stripper for the fourth and fifth compression stages. The primary function of this column is propylene and propane recovery from these two feeds. The column operating pressure typically will be 240 to 340 psia, which is sufficient to condense the overhead vapor with cooling water or ambient air. The overhead from this column contains all of the propane present in the feed, as well as some butane that are stripped to provide complete recovery of propane. Ethane content in the distillate is negligible due to the over stripping commonly carried out in the preceding de-ethanizer.

A depropanizer normally contains 35 to 55 actual trays and has a uniform diameter. In these systems, the surface tension of the liquid phase is below 6 dyne/cm, the liquid density is near 30 lb/ft^3, and the vapor density is about 8% of the liquid density. Under such conditions, the downcomer residence time required can be a significant factor in the specification of the tower diameter for a tray column. The downcomer normally is designed so that the froth height is no more than 70% of the tray spacing; therefore, downcomer area must be large to avoid the need for excessively tall columns.

1.6 ADVANTAGES OF TRAY COLUMN

1.6.1 Factors Favoring Tray Columns

The following factors generally favor trays compared to either random or structured packings:

- **Solids:**

  Trays can handle solids a lot easier than packed columns. Both gas and liquid velocities are often an order of magnitude higher on a tray than through packings. These high liquid and gas velocities provide a sweeping action that keeps tray openings and perforations clear. Solids tend to accumulate in the void of packed column. There are fewer locations where solids can be deposited in a tray column.

  Further, packed towers need liquid distributors and plugging in these has been a common trouble spot. Cleaning trays is easier than cleaning random packings, while cleaning structured packings is practically impossible.

- **High Liquid rates:**

  Multipass trays effectively lower the liquid load "seen" by each part of the tray. The capacity of packings, especially structured, tends to rapidly fall off at high liquid rates. It is often more economical to handle high liquid rates in tray columns.

- **Large diameter**:

  Packings are prone to maldistribution problems in large diameter columns. These problems are far less in plate columns.

- **Complex columns**:

  Inter-reboilers, inter-condensers, cooling coils, and side drawoffs are more easily incorporated in tray than in packed columns. In packed columns, every complexity requires additional distribution and/or liquid collection equipment.

- **Feed composition variation**:

  One way of allowing for design uncertainties and feedstock variation is by installing alternate feed points. In packed columns, every alternate feed point requires expensive distribution equipment.

- **Performance prediction**:

  There is greater uncertainty in predicting packed column performance. Greater overdesign is often required.

- **Chemical reaction/absorption**:

  By using high weirs, trays are capable of providing greater residence time for absorption or chemical reaction than packing.

- **Weight**:

  Tray columns usually weigh less than packed columns. This saves on the cost of foundations, supports, and column shell.

- **Intermittent operation**:

  When temperature is either lower or higher than atmospheric, intermittent operation repeatedly expands and contracts the shell. This may crush the packings or damage the shell in a packed column, but is easy to accommodate for in tray columns.

CHAPTER 2

LITERATURE REVIEW

2. LITERATURE REVIEW

2.1 MULTICOMPONENT DISTILLATION METHOD

Some approximate calculation methods for the solution of multicomponent, multistage separation problems continue to serve useful purposes even though computers are available to provide more rigorous solutions. The available phase equilibrium and enthalpy data may not be accurate enough to justify the longer rigorous methods. In design and optimization studies, a large number of cases can be worked quickly and cheaply by an approximate method to define roughly the optimum specifications, which can then be investigated more exactly with a rigorous method.

Two approximate multicomponent shortcut methods for simple distillation are the Smith-Brinkley (SB) method, which is based on an analytical solution of the finite-difference equations that can be written for staged separation processes when stages and interstage flow rates are known or assumed and the Fenske-Underwood-Gilliland (FUG) method, which combines Fenske's total-reflux equation and Underwood's minimum-reflux equation with a graphical correlation by Gilliland that relates actual column performance to total- and minimum- reflux conditions for a specified separation between two key components. Both methods work best when mixtures are nearly ideal.

2.1.1 Fenske-Underwood-Gilliland (FUG) Method

The first step in the design of distillation equipment is specification of light and heavy key components. Then the specific operating conditions and equipment size are established, ultimately on the basis of an economic balance or simply by exercise of judgment derived from experience. The design parameters that need to be determined include intermediate ones such as limiting reflux and trays that are needed for establishing a working design. These design parameters are the following:

- Minimum number of theoretical trays
- Distribution of nonkeys between the overhead and bottoms products
- Minimum reflux
- Operating reflux

- Number of theoretical trays
- Location of the feed tray
- Tray efficiencies

These shortcut methods assume constant molar overflow in the rectifying and stripping zones and constant relative volatilities, which may be taken at the conditions of the feed tray or as a geometric mean of the values at the top and bottom of the column. Since the top conditions are not known completely in advance, evaluation of a mean relative volatility is an iterative process that can be started with the value at the feed tray or at the feed condition. Particular modes of variation of α sometimes are assumed.

2.1.1.1 Minimum Number of Stages

The Fenske equation (Fenske, 1932) can be used to estimate the minimum stages required at total reflux. The equation applies equally to multicomponent systems and can be written as:

$$\left[\frac{x_i}{x_r}\right]_d = \alpha_i^{N_m} \left[\frac{x_i}{x_r}\right]_b$$

Where,

$[x_i / x_r]$ = the ratio of the concentration of any component i to the concentration of a reference component r, and the suffixes d and b denote the distillate (d) and the bottoms (b),

N_m = minimum number of stages at total reflux, including the reboiler

α_i = average relative volatility of the component / with respect to the reference component.

Normally the separation required will be specified in terms of the key components, and the above equation can be rearranged to give an estimate of the number of stages:

$$N_m = \frac{\log \left[\frac{x_{LK}}{x_{HK}}\right]_d \left[\frac{x_{HK}}{x_{LK}}\right]_b}{\log \alpha_{LK}}$$

Where, α_{LK} is the average relative volatility of the light key with respect to the heavy key, and x_{LK} and x_{HK} are the light and heavy key concentrations. The relative volatility is taken as the geometric mean of the values at the column top and bottom temperatures. To calculate these temperatures initial estimates of the compositions must be made, so the calculation of the minimum number of stages by the Fenske equation is a trialand – error procedure. If there is a wide difference between the relative volatilities at the top and bottom of the column the use of the average value in the Fenske equation will underestimate the number of stages. In these circumstances, a better estimate can be made by calculating the number of stages in the rectifying and stripping sections separately; taking the concentration as the base concentration for the rectifying section and as the top concentration for the stripping section, and estimating the average relative volatilities separately for each section. This procedure will also give an estimate of the feed point location.

2.1.1.2 Minimum Reflux

The method of Underwood employs auxiliary parameters derived from the equation:

$$\sum_{i=1}^{C} \frac{\alpha_i x_{Fi}}{\alpha_i - \theta} = 1 - q$$

Where, q is the thermal condition of the feed and the summation extends over all the components in the feed. The only roots required are those in numerical value between the relative volatilities of the light and heavy keys. For instance, if there is one distributed component, subscript dk, the required roots θ_1 & θ_2

$$\alpha_{lk} > \theta_1 > \alpha_{dk},$$
$$\alpha_{dk} > \theta_2 > \alpha_{hk}.$$

Then the minimum reflux and the distribution of the intermediate component are found from the two equations that result from substitution of the two values of θ into Underwood's second equation:

$$R_m + 1 = \frac{1}{D}\sum \frac{\alpha_i d_i}{\alpha_i - \theta}.$$

The number of values of θ=1 plus the number of components with relative volatilities between those of the light and heavy keys. When there is no distributed component, the equation below may be used in terms of mole fractions and only a single form is needed for finding the minimum reflux,

$$R_m + 1 = \sum \frac{\alpha_i x_{iD}}{\alpha_i - \theta}.$$

Occasionally the minimum reflux calculated by this method comes out a negative number. That, of course, is a signal that some other method should be tried, or it may mean that the separation between feed and overhead can be accomplished in less than one equilibrium stage.

2.1.1.3 Operating Reflux

The operating reflux is an amount in excess of the minimum that ultimately should be established by an economic balance between operating and capital costs for the operation. In many cases, however, the assumptions $R = 1.5R_m$, often is close to the optimum and is used without further study unless the installation is quite a large one.

2.1.1.4 Actual Number of Theoretical Plates

An early observation by of the plate-reflux relation was:

$$(R - R_m)(N - N_m) = \text{const},$$

But no general value for the constant was possible. Several correlations of calculated data between these same variables have since been made. A graphical correlation made by Gilliland has found wide acceptance because of its fair accuracy and simplicity of use. Of the several representations of the plot by equations that of Molokanov et al. is accurate and easy to use:

$$Y = \frac{N - N_{\min}}{N+1} = 1 - \exp\left[\left(\frac{1 + 54.4X}{11 + 117.2X}\right)\left(\frac{X - 1}{X^{0.5}}\right)\right],$$

Where,

$$X = \frac{R - R_{\min}}{R+1},$$

From which the number of theoretical trays is

$$N = \frac{N_m + Y}{1 - Y}$$

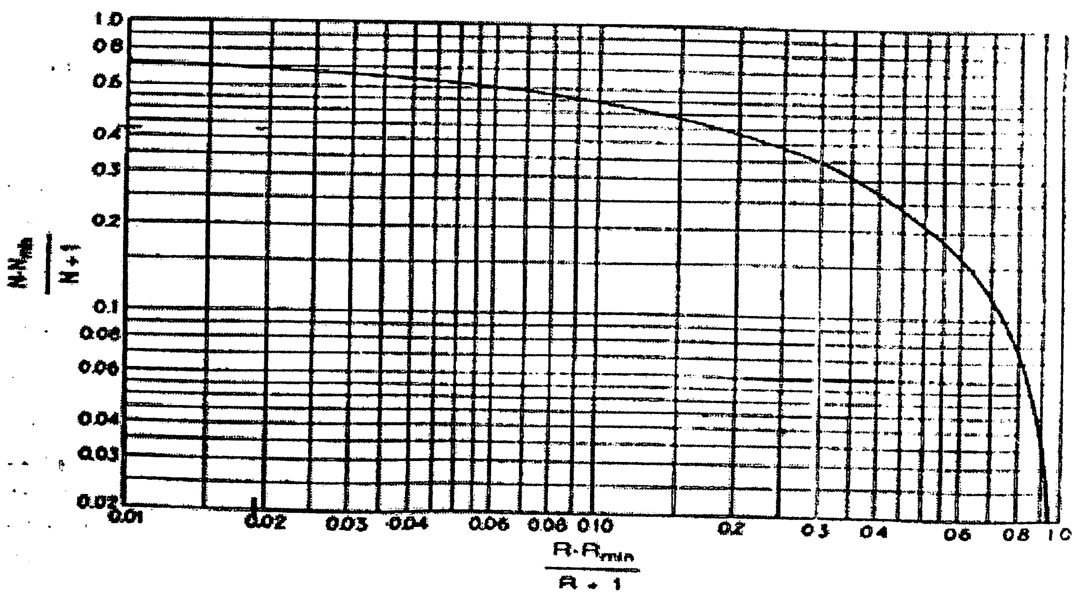


Figure 2: The Gilliland Correlation

The Gilliland correlation appears to be conservative for feeds with values of **q** (the thermal condition of the feed), and can be in error when there is a large difference in tray requirements above and below the feed. The principal value of the correlation appears to be for preliminary exploration of design variables which can be refined by computer calculations. Although it is often used for final design, that should be done with caution.

2.1.1.5 Feed Tray Location

An estimate can be made by using the Fenske equation to calculate the number of stages in the rectifying and stripping sections separately, but this requires an estimate of the feed-point temperature. An alternative approach is to use the empirical equation given by Kirkbride:

$$\log\left[\frac{N_r}{N_s}\right] = 0.206 \log\left[\left(\frac{B}{D}\right)\left(\frac{x_{f,\mathrm{HK}}}{x_{f,\mathrm{LK}}}\right)\left(\frac{x_{b,\mathrm{LK}}}{x_{d,\mathrm{HK}}}\right)^2\right]$$

Where,

N_r = number of stages above the feed, including any partial condenser,

N_s = number of stages below the feed, including the reboiler,

B = molar flow bottom product,

D = molar flow top product,

$X_{f,\,HK}$ = concentration of the heavy key in the feed,

$X_{f,\,LK}$ = concentration of the light key in the feed,

$X_{d,\,HK}$ = concentration of the heavy key in the top product,

$X_{b,\,LK}$ = concentration of the light key if in the bottom product.

2.2 DESIGNING PARAMETERS FOR DEPROPANIZER COLUMN

The following data needs to be calculated to design a depropanizer column:

- Distillates flow rate
- Bottoms flow rate
- Composition of the various components obtained from distillates and bottoms.
- Minimum number of stages

- Minimum reflux
- Number of Theoretical stages
- Actual Number of stages
- Stripping and Rectifying feed tray location
- Optimum Reflux
- Column Internals
- Reboiler and Condenser Loads

However, the following specifications are needed to be given in order to design a distillation column:

Separation Specification:

$F = B + D$ ------------------------------------(A)

$Fx_F = Bx_B + Dx_D$ --------------------------------------(B)

At a given feed flow rate and feed composition, there are two equations [Eqs. (A) & (B)] and four unknowns: B, D, x_B and x_D. There fore, only two variables can be specified for the separation. Further, at least one of the two specified variables must be a composition.

Composition specification

If one product flow is specified, the concentration of one component either in the distillate or in the bottom (but not both) can be specified. If neither a recovery nor a product rate is specified, the concentration of one component in the distillate and one component in the bottom can be specified.

The above applies to both binary and multicomponent distillation. In multicomponent distillation, once the above are specified, other components will distribute according to the equilibrium relationship. Frequently, a product specification sets the maximum concentration of impurities that can be tolerated in the product. The one impurity which is dependent on the column separation and is most difficult to achieve sets the composition specification in the column.

Physical Property Specification

A product composition can often be specified in terms of a physical property that is a direct function of composition. For instance, the vapor pressure of a bottom product is often a good measure of the concentration of lights in the bottoms, and may be specified instead. Other physical properties include the Reid vapor pressure (RVP), viscosity, refractive index, freezing point, molecular weight & others. A physical property specification is often preferred either when it is easy to monitor (e.g., refractive index), or when it provides a good functional specification of product purity.

Heat Duty Specifications

A composition or product rate specification may be substituted by a heat duty or internal flow (e.g., reflux) specification. This is done either to improve convergence in a computer simulation (especially if compositions are in the part per million levels), or in a revamp when the column or its exchangers are at a capacity limits. The mass, component, and energy balance equations translate this specification into a composition or product rate specification.

Side product

For each side product, one additional specification is required. This specification is either a product rate (e.g., the side product rate) or a product composition.

Heat addition or removal

For each point of heat addition or removal, an additional specification is required. This specification is usually a heat duty or an internal product flow.

2.3 IMPORTANT CONSIDERATIONS IN OPTIMIZATION OF DISTILLATION COLUMN

2.3.1 Pressure Considerations

As the pressure of a column is raised:

- Separation becomes more difficult since the relative volatility decreases - more plates and reflux are required to achieve the separation.
- The latent heat of vaporization decreases, reducing the duties of the reboiler and condenser.
- The vapor density increases, resulting in a smaller column diameter.
- The reboiler temperature increases. This is usually limited by the decomposition temperature of the material being vaporized.
- Condenser temperature increases.

As the pressure is lowered, these effects are reversed. A lower pressure limit is usually encountered by a desire to avoid vacuum operation and / or refrigeration in the condenser. For an initial design, it is adequate to set the distillation pressure above ambient and less as allowed by cooling water or air cooling in the condenser. An initial starting value might be selected so that the bubble point of the overhead product is 10°C above the summer cooling water temperature or to atmospheric pressure if vacuum operation is suggested.

2.3.2 Reflux Ratio Considerations

We have several trade-offs in the selection of a reflux ratio. As the reflux ratio is increased:

- The purity of the product is increased.
- The capital costs decrease since the number of trays is decreased.
- The energy costs increase as more reboiling and condensing are required.

If the optimal reflux ratio is less than 1.1 times the minimum reflux, select 1.1 times the minimum reflux since a small error in design data or operating conditions might lead to a column that does not work.

2.3.3 Feed Considerations

The feed consideration is more of an afterthought rather than a critical design parameter. The question is whether the feed is at the bubble point, sub cooled, partial vapor, or all vapor. In general, a sub cooled feed:

- Decreases the number of tray in the rectifying section but increases the trays in the stripping section.
- Increases the size of the reboiler but decreases the size of the condenser.

Partially vaporized feed reverses this.

2.4 PROCESS DESIGN AND ADOPTED PROCEDURE

Process design proceeds in the following steps:

1. Specify separation. If product composition or product flow requirements are not defined, determine them by material and energy balance optimization.
2. Set column pressure.
3. Determine the minimum reflux and minimum number of stages.
4. Find the optimum feed stage.
5. Select three ratios of actual to minimum reflux. For each, calculate the number of stages and size the column and auxiliaries. Determine which is the most economical. This optimization procedure can be bypassed by selecting a single ratio of reflux to minimum reflux.
6. The calculations so far can be shortcut or rigorous.
7. Re examine steps 3 and 4, refining earlier estimates as necessary. If the refinements are large, steps 5 and 6 may need repeating.
8. Analyze the design graphically to ensure optimum design and absence of pinched regions.

2.4.1 Approximate Column Sizing

An approximate estimate of the overall column size can be made once the number of real stages required for the separation is known. This is often needed to make a rough estimate of the capital cost for project evaluation.

2.4.1.1 Plate Spacing

The overall height of the column will depend on the plate spacing. Plate spacing's from 0.15 m (6 in.) to 1 m (36 in.) are normally used. The spacing chosen will depend on the column diameter and operating conditions. Close spacing is used with small-diameter columns, and where head room is restricted; as it will be when a column is installed in a building. For columns above 1 m diameter, plate spacing's of 0.3 to 0.6 m will normally be used, and 0.5 m (18 in.) can be taken as an initial estimate. This would be revised, as necessary, when the detailed plate design is made. A larger spacing will be needed between certain plates to accommodate feed and side streams arrangements, and for manways.

2.4.1.2 Column Diameter

The principal factor that determines the column diameter is the vapor flow-rate. The vapor velocity must be below that which would cause excessive liquid entrainment or a high-pressure drop. The equation given below, which is based on the well-known Souders and Brown equation, Lowenstein (1961), can be used to estimate the maximum allowable superficial vapor velocity, and hence the column area and diameter:

$$\hat{u}_v = (-0.171 l_t^2 + 0.27 l_t - 0.047)\left[\frac{(\rho_L - \rho_v)}{\rho_v}\right]^{1/2}$$

Where,

u_v = maximum allowable vapor velocity, based on the gross (total) column cross-sectional area, m/s,

l_t = plate spacing, m, (range 0.5-1.5).

The column diameter, D_c, can then be calculated:

$$D_c = \sqrt{\frac{4\hat{V}_w}{\pi \rho_v \hat{u}_v}}$$

Where, V_w is the maximum vapor rate, kg/s.

This approximate estimate of the diameter would be revised when the detailed plate design is undertaken.

2.4.2 Tray Design

Once the process design is completed, the equipment design begins. This phase of the design translates the process requirements (i.e., the vapor and liquid loads in each section of the column) into actual hardware. The hardware design proceeds in two phases: primary (basic) and secondary (detailed layout). The primary phase sets column diameter, type of tray, and split of tray area into bubbling and downcomer areas.

This phase also provides a preliminary (and usually close) estimate of tray spacing, number of passes, and other features of tray and downcomer layout such as weir height, fractional hole area, hole diameter, and clearance under the downcomer. These estimates are later firmed up in the secondary phase. Functionally, the primary phase sets the major equipment requirements, while the secondary phase engineers the finer details. The primary phase has a major impact on column costs, but a relatively small influence on achieving trouble-free operation. These roles are reversed in the secondary phase: it has a relatively small impact on column costs, but a major impact on achieving trouble-free operation.

2.4.2.1 Common Types of Tray

- Bubble cap tray
- Sieve tray
- Dual flow tray
- Valve tray

Bubble Cap Tray:

The bubble cap tray was the workhorse of distillation before the 1960s. Presently, bubble-cap trays are specified only for special applications, while sieve and valve trays are the most popular types. The bubble-cap tray is a flat perforated plate with risers (chimney like pipes) around the holes, and caps in the form of inverted cups over the risers. The caps are usually (but not always) equipped with slots or holes through which the vapor comes out. Liquid and froth are trapped on the tray to a depth at least equal to the weir height or riser height, giving the bubble-cap tray a unique ability to operate at low vapor and liquid rates.

They offer the distinct advantage of being able to handle very wide ranges of liquid and gas flow rates satisfactorily. They have now been abandoned for new installations because of their cost, which is roughly, double that for sieve, counter flow, and valve tray.

Sieve Tray:

The sieve tray is a flat perforated plate. Vapor issues from the holes to give a multi-orifice effect. The gas dispersed by the perforations, expands the liquid into a turbulent froth, characterized by a very large interfacial surface for mass transfer. The vapor velocity keeps the liquid from flowing down through the holes (weeping). At low velocities, liquid weeps through the holes, bypassing some of the tray and reducing efficiency, giving sieve trays relatively poor turndown. Sieve trays are simple and easy to fabricate, and are therefore relatively inexpensive.

Dual Flow Tray:

A dual-flow tray is a sieve tray with no downcomers. This tray operates with liquid continuously weeping through the holes, hence its low efficiency. Tray froth height diminishes rapidly when vapor velocity is reduced, causing further efficiency deterioration upon turndown.

Turndown of a dual-flow tray is even poorer than that of a sieve tray with downcomers. Large-diameter (>8 ft) dual flow trays are known to sometimes experience instability. Dual flow trays are prone to channeling, and are therefore sensitive to out of levelness and to liquid distribution. Due to the absence of downcomers, dual flow trays give more tray area, and therefore have a greater capacity than any of the common tray types. The absence of downcomers, and the larger open areas, renders dual flow trays the most suitable to handle highly fouling services, slurries, and corrosive services. Dual-flow trays are also the least expensive to make, and easiest to instal1 and maintain.

Valve Tray:

These are the sieve trays with large (roughly 35 to 40 mm diameter) variable openings for gas flow. The perforations are covered with movable caps which rise as the flow rate of gas increases. Valves can be round or rectangular, with or without caging. The upper limit of opening is controlled by a caging structure or by restrictive legs at the bottom of the valve unit. As vapor rate falls, the disk openings are reduced, or they may settle intermittently over the holes, this stops the liquid from weeping and gives the valve tray its main advantage-good operation at low flow rates, and therefore, a high turndown.

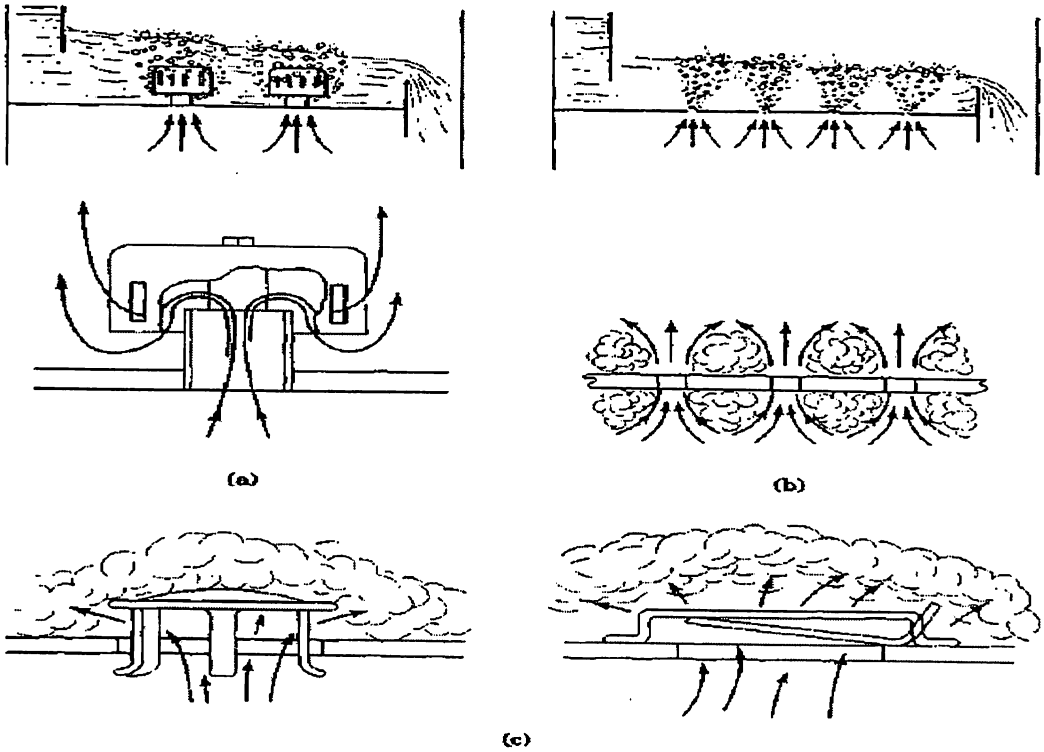


Figure 3. Flow Through Tray Vapor Passages (a) Bubble Cap (b) Sieve (c) Valve

2.4.2.2 Comparison of Common Types of Trays

The principal factors to consider when comparing the performance of bubble-cap, sieve and valve plates are: cost, capacity, operating range, efficiency and pressure drop.

Cost:

Bubble-cap plates are appreciably more expensive than sieve or valve plates. The relative cost will depend on the material of construction used; for mild steel the ratios, bubble-cap: valve: sieve, are approximately 3.0: 1.5: 1.0.

Capacity:

There is little difference in the capacity rating of the three types (the diameter of the column required for a given flow-rate); the ranking is sieve, valve and bubble-cap.

Operating range:

This is the most significant factor. By operating range is meant the range of vapor and liquid rates over which the plate will operate satisfactorily (the stable operating range). Some flexibility will always be required in an operating plant to allow for changes in production rate, and to cover start-up and shut-down conditions. The ratio of the highest to the lowest flow rates is often referred to as the "turn-down" ratio. Bubble-cap plates have a positive liquid seal and can therefore operate efficiently at very low vapor rates. Sieve plates rely on the flow of vapor through the holes to hold the liquid on the plate, and cannot operate at very low vapor rates. But, with good design, sieve plates can be designed to give a satisfactory operating range; typically, from 50 per cent to 120 per cent of design capacity. Valve plates are intended to give greater flexibility than sieve plates at a lower cost than bubble-caps.

Efficiency:

The Murphree efficiency of the three types of plate will be virtually the same when operating over their design flow range and no real distinction can be made between them

Pressure drop:

The pressure drop over the plates can be an important design consideration, particularly for vacuum columns. The plate pressure drop will depend on the detailed design of the plate but, in general, sieve plates give the lowest pressure drop, followed by valves, with bubble-caps giving the highest.

Summary:

Sieve plates are the cheapest and are satisfactory for most applications. Valve plates should be considered if the specified turn-down ratio cannot be met with sieve plates. Bubble-caps should only be used where very low vapor (gas) rates have to be handled and a positive liquid seal is essential at all flow-rates.

Table 1. Comparison of Common Type of Trays

Type	Sieve Tray	Valve Tray	Bubble Cap Tray	Dual Flow Trays
Maintenance	Low	Low to moderate	Relatively high	Low
Fouling tendency	Low	Low to moderate	High, tends to collect solids	Extremely low. Suitable where fouling is extensive and for slurry handling.
Effect of corrosion	Low	Low to moderate	High	Very low
Availability of design information	Well known	Proprietary but information readily available	Well known	Some information available
Capacity	High	High to very high	Moderately high	Very high
Efficiency	High	High	Moderately high	Lower than other types
Turndown	About 2:1, not suitable for operation under variable loads	About 4-5:1, some special designs achieve 10:1 or more	Excellent, better than valve trays. Good at extremely low liquid rates	Low, even lower than sieve trays. Unsuitable for variable load operations
Entrainment	Moderate	Moderate	High, about 3 times higher tan sieve trays	Low to moderate
Pressure drop	Moderate	Moderate, early designs some what higher, recent designs same as sieve trays	High	Low to moderate
Cost	Low	About 20% higher than sieve trays	High, about 2 to 3 times the cost of sieve trays	Low

Main applications	Most columns when turndown is not critical	Most columns, services where turndown is important	Extremely low flow conditions, where leakages must be minimized	Capacity revamps where efficiency and turndown can be sacrificed, Highly fouling and corrosive services
Share of market	25%	70%	5%	No information

2.4.3 Plate Hydraulic Design

The basic requirements of a plate contacting stage are that it should:

- Provide good vapor-liquid contact.
- Provide sufficient liquid hold-up for good mass transfer (high efficiency).
- Have sufficient area and spacing to keep the entrainment and pressure drop within acceptable limits.
- Have sufficient downcomer area for the liquid to flow freely from plate to plate.

The plate design methods use semi-empirical correlations derived from fundamental research work combined with practical experience obtained from the operation of commercial columns. Proven layouts are used, and the plate dimensions are kept within the range of values known to give satisfactory performance.

2.4.3.1 Operating Range

Satisfactory operation will only be achieved over a limited range of vapor and liquid flow rates. A typical performance diagram for a sieve plate is shown in Figure given below:

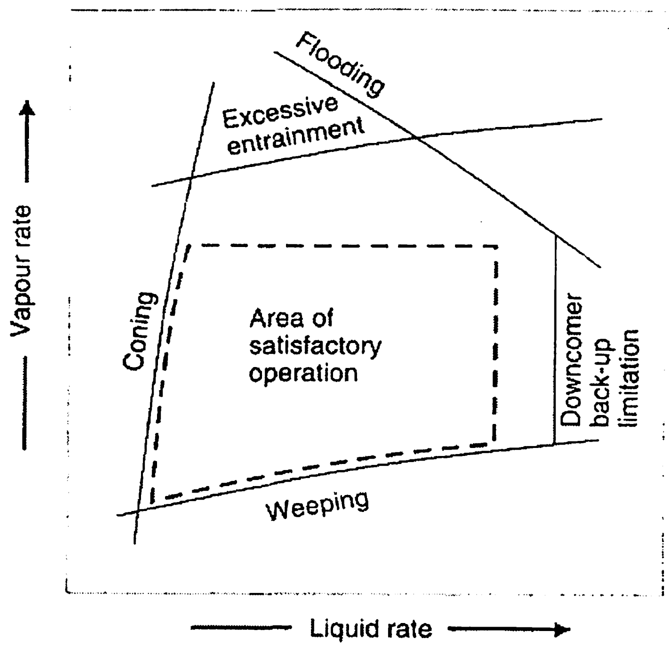


Figure 4. Sieve Tray Performance Diagram

The upper limit to vapor flow is set by the condition of flooding. At flooding there is a sharp drop in plate efficiency and increase in pressure drop. Flooding is caused by either the excessive carryover of liquid to the next plate by entrainment, or by liquid backing-up in the downcomers. The lower limit of the vapor flow is set by the condition of weeping. Weeping occurs when the vapor flow is insufficient to maintain a level of liquid on the plate. "Coning" occurs at low liquid rates, and is the term given to the condition where the vapor pushes the liquid back from the holes and jets upward, with poor liquid contact. In the following sections gas can be taken as synonymous with vapor when applying the method to the design of plates for absorption columns.

2.4.3.2 Plate-Design Procedure

A trial-and-error approach is necessary in plate design: starting with a rough plate layout, checking key performance factors and revising the design, as necessary, until a satisfactory design is achieved, a typical design procedure is set out below and discussed in the following sections. The normal range of each design variable is given in the discussion, together with recommended values which can be used to start the design.

1. Calculate the maximum and minimum vapor and liquid flow-rates, for the turn down ratio required.
2. Collect, or estimate, the system physical properties.
3. Select trial plate spacing.
4. Estimate the column diameter, based on flooding considerations.
5. Decide the liquid flow arrangement.
6. Make a trial plate layout: downcomer area, active area, hole area, hole size, weir height.
7. Check the weeping rate, if unsatisfactory return to step 6.
8. Check the plate pressure drop, if too high return to step 6.
9. Check downcomer back-up, if too high return to step 6 or 3.
10. Decide plate layout details: calming zones, unperforated areas. Check hole pitch, if unsatisfactory return to step 6.
11. Recalculate the percentage flooding based on chosen column diameter.
12. Check entrainment, if too high return to step 4.
13. Optimize design: repeat steps 3 to 12 to find smallest diameter and plate spacing acceptable (lowest cost).
14. Finalize design: draw up the plate specification and sketch the layout.

2.4.3.3 Plate Areas

The following areas terms are used in the plate design procedure:

A_c = total column cross-sectional area,

A_d = cross-sectional area of downcomer,

A_n = net area available for vapor-liquid disengagement, normally equal to $A_c - A_d$ for a single pass plate,

A_a = active, or bubbling, area, equal to $A_c — 2A_d$ for single-pass plates,

A_h = hole area, the total area of all the active holes,

A_p = perforated area (including blanked areas),

A_{ap} = the clearance area under the downcomer apron.

2.4.3.4 Diameter

The flooding condition fixes the upper limit of vapor velocity. A high vapor velocity is needed for high plate efficiencies, and the velocity will normally be between 70 to 90 per cent of that which would cause flooding. For design, a value of 80 to 85 per cent of the flooding velocity should be used.

The flooding velocity can be estimated from the correlation given by Fair (1961):

$$u_f = K_1 \sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

Where,

u_f = flooding vapor velocity, m/s, based on the net column cross-sectional area An

K_1 = a constant obtained from Figure given below:

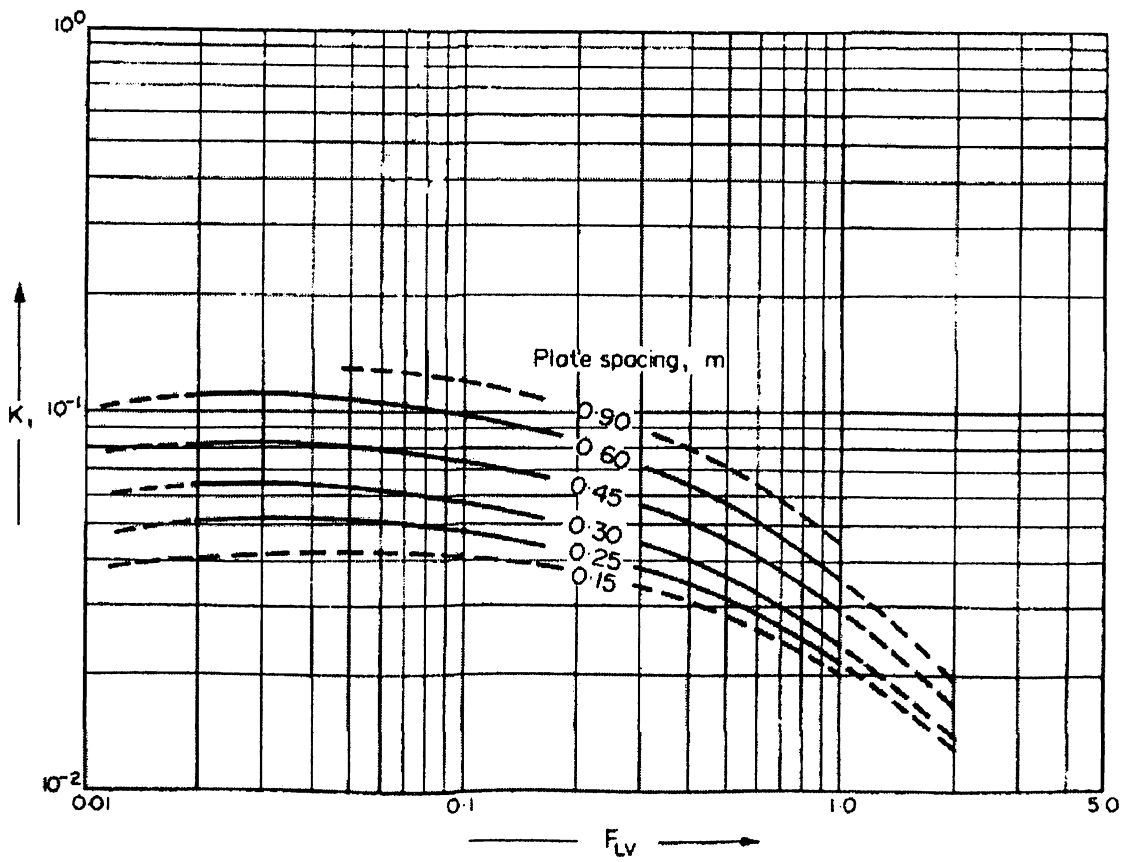


Figure 5: Flooding Velocity for Sieve Trays

The liquid-vapor flow factor F_{LV} in Figure 5 is given by:

$$F_{LV} = \frac{L_W}{V_W}\sqrt{\frac{\rho_v}{\rho_L}}$$

Where,

L_w = liquid mass flow-rate, kg/s,

V_w = vapor mass flow-rate, kg/s.

The following restrictions apply to the use of Figure 5:

1. Hole size less than 6.5 mm. Entrainment may be greater with larger hole sizes.
2. Weir height less than 15 per cent of the plate spacing.
3. Non-foaming systems.
4. Hole: active area ratio greater than 0.10; for other ratios apply the following corrections:

hole: active area	**multiply K_1 by**
0.10	1.0
0.08	0.9
0.06	0.8

5. Liquid surface tension 0.02 N/m, for other surface tensions ρ multiply the value of K1 by $[\rho/0.02]^{0.2}$.

To calculate the column diameter an estimate of the net area A_n is required. As a first trial take the downcomer area as 12 per cent of the total, and assume that the hole-active area is 10 per cent. Where the vapor and liquid flow-rates, or physical properties, vary significantly throughout the column a plate design should be made for several points up the column. For distillation it will usually be sufficient to design for the conditions above and below the feed points, Changes in the vapor flow-rate will normally be accommodated by adjusting the hole area; often by blanking off some rows of holes. Different column diameters would only be used where there is a considerable change in flow-rate. Changes in liquid rate can be allowed for by adjusting the liquid downcomer areas.

2.4.3.5 Liquid-Flow Arrangement

The choice of plate type (reverse, single pass or multiple pass) will depend on the liquid flow-rate and column diameter. An initial selection can be made using Figure-6, which has been adapted from a similar figure given by Huang and Hodson (1958). The selection of plate can be done by using the figure five.

2.4.3.6 Entrainment

Entrainment can be estimated from the correlation given by Fair (1961), Figure 7, which gives the fractional entrainment ψ (kg/kg gross liquid flow) as a function of the liquid-vapor factor F_{LV} with the percentage approach to flooding as a parameter. The percentage flooding is given by:

$$\text{percentage flooding} = \frac{u_n \text{ actual velocity (based on net area)}}{u_f \text{ (from equation 11.81)}}$$

As a rough guide the upper limit of ψ can be taken as 0.1; below this figure the effect on efficiency will be small. The optimum design value may be above this figure.

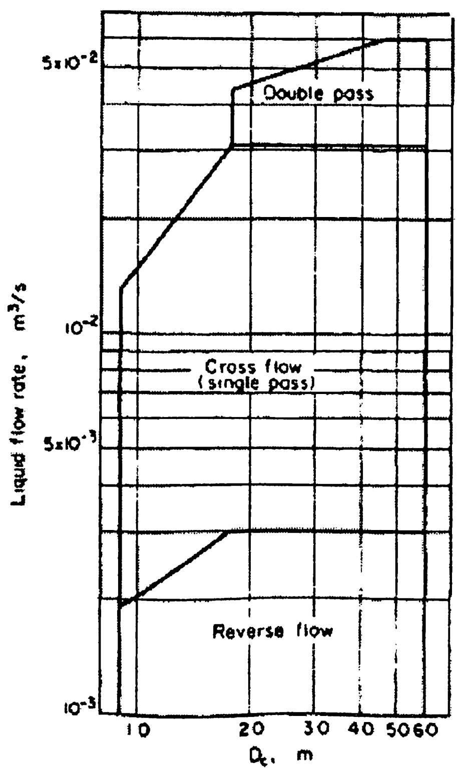


Figure 6. Selection of Liquid Flow Arrangement

2.4.3.7 Weep Point

The lower limit of the operating range occurs when liquid leakage through the plate holes becomes excessive. This is known as the weep point. The vapor velocity at the weep point is the minimum value for stable operation. The hole area must be chosen so that at the lowest operating rate the vapor flow velocity is still well above the weep point. Several correlations have been proposed for predicting the vapor velocity at the weep point; see Chase (1967). That given by Eduljee (1959) is one of the simplest to use, and has been shown to be reliable. The minimum design vapor velocity is given by:

$$\check{u}_h = \frac{[K_2 - 0.90(25.4 - d_h)]}{(\rho_v)^{1/2}}$$

Where,

u_h = minimum vapor velocity through the holes(based on the hole area), m/s,

d_h = hole diameter, mm,

K_2 = a constant, dependent on the depth of clear liquid on the plate, obtained from Figure 8.

The clear liquid depth is equal to the height of the weir h_w plus the depth of the crest of liquid over the weir h_{ow}.

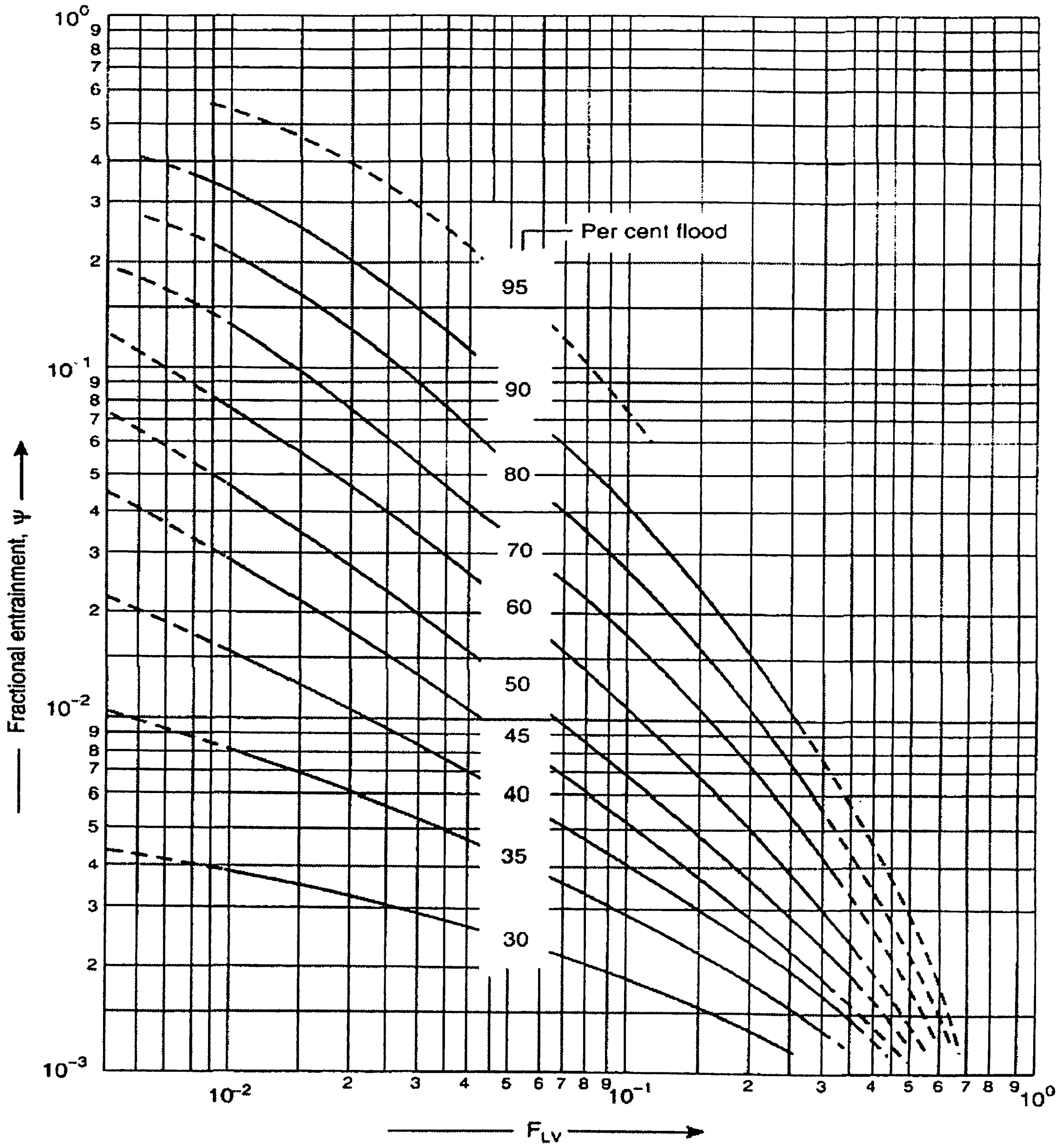


Figure 7. Entrainment Correlation for Sieve Tray

2.4.3.8 Weir Liquid Crest

The height of the liquid crest over the weir can be estimated using the Francis weir formula. For a segmental downcomer this can be written as:

$$h_{ow} = 750\left[\frac{L_w}{\rho_L l_w}\right]^{2/3}$$

Where,

l_w =weir length, m,

h_{ow} = weir crest, mm liquid,

L_w = liquid flow-rate, kg/s.

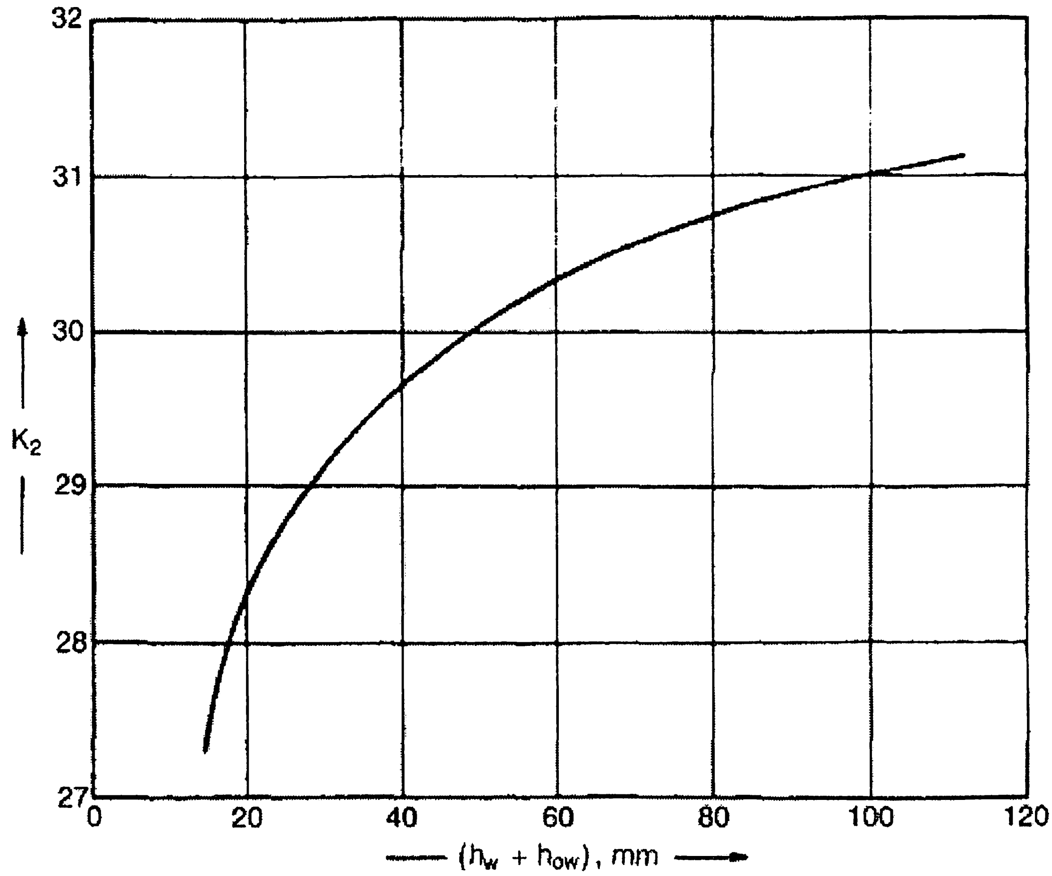


Figure 8. Weep Point Correlation

With segmental downcomers the column wall constricts the liquid flow, and the weir crest will be higher than that predicted by the Francis formula for flow over an open weir. To ensure an even flow of liquid along the weir, the crest should be at least 10 mm at the lowest liquid rate. Serrated weirs are sometimes used for very low liquid rates.

2.4.3.9 Weir Dimensions

Weir Height

The height of the weir determines the volume of liquid on the plate and is an important factor in determining the plate efficiency. A high weir will increase the plate efficiency but at the expense of a higher plate pressure drop. For columns operating above atmospheric pressure the weir heights will normally be between 40 mm to 90 mm (1.5 to 3.5 in.); 40 to 50 mm is recommended. For vacuum operation lower weir heights are used to reduce the pressure drop; 6 to 12 mm (0.25 to 0.5 in.) is recommended.

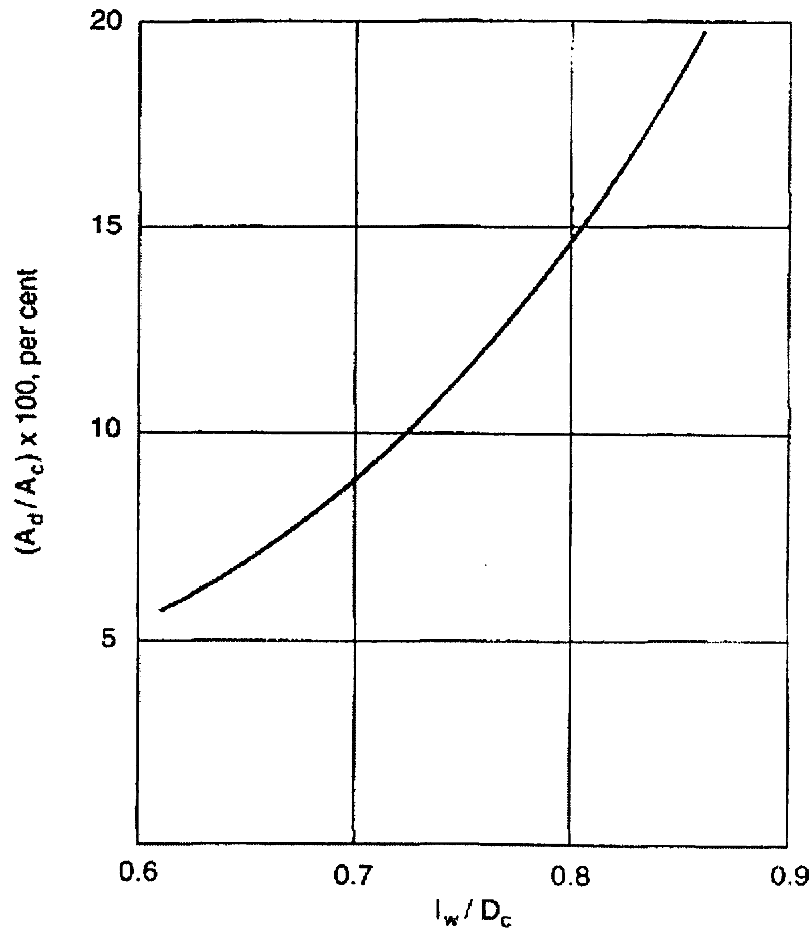


Figure 9. Relation between Downcomer Area & Weir Height

Inlet Weirs

Inlet weirs, or recessed pans, are sometimes used to improve the distribution of liquid across the plate; but are seldom needed with segmental downcomers.

Weir Length

With segmental downcomers the length of the weir fixes the area of the downcomer. The chord length will normally be between 0.6 to 0.85 of the column diameter. A good initial value to use is 0.77, equivalent to a downcomer area of 12 per cent. The relationship between weir length and downcomer area is given in Figure 8. For double-pass plates the width of the central downcomer is normally 200-250 mm (8-10 in.).

2.4.3.10 Perforated Area

The area available for perforation will be reduced by the obstruction caused by structural members (the support rings and beams), and by the use of calming zones. Calming zones are unperforated strips of plate at the inlet and outlet sides of the plate. The width of each zone is usually made the same; recommended values are: below 1.5 m diameter, 75 mm; above, 100 mm. The width of the support ring for sectional plates will normally be 50 to 75 mm; the support ring should not extend into the downcomer area. A strip of unperforated plate will be left round the edge of cartridge-type trays to stiffen the plate.

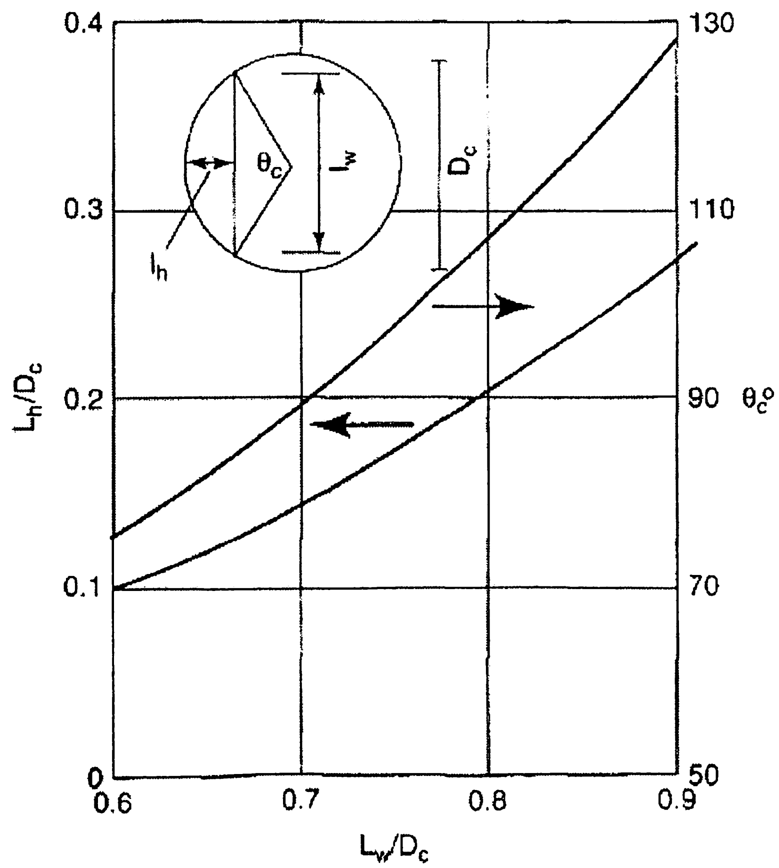


Figure 10. Relation between Angle Subtended by Chord, Chord Height and Chord Length

The unperforated area can be calculated from the plate geometry. The relationship between the weir chord length, chord height and the angle subtended by the chord is given in Figure 10.

2.4.3.11 Hole Size

The hole sizes used vary from 2.5 to 12 mm; 5 mm is the preferred size. Larger holes are occasionally used for fouling systems. The holes are drilled or punched. Punching is cheaper, but the minimum size of hole that can be punched will depend on the plate thickness, For carbon steel, hole sizes approximately equal to the plate thickness can be punched, but for stainless steel the minimum hole size that can be punched is about twice the plate thickness. Typical plate thicknesses used are: 5 mm (3/16 in.) for carbon steel, and 3 mm (12 gauges) for stainless steel. When punched plates are used they should be installed with the direction of punching upward. Punching forms a slight nozzle, and reversing the plate will increase the pressure drop.

2.4.3.12 Hole Pitch

The hole pitch (distance between the hole centers) l_p should not be less than 2.0 hole diameters, and the normal range will be 2.5 to 4.0 diameters. Within this range the pitch can be selected to give the number of active holes required for the total hole area specified. Square and equilateral triangular patterns are used; triangular is preferred. The total hole area as a fraction of the perforated area A_p is given by the following expression, for an equilateral triangular pitch:

$$\frac{A_h}{A_p} = 0.9\left[\frac{d_h}{l_p}\right]^2$$

This equation is plotted in Figure 11.

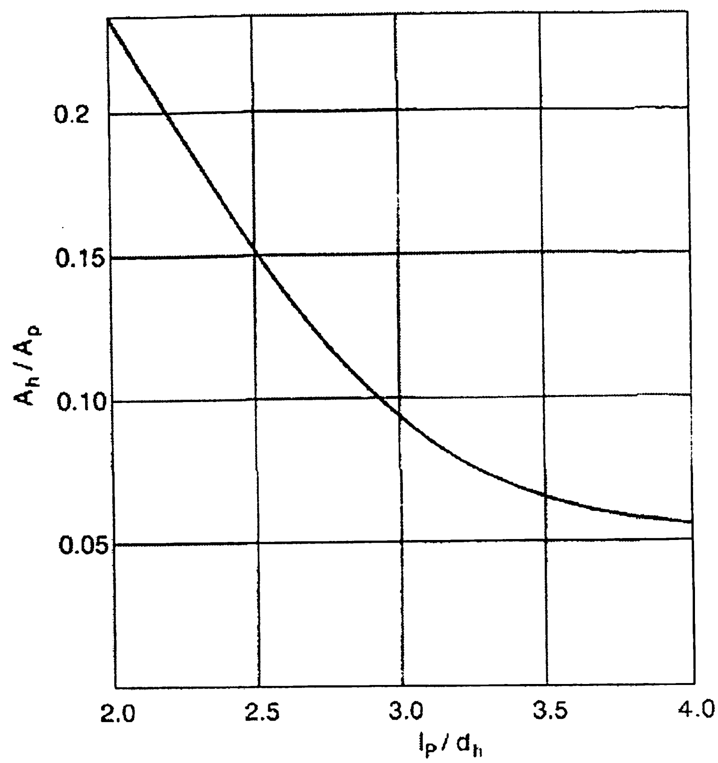


Figure 11. Relation between Hole Area & Pitch

2.4..3.13 Hydraulic Gradient

The hydraulic gradient is the difference in liquid level needed to drive the liquid flow across the plate. On sieve plates, unlike bubble-cap plates, the resistance to liquid flow will be small, and the hydraulic gradient is usually ignored in sieve-plate design. It can be significant in vacuum operation, as with the low weir heights used the hydraulic gradient can be a significant fraction of the total liquid depth. Methods for estimating the hydraulic gradient are given by Fair (1963).

2.4.3.14 Liquid Throw

The liquid throw is the horizontal distance travelled by the liquid stream flowing over the downcomer weir. It is only an important consideration in the design of multiple-pass plates. Bolles (1963) gives a method for estimating the liquid throw.

2.4.3.15 Rate Pressure Drop

The pressure drop over the plates is an important design consideration. There are two main sources of pressure loss: that due to vapor flow through the holes (an orifice loss), and that due to the static head of liquid on the plate.

A simple additive model is normally used to predict the total pressure drop. The total is taken as the sum of the pressure drop calculated for the flow of vapor through the dry plate (the dry plate drop h_d); the head of clear liquid on the plate ($h_w + h_{gw}$) and a term to account for other, minor, sources of pressure loss, the so-called residual loss h_r. The residual loss is the difference between the observed experimental pressure drop and the simple sum of the dry-plate drop and the clear-liquid height. It accounts for the two effects: the energy to form the vapor bubbles and the fact that on an operating plate the liquid head will not be clear liquid but a head of "aerated" liquid froth, and the froth density and height will be different from that of the clear liquid. It is convenient to express the pressure drops in terms of millimeters of liquid. In pressure units:

$$\Delta P_t = 9.81 \times 10^{-3} h_t \rho_L$$

Where,

ΔP_t = total plate pressure drop, Pa (N/m^2),

h_t = total plate pressure drop, mm liquid.

2.4.3.16 Dry Plate Drop

The pressure drop through the dry plate can be estimated using expressions derived for flow through orifices.

$$h_d = 51 \left[\frac{u_h}{C_0}\right]^2 \frac{\rho_v}{\rho_L}$$

Where the orifice coefficient C_o is a function of the plate thickness, hole diameter, and the hole to perforated area ratio. C_o can be obtained from Figure 12; which has been adapted from a similar figure by Liebson et al. (1957). U_f, is the velocity through the holes, m/s.

2.4.3.17 Residual Head

Methods have been proposed for estimating the residual head as a function of liquid surface tension, froth density and froth height. However, as this correction term is small the use of an elaborate method for its estimation is not justified, and the simple equation proposed by Hunt et al. (1955) can be used:

$$h_r = \frac{12.5 \times 10^3}{\rho_L}$$

Above equation is equivalent to taking the residual drop as a fixed value of 12.5 mm of water (0.5 in.).

2.4.3.18 Total Drop

The total plate drop is given by:

$$h_t = h_d + (h_w + h_{ow}) + h_r$$

If the hydraulic gradient is significant, half its value is added to the clear liquid height.

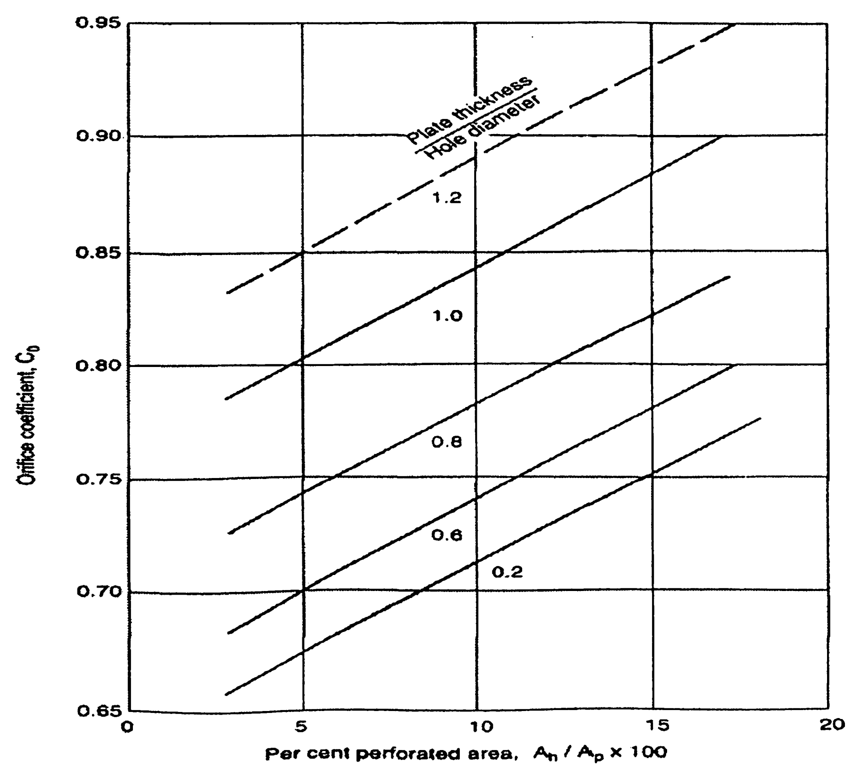


Figure 12. Discharge Coefficient, Sieve Plate

2.4.3.19 Downcomer Design [Back-Up]

The downcomer area and plate spacing must be such that the level of the liquid and froth in the downcomer is well below the top of the outlet weir on the plate above. If the level rises above the outlet weir the column will flood. The back-up of liquid in the downcomer is caused by the pressure drop over the plate (the downcomer in effect forms one leg of a U-tube) and the resistance to flow in the downcomer itself; see Figure 13. In terms of clear liquid the downcomer back-up is given by:

$$h_b = (h_w + h_{ow}) + h_t + h_{dc}$$

Where,

h_b = downcomer back-up, measured from plate surface, mm,

h_{dc} = head loss in the downcomer, mm.

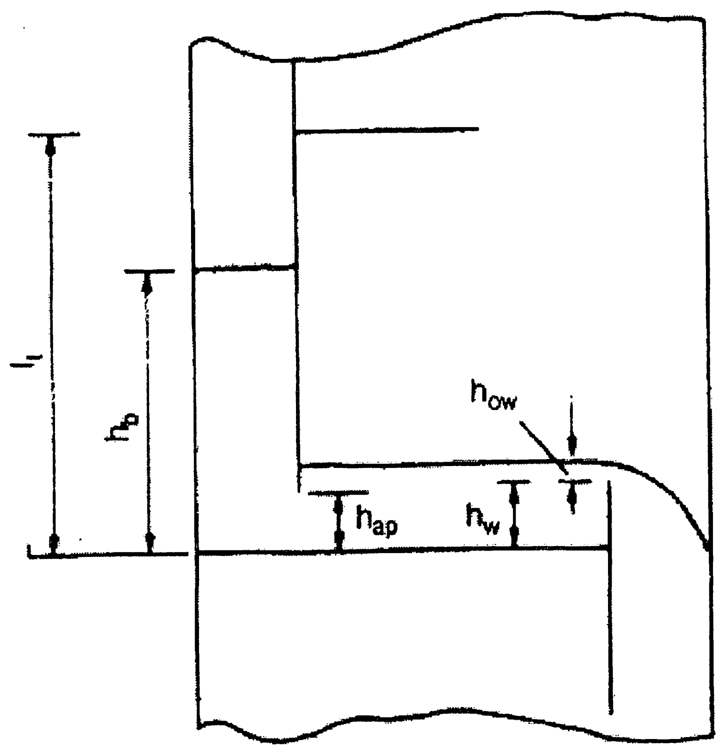


Figure 13. Downcomer Backup

The main resistance to flow will be caused by the constriction at the downcomer outlet, and the head loss in the downcomer can be estimated using the equation given by Cicalese etal. (1947):

$$h_{dc} = 166\left[\frac{L_{wd}}{\rho_L A_m}\right]^2$$

Where,

L_{wd} = liquid flow rate in downcomer, kg/s,

A_m = either the downcomer area Ad or the clearance area under the downcomer A_{sp} whichever is the smaller, m^2.

The clearance area under the downcomer is given by:

$$A_{ap} = h_{ap} l_w$$

Where, h_{ap} is height of the bottom edge of the apron above the plate. This height is normally set at 5 to 10 mm (0.25 to 0.5 in.) below the outlet weir height:

$$h_{ap} = h_w - (5 \text{ to } 10 \text{ mm})$$

2.4.3.20 Froth Height

To predict the height of "aerated" liquid on the plate, and the height of froth in the downcomer, some means of estimating the froth density is required. The density of the "aerated" liquid will normally be between 0.4 to 0.7 times that of the clear liquid. A number of correlations have been proposed for estimating froth density as a function of the vapor flow-rate and the liquid physical properties; however, none is particularly reliable, and for design purposes it is usually satisfactory to assume an average value of 0.5 of the liquid density. This value is also taken as the mean density of the fluid in the downcomer; which means that for safe design the clear liquid back-up should not exceed half the plate spacing l_t to avoid flooding.

Allowing for the weir height:

$$h_b \not> \frac{1}{2}(l_t + h_w)$$

This criterion is, if anything, oversafe, and where close plate spacing is desired a better estimate of the froth density in the downcomer should be made. The method proposed by Thomas and Shah (1964) is recommended.

2.4.3.21 Downcomer Residence Time

Sufficient residence time must be allowed in the downcomer for the entrained vapour to disengage from the liquid stream; to prevent heavily "aerated" liquid being carried under the downcomer. A time of at least 3 seconds is recommended.

The downcomer residence time is given by:

$$t_r = \frac{A_d h_{bc} \rho_L}{L_{wd}}$$

Where,

t_r = residence time, s,

h_{bc} = clear liquid back-up, m

2.5 USE OF ASPEN HYSYS FOR SIMULATING DEPROPANIZER COLUMN

2.5.1 Use of Process Simulators

Process simulation allows us to predict the behavior of a process by using basic engineering relationships, such as mass and energy balances, and phase and chemical equilibrium. Given reliable thermodynamic data, realistic operating conditions, and rigorous equipment models, we can simulate actual plant behavior. Process simulation enables you to run many cases, conduct "what if" analyses, and perform sensitivity studies and optimization runs. With simulation, we can design better plants and increase profitability in existing plants. Process simulation is useful throughout the entire lifecycle of a process, from research and development through process design to production.

2.5.2 Aspen Hysys as a Process Simulator

A process consists of chemical components being mixed, separated, heated, cooled, and converted by unit operations. These components are transferred from unit to unit through process streams.

We can translate a process into an Aspen Tech process simulation model by performing the following steps:

1. Define the process flow sheet:
   - Define the unit operations in the process.
   - Define the process streams that flow to and from the unit operations.
   - Select models from the Aspen Tech Model Library to describe each unit operation and place them on the process flow sheet.
   - Place labeled streams on the process flow sheet and connect them to the unit operation models.
2. Specify the chemical components in the process. We can take these components from the Aspen Tech databanks, or we can define them.
3. Specify thermodynamic models to represent the physical properties of the components and mixtures in the process. These models are built into Aspen Tech.
4. Specify the component flow rates and the thermodynamic conditions (for example, temperature and pressure) of feed streams.
5. Specify the operating conditions for the unit operation models.

With Aspen Tech we can interactively change specifications such as, flow sheet configuration; operating conditions; and feed compositions, to run new cases and analyze process alternatives. In addition to process simulation, Aspen Tech allows us to perform a wide range of other tasks such as estimating and regressing physical properties, generating custom graphical and tabular output results, fitting plant data to simulation models, optimizing our process, and interfacing results to spreadsheets.

Aspen Hysys version 3.1.3 is used here for simulation purpose. A complete understanding of software is done under the able guidance of mentor before using it. We have used Peng

Robinson fluid package to estimate the various fluid properties. This fluid package is termed as "basis-1" in our hysys file.

2.5.3 Peng-Robinson Equation of State

The Peng Robinson (1976) equation of state (EOS) is a modification of the RK equation to better represent VLE calculations. The densities for the liquid phase in the SRK did not accurately represent the experimental values due to a high universal critical compressibility factor of 0.3333. The PR is a modification of the RK equation of state which corresponds to a lower critical compressibility of about 0.307 thus representing the VLE of natural gas systems accurately. The PR equation is represented by:

Property Class Name Applicable Phase

$$P = \frac{RT}{V-b} - \frac{a}{V(V+b)+b(V-b)} \qquad (4.8)$$

where:

$$a = a_c\alpha$$

$$a_c = 0.45724\frac{R^2T_c^2}{P_c} \qquad (4.9)$$

$$b = 0.077480\frac{RT_c}{P_c}$$

The functional dependency of the "*a*" term is shown in the following relation.

$$\sqrt{\alpha} = 1+\kappa(1-T_r^{0.5}) \qquad (4.10)$$

$$\kappa = 0.37464 + 1.54226\omega - 0.26992\omega^2$$

Where: The functional dependency of the "a" term is shown in the following relation.

The accuracy of the PR and SRK equations of state are approximately the same. However, the PR EOS represents the density of the liquid phase more accurately due to the lower critical compressibility factor. These equations were originally developed for pure components. To apply the PR EOS to mixtures, mixing rules are required for the "a" and "b" terms in Equation (4.2).

Property Methods

A quick reference of calculation methods is shown in the table below for the PR EOS.

Calculation Method	Applicable Phase	Property Class Name
Z Factor	Vapour and Liquid	COTHPRZFactor Class
Molar Volume	Vapour and Liquid	COTHPRVolume Class
Enthalpy	Vapour and Liquid	COTHPREnthalpy Class
Entropy	Vapour and Liquid	COTHPREntropy Class
Isobaric heat capacity	Vapour and Liquid	COTHPRCp Class
Fugacity coefficient calculation	Vapour and Liquid	COTHPRLnFugacityCoeff Class
Fugacity calculation	Vapour and Liquid	COTHPRLnFugacity Class
Isochoric heat capacity	Vapour and Liquid	COTHPRCv Class
Mixing Rule 1	Vapour and Liquid	COTHPRab_1 Class
Mixing Rule 2	Vapour and Liquid	COTHPRab_2 Class
Mixing Rule 3	Vapour and Liquid	COTHPRab_3 Class
Mixing Rule 4	Vapour and Liquid	COTHPRab_4 Class
Mixing Rule 5	Vapour and Liquid	COTHPRab_5 Class
Mixing Rule 6	Vapour and Liquid	COTHPRab_6 Class

PR Z Factor

The compressibility factor, Z, is calculated as the root for the following equation:

$$Z^3 - (1 - B)Z^2 + Z(A - 3B^2 - 2B) - (AB - B^2 - B^3) = 0 \quad (4.11)$$

$$A = \frac{aP}{R^2T^2} \quad (4.12)$$

$$B = \frac{bP}{RT} \quad (4.13)$$

There are three roots for the above equation. It is considered that the smallest root is for the liquid phase and the largest root is for the vapour phase. The third root has no physical meaning.

PR Molar Volume

The following relation calculates the molar volume for a specific phase.

$$V = \frac{ZRT}{P} \quad (4.14)$$

Property Class Name and Applicable Phases

Property Class Name	Applicable Phase
COTHPRVolume Class	Vapour and Liquid

Notes

The compressibility factor, Z, is calculated using PR Z Factor. For consistency, the PR molar volume always calls the PR Z Factor for the calculation of Z.

PR Enthalpy

The following relation calculates the enthalpy.

$$H - H^{IG} = PV - RT - \left(a - \left(\frac{da}{dT}\right)T\right)\frac{1}{2\sqrt{2}b}\ln\frac{V + b(1+\sqrt{2})}{V - b(1-\sqrt{2})} \tag{4.15}$$

where: H^{IG} *is the ideal gas enthalpy calculated at temperature,* T

Property Class Name and Applicable Phases

Property Class Name	Applicable Phase
COTHPREnthalpy Class	Vapour and Liquid

Notes

The volume, V, is calculated using PR Molar Volume. For consistency, the PR Enthalpy always calls the PR Volume for the calculation of V.

PR Entropy

The following relation calculates the entropy.

$$S - S^{IG} = R\ln\left(\frac{V-b}{RT}\right) - \frac{1}{2b\sqrt{2}}\ln\left(\frac{V + b(1+\sqrt{2})}{V + b(1-\sqrt{2})}\right)\frac{da}{dT} \tag{4.16}$$

where: S^{IG} *is the ideal gas entropy calculated at temperature,* T

Property Class Name and Applicable Phases

Property Class Name	Applicable Phase
COTHPREntropy Class	Vapour and Liquid

Notes

The volume, V, is calculated using PR Molar Volume. For consistency, the PR Entropy always calls the PR Volume for the calculation of V.

PR Cp [Heat Capacity]

The following relation calculates the isobaric heat capacity.

$$C_p - C_p^{IG} = -T \int_{\infty}^{V} \left(\frac{\partial^2 P}{\partial T^2}\right)_V dV + R + \frac{T\left(\frac{\partial V}{\partial T}\right)_P^2}{\left(\frac{\partial V}{\partial P}\right)_T} \tag{4.17}$$

where: Cp^{IG} *is the ideal gas heat capacity calculated at temperature, T*

Property Class Name and Applicable Phases

Property Class Name	Applicable Phase
COTHPRCp Class	Vapour and Liquid

Notes

The volume, V, is calculated using PR Molar Volume. For consistency, the PR Entropy always calls the PR Volume for the calculation of V.

PR Fugacity Coefficient

The following relation calculates the fugacity coefficient.

$$\ln\phi_i = -\ln(V-b) + \frac{\bar{b}}{V-b} + \frac{a}{2\sqrt{2}b}\ln\left(\frac{V+b(1+\sqrt{2})}{V+b(1-\sqrt{2})}\right)\left(-1+\frac{\bar{a}}{a}+\frac{\bar{b}}{b}\right) \tag{4.18}$$

$$\bar{a} = \frac{\partial n^2 a}{\partial n} \tag{4.19}$$

$$\bar{b} = \frac{\partial n b}{\partial n} \tag{4.20}$$

Property Class Name and Applicable Phases

Property Class Name	Applicable Phase
COTHPRLnFugacityCoeff Class	Vapour and Liquid

Notes

The volume, V, is calculated using PR Molar Volume. For consistency, the PR Fugacity Coefficient always calls the PR Volume for the calculation of V. The parameters *a* and *b* are calculated from the Mixing Rules.

PR Fugacity

The following relation calculates the fugacity for a specific phase.

$$f_i = \phi_i y_i P \tag{4.21}$$

Property Class Name and Applicable Phases

Property Class Name	Applicable Phase
COTHPRLnFugacity Class	Vapour and Liquid

PR Cv [isochoric]

The following relation calculates the isochoric heat capacity.

$$C_v = C_p + \frac{T\left(\frac{\partial P}{\partial T}\right)_V^2}{\left(\frac{\partial P}{\partial V}\right)_T} \quad (4.22)$$

Property Class Name and Applicable Phases

Property Class Name	Applicable Phase
COTHPRCv Class	Vapour and Liquid

2.5.4 Steps for Simulating Depropanizer Column on Aspen Hysys

1. In file menu, go to new case (ctrl+ N) to enter into the simulation basis manager. Under the component list, highlight the "master component" list and add the component used in unit operation. Click on the "fluid packages" tab and select the appropriate property package to all required properties. Here, we have used Peng Robinson property package. Now, enter the simulation environment. The window shown below or figure 14 will be seen:

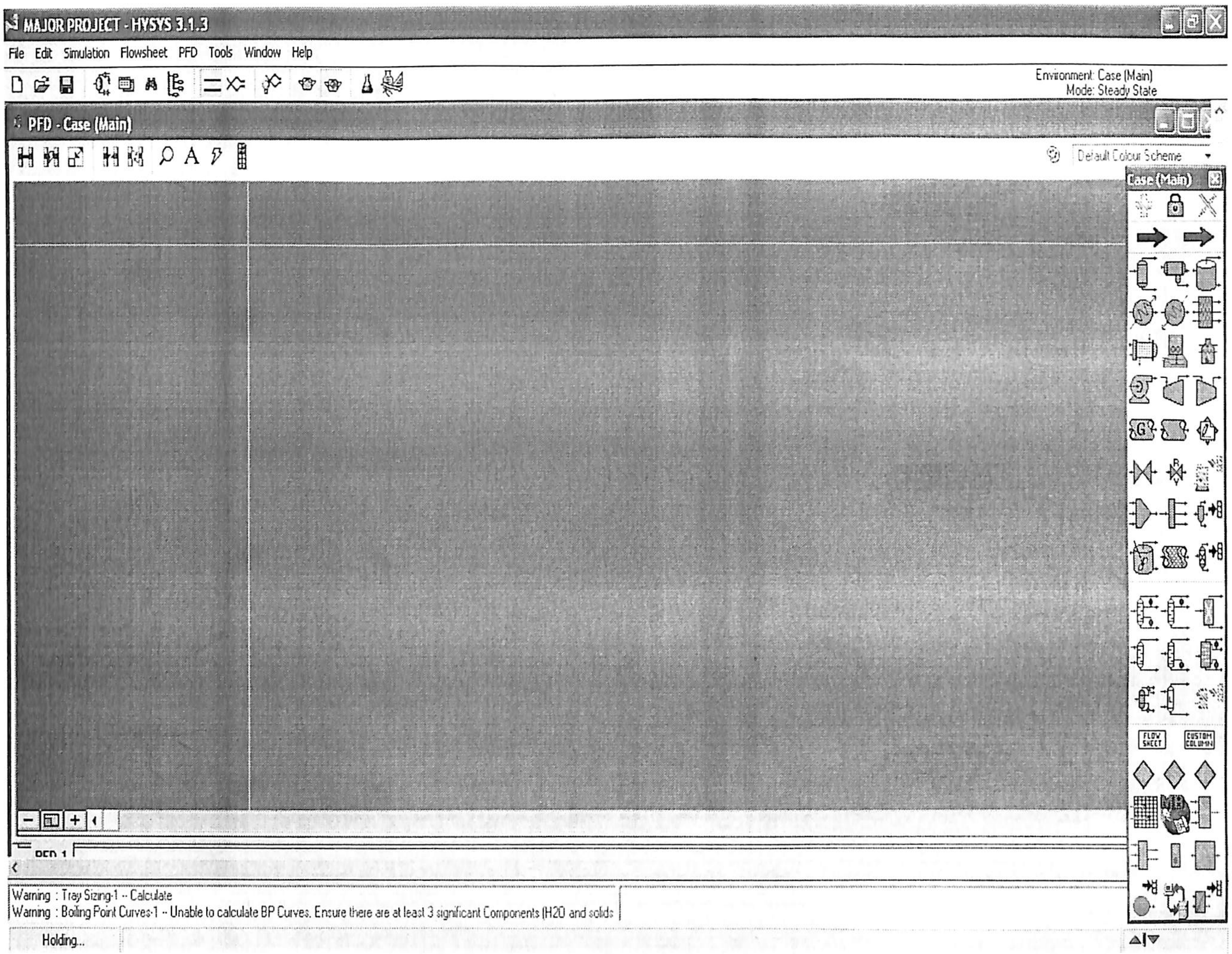


Figure 14

2. In the object palette, click on "shortcut distillation column". In PFD window, a short cut column in the name of T-100-2 will come. On double clicking T – 100-2, a pop up will appear as shown in figure 15. Under "design" tab, in "connections" bar, give the stream name, and fluid package of inlet, condenser duty, distillate, reboiler duty and bottoms. Now in "parameters" bar, put the values of the mole fraction of light key in bottoms and heavy key in distillate. Put the value of the condenser and reboiler pressure. Set external reflux ratio to any value. This will automatically calculate the minimum reflux ratio and hence, the value of external reflux should be made 1.2 – 1.5 times of minimum reflux.

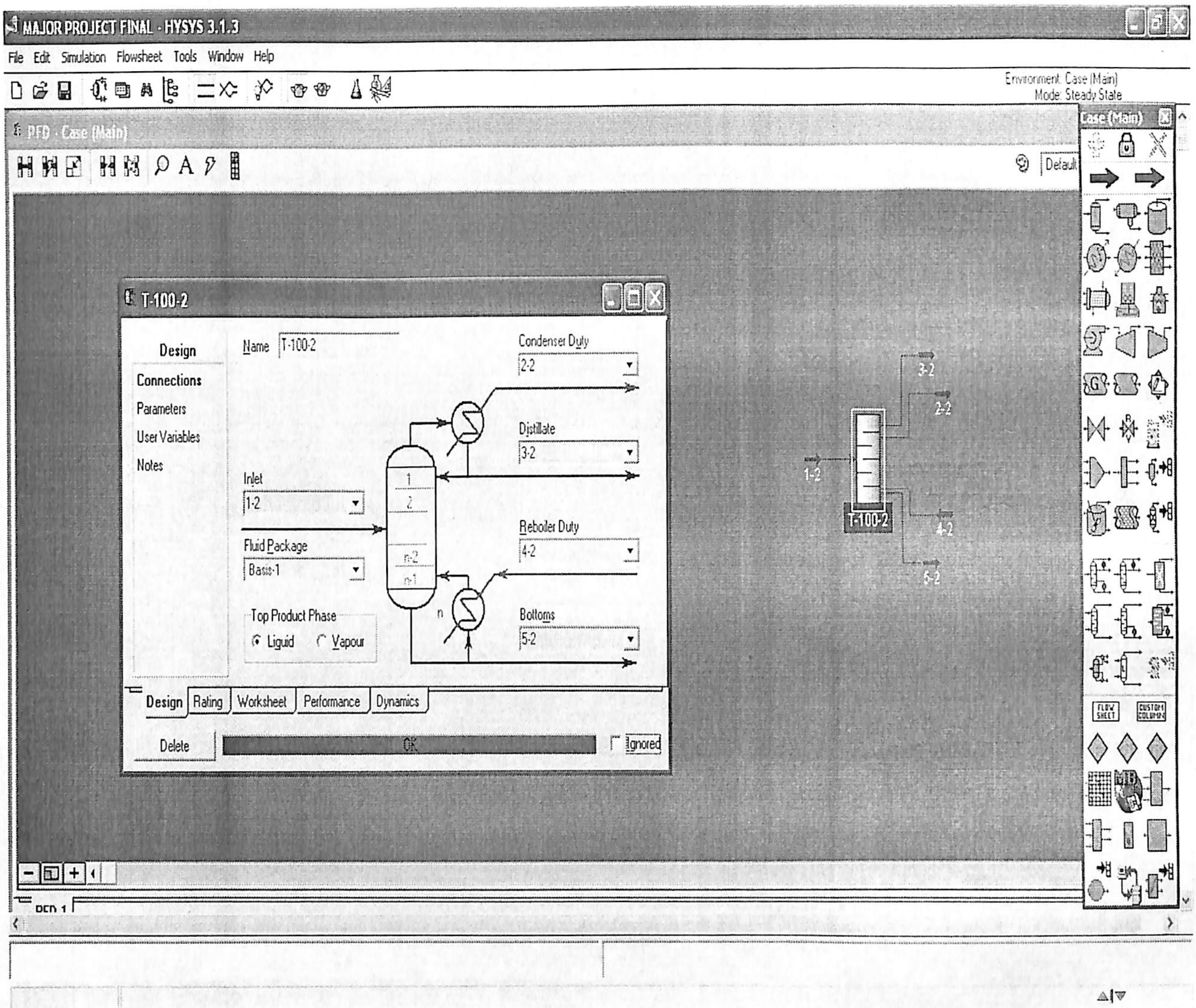


Figure 15

3. Double click on stream 1-2 and the temperature, pressure and molar flow (conditions) of the stream. Enter the compositions of each component in mole fraction. Click on the green button of "solver active" and this will aromatically calculate the following design parameters:
   - Conditions and compositions of distillate and bottoms.
   - Minimum number of trays, actual number of trays and feed tray location
   - Temperature & duty of condenser and reboiler

The shortcut column calculates the parameters at 100% efficiency. The efficiency of distillation column is ideally 70%, so we have to make the corrections accordingly.

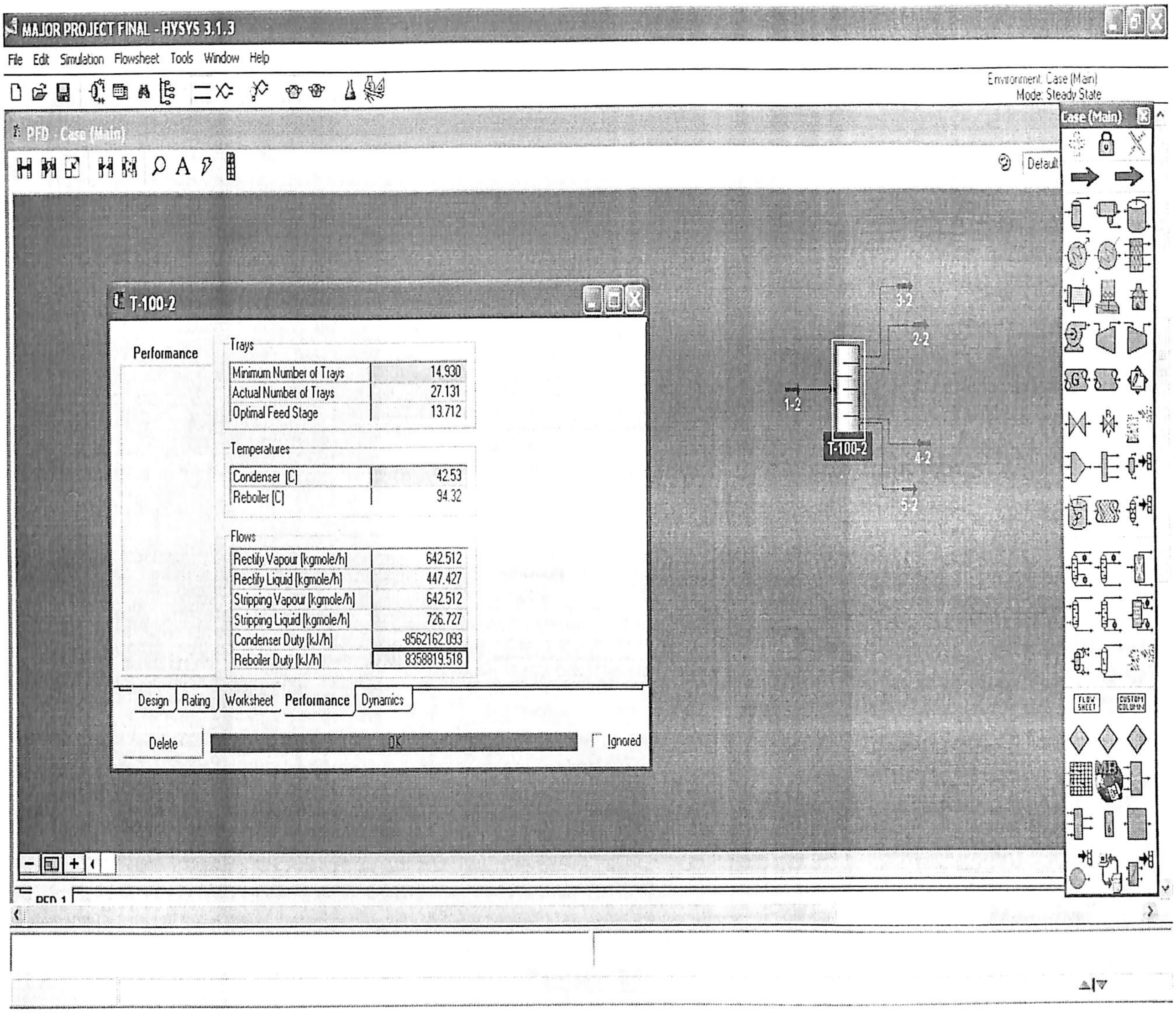


Figure 16

4. In the object palette, click on the distillation column. In PFD window, a distillation column in the name of T-100 will come. On double clicking T – 100, a pop up will appear as shown in figure 17. Under "design" tab, in "connections" bar, give the stream names of inlet stream, condenser energy stream, distillate liquid outlet, reboiler energy stream and bottom liquid outlet. Set the value of Delta P of condenser and reboiler and put the value of pressure of condenser & reboiler with the pressure difference of 80 KPa. Now, enter the value of number of stages.

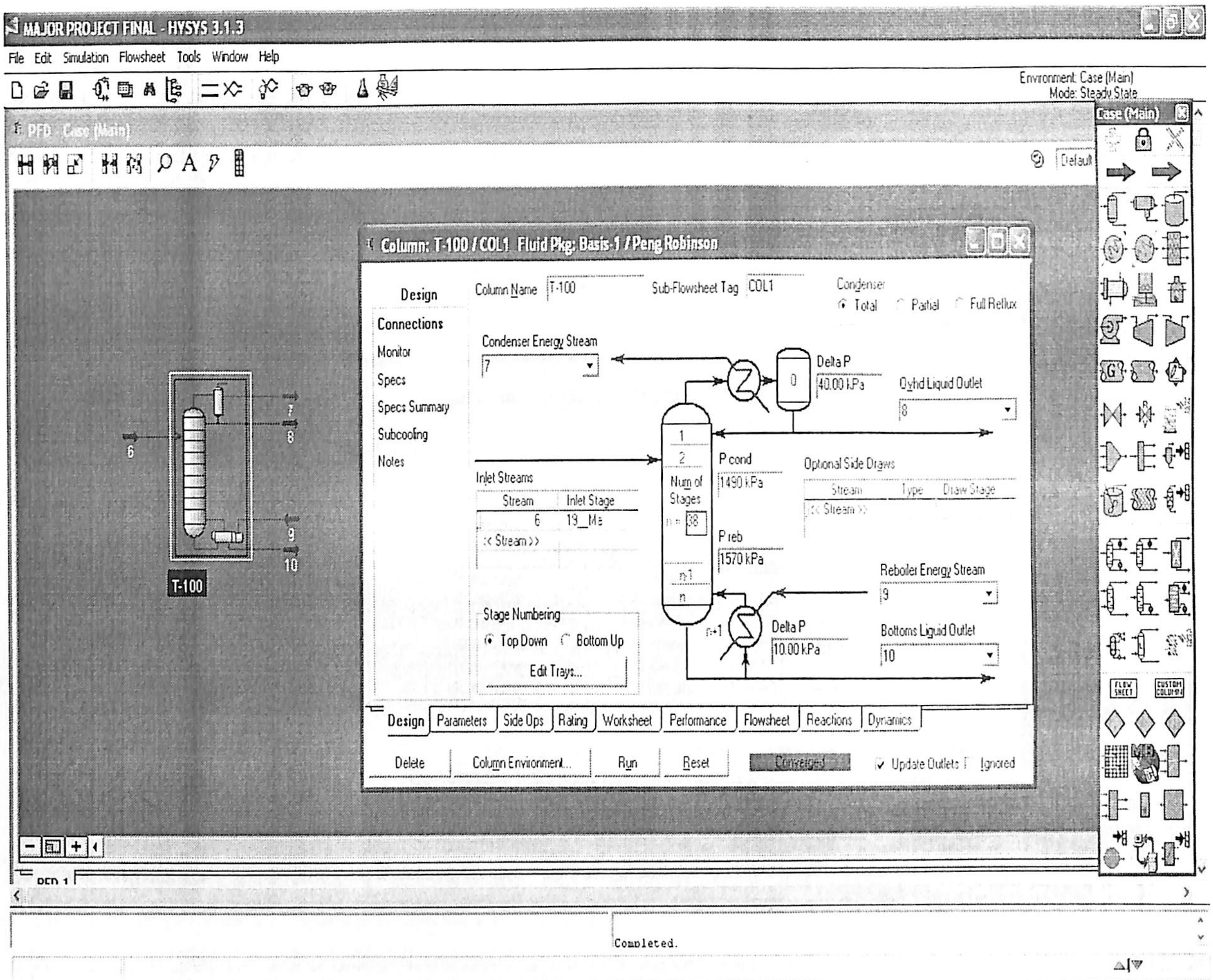


Figure 17

5. In the monitor tab, cick on “add spec” and add column component fraction with stage as “condenser” , flow basis as “mole fraction”, phase as “liquid” and key component obtained in the distillate with its mole fraction. Similarly, add another column component fraction for reboiler. Check the degree of freedom which should be zero as in figure 18.

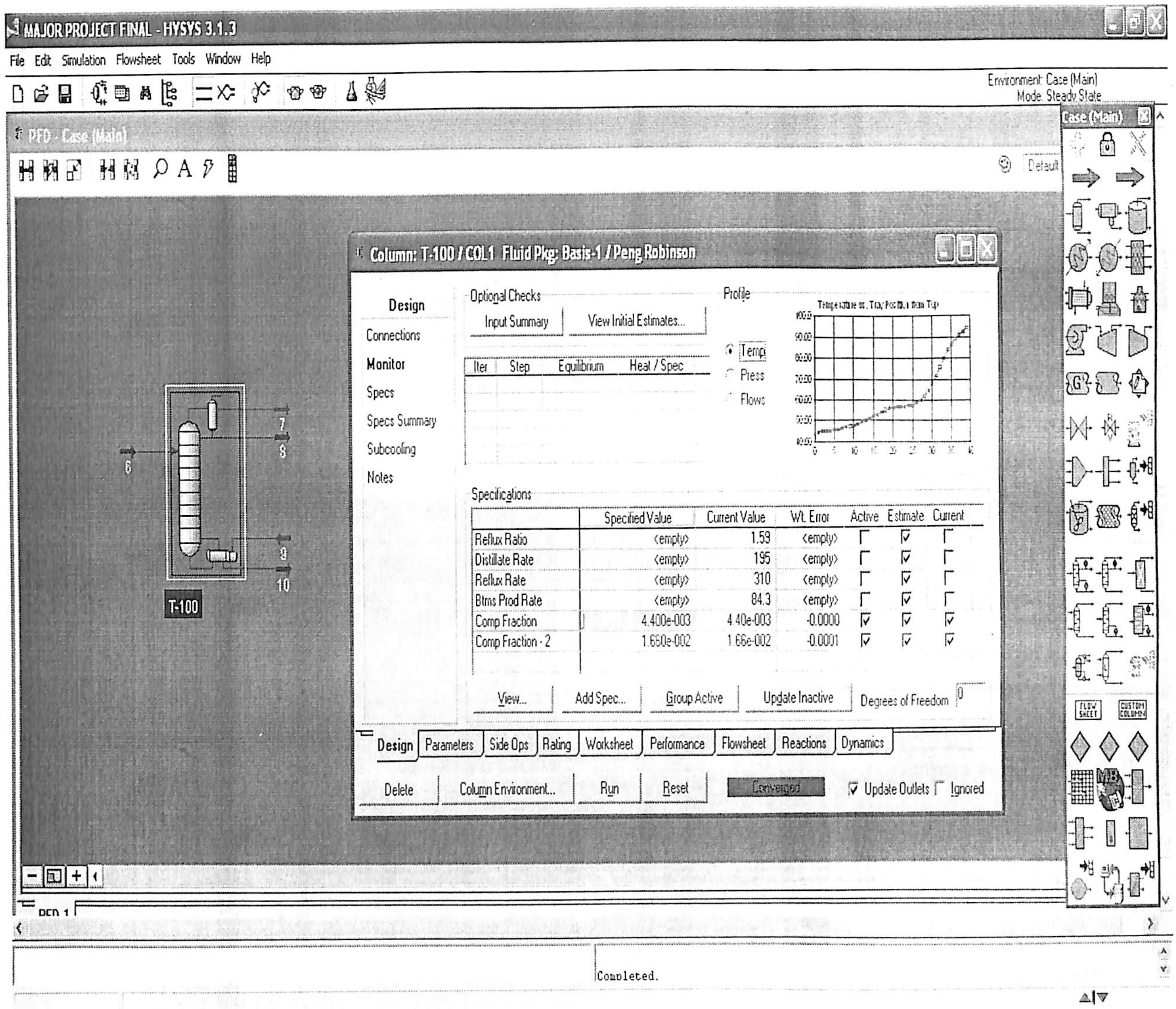


Figure 18

6. Again, click on the green button of "solver active". In this way, column T-100 can be converged. Thus, we obtain the pressure, temperature, net liquid flow & net vapour flow on the individual trays. Now, in tools menu, click on "utilities" (ctrl+U), add a utility of tray sizing and view tray sizing -1 utility. Click on "select TS" in setup bar under "design" tab & highlight "T-100". To obtain the values of tray internals, click on "results" bar under the "performance" tab.

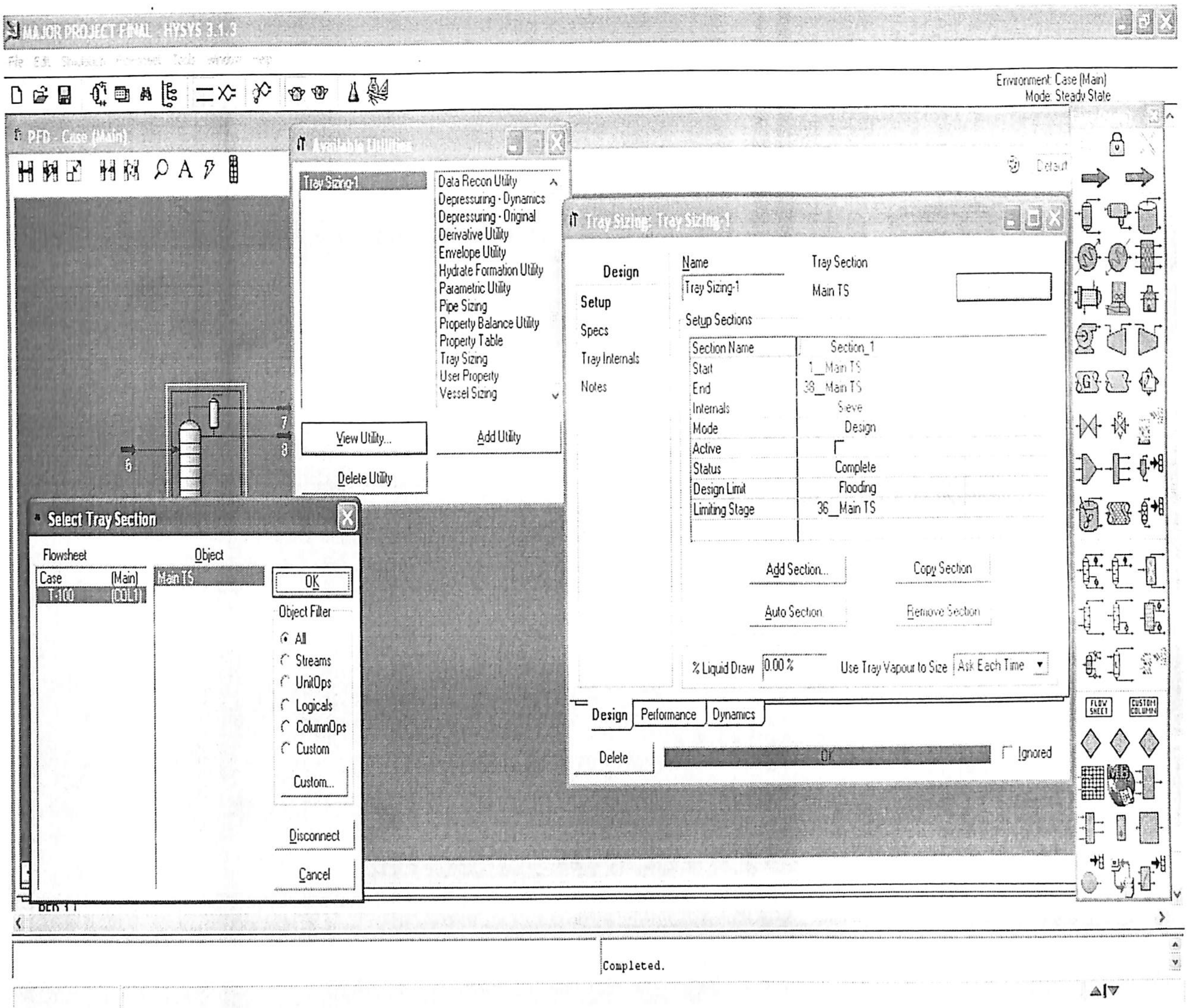


Figure 19

In this way, we can simulate a depropanizer column.

CHAPTER 3

DESIGNING OF DEPROPANIZER TRAY COLUMN

3. DESIGNING OF DEPROPANIZER TRAY COLUMN

3.1 CALCULATION OF BASIC PARAMETERS

The client has given us the following data.

Feed Flow rate	**279.3 Kgmole / hr**
Temperature	**65.60 °C**
Pressure	**20.40 bar**
Vapor Feed	**1**
Flooding	**85%**
Turndown	**50%**
Efficiency	**70%**

COMPOSION OF FEED

		Mole %	Mole Fraction
Ethane	**Light non-key (C2)**	**00.85**	**0.0085**
Propane	**Light key (C3)**	**69.17**	**0.6917**
i- Butane	**Heavy Key (C4)**	**12.51**	**0.1251**
n- Butane	**Heavy non-Key (C4)**	**17.39**	**0.1739**
i- Pentane	**Heavy non-Key (C5)**	**00.06**	**0.0006**
n- Pentane	**Heavy non-Key (C5)**	**00.02**	**0.0002**

PURITY THAT SHOULD BE OBTAINED FROM THE DISTILLATES

Propane	**Light key (C3)**	**98.3**	**0.9830**

PURITY THAT SHOULD BE OBTAINED FROM THE BOTTOMS

i- Butane	**Heavy Key (C4)**	**40.49**	**0.4049**
n- Butane	**Heavy non-Key (C4)**	**57.58**	**0.5758**

Writing,

Flow rate of Feed with F

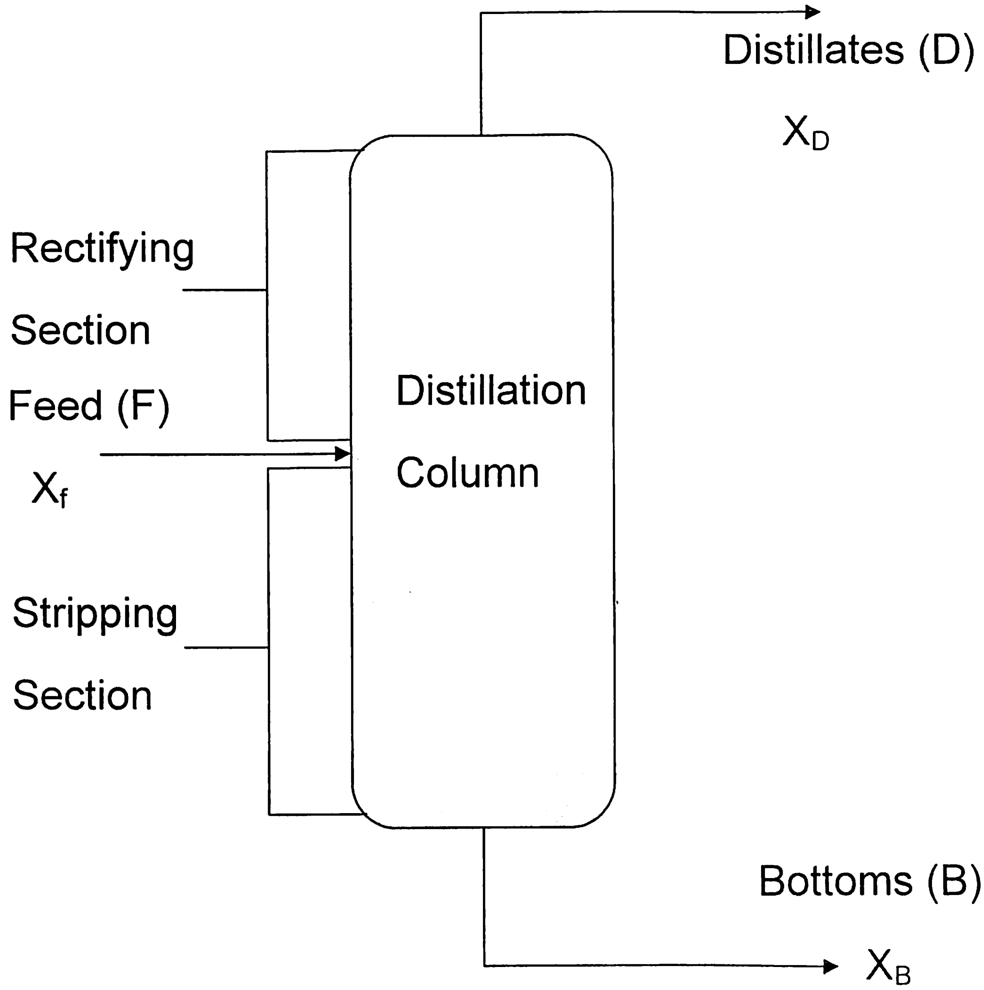
Mole fraction of Feed with X_f

Flow Rate of Distillates with D

Mole fraction of Distillates with X_D

Flow Rate of Bottoms with B

Mole fraction of Bottoms with X_B



APPLYING MASS/ MATERIAL BALANCE

To obtain the flow rate of Distillates and Bottoms and the mole fraction of non key components in Distillates and Bottoms

$D + B = F$

$D + B = 279.3$

APPLYING BUTANE BALANCE

$F * X_f = D * X_D + B * X_B$

$279.3 * (0.1251 + 0.1739) = D * (0.4049 + 0.5758) + B * (1 - 0.983 - X_{D,c})$

$279.3 * (0.1251 + 0.1739) = D * (0.4049 + 0.5758) + (279.3 - D) * (1 - 0.983 - X_{D,c})$ (1)

APPLYING ETHANE BALANCE

$F * X_f = D * X_D + B * X_B$

$279.3 * 0.0085 = D * X_{D,e}$ (2)

Solving equations (1) and (2), simultaneously, we get;

D = 195.1071 Kgmole / hr

B = 84.1928 Kgmole / hr

$X_{D,e} = 0.0122$

Similarly applying material balances for various components we get the composition of heavy and light key and non- key components.

		Distillate	**Bottoms**
Ethane	**Light non-key (C2)**	**0.0122**	**0.0000**
Propane	**Light key (C3)**	**0.9831**	**0.0166**
i- Butane	**Heavy Key (C4)**	**0.0044**	**0.4049**
n- Butane	**Heavy non-Key (C4)**	**0.0003**	**0.5758**
i- Pentane	**Heavy non-Key (C5)**	**0.0000**	**0.0020**
n- Pentane	**Heavy non-Key (C5)**	**0.0000**	**0.0007**
Total		**1.0000**	**1.0000**

From the values of Temperature and Pressure we can determine the values of K for different components using the K-value graph below.

α = Volatility = (K-value of ith component) / (K-value of Heavy Key)

z = compressibility factor = Mole fraction of feed

$$\text{Relative volatility} = \frac{\text{Volatility of i}^{\text{th}}\text{ term} \times \text{Compressibility of i}^{\text{th}}\text{ term}}{\text{Volatility of i}^{\text{th}}\text{ term} - \theta\text{ determined by trial method}}$$

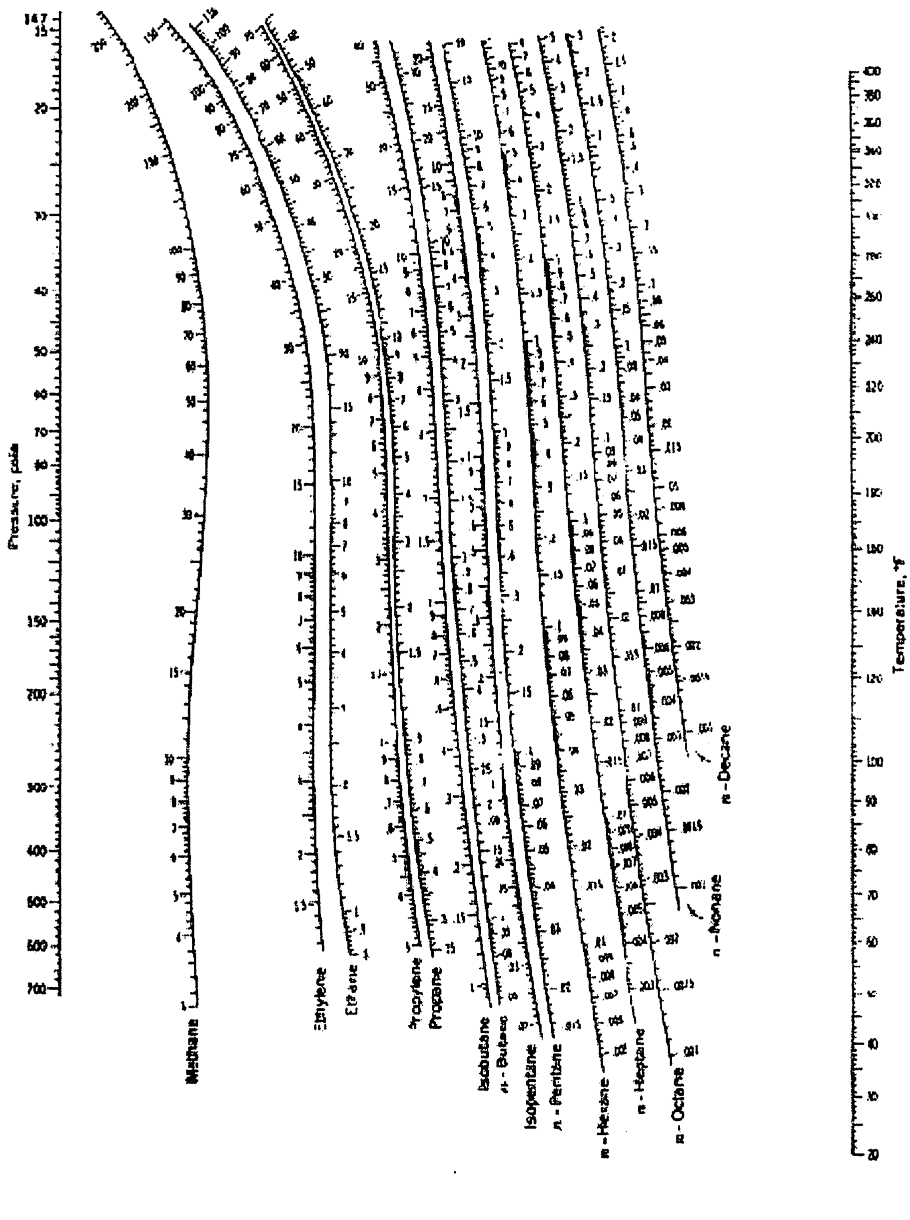


Figure 20. K-value Correlation Graph

DETERMINING K - VALUES FROM GRAPH

	K	α	z	Relative Volatilities	(α*Xd) / (α-θ)
Ethane	3.00	4.8387	0.0085	0.01	0.02
Propane	1.26	2.0323	0.6917	1.78	2.53
i- Butane	0.62	1.0000	0.1251	-0.52	-0.02
n- Butane	0.47	0.7581	0.1739	-0.27	0.00
i- Pentane	0.22	0.3548	0.0006	0.00	0.00
n- Pentane	0.19	0.3065	0.0002	0.00	0.00
Total				1.00	2.53

	Top	Middle	Bottom
K (Propane Light Key)	0.95	1.26	2.24
K (Butane Heavy Key)	0.4633	0.62	1.276
$\alpha = K_{LK} / K_{HK}$	2.0505	2.0322	1.7554

$$\beta_{LK/HK} * (K_{top}\ HK) \wedge \theta_{LK} = K_{top}\ LK$$

$$\beta_{LK/HK} * (K_{bottom}\ HK) \wedge \theta_{LK} = K_{bottom}\ LK$$

$$\beta_{LK/HK} * (0.4633) \wedge \theta_{LK} = 0.95$$

$$\beta_{LK/HK} * (1.276) \wedge \theta_{LK} = 2.24$$

Taking logarithm on both sides

$LOG_{10}(\beta_{LK/HK}) + \theta_{LK} LOG_{10}(K_{top} HK) = LOG_{10}(K_{top} LK)$

$LOG_{10}(\beta_{LK/HK}) + \theta_{LK} LOG_{10}(K_{bottom} HK) = LOG_{10}(K_{bottom} LK)$

$LOG_{10}(\beta_{LK/HK}) + \theta_{LK} LOG_{10}(0.4633) = LOG_{10}(0.95)$ (3)

$LOG_{10}(\beta_{LK/HK}) + \theta_{LK} LOG_{10}(1.276) = LOG_{10}(2.24)$ (4)

Solving (3) and (4) simultaneously, we get;

θ_{LK}	0.8466
$\beta_{LK/HK}$	1.8223

θ is determined such that the sum of the relative volatilities from the above table is as close to vapor feed which is 1.

θ (by trial method) 1.2425

A' and B' are any Gilliand Constants

B' = 0.48 from Gilliand Curve i.e. figure 2.

$$A' = \frac{\text{Reflux ratio} - \text{Minimum reflux}}{\text{Reflux ratio} + 1}$$

A' = 0.2320

$N_r / N_s = ((z_{LK} * (X_{B,LK})^2 * D) / (z_{HK} * (X_{D,HK})^2 * B))^{0.206}$

$N_r / N_s = 1.0268$

$$\text{Minimum number of stages} = \frac{\ln \dfrac{(X_{D,LK})(X_{B,HK})\theta_{LK}}{(X_{B,LK})(X_{D,HK})\theta_{LK}}}{\ln \beta_{\frac{LK}{HK}}}$$

$$= 13.1886$$

Minimum number of stages = 13

Minimum Reflux = $\sum (\alpha * Xd) / (\alpha - \theta) - 1$

= 2.53 – 1

Minimum reflux = 1.53

$$\text{Number of theoretical stages} = \frac{B' + \text{Minimum number of stages}}{(1 - B')}$$

= 26.2859

Number of theoretical stages = 26

Actual Number of Stages = Number of Theoretical Stages / Efficiency

= Number of Theoretical Stages / 0.7

= 37.5513

Actual number of stages = 38

Stripping Feed Tray Location = Actual Number of stages / (N_r/ N_s + 1)

= 18.5266

Stripping feed tray location = 19

Rectifying Feed Tray Location = Actual Number of Stages – Stripping Feed Tray Location

= 19.0246

Rectifying feed tray location = 19

Reflux Ratio = Minimum Reflux *1.5

Optimum reflux or Reflux ratio = 2.2935

3.2 CALCULATION OF COLUMN INTERNALS

Plate spacing (l_t) = 24 inches **Within the standard values**

From Hysys,

First stage ρ_L = 459.7 Kg/m^3

First stage ρ_V = 34.18 Kg/m^3

Last stage ρ_L = 462.6 Kg/m^3

Last stage ρ_V = 40.99 Kg/m^3

Top Section Liquid flow rate (L) = 13865 Kg/hr

Top Section Vapor flow rate (V) = 22445 Kg/hr

Bottom Section L = 30558 Kg/hr

Bottom Section V = 25677 Kg/hr

Surface Tension First stage = 4.685$\times 10^{-3}$ N/m

Surface Tension Last stage = 3.881$\times 10^{-3}$ N/m

Top Molecular Weight = 44.15

Bottom Molecular Weight = 57.75

3.2.1 Calculation of Column Diameter

$$F_{LV} = \frac{L}{V}\sqrt{\frac{\rho_V}{\rho_L}}$$

$$F_{LV\,top} = \frac{L_{top\ section}}{V_{top\ section}}\sqrt{\frac{\rho_{V\,top}}{\rho_{L\,top}}}$$

$$F_{LV\,top} = \frac{13865}{22445}\sqrt{\frac{34.18}{459.7}} = 0.19$$

$$F_{LV\,bottom} = \frac{L_{bottom\ section}}{V_{bottom\ section}}\sqrt{\frac{\rho_{V\,bottom}}{\rho_{L\,bottom}}}$$

$$F_{LV\,bottom} = \frac{30558}{25677}\sqrt{\frac{40.99}{462.6}} = 0.38$$

By using figure 5,

$$K_{1\,top}\ \text{after correction} = 8.1 \times 10^{-2} \times \left(\frac{4.685}{20}\right)^{0.2} = 0.0606$$

$$K_{1\,bottom}\ \text{after correction} = 5.5\times 10^{-2} \times \left(\frac{3.881}{20}\right)^{0.2} = 0.0397$$

$$u_f = K_1\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

$$u_{f\,top} = K_{1\,top}\sqrt{\frac{\rho_{L\,top} - \rho_{V\,top}}{\rho_{V\,top}}}$$

$$u_{f\,top} = 0.0606\sqrt{\frac{459.7-34.18}{34.18}} = 0.0606\times 3.5284 = 0.2138\ \text{m/sec}$$

$$u_{f\,bottom} = K_{1\,bottom}\sqrt{\frac{\rho_{L\,bottom} - \rho_{V\,bottom}}{\rho_{V\,bottom}}}$$

$$u_{f\,bottom} = 0.0396\sqrt{\frac{462.6-40.99}{40.99}} = 0.0396 \times 3.2071 = 0.1274 \text{ m/sec}$$

$\hat{u}_{top}$ = 85% of $u_{f\,top}$ = 0.1817 m/sec

$\hat{u}_{bottom}$ = 85% of $u_{f\,bottom}$ = 0.1083 m/sec

Maximum Volumetric Flow Rate,

$$\text{Top} = \frac{V_{top\,section} \text{ in Kg/hr}}{\rho_{V\,top} \text{ in Kg/m}^3 \times 3600}$$

$$\text{Top} = \frac{22445}{34.18 \times 3600} = 0.1824 \text{ m}^3\text{/sec}$$

$$\text{Bottom} = \frac{V_{bottom\,section} \text{ in Kg/hr}}{\rho_{V\,bottom} \text{ in Kg/m}^3 \times 3600}$$

$$\text{Bottom} = \frac{25677}{40.99 \times 3600} = 0.1739 \text{ m}^3\text{/sec}$$

Net Area (A_n),

$$\text{Top} = \frac{\text{Maximum Volumetric Flow Rate of Top}}{\hat{u}_{top}}$$

$$\text{Top} = \frac{0.1824}{0.1817} = 1.004 \text{ m}^2$$

$$\text{Bottom} = \frac{\text{Maximum Volumetric Flow Rate of Bottom}}{\hat{u}_{bottom}}$$

$$\text{Bottom} = \frac{0.1739}{0.1083} = 1.6057 \text{ m}^2$$

$A_d = 0.12\ A_C$

$A_n = A_C - A_d$

$A_n = A_C - 0.12\ A_C$

$$A_C = \frac{A_n}{0.88}$$

$$A_{C\ top} = \frac{A_{n\ top}}{0.88}$$

$$A_{C\ top} = \frac{1.004}{0.88} = 1.141\ m^2$$

$$A_{C\ bottom} = \frac{A_{n\ bottom}}{0.88}$$

$$A_{C\ bottom} = \frac{1.6057}{0.88} = 1.824\ m^2$$

$$D_C = \sqrt{\frac{A_C \times 4}{\pi}}$$

$$D_{C\ top} = \sqrt{\frac{A_{C\ top} \times 4}{\pi}}$$

$$D_{C\ top} = \sqrt{\frac{1.141 \times 4}{\pi}} = 1.205\ m$$

$$D_{C\ bottom} = \sqrt{\frac{A_{C\ bottom} \times 4}{\pi}}$$

$$D_{C\ bottom} = \sqrt{\frac{1.824 \times 4}{\pi}} = 1.524\ m$$

Diameter of column = 1.524 m

3.2.2 Calculation of Liquid Flow Arrangement

$$\text{Maximum Volumetric Liquid Rate} = \frac{L_{\text{bottom section}} \text{ in Kg/hr}}{\rho_{L\text{ bottom}} \times 3600}$$

$$\text{Maximum Volumetric Liquid Rate} = \frac{30558}{462.6 \times 3600} = 0.0183 \text{ m}^3/\text{sec}$$

From figure 6, at D_C = 1.524 m and Liquid Rate = 0.0183 m^3/sec.

We obtain Cross Flow (Single Pass) Liquid flow arrangement.

3.2.3 Calculation of Provisional Plate Design

Column Diameter (D_C) = 1.524 m

Column Area (A_C) = 1.824 m^2

Downcomer Area (A_d) =0.2189 m^2 (12% of Column Area)

Note: As per Hysys Downcomer Area = 0.3153 m^2 (17.3% of Column Area). We will be using this value for further calculation as Hysys is a simulator so as to obtain the desired result.

$A_n = A_C - A_d$ = 1.824 – 0.3153

Net Area (A_n) = 1.5087 m^2

$A_a = A_C - 2A_d$ = 1.824 – 0.6306

Active Area (A_a) = 1.194 m^2

Hole Area (A_h) = 0.1194 m^2 (10% of Active Area)

Note: As per Hysys Hole Area = 0.1678m^2 (14.05% of Active Area in Hysys)

From the figure 9, Weir Length (l_W) = 50.44 inches

Weir height (h_W) = 2 inches	**Within the standard values**
Hole size (d_h) = 0.1969 inches	**Within the standard values**
Hole pitch (l_p) = 0.5 inches	**Within the standard values**
Plate thickness= 0.1969 inches	**Within the standard values**

The Material Used For Plates is Stainless Steel.

3.2.4 Calculation to Check Weeping

$$\text{Maximum Liquid Rate} = \frac{L_{\text{bottom section}} \text{ in Kg/hr}}{3600}$$

$$\text{Maximum Liquid Rate} = \frac{30558}{3600} = 8.4883 \text{ Kg/sec}$$

$$\text{Minimum Liquid Rate} = \text{Turndown} \times \text{Maximum Liquid Rate}$$

$$\text{Minimum Liquid Rate} = 0.50 \times 8.4883 = 4.2441 \text{ Kg/sec}$$

$$h_{ow} = 750 \left[\frac{\text{Liquid Rate}}{\rho_{L\text{ bottom}} \times l_W}\right]^{2/3}$$

$$\text{Maximum } h_{ow} = 750 \left[\frac{\text{Maximum Liquid Rate}}{\rho_{L\text{ bottom}} \times l_W}\right]^{2/3}$$

$$\text{Maximum } h_{ow} = 750 \left[\frac{8.4883}{462.6 \times 1.2812}\right]^{2/3} = 44 \text{ mm liquid}$$

$$\text{Minimum } h_{ow} = 750 \left[\frac{4.2441}{462.6 \times 1.2812}\right]^{2/3} = 28 \text{ mm liquid}$$

$$\text{Minimum } h_W + h_{ow} = 51 + 28 = 79 \text{ mm liquid}$$

From figure 8, $K_2 = 30.6$

$$\breve{u}h_{(min)} = \frac{[K_2 - 0.90(25.4 - d_h)]}{(\rho_{V\ bottom})^{1/2}}$$

$$\breve{u}h_{(min)} = \frac{[30.6 - 0.90(25.4 - 5)]}{(40.99)^{1/2}} = 1.9118 \text{ m/sec}$$

$$\text{Actual Minimum Vapor Velocity} = \frac{\text{Turndown} \times \text{Maximum Volumetric flow rate}}{A_h}$$

$$\text{Actual Minimum Vapor Velocity} = \frac{0.50 \times 0.1739}{0.1678} = 0.5181 \text{ m/sec}$$

So minimum operating rate will be well above Weep Point

3.2.5 Calculation for Plate Pressure Drop

$$\text{Maximum Vapor Velocity} = \frac{\text{Maximum Volumetric flow rate in } m^3/\text{sec}}{A_h}$$

$$\text{Maximum Vapor Velocity} = \frac{0.1739}{0.1678} = 1.036 \text{ m/sec}$$

From figure 12, using the values of

Plate Thickness / Hole Diameter = 1

$A_h / A_p = A_h / A_a = 14.05$

We obtain $C_O = 0.87$

$$\text{Dry Plate Drop } (h_d) = 51 \left[\frac{u_h}{C_O}\right]^2 \frac{\rho_{v\ bottom}}{\rho_{L\ bottom}}$$

$$h_d = 51 \left[\frac{1.036}{0.87}\right]^2 \frac{40.99}{462.6}$$

h_d = 6 mm

$$\text{Residual Head } (h_r) = \frac{12.5 \times 10^3}{\rho_{L\ bottom}}$$

$$h_r = \frac{12.5 \times 10^3}{462.6}$$

$\mathbf{h_r = 27\ mm}$

$$\text{Total Drop } (h_t) = h_d + (h_w + h_{ow}) + h_r$$

$$h_t = 6 + (51 + 44) + 27$$

$\mathbf{h_t = 128\ mm}$

$$\text{Total Plate Pressure Drop } (\Delta P_t) = 9.81 \times 10^{-3} h_t \rho_{L\,bottom}$$

$$\Delta P_t = 9.81 \times 10^{-3} \times 128 \times 462.6$$

$\mathbf{\Delta P_t = 580.8776\ Pa}$

3.2.6 Calculation for Downcomer

$$\text{Downcomer Clearance } (h_{ap}) = h_w - 5 = 50.8 - 5$$

$\mathbf{h_{ap} = 45.8\ mm}$

$A_d = 0.3153\ m^2$

$$\text{Downcomer Clearance Area } (A_{ap}) = l_w \times h_{ap} = 1.281 \times 0.0458$$

$\mathbf{A_{ap} = 0.0587\ m^2}$

A_{ap} is less than A_d hence we will use this in the formula to calculate head loss in the downcomer

$$\text{Downcomer head loss } (h_{dc}) = 166 \left[\frac{\text{Maximum liquid rate in Kg/sec}}{\rho_{L\,bottom} A_{ap}}\right]^2$$

$$h_{dc} = 166 \left[\frac{8.4883}{462.6 \times 0.0587}\right]^2 = 16\ mm$$

$$\text{Downcomer Backup } (h_b) = h_t + (h_w + h_{ow}) + h_{dc}$$

$$h_b = 128 + (51 + 44) + 16$$

$\mathbf{h_b = 239}$ **mm**

$$h_b < \frac{1}{2}(\text{plate spacing} + \text{weir height})$$

$239 < \frac{1}{2}(609.6 + 50.8) = 239<330.2$

Hence tray spacing is acceptable

$$\text{Downcomer Residence Time } (t_r) = \frac{A_d h_b \rho_{L\,bottom}}{L_{wd}}$$

$$t_r = \frac{0.3153 \times 0.239 \times 462.6}{8.4883}$$

$\mathbf{t_r = 4.1068}$ **sec (Which is greater than 3 sec)**

3.2.7 Calculation of Entrainment

For $F_{LV} = 0.38$ from figure 7,

$\Psi = 0.005$ which is well below 0.1

3.2.8 Calculation for Number of Holes

$$\text{Area of One Hole} = \frac{\pi}{4} \times {d_h}^2 = 1.9637 \times 10^{-5}\ m^2$$

$$\text{Number Of Holes} = \frac{\text{Hole Area}}{\text{Area of One Hole}}$$

$$\text{Number Of Holes} = \frac{0.1678}{1.9637 \times 10^{-5}}$$

Number Of Holes = 8545

3.3 DATASHEETS OBTAINED FROM ASPEN HYSYS

HYPROTECH	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
TEAM LND Calgary, Alberta CANADA	Unit Set:	SI
	Date/Time:	Thu Apr 23 01:51:50 2009

Shortcut Column: T-100-2

Connections

Feed	Condensor Duty	Distillate	Reboiler Duty	Bottoms
1-2	2-2	3-2	4-2	5-2

Parameters

	Component	Mole Fraction
Light Key	Propane	1.660e-002 *
Heavy Key	i-Butane	4.400e-003 *

Pressures (kPa)		Reflux Ratios	
Condenser Pressure	1490 *	External Reflux Ratio	2.293 *
Reboiler Pressure	1570 *	Minimum Reflux Ratio	1.529

User Variables

Results Summary

Trays / Temperatures		Flows		
Minimum # of Trays	14.93	Rectify Vapour	(kgmole/h)	642.5
Actual # of Trays	27.13	Rectify Liquid	(kgmole/h)	447.4
Optimal Feed Stage	13.71	Stripping Vapour	(kgmole/h)	642.5
Condenser Temperature (C)	42.53	Stripping Liquid	(kgmole/h)	726.7
Reboiler Temperature (C)	94.32	Condenser Duty	[illegible]	[illegible]e+006
		Reboiler Duty	[illegible]	[illegible]e+006

Hyprotech Ltd. | HYSYS v3.1.3 (Build 4827) | Page 1 of 1

Licensed to: TEAM LND

* Specified by user

HYPROTECH — TEAM LND, Calgary, Alberta, CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\…
	Unit Set	SI
	Date/Time	Thu Apr 23 02 14:21 2…

Distillation: T-100

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION
9	Reboiler	
6	19__Main TS	

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
7	Condenser	
8	Condenser	
10	Reboiler	

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
Reflux Ratio	---	1.585	---	1.000e-002 *	1.000e-002 *	Off	On	Off
Distillate Rate	---	195.0 kgmole/h	---	1.000e-002 *	1.000 kgmole/h *	Off	On	Off
Reflux Rate	---	309.8 kgmole/h	---	1.000e-002 *	1.000 kgmole/h *	Off	On	Off
Btms Prod Rate	---	84.25 kgmole/h	---	1.000e-002 *	1.000 kgmole/h *	Off	On	Off
Comp Fraction	4.400e-003 *	4.400e-003	-3.710e-006	1.000e-002 *	1.000e-003 *	On	On	On
Comp Fraction - 2	1.660e-002 *	1.660e-002	-1.051e-004	1.000e-002 *	1.000e-003 *	On	On	On

SPECS

Column Specification Parameters

Reflux Ratio

Fixed / Ranged	Fixed	Primary / Alternate	Primary	Lower Bound	---	Upper Bound	---
Stage	Condenser	Flow Basis	Molar	Liquid Specification	---		

Distillate Rate

Fixed / Ranged	Fixed	Primary / Alternate	Primary	Lower Bound	---	Upper Bound	---
Stream	8 @COL1	Flow Basis	Molar				

Reflux Rate

Fixed / Ranged	Fixed	Primary / Alternate	Primary	Lower Bound	---	Upper Bound	---
Stage	Condenser	Flow Basis	Molar	Liquid Specification	---		

Btms Prod Rate

Fixed / Ranged	Fixed	Primary / Alternate	Primary	Lower Bound	---	Upper Bound	---
Stream	10 @COL1	Flow Basis	Molar				

Comp Fraction

Fixed / Ranged	Fixed	Primary / Alternate	Primary	Lower Bound	---	Upper Bound	---
Stage	Condenser	Flow Basis	Mole Fraction	Phase	Liquid		
Components		i-Butane					

Comp Fraction - 2

Fixed / Ranged	Fixed	Primary / Alternate	Primary	Lower Bound	---	Upper Bound	---
Stage	Reboiler	Flow Basis	Mole Fraction	Phase	Liquid		
Components		Propane					

SUBCOOLING

	Condenser	
Degrees of Subcooling	---	

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 1 of 23

Licensed to: TEAM LND

* Specified by user

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set:	SI
		Date/Time:	Thu Apr 23 02 14:21 2009

Distillation: T-100 (continued)

SUBCOOLING

Subcool to		

User Variables

PROFILES

General Parameters

Sub-Flow Sheet	T-100 (COL1)	Number of Stages	38 *

Profile Estimates

	Temperature (C)	Net Liquid (kgmole/h)	Net Vapour (kgmole/h)
Condenser	42.93 *	[illegible]	[illegible]e-020
1__Main TS	44.5[illegible]	[illegible]	504.9
2__Main TS	44.80	[illegible]	509.1
3__Main TS	45.15	313.2	508.8
4__Main TS	45.42	312.5	508.3
5__Main TS	45.72	311.7	507.6
6__Main TS	46.07	310.7	506.7
7__Main TS	46.47	309.6	505.8
8__Main TS	46.93	[illegible]	504.6
9__Main TS	47.4[illegible]	[illegible]	[illegible]
10__Main TS	[illegible]	305.0	501.8
11__Main TS	48.73	303.1	500.0
12__Main TS	49.47	301.0	498.1
13__Main TS	50.28	298.7	496.0
14__Main TS	51.16	296.3	493.8
15__Main TS	52.09	293.8	491.4
16__Main TS	53.07	291.2	488.8
17__Main TS	54.08	288.6	486.2
18__Main TS	55.10	[illegible]	483.6
19__Main TS	56.11	[illegible]	481.0
20__Main TS	56.33	537.1	452.5
21__Main TS	56.50	537.2	452.9
22__Main TS	56.67	537.0	452.9
23__Main TS	56.91	536.6	452.8
24__Main TS	57.27	535.8	452.3
25__Main TS	57.82	534.5	451.5
26__Main TS	[illegible]	[illegible]	450.2
27__Main TS	[illegible]	[illegible]	448.3
28__Main TS	61.81	[illegible]	445.6
29__Main TS	64.40	522.6	442.1
30__Main TS	67.75	519.3	438.3
31__Main TS	71.72	517.3	435.0
32__Main TS	75.97	517.4	433.1
33__Main TS	80.10	519.1	433.1
34__Main TS	83.77	521.9	434.9
35__Main TS	[illegible]	[illegible]	437.7
36__Main TS	[illegible]	[illegible]	440.6
37__Main TS	[illegible]	[illegible]	443.0
38__Main TS	92.73	529.1	444.6
Reboiler	94.32 *	84.25	444.9

EFFICIENCIES

Hyprotech Ltd.	HYSYS v3.1.3 (Build 4827)	Page 2 of 23

Licensed to: TEAM LND

* Specified by user

HYPROTECH	TEAM LnD Calgary, Alberta CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\...
		Unit Set	SI
		Date/Time	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

Stage Efficiencies

Stages	Overall Efficiency	Ethane	Propane	i-Butane	n-Butane	i-Pentane
Condenser	1.000	1.000	1.000	1.000	[illegible]	1.000
1__Main TS	1.000	1.000	1.000	[illegible]	[illegible]	1.000
2__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
3__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
4__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
5__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
6__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
7__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
8__Main TS	1.000	1.000	1.000	1.000	[illegible]	1.000
9__Main TS	1.000	1.000	1.000	1.000	[illegible]	1.000
10__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
11__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
12__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
13__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
14__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
15__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
16__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
17__Main TS	1.000	1.000	1.000	1.000	[illegible]	1.000
18__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
19__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
20__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
21__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
22__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
23__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
24__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
25__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
26__Main TS	1.000	1.000	1.000	1.000	[illegible]	1.000
27__Main TS	1.000	1.000	1.000	1.000	[illegible]	1.000
28__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
29__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
30__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
31__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
32__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
33__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
34__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
35__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
36__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
37__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
38__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
Reboiler	1.000	1.000	1.000	1.000	1.000	1.000

Stages	Overall Efficiency	n-Pentane
Condenser	1.000	1.000
1__Main TS	1.000	1.000
2__Main TS	1.000	1.000
3__Main TS	1.000	1.000
4__Main TS	1.000	1.000
5__Main TS	1.000	1.000
6__Main TS	1.000	1.000
7__Main TS	1.000	1.000
8__Main TS	1.000	1.000
9__Main TS	1.000	1.000
10__Main TS	1.000	1.000

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 3 of 23

Licensed to: TEAM LnD Specified by user.

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\
		Unit Set	SI
		Date/Time	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

Stage Efficiencies

Stages	Overall Efficiency	n-Pentane				
11__Main TS	1.000	1.000				
12__Main TS	1.000	1.000				
13__Main TS	1.000	1.000				
14__Main TS	1.000	1.000				
15__Main TS	1.000	1.000				
16__Main TS	1.000	1.000				
17__Main TS	1.000	1.000				
18__Main TS	1.000	1.000				
19__Main TS	1.000	1.000				
20__Main TS	1.000	1.000				
21__Main TS	1.000	1.000				
22__Main TS	1.000	1.000				
23__Main TS	1.000	1.000				
24__Main TS	1.000	1.000				
25__Main TS	1.000	1.000				
26__Main TS	1.000	1.000				
27__Main TS	1.000	1.000				
28__Main TS	1.000	1.000				
29__Main TS	1.000	1.000				
30__Main TS	1.000	1.000				
31__Main TS	1.000	1.000				
32__Main TS	1.000	1.000				
33__Main TS	1.000	1.000				
34__Main TS	1.000	1.000				
35__Main TS	1.000	1.000				
36__Main TS	1.000	1.000				
37__Main TS	1.000	1.000				
38__Main TS	1.000	1.000				
Reboiler	1.000	1.000				

SOLVER

Column Solving Algorithm: HYSIM Inside-Out

Solving Options		Acceleration Parameters	
Maximum Iterations	10000	Accelerate K Value & H Model Parameters	Off
Equilibrium Error Tolerance	1.000e-05		
Heat/Spec Error Tolerance	5.000e-004		
Save Solutions as Initial Estimate	On		
Super Critical Handling Model	Simple K		
Trace Level	Low		
Init from Ideal K's	Off	**Damping Parameters**	
Initial Estimate Generator Parameters		Azeotrope Check	Off
Iterative IEG (Good for Chemicals)	Off	Fixed Damping Factor	1

SIDE STRIPPERS

SIDE RECTIFIERS

PUMP AROUNDS

VAP BYPASSES

* Specified by user

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set:	SI
		Date/Time:	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

RATING

Tray Sections

Tray Section		Main TS @COL1			
Tray Diameter	(m)	1.500 *			
Weir Height	(m)	5.000e-002 *			
Weir Length	(m)	1.200 *			
Tray Space	(m)	0.5500 *			
Tray Volume	(m3)	0.9719			
Disable Heat Loss Calculations		No			
Heat Model		None			
Rating Calculations		No			
Tray Hold Up	(m3)	8.836e-002			

Vessels

Vessel		Reboiler @COL1	Condenser @COL1		
Diameter	(m)	1.193	1.193		
Length	(m)	1.789	1.789		
Volume	(m3)	2.000 *	2.000 *		
Orientation		Horizontal	Horizontal		
Vessel has a Boot		No	No		
Boot Diameter	(m)	---	---		
Boot Length	(m)	---	---		
Hold Up	(m3)	1.000	1.000		

Other Equipment In Column Flowsheet

Pressure Profile

	Pressure (kPa)	Pressure Drop (kPa)
Condenser	1490 kPa *	40.00 kPa
1__Main TS	1530 kPa	0.8108 kPa
2__Main TS	1531 kPa	0.8108 kPa
3__Main TS	1532 kPa	[illegible]
4__Main TS	1532 kPa	0.8108 kPa
5__Main TS	1533 kPa	0.8108 kPa
6__Main TS	1534 kPa	0.8108 kPa
7__Main TS	1535 kPa	0.8108 kPa
8__Main TS	1536 kPa	0.8108 kPa
9__Main TS	1536 kPa	0.8108 kPa
10__Main TS	1537 kPa	0.8108 kPa
11__Main TS	1538 kPa	[illegible]
12__Main TS	1539 kPa	[illegible]
13__Main TS	1540 kPa	[illegible]
14__Main TS	1541 kPa	0.8108 kPa
15__Main TS	1541 kPa	0.8108 kPa
16__Main TS	1542 kPa	0.8108 kPa
17__Main TS	1543 kPa	0.8108 kPa
18__Main TS	1544 kPa	0.8108 kPa
19__Main TS	1545 kPa	0.8108 kPa
20__Main TS	1545 kPa	[illegible]
21__Main TS	1546 kPa	[illegible]
22__Main TS	1547 kPa	[illegible]

Hyprotech Ltd. | HYSYS v3.1.3 (Build 4827) | Page 5 of 23

Licensed to: TEAM LND

* Specified by user

HYPROTECH TEAM LND Calgary, Alberta CANADA	Case Name: C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
	Unit Set: SI
	Date/Time: Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

23__Main TS	1548 kPa	0.8109 kPa
24__Main TS	1549 kPa	0.8106 kPa
25__Main TS	1549 kPa	0.8109 kPa
26__Main TS	1550 kPa	0.8108 kPa
27__Main TS	1551 kPa	0.8108 kPa
28__Main TS	1552 kPa	0.8109 kPa
29__Main TS	1553 kPa	[illegible] kPa
30__Main TS	1554 kPa	[illegible] kPa
31__Main TS	1554 kPa	0.8109 kPa
32__Main TS	1555 kPa	0.8108 kPa
33__Main TS	1556 kPa	0.8108 kPa
34__Main TS	1557 kPa	0.8108 kPa
35__Main TS	1558 kPa	0.8108 kPa
36__Main TS	1558 kPa	0.8108 kPa
37__Main TS	1559 kPa	0.8109 kPa
38__Main TS	1560 kPa	[illegible]
Reboiler	1570 kPa	[illegible] kPa

Pressure Solving Options

Pressure Tolerance	1.000e-004 *	Pressure Drop Tolerance	1.000e-004 *	Damping Factor	1.000 *	Max Press Iterations	100 *

PROPERTIES

Properties : 6 @COL1

	Overall	Liquid Phase			
Vapour/Phase Fraction	0.0000	1.0000			
Temperature (C)	65.60	65.60			
Pressure (kPa)	2040	2040			
Molar Flow (kgmole/h)	279.3	279.3			
Mass Flow (kg/h)	1.346e+004	1.346e+004			
Std Ideal Liq Vol Flow (m3/h)	25.49	25.49			
Molar Enthalpy (kJ/kgmole)	-1.233e+005	-1.233e+005			
Mass Enthalpy (kJ/kg)	-2559	-2559			
Molar Entropy (kJ/kgmole-C)	107.4	107.4			
Mass Entropy (kJ/kg-C)	2.229	2.229			
Heat Flow (kJ/h)	-3.444e+007	-3.444e+007			
Molar Density (kgmole/m3)	9.358	9.358			
Mass Density (kg/m3)	451.0	451.0			
Std Ideal Liq Mass Density (kg/m3)	528.0	528.0			
Liq Mass Density @Std Cond (kg/m3)	531.6	531.6			
Molar Heat Capacity (kJ/kgmole-C)	161.9	161.9			
Mass Heat Capacity (kJ/kg-C)	3.360	3.360			
Thermal Conductivity (W/m-K)	7.229e-002	7.229e-002			
Viscosity (cP)	8.048e-002	8.048e-002			
Surface Tension (dyne/cm)	3.736	3.736			
Molecular Weight	48.19	48.19			
Z Factor	7.740e-002	7.740e-002			

Properties : 8 @COL1

	Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction	0.0000	0.0000	1.0000		
Temperature (C)	42.52	42.52	42.52		
Pressure (kPa)	1490 *	1490	1490		
Molar Flow (kgmole/h)	195.0	0.0000	195.0		
Mass Flow (kg/h)	8580	0.0000	8580		
Std Ideal Liq Vol Flow (m3/h)	16.98	0.0000	16.98		

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 6 of 23

Licensed to: TEAM LND * Specified by user

HYPROTECH	TEAM LND Calgary Alberta CANADA	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set:	SI
		Date/Time:	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

Properties : 8 @COL1

		Overall	Vapour Phase	Liquid Phase
Molar Enthalpy	(kJ/kgmole)	-1.176e+005	-1.035e+005	-1.176e+005
Mass Enthalpy	(kJ/kg)	-2073	-2376	-2073
Molar Entropy	(kJ/kgmole-C)	99.09	142.1	99.09
Mass Entropy	(kJ/kg-C)	2.252	3.252	[illegible]
Heat Flow	(kJ/h)	-2.204e+007	0.0000	-2.204e+007
Molar Density	(kgmole/m3)	[illegible]	[illegible]	[illegible]
Mass Density	(kg/m3)	462.1	32.92	462.1
Std Ideal Liq Mass Density	(kg/m3)	505.2	502.3	505.2
Liq Mass Density @Std Cond	(kg/m3)	506.9	506.0	506.9
Molar Heat Capacity	(kJ/kgmole-C)	140.3	92.72	140.3
Mass Heat Capacity	(kJ/kg-C)	3.190	2.122	3.190
Thermal Conductivity	(W/m-K)	8.432e-002	2.155e-002	8.432e-002
Viscosity	(cP)	8.368e-002	9.553e-003	8.368e-002
Surface Tension	(dyne/cm)	4.856	---	4.856
Molecular Weight		43.99	43.70	43.99
Z Factor		5.404e-002	0.7536	5.404e-002

Properties : 10 @COL1

		Overall	Vapour Phase	Liquid Phase
Vapour/Phase Fraction		0.0000	0.0000	1.0000
Temperature	(C)	94.32	94.32	94.32
Pressure	(kPa)	1570 *	1570	1570
Molar Flow	(kgmole/h)	84.26	0.0000	[illegible]
Mass Flow	(kg/h)	[illegible]	0.0000	[illegible]
Std Ideal Liq Vol Flow	(m3/h)	8.511	0.0000	8.511
Molar Enthalpy	(kJ/kgmole)	-1.390e+005	-1.245e+005	-1.390e+005
Mass Enthalpy	(kJ/kg)	-2400	-2157	-2400
Molar Entropy	(kJ/kgmole-C)	103.3	144.7	103.3
Mass Entropy	(kJ/kg-C)	1.784	2.509	1.784
Heat Flow	(kJ/h)	-1.171e+007	0.0000	-1.171e+007
Molar Density	(kgmole/m3)	7.975	0.7163	7.975
Mass Density	(kg/m3)	462.0	41.34	462.0
Std Ideal Liq Mass Density	(kg/m3)	573.5	571.7	573.5
Liq Mass Density @Std Cond	(kg/m3)	575.0	573.3	575.0
Molar Heat Capacity	(kJ/kgmole-C)	192.5	136.5	192.5
Mass Heat Capacity	(kJ/kg-C)	3.323	2.365	3.323
Thermal Conductivity	(W/m-K)	6.295e-002	2.462e-002	6.295e-002
Viscosity	(cP)	8.844e-002	1.028e-002	8.844e-002
Surface Tension	(dyne/cm)	3.639	---	3.639
Molecular Weight		57.93	57.72	57.93
Z Factor		6.444e-002	0.7174	6.444e-002

SUMMARY

Flow Basis:	Molar	The composition option is selected

Feed Composition

	6
Flow Rate (kgmole/h)	279.3000
	***
Ethane	0.0085
Propane	0.691[illegible]
i-Butane	0.1251
n-Butane	0.1739
i-Pentane	0.0006

Hyprotech Ltd. | HYSYS v3.1.3 (Build 4827) | Page 7 of 23

Licensed to: TEAM LND | * Specified by user

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name: C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set: SI
		Date/Time: Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

SUMMARY

	6				
n-Pentane	0.0002				

Flow Basis: Molar — The composition option is selected

Feed Flows

	6				
Flow Rate (kgmole/h)	279.3000				

Ethane (kgmole/h)	2.3741				
Propane (kgmole/h)	193.1918				
i-Butane (kgmole/h)	34.9404				
n-Butane (kgmole/h)	48.5703				
i-Pentane (kgmole/h)	0.1676				
n-Pentane (kgmole/h)	0.0559				

Products

Flow Basis: Molar — The composition option is selected

Product Compositions

	8	10			
Flow Rate (kgmole/h)	195.0472	84.2528			
	---	---			
Ethane	0.0122	0.0000			
Propane	0.9833	0.0166			
i-Butane	0.0044	0.4045			
n-Butane	0.0001	0.5762			
i-Pentane	0.0000	0.0020			
n-Pentane	0.0000	0.0007			

Flow Basis: Molar — The composition option is selected

Product Flows

	8	10			
Flow Rate (kgmole/h)	195.0472 *	84.2528 *			
	---	---			
Ethane (kgmole/h)	2.3740 *	[illegible] *			
Propane (kgmole/h)	191.7935 *	[illegible] *			
i-Butane (kgmole/h)	0.8581 *	34.0823 *			
n-Butane (kgmole/h)	0.0215 *	48.5488 *			
i-Pentane (kgmole/h)	0.0000 *	0.1676 *			
n-Pentane (kgmole/h)	0.0000 *	0.0559 *			

Flow Basis: Molar — The composition option is selected

Product Recoveries

	8	10			
Flow Rate (kgmole/h)	195.0472	84.2528			
	---	---			
Ethane (%)	100.0000	0.0000			
Propane (%)	99.2762	0.7238			
i-Butane (%)	2.4560	97.5440			
n-Butane (%)	0.0443	99.9557			
i-Pentane (%)	0.0000	100.0000			
n-Pentane (%)	0.0000	100.0000			

COLUMN PROFILES

Reflux Ratio	1.588	Reboil Ratio	5.283	The Flows Option is Selected	Flow Basis: Molar

Column Profiles Flows

* Specified by user

HYPROTECH	TEAM LND Calgary, Alberta CANADA	
	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\[illegible]
	Unit Set:	SI
	Date/Time:	Thu Apr [illegible]

Distillation: T-100 (continued)

COLUMN PROFILES

	Temperature (C)	Pressure (kPa)	Net Liq (kgmole/h)	Net Vap (kgmole/h)	Net Feed (kgmole/h)	Net Draws (kgmole/h)
Condenser	42.52	[illegible]	[illegible]	---		[illegible]
1__Main TS	44.53	[illegible]	[illegible]	[illegible]		
2__Main TS	44.89	[illegible]	[illegible]	[illegible]	---	
3__Main TS	45.15	1532	[illegible]	508.8	---	---
4__Main TS	45.42	1532	312.5	508.3	---	---
5__Main TS	45.72	1533	311.7	507.6	---	---
6__Main TS	46.07	1534	310.7	506.7	---	---
7__Main TS	46.47	1535	309.6	505.6	---	---
8__Main TS	46.93	1536	308.2	504.6	---	---
9__Main TS	47.46	[illegible]	[illegible]	[illegible]	---	---
10__Main TS	48.06	[illegible]	[illegible]	[illegible]	---	---
11__Main TS	48.73	[illegible]	[illegible]	[illegible]	---	---
12__Main TS	49.47	1539	301.9	[illegible]	---	---
13__Main TS	50.28	1540	298.7	496.0	---	---
14__Main TS	51.16	1541	[illegible]	493.8	---	---
15__Main TS	52.09	1541	293.8	491.4	---	---
16__Main TS	53.07	1542	291.2	489.8	---	---
17__Main TS	54.08	1543	288.6	486.2	---	---
18__Main TS	55.10	1544	[illegible]	[illegible]	---	---
19__Main TS	56.11	1545	[illegible]	[illegible]	[illegible]	---
20__Main TS	56.33	1545	[illegible]	452.5	---	---
21__Main TS	56.50	1546	537.2	452.9	---	---
22__Main TS	56.67	1547	537.0	452.9	---	---
23__Main TS	56.91	1548	536.6	452.8	---	---
24__Main TS	57.27	1549	535.8	452.3	---	---
25__Main TS	57.82	1549	534.5	451.5	---	---
26__Main TS	58.66	1550	532.5	450.2	---	---
27__Main TS	59.94	1551	529.8	[illegible]	---	---
28__Main TS	61.81	1552	526.4	445.6	---	---
29__Main TS	64.40	1553	522.6	442.1	---	---
30__Main TS	67.75	1554	519.3	438.3	---	---
31__Main TS	71.72	1554	517.3	435.0	---	---
32__Main TS	75.97	1555	517.4	433.1	---	---
33__Main TS	80.10	1556	519.1	433.1	---	---
34__Main TS	83.77	1557	521.9	434.9	---	---
35__Main TS	86.83	1558	524.9	[illegible]	---	---
36__Main TS	89.27	1558	[illegible]	[illegible]		
37__Main TS	91.10	[illegible]	[illegible]	[illegible]		
38__Main TS	92.73	[illegible]	[illegible]	444.6	---	---
Reboiler	94.32	1570	---	444.0	---	[illegible]

Column Profiles Energy

	Temperature (C)	Liquid Enthalpy (kJ/kgmole)	Vapour Enthalpy (kJ/kgmole)	Heat Loss (kJ/h)
Condenser	42.52	-1.176e+005	-1.033e+005	---
1__Main TS	44.53	-1.176e+005	-1.042e+005	---
2__Main TS	44.89	-1.178e+005	-1.044e+005	---
3__Main TS	45.15	[illegible]	[illegible]	---
4__Main TS	45.42	[illegible]	[illegible]	---
5__Main TS	45.72	-1.156e+005	-1.047e+005	---
6__Main TS	46.07	-1.187e+005	-1.048e+005	---
7__Main TS	46.47	-1.191e+005	-1.050e+005	---
8__Main TS	46.93	-1.194e+005	-1.051e+005	---

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 9 of 23

Licensed to: TEAM LND * Specified by user

HYPROTECH

TEAM LND
Calgary, Alberta
CANADA

Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
Unit Set	SI
Date/Time	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

COLUMN PROFILES

	Temperature (C)	Liquid Enthalpy (kJ/kgmole)	Vapour Enthalpy (kJ/kgmole)	Heat Loss (kJ/h)
9__Main TS	47.4[illegible]	[illegible]	[illegible]	
10__Main TS	48.0[illegible]	-1.20[illegible]	[illegible]	
11__Main TS	48.73	-1.208e+005	[illegible]	---
12__Main TS	49.47	-1.213e+005	-1.060e+005	---
13__Main TS	50.28	-1.218e+005	-1.062e+005	---
14__Main TS	51.16	-1.224e+005	-1.065e+005	---
15__Main TS	52.09	-1.229e+005	-1.068e+005	---
16__Main TS	53.07	-1.234e+005	-1.070e+005	---
17__Main TS	54.08	-1.23[illegible]e+005	-1.07[illegible]e+005	---
18__Main TS	55.10	-1.244e+005	-1.074e+005	
19__Main TS	56.11	-1.248e+005	-1.076e+005	
20__Main TS	56.33	-1.248e+005	-1.077e+005	
21__Main TS	56.50	-1.249e+005	-1.078e+005	---
22__Main TS	56.67	-1.250e+005	-1.079e+005	---
23__Main TS	56.91	-1.251e+005	-1.079e+005	---
24__Main TS	57.27	-1.252e+005	-1.080e+005	---
25__Main TS	57.82	-1.256e+005	-1.083e+005	---
26__Main TS	58.66	-1.261e+005	-1.08[illegible]e+005	---
27__Main TS	59.94	-1.269e+005	-1.091e+005	
28__Main TS	61.61	-1.280e+005	-1.09[illegible]e+005	---
29__Main TS	64.40	-1.294e+005	-1.111e+005	---
30__Main TS	67.75	-1.311e+005	-1.126e+005	---
31__Main TS	71.72	-1.330e+005	-1.145e+005	---
32__Main TS	75.97	-1.348e+005	-1.167e+005	---
33__Main TS	80.10	-1.363e+005	-1.186e+005	---
34__Main TS	83.7[illegible]	-1.376e+005	-1.207e+005	---
35__Main TS	86.83	-1.383e+005	-1.22[illegible]e+005	
36__Main TS	89.37	-1.38[illegible]e+005	-1.234e+005	
37__Main TS	91.19	-1.391e+005	-1.241e+005	---
38__Main TS	92.73	-1.392e+005	-1.244e+005	---
Reboiler	94.32	-1.390e+005	-1.245e+005	---

FEEDS / PRODUCTS

Flow Basis: Molar

	Stream	Type	Duty (kJ/h)	State	Flows (kgmole/h)	Enthalpy (kJ/kgmole)	Temp (C)
Condenser	7	Energy	6.748e+006 *				
	8	Draw	---	Liquid	156.0	-1.17[illegible]e+005	4[illegible] *
1__Main TS							
2__Main TS							
3__Main TS							
4__Main TS							
5__Main TS							
6__Main TS							
7__Main TS							
8__Main TS							
9__Main TS							
10__Main TS							
11__Main TS							
12__Main TS							
13__Main TS							
14__Main TS							

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 10 of 23

Licensed to: TEAM LND * Specified by user.

HYPROTECH	TEAM LnD Calgary, Alberta CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\[illegible]\DESKTOP\M
		Unit Set	SI
		Date/Time	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

FEEDS / PRODUCTS

15__Main TS							
16__Main TS							
17__Main TS							
18__Main TS							
19__Main TS	6	Feed	---	Liquid	279.3	-1.253e+005	65.60
20__Main TS							
21__Main TS							
22__Main TS							
23__Main TS							
24__Main TS							
25__Main TS							
26__Main TS							
27__Main TS							
28__Main TS							
29__Main TS							
30__Main TS							
31__Main TS							
32__Main TS							
33__Main TS							
34__Main TS							
35__Main TS							
36__Main TS							
37__Main TS							
38__Main TS							
Reboiler	9	Energy	6.545e+006		---	---	---
	10	Draw	---	Liquid	84.25	-1.390e+005	94.32

SETUP

Sub-Flowsheet

Feed Streams			Product Streams		
Internal Stream	External Stream	Transfer Basis	Internal Stream	External Stream	Transfer Basis
9	9	None Req'd	7	7	None Req'd
6	6	P-H Flash	8	[illegible]	P-H Flash
			10	10	P-H Flash

VARIABLES

Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
9		
6		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
7		
8		
10		

DYNAMICS

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 11 of 23

Licensed to: TEAM LND

HYPROTECH	TEAM LND Calgary, Alberta CANADA	
	Case Name	C:\DOCUMENTS AND SETTINGS\[illegible]\DESKTOP\M
	Unit Set	SI
	Date/Time	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

DYNAMICS

Vessel Dynamic Specifications

Vessel		Reboiler @COL1	Condenser @COL1
Diameter	(m)	1.193	1.193
Height 0	(m)	1.789	1.789
Volume 0	(m3)	2.000 *	2.000 *
Liquid Volume Percent	(%)	50.00 *	50.00 *
Level Calculator		Horizontal cylinder	Horizontal cylinder
Fraction Calculator		Use levels and nozzles	Use levels and nozzles
Vessel Delta P	(kPa)	[illegible]	[illegible]
Fixed Vessel P Spec	(kPa)	[illegible]	1440
Fixed P Spec Active		Not Active	Not Active

Other Equipment in Column Flowsheet

Holdup Details

	Pressure (kPa)	Volume (m3)	[illegible] (m3)
Condenser	0.0000	0.0000 *	---
1__Main TS	0.0000	0.0000 *	---
2__Main TS	0.0000	0.0000 *	---
3__Main TS	0.0000	0.0000 *	---
4__Main TS	0.0000	0.0000 *	---
5__Main TS	0.0000	0.0000 *	---
6__Main TS	0.0000	0.0000 *	---
7__Main TS	0.0000	0.0000 *	---
8__Main TS	0.0000	0.0000 *	---
9__Main TS	0.0000	0.0000 *	---
10__Main TS	0.0000	0.0000 *	---
11__Main TS	0.0000	0.0000 *	---
12__Main TS	0.0000	0.0000 *	---
13__Main TS	0.0000	0.0000 *	---
14__Main TS	0.0000	0.0000 *	---
15__Main TS	0.0000	0.0000 *	---
16__Main TS	0.0000	0.0000 *	---
17__Main TS	0.0000	0.0000 *	---
18__Main TS	0.0000	0.0000 *	---
19__Main TS	0.0000	0.0000 *	---
20__Main TS	0.0000	0.0000 *	---
21__Main TS	0.0000	0.0000 *	---
22__Main TS	0.0000	0.0000 *	---
23__Main TS	0.0000	0.0000 *	---
24__Main TS	0.0000	0.0000 *	---
25__Main TS	0.0000	0.0000 *	---
26__Main TS	0.0000	0.0000 *	---
27__Main TS	0.0000	0.0000 *	---
28__Main TS	0.0000	0.0000 *	---
29__Main TS	0.0000	0.0000 *	---
30__Main TS	0.0000	0.0000 *	---
31__Main TS	0.0000	0.0000 *	---
32__Main TS	0.0000	0.0000 *	---
33__Main TS	0.0000	0.0000 *	---

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 12 of 23

Licensed to: TEAM LND Specified by user

HYPROTECH	TEAM LNG Calgary, Alberta CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANOJIA\DESKTOP\M
		Unit Set	SI
		Date/Time	Thu Apr 23 02:14:21 2009

Distillation: T-100 (continued)

	Pressure (kPa)	Volume (m3)	[illegible] (%)
34__Main TS	0.0000	0.0000	---
35__Main TS	0.0000	0.0000	
36__Main TS	0.0000	0.0000	---
37__Main TS	0.0000	0.0000	---
38__Main TS	0.0000	0.0000	---
Reboiler	0.0000	0.0000	

NOTES

HYPROTECH TEAM LND Calgary, Alberta CANADA	Case Name: C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
	Unit Set: SI
	Date/Time: Thu Apr 23 02:14:21 2009

Tray Sizing: Tray Sizing-1

SETUP

Tray Section: Main TS @COL1 | Liquid Draw: 0.00 % * | Sieve Tray Flooding Method: Minimum Csb

	Section_1			
Section Start	1__Main TS			
Section End	38__Main TS			
Internals	Sieve			
Mode	Design			
Active	On			
Status	Complete			
Design Limit	Flooding			
Limiting Stage	36__Main TS			

SPECIFICATIONS

	Section_1		
Section Start	1__Main TS		
Section End	38__Main TS		
Internals	Sieve		
Mode	Design		
Number of Flow Paths	---		
Section Diameter (m)	---		
Tray for Properties	---		
Tray Spacing (mm)	609.6 *		
Tray Thickness (mm)	5.000 *		
Foaming Factor	1.000 *		
Max Delta P (ht of liq)	203.2 mm *		
Max Flooding (%)	85.00 *		
Packing Correlation	---		
HETP (m)	---		
Packing Type			

TRAY INTERNALS

	Section_1			
Section Start	1__Main TS			
Section End	38__Main TS			
Internals	Sieve			
Sieve Hole Pitch (mm)	12.70 *			
Sieve Hole Diameter (mm)	5.000 *			
Valve Matl Density (kg/m3)	---			
Valve Matl Thickness (mm)	---			
Hole Area (% of AA) (%)	---			
Valve Orifice Type	---			
Valve Design Manual	---			
Bubble Cap Slot Height (mm)	---			
Side Weir Type	Straight			
Weir Height (mm)	50.80 *			
Max Weir Loading (m3/h-m)	89.42 *			
Downcomer Type	Vertical			
Downcomer Clearance (mm)	45.80 *			
Max DC Backup (%)	50.00 *			
Side DC Top Width (mm)	---			
Side DC Bottom Width (mm)	---			
Centre DC Top Width (mm)	---			

Hyprotech Ltd. | HYSYS v3.1.3 (Build 4827) | Page 14 of 23

Licensed to: TEAM LND

* Specified by user

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name: C:\DOCUMENTS AND SETTINGS\MAYANK KANOJIA\DESKTOP\
		Unit Set: SI
		Date/Time: Thu Apr 23 02:14:21 2009

Tray Sizing: Tray Sizing-1 (continued)

Centre DC Bottom Width	(mm)	---
O.C. DC Top Width	(mm)	---
O.C. DC Bottom Width	(mm)	---
O.S. DC Top Width	(mm)	---
O.S. DC Bottom Width	(mm)	---

TRAY RESULTS

		Section_1
Section Start		1__Main TS
Section End		33__Main TS
Internals		Sieve
Section Diameter	(m)	1.524
Max Flooding	(%)	72.05
X-Sectional Area	(m2)	1.824
Section Height	(m)	23.16
Section DeltaP	(kPa)	9.625
NFP		1
Flow Length	(mm)	925.5
Flow Width	(mm)	1446
Max DC Backup	(%)	26.39
Max Weir Load	(m3/h-m)	81.56
Max DP/Tray	(kPa)	0.2480
Tray Spacing	(mm)	609.6 *
Total Weir Length	(mm)	1281
Weir Height	(mm)	50.80 *
Active Area	(m2)	1.194
DC Clearance	(mm)	45.80 *
DC Area	(m2)	0.3153
Side Weir Length	(m)	1.281
Hole Area	(m2)	0.1678
Estimated # of Holes/Valves		8545
Relief Area	(m2)	0.0000
Relief - S	(mm)	---
Relief - A	(mm)	---
Relief - B	(mm)	---
Side DC Top Width	(mm)	349.3
Side DC Btm Width	(mm)	349.3
Side DC Top Length	(m)	1.281
Side DC Btm Length	(m)	1.281
Side DC Top Area	(m2)	0.3153
Side DC Btm Area	(m2)	0.3153
Centre DC Top Width	(mm)	0.0000
Centre DC Btm Width	(mm)	0.0000
Centre DC Top Length	(m)	0.0000
Centre DC Btm Length	(m)	0.0000
Centre DC Top Area	(m2)	0.0000
Centre DC Btm Area	(m2)	0.0000
O.C. DC Top Width	(mm)	0.0000
O.C. DC Btm Width	(mm)	0.0000
O.C. DC Top Length	(m)	0.0000
O.C. DC Btm Length	(m)	0.0000
O.C. DC Top Area	(m2)	0.0000
O.C. DC Btm Area	(m2)	0.0000
O.S. DC Top Width	(mm)	0.0000

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 15 of 23

Licensed to: TEAM LND * Specified by user

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set:	SI
		Date/Time:	Thu Apr 23 02:14:21 2009

Tray Sizing: Tray Sizing-1 (continued)

O.S. DC Btm Width	(mm)	0.0000			
O.S. DC Top Length	(m)	0.0000			
O.S. DC Btm Length	(m)	0.0000			
O.S. DC Top Area	(m2)	0.0000			
O.S. DC Btm Area	(m2)	0.0000			

PACKED RESULTS

		Section_1		
Section Start		1__Main TS		
Section End		38__Main TS		
Internals		Sieve		
Section Diameter	(m)	1.524		
Max Flooding	(%)	72.05		
X-Sectional Area	(m2)	1.824		
Section Height	(m)	23.16		
Section DeltaP	(kPa)	[illegible]		
DP per Length	(kPa/m)	---		
Flood Gas Velocity	(m3/h-m2)	---		
Flood Gas Velocity	(m/s)	---		
Estimated # Pieces of Packing		---		
Estimated Mass of Packing	(kg)	---		
Estimated Packing Cost (US$)		---		
HETP	(m)	---		
HETP Correlation		---		
Packing Correlation		---		
Packing Type				

TRAY PRESSURE DROPS

Section_1

	Delta P (kPa)	Delta P (ht of liq) (mm)	Dry Delta P (ht of liq) (mm)
1__Main TS	0.2092	46.41	5.939
2__Main TS	0.2092	46.41	5.938
3__Main TS	0.2093	46.41	5.933
4__Main TS	0.2093	46.41	5.925
5__Main TS	0.2094	46.40	5.916
6__Main TS	0.2095	46.39	5.909
7__Main TS	0.2095	46.38	5.894
8__Main TS	0.2096	46.36	5.881
9__Main TS	0.2098	[illegible]	5.866
10__Main TS	0.2099	[illegible]	5.848
11__Main TS	0.2100	[illegible]	5.829
12__Main TS	0.2102	46.28	5.808
13__Main TS	0.2103	46.25	5.784
14__Main TS	0.2105	46.22	5.758
15__Main TS	0.2106	46.19	5.731
16__Main TS	0.2108	46.15	5.702
17__Main TS	0.2109	46.11	5.672
18__Main TS	0.2111	[illegible]	5.641
19__Main TS	0.2393	[illegible]	[illegible]
20__Main TS	0.2393	52.18	5.004
21__Main TS	0.2394	52.20	5.009
22__Main TS	0.2394	52.20	5.011

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 16 of 23

Licensed to: TEAM LND * Specified by user.

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set:	SI
		Date/Time:	Thu Apr 23 02:14:21 [illegible]

Tray Sizing: Tray Sizing-1 (continued)

Section_1

23__Main TS	0.2394	[illegible]	5.012
24__Main TS	[illegible]	[illegible]	5.013
25__Main TS	[illegible]	[illegible]	[illegible]
26__Main TS	[illegible]	[illegible]	[illegible]
27__Main TS	0.2401	[illegible]	5.027
28__Main TS	0.2405	52.32	5.048
29__Main TS	0.2411	52.41	5.093
30__Main TS	0.2419	[illegible]	5.175
31__Main TS	0.2428	[illegible]	5.306
32__Main TS	0.2438	53.12	5.486
33__Main TS	0.2449	[illegible]	5.697
34__Main TS	0.2458	[illegible]	5.912
35__Main TS	0.2466	[illegible]	[illegible]
36__Main TS	0.2473	[illegible]	6.258
37__Main TS	0.2477	54.56	6.369
38__Main TS	0.2480	54.65	6.434

DOWNCOMER RESULTS

Section_1

	DC Backup (%)	DC Backup (ht of liq) (mm)	DC Head Loss (mm)	DC Res Time (seconds)	[illegible] Res Time (seconds)
1__Main TS	20.34	124.0	3.119	[illegible]	---
2__Main TS	20.35	124.0	3.127	[illegible]	---
3__Main TS	20.35	124.0	3.128	4.662	---
4__Main TS	20.35	124.0	3.127	[illegible]	---
5__Main TS	20.34	124.0	3.124	4.664	---
6__Main TS	20.34	124.0	3.120	[illegible]	---
7__Main TS	20.34	124.0	3.115	[illegible]	---
8__Main TS	20.33	[illegible]	[illegible]	[illegible]	---
9__Main TS	20.32	[illegible]	3.102	[illegible]	---
10__Main TS	20.31	123.8	3.093	[illegible]	---
11__Main TS	20.30	123.8	3.082	4.686	---
12__Main TS	20.29	123.7	3.068	4.694	---
13__Main TS	20.28	123.6	3.053	4.702	---
14__Main TS	20.26	123.5	3.036	4.712	---
15__Main TS	20.25	123.4	3.016	[illegible]	---
16__Main TS	20.23	123.3	2.996	[illegible]	---
17__Main TS	20.21	123.2	[illegible]	[illegible]	---
18__Main TS	20.19	123.1	2.947	[illegible]	---
19__Main TS	24.48	149.3	10.48	[illegible]	---
20__Main TS	24.50	149.4	10.52	3.061	---
21__Main TS	24.51	149.4	10.54	3.059	---
22__Main TS	24.52	149.4	10.55	[illegible]	---
23__Main TS	24.52	149.5	10.57	[illegible]	---
24__Main TS	24.53	149.5	10.58	[illegible]	[illegible]
25__Main TS	24.54	149.6	[illegible]	[illegible]	[illegible]
26__Main TS	24.55	[illegible]	[illegible]	[illegible]	[illegible]
27__Main TS	24.58	[illegible]	[illegible]	[illegible]	[illegible]
28__Main TS	24.62	150.1	10.80	[illegible]	---
29__Main TS	24.70	150.6	10.97	3.021	---
30__Main TS	24.83	151.3	11.26	2.997	---

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 17 of 23

Licensed to: TEAM LND

* Specified by user

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set:	SI
		Date/Time:	Thu Apr 23 02:14:[illegible]

Tray Sizing: Tray Sizing-1 (continued)

Section_1

31__Main TS	25.61	152.6	13.69	2.444	---
32__Main TS	25.25	153.9	12.24	[illegible]	---
33__Main TS	25.52	155.0	[illegible]	[illegible]	---
34__Main TS	[illegible]	[illegible]	[illegible]	[illegible]	---
35__Main TS	26.02	158.6	14.05	2.012	---
36__Main TS	26.20	159.7	14.46	2.761	---
37__Main TS	26.32	160.5	14.79	2.771	---
38__Main TS	26.39	160.9	14.96	2.765	---

	DC Velocity (m3/h-m2)	DC Velocity (m/s)	DC Design Vel (m3/h-m2)	DC Design Vel (m/s)	DC Load Factor
1__Main TS	95.65	2.657e-002	463.0	[illegible]	2.570
2__Main TS	95.78	2.661e-002	[illegible]	[illegible]	2.574
3__Main TS	95.60	2.661e-002	[illegible]	[illegible]	2.573
4__Main TS	95.77	2.660e-002	463.0	0.1286	2.574
5__Main TS	95.73	2.659e-002	463.1	0.1287	2.575
6__Main TS	95.67	2.658e-002	463.3	0.1287	2.576
7__Main TS	95.60	2.656e-002	463.4	0.1287	2.576
8__Main TS	95.51	2.653e-002	463.6	0.1288	2.580
9__Main TS	95.39	2.650e-002	463.8	0.1288	2.583
10__Main TS	95.25	2.646e-002	464.0	[illegible]	2.586
11__Main TS	95.08	2.641e-002	464.2	[illegible]	2.589
12__Main TS	94.88	2.636e-002	[illegible]	0.129[illegible]	2.594
13__Main TS	94.64	2.629e-002	464.8	0.1291	2.599
14__Main TS	94.37	2.621e-002	465.2	0.1292	2.604
15__Main TS	94.07	2.613e-002	465.5	0.1293	2.610
16__Main TS	93.73	2.604e-002	465.9	0.1294	2.617
17__Main TS	93.37	2.594e-002	466.2	0.1295	2.624
18__Main TS	92.99	2.583e-002	466.6	0.1296	2.632
19__Main TS	175.4	4.871e-002	467.0	[illegible]	1.800
20__Main TS	175.7	4.880e-002	466.9	[illegible]	1.798
21__Main TS	175.9	4.884e-002	466.9	[illegible]	1.796
22__Main TS	176.0	4.888e-002	466.8	0.1297	1.796
23__Main TS	176.1	4.891e-002	466.8	0.1297	1.795
24__Main TS	176.2	4.894e-002	466.8	0.1297	1.794
25__Main TS	176.4	4.899e-002	466.8	0.1297	1.793
26__Main TS	176.8	4.907e-002	466.8	0.1297	1.792
27__Main TS	177.1	4.920e-002	466.9	0.1297	1.789
28__Main TS	178.0	4.944e-002	466.9	[illegible]	1.784
29__Main TS	179.4	4.984e-002	[illegible]	[illegible]	1.775
30__Main TS	181.8	5.050e-002	466.6	[illegible]	1.760
31__Main TS	185.2	5.144e-002	466.0	0.1294	1.740
32__Main TS	189.5	5.264e-002	465.2	0.1292	1.714
33__Main TS	194.3	5.398e-002	464.1	0.1289	1.686
34__Main TS	199.0	5.527e-002	463.1	0.1286	1.660
35__Main TS	203.0	5.639e-002	462.2	0.1284	1.638
36__Main TS	206.1	5.725e-002	461.5	0.1282	1.622
37__Main TS	208.3	5.787e-002	461.1	[illegible]	1.611
38__Main TS	209.5	5.819e-002	460.9	[illegible]	1.605

TRAY FLOODING RESULTS

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name:	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\...
		Unit Set:	SI
		Date/Time:	Thu Apr 23 02:14:21 [illegible]

Tray Sizing: Tray Sizing-1 (continued)

Section_1

	Flooding (%)	Vapour Load (ACT_m3/h)	Weir Load (m3/h-m)	Liq Ht over Weir (mm)
1__Main TS	[illegible]	[illegible]	[illegible]	[illegible]
2__Main TS	[illegible]	[illegible]	[illegible]	[illegible]
3__Main TS	52.77	186.1	[illegible]	[illegible]
4__Main TS	52.75	186.0	[illegible]	23.73
5__Main TS	52.73	185.8	[illegible]	23.70
6__Main TS	52.70	[illegible]	[illegible]	23.69
7__Main TS	52.66	185.5	[illegible]	23.68
8__Main TS	52.61	185.3	[illegible]	[illegible]
9__Main TS	52.55	[illegible]	[illegible]	23.64
10__Main TS	52.48	[illegible]	[illegible]	[illegible]
11__Main TS	52.40	184.5	[illegible]	[illegible]
12__Main TS	52.30	184.1	23.35	23.56
13__Main TS	52.19	183.8	23.30	23.52
14__Main TS	52.06	183.4	23.23	23.47
15__Main TS	51.92	182.9	23.15	23.42
16__Main TS	51.76	182.4	[illegible]	23.36
17__Main TS	51.59	182.0	[illegible]	23.30
18__Main TS	51.42	181.5	[illegible]	23.24
19__Main TS	59.66	[illegible]	[illegible]	35.81
20__Main TS	59.76	170.9	43.24	35.85
21__Main TS	59.85	171.0	43.28	35.88
22__Main TS	59.89	171.0	43.31	35.89
23__Main TS	59.92	171.1	43.34	35.91
24__Main TS	59.96	171.1	43.37	35.93
25__Main TS	60.01	171.1	43.41	35.95
26__Main TS	60.09	171.2	[illegible]	35.99
27__Main TS	60.24	171.4	[illegible]	[illegible]
28__Main TS	60.52	171.8	[illegible]	36.17
29__Main TS	61.00	172.7	[illegible]	36.38
30__Main TS	61.81	174.2	44.75	36.70
31__Main TS	62.99	176.5	45.54	37.18
32__Main TS	64.54	179.7	46.65	37.77
33__Main TS	66.29	183.3	47.83	38.42
34__Main TS	68.01	186.9	[illegible]	39.06
35__Main TS	69.53	190.0	[illegible]	39.60
36__Main TS	70.72	192.5	[illegible]	40.02
37__Main TS	71.57	[illegible]	[illegible]	40.31
38__Main TS	72.05	195.3	[illegible]	40.46

	Flood Capacity	Entrainment (%)	Entrainment (kg/h)	Weep Velocity (m/s)
1__Main TS	0.4265	0.51	[illegible]	0.1232
2__Main TS	0.4265	0.51	[illegible]	0.1229
3__Main TS	0.4265	0.51	70.28	0.1227
4__Main TS	0.4264	0.51	[illegible]	0.1225
5__Main TS	0.4264	0.50	[illegible]	0.1223
6__Main TS	0.4264	0.50	[illegible]	0.1222
7__Main TS	0.4263	0.50	[illegible]	0.1220
8__Main TS	0.4263	0.50	[illegible]	0.1218
9__Main TS	0.4262	0.50	69.40	0.1216
10__Main TS	0.4262	0.50	69.14	0.1215

Hyprotech Ltd.	HYSYS v3.1.3 (Build 4827)	Page 19 of 23
Licensed to: TEAM LND		* Specified by user.

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set	SI
		Date/Time	[illegible]

Tray Sizing: Tray Sizing-1 (continued)

Section_1

11__Main TS	0.4262	[illegible]	[illegible]	[illegible]
12__Main TS	0.4261	[illegible]	[illegible]	[illegible]
13__Main TS	0.4261	[illegible]	[illegible]	0.1210
14__Main TS	0.4260	[illegible]	[illegible]	0.1210
15__Main TS	0.4260	0.49	67.05	0.1209
16__Main TS	0.4259	0.48	66.47	0.1209
17__Main TS	0.4259	0.48	65.86	0.1209
18__Main TS	0.4259	0.48	65.20	0.1210
19__Main TS	0.4258	0.29	[illegible]	0.1210
20__Main TS	0.4258	0.27	[illegible]	0.1207
21__Main TS	0.4257	0.27	[illegible]	0.1204
22__Main TS	0.4257	[illegible]	[illegible]	[illegible]
23__Main TS	0.4257	0.27	[illegible]	0.1200
24__Main TS	0.4256	0.27	69.10	0.1197
25__Main TS	0.4255	0.27	69.19	0.1193
26__Main TS	0.4254	0.27	69.36	0.1187
27__Main TS	0.4252	0.27	69.63	0.1178
28__Main TS	0.4249	0.27	70.31	0.1166
29__Main TS	0.4245	0.27	71.46	0.1147
30__Main TS	0.4239	[illegible]	[illegible]	[illegible]
31__Main TS	0.4232	[illegible]	[illegible]	[illegible]
32__Main TS	0.4225	0.29	[illegible]	0.1050
33__Main TS	0.4218	0.30	85.11	0.1028
34__Main TS	0.4212	0.31	90.02	9.990e-002
35__Main TS	0.4207	0.32	94.52	9.759e-002
36__Main TS	0.4204	0.33	98.20	9.602e-002
37__Main TS	0.4202	0.33	100.6	9.495e-002
38__Main TS	0.4201	0.34	[illegible]	[illegible]

LIQUID PROFILE (FROM TRAY)

	1__Main TS	2__Main TS	3__Main TS	4__Main TS
Mass Flow (kg/h)	1.386e+004	1.388e+004	1.389e+004	1.389e+004
Liquid Flow (m3/s)	8.378e-003	8.390e-003	8.391e-003	8.395e-003
Molecular Weight	44.15	44.25	44.34	44.45
Temperature (C)	44.53	44.89	45.15	45.42
Density (kg/m3)	459.7	459.7	459.8	460.0
Viscosity (cP)	8.269e-002	8.277e-002	[illegible]	[illegible]
Surface Tension (dyne/cm)	4.685	[illegible]	[illegible]	4.660
	5__Main TS	6__Main TS	7__Main TS	8__Main TS
Mass Flow (kg/h)	1.389e+004	1.389e+004	1.389e+004	1.389e+004
Liquid Flow (m3/s)	8.385e-003	8.380e-003	8.373e-003	8.365e-003
Molecular Weight	44.57	44.71	44.87	45.05
Temperature (C)	45.72	46.07	46.47	46.93
Density (kg/m3)	460.2	460.4	460.7	461.1
Viscosity (cP)	8.331e-002	8.356e-002	8.384e-002	8.416e-002
Surface Tension (dyne/cm)	4.658	4.652	4.648	4.644
	9__Main TS	10__Main TS	11__Main TS	12__Main TS
Mass Flow (kg/h)	1.388e+004	1.387e+004	1.387e+004	1.385e+004
Liquid Flow (m3/s)	8.355e-003	8.343e-003	8.328e-003	8.310e-003
Molecular Weight	45.28	45.49	45.75	46.02
Temperature (C)	47.46	48.06	48.73	49.47
Density (kg/m3)	461.5	462.0	462.5	463.0

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 20 of 23

Licensed to: TEAM LND

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DEPKTOPM
		Unit Set	SI
		Date/Time	Thu Apr 23 02:14:21 2009

Tray Sizing: Tray Sizing-1 (continued)

Viscosity (cP)	8.451e-002	8.490e-002	[illegible]	8.572e-002
Surface Tension (dyne/cm)	4.640	4.637	[illegible]	4.637
	13__Main TS	14__Main TS	[illegible]__Main TS	[illegible]__Main TS
Mass Flow (kg/h)	1.334e+004	[illegible]	[illegible]	[illegible]
Liquid Flow (m3/s)	8.290e-003	8.296e-003	[illegible]	9.210e-003
Molecular Weight	45.32	45.63	46.94	47.27
Temperature (C)	50.28	51.16	52.09	53.07
Density (kg/m3)	463.7	464.3	[illegible]	465.7
Viscosity (cP)	8.615e-002	8.657e-002	[illegible]	8.724e-002
Surface Tension (dyne/cm)	4.632	4.634	[illegible]	4.641
	17__Main TS	18__Main TS	[illegible]__Main TS	20__Main TS
Mass Flow (kg/h)	1.373e+004	1.370e+004	[illegible]	[illegible]
Liquid Flow (m3/s)	8.178e-003	8.145e-003	[illegible]	1.50[illegible]e-002
Molecular Weight	47.58	47.89	48.16	48.23
Temperature (C)	54.08	55.10	56.11	56.33
Density (kg/m3)	466.4	467.1	467.6	467.7
Viscosity (cP)	8.767e-002	8.797e-002	8.622e-002	8.810e-002
Surface Tension (dyne/cm)	4.647	4.655	4.664	4.650
	21__Main TS	22__Main TS	23__Main TS	24__Main TS
Mass Flow (kg/h)	[illegible]	2.584e+004	[illegible]	[illegible]
Liquid Flow (m3/s)	1.540e-002	1.541e-002	[illegible]	1.54[illegible]e-002
Molecular Weight	48.20	48.31	[illegible]	48.50
Temperature (C)	56.56	56.67	[illegible]	57.27
Density (kg/m3)	467.6	467.6	467.6	467.7
Viscosity (cP)	8.614e-002	8.615e-002	[illegible]	8.812e-002
Surface Tension (dyne/cm)	4.642	4.637	4.631	4.625
	25__Main TS	26__Main TS	27__Main TS	28__Main TS
Mass Flow (kg/h)	2.602e+004	2.607e+004	2.616e+004	2.631e+004
Liquid Flow (m3/s)	1.541e-002	1.547e-002	[illegible]	[illegible]
Molecular Weight	48.69	[illegible]	[illegible]	49.98
Temperature (C)	57.82	58.66	[illegible]	61.81
Density (kg/m3)	467.8	468.1	468.4	468.6
Viscosity (cP)	8.849e-002	8.877e-002	8.910e-002	8.967e-002
Surface Tension (dyne/cm)	4.617	4.605	4.587	4.560
	29__Main TS	30__Main TS	31__Main TS	32__Main TS
Mass Flow (kg/h)	2.654e+004	2.689e+004	2.736e+004	[illegible]
Liquid Flow (m3/s)	1.572e-002	1.592e-002	1.622e-002	1.660e-002
Molecular Weight	50.79	51.79	[illegible]	54.05
Temperature (C)	64.40	[illegible]	[illegible]	75.97
Density (kg/m3)	469.1	469.1	[illegible]	468.0
Viscosity (cP)	9.025e-002	9.078e-002	9.111e-002	9.117e-002
Surface Tension (dyne/cm)	4.519	4.459	4.376	4.284
	33__Main TS	34__Main TS	35__Main TS	36__Main TS
Mass Flow (kg/h)	2.860e+004	2.921e+004	[illegible]	[illegible]
Liquid Flow (m3/s)	1.702e-002	1.743e-002	1.778e-002	[illegible]
Molecular Weight	55.10	55.97	[illegible]	[illegible]
Temperature (C)	80.10	83.77	[illegible]	[illegible]
Density (kg/m3)	466.8	465.6	[illegible]	463.6
Viscosity (cP)	9.097e-002	9.061e-002	[illegible]	[illegible]
Surface Tension (dyne/cm)	4.184	4.000	4.011	3.951
	37__Main TS	38__Main TS		
Mass Flow (kg/h)	3.041e+004	3.056e+004		
Liquid Flow (m3/s)	1.825e-002	1.835e-002		
Molecular Weight	57.50	57.75		

HYPROTECH	TEAM LND Calgary Alberta CANADA	Case Name	[illegible]
		Unit Set	SI
		Date/Time	Thu Apr 23 02 14 21 2009

Tray Sizing: Tray Sizing-1 (continued)

Temperature (C)	51 16	[illegible]		
Density (kg/m3)	463 5	[illegible]		
Viscosity (cP)	8 935e-002	[illegible]		
Surface Tension (dyne/cm)	3 908	[illegible]		

VAPOUR PROFILE (TO TRAY)

	1__Main TS	2__Main TS	3__Main TS	4__Main TS
Mass Flow (kg/h)	2 244e+004	2 246e+004	2 247e+004	2 247e+004
Gas Flow (ACT_m3/h)	656 6	656 4	[illegible]	655 2
Molecular Weight	44 09	44 16	[illegible]	44 27
Temperature (C)	44 63	[illegible]	[illegible]	45 72
Density (kg/m3)	34 18	34 22	[illegible]	34 29
Viscosity (cP)	9 626e-003	9 631e-003	9 636e-003	9 643e-003
Fluid Pressure (kPa)	1530	1531	1532	1532
	5__Main TS	6__Main TS	7__Main TS	8__Main TS
Mass Flow (kg/h)	2 247e+004	2 247e+004	2 247e+004	2 247e+004
Gas Flow (ACT_m3/h)	654 5	653 6	652 9	652 0
Molecular Weight	44 34	44 43	44 53	44 64
Temperature (C)	46 07	46 47	[illegible]	47 49
Density (kg/m3)	34 33	34 37	[illegible]	34 46
Viscosity (cP)	9 651e-003	9 659e-003	9 669e-003	9 679e-003
Fluid Pressure (kPa)	1533	1534	1535	1536
	9__Main TS	10__Main TS	11__Main TS	12__Main TS
Mass Flow (kg/h)	2 246e+004	2 245e+004	2 245e+004	2 243e+004
Gas Flow (ACT_m3/h)	651 0	650 0	[illegible]	647 5
Molecular Weight	44 77	44 91	[illegible]	45 22
Temperature (C)	48 06	48 73	[illegible]	50 28
Density (kg/m3)	34 56	34 5?	[illegible]	34 64
Viscosity (cP)	9 691e-003	[illegible]	[illegible]	[illegible]
Fluid Pressure (kPa)	1536	1537	1538	1539
	13__Main TS	14__Main TS	15__Main TS	16__Main TS
Mass Flow (kg/h)	2 242e+004	2 240e+004	2 237e+004	2 234e+004
Gas Flow (ACT_m3/h)	646 2	644 8	[illegible]	641 5
Molecular Weight	45 40	45 56	[illegible]	45 95
Temperature (C)	51 16	52 09	[illegible]	54 08
Density (kg/m3)	34 69	34 73	[illegible]	34 8?
Viscosity (cP)	9 751e-003	9 766e-003	[illegible]	[illegible]
Fluid Pressure (kPa)	1540	1541	1541	1542
	17__Main TS	18__Main TS	19__Main TS	20__Main TS
Mass Flow (kg/h)	2 231e+004	2 228e+004	2 099e+004	2 103e+004
Gas Flow (ACT_m3/h)	640 4	638 9	600 8	601 5
Molecular Weight	46 13	46 31	46 38	46 42
Temperature (C)	55 10	56 11	[illegible]	56 50
Density (kg/m3)	34 84	34 87	34 94	34 98
Viscosity (cP)	9 821e-003	9 836e-003	[illegible]	[illegible]
Fluid Pressure (kPa)	1543	1544	[illegible]	1545
	21__Main TS	22__Main TS	23__Main TS	24__Main TS
Mass Flow (kg/h)	2 105e+004	2 106e+004	2 105e+004	2 110e+004
Gas Flow (ACT_m3/h)	601 0	600 8	600 4	599 9
Molecular Weight	46 47	46 52	46 61	46 74
Temperature (C)	56 67	56 93	[illegible]	[illegible]
Density (kg/m3)	35 02	35 06	[illegible]	35 18
Viscosity (cP)	9 846e-003	9 851e-003	[illegible]	[illegible]
Fluid Pressure (kPa)	1546	1547	[illegible]	1549

Hyprotech Ltd. HYSYS v3.1.3 (Build 4827) Page 22 of 23

Licensed to: TEAM LND Specified by user.

HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name	C:\DOCUMENTS AND SETTINGS\MAYANK KANODIA\DESKTOP\M
		Unit Set	SI
		Date/Time	Thu Apr 23 02:14:21 2009

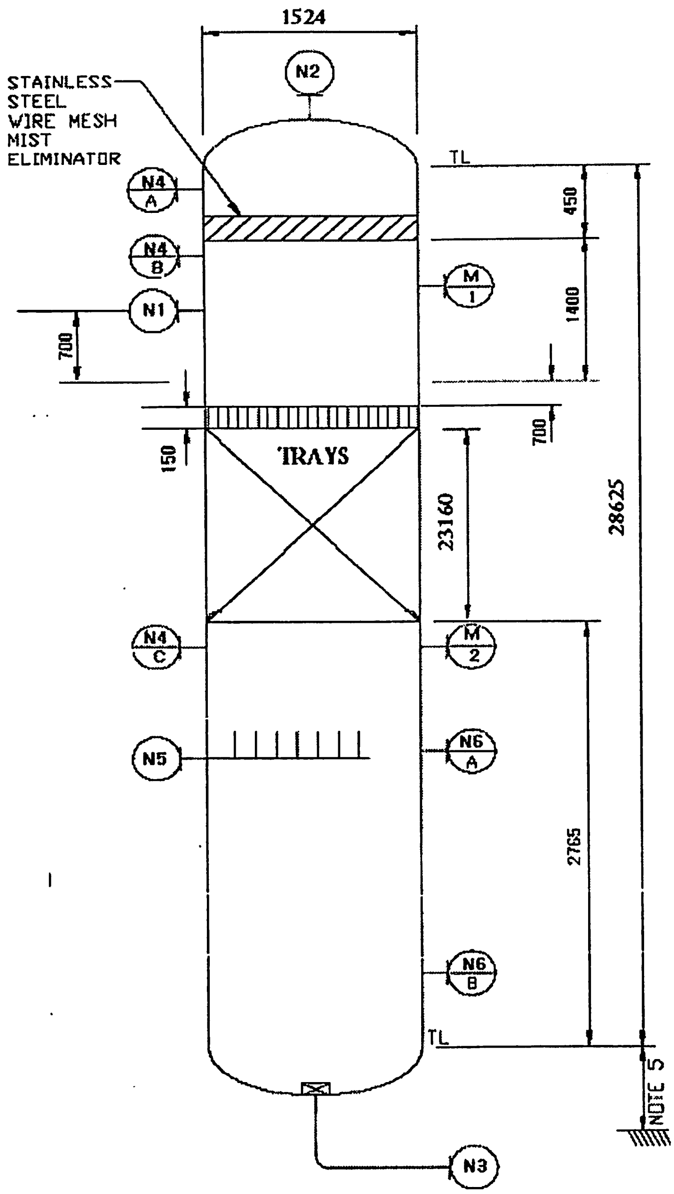
Tray Sizing: Tray Sizing-1 (continued)

	25__Main TS	2[illegible]__Main TS	2[illegible]__Main TS	2[illegible]__Main TS
Mass Flow (kg/h)	2.114e+004	2.119e+004	[illegible]	2.14[illegible]e+004
Gas Flow (ACT_m3/h)	599.3	598.5	[illegible]	596.4
Molecular Weight	46.95	47.27	47.7[illegible]	48.46
Temperature (C)	58.66	59.94	[illegible]	64.40
Density (kg/m3)	35.27	35.41	[illegible]	35.92
Viscosity (cP)	9.882e-003	9.904e-003	[illegible]	[illegible]
Fluid Pressure (kPa)	1549	1550	[illegible]	1552
	29__Main TS	30__Main TS	31__Main TS	3[illegible]__Main TS
Mass Flow (kg/h)	2.1[illegible]e+004	2.201e+004	[illegible]	[illegible]
Gas Flow (ACT_m3/h)	595.6	595.7	[illegible]	600.7
Molecular Weight	49.41	50.60	51.94	53.30
Temperature (C)	67.75	71.72	7[illegible]	80.10
Density (kg/m3)	36.36	36.95	[illegible]	38.43
Viscosity (cP)	1.002e-002	1.006e-002	1.010e-002	1.015e-002
Fluid Pressure (kPa)	1553	1554	1554	1555
	33__Main TS	34__Main TS	35__Main TS	3[illegible]__Main TS
Mass Flow (kg/h)	2.372e+004	2.438e+004	[illegible]	[illegible]
Gas Flow (ACT_m3/h)	606.6	611.2	[illegible]	620.6
Molecular Weight	54.55	55.56	[illegible]	57.08
Temperature (C)	83.77	86.83	89.27	91.16
Density (kg/m3)	39.17	39.81	40.31	40.67
Viscosity (cP)	1.019e-002	1.021e-002	1.023e-002	1.025e-002
Fluid Pressure (kPa)	1556	1557	1558	1558
	37__Main TS	38__Main TS		
Mass Flow (kg/h)	2.553e+004	2.566e+004		
Gas Flow (ACT_m3/h)	624.0	626.4		
Molecular Weight	57.42	57.72		
Temperature (C)	92.73	94.15		
Density (kg/m3)	40.91	40.99		
Viscosity (cP)	1.026e-002	1.027e-002		
Fluid Pressure (kPa)	1559	1560		

Hyprotech Ltd. | HYSYS v3.1.3 (Build 4827) | Page 23 of 23

Licensed to: TEAM LND | * Specified by user

3.4 LAYOUT OF THE DESIGNED COLUMN



Note: All the dimensions are in mm.

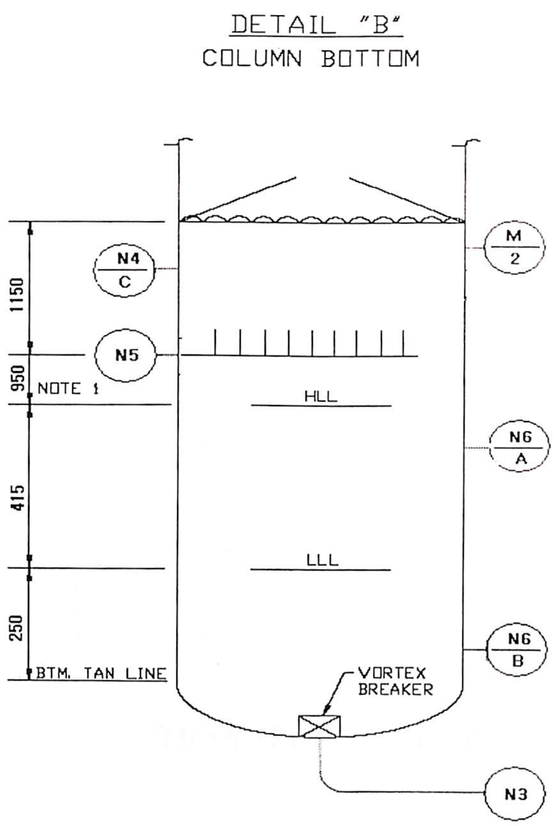
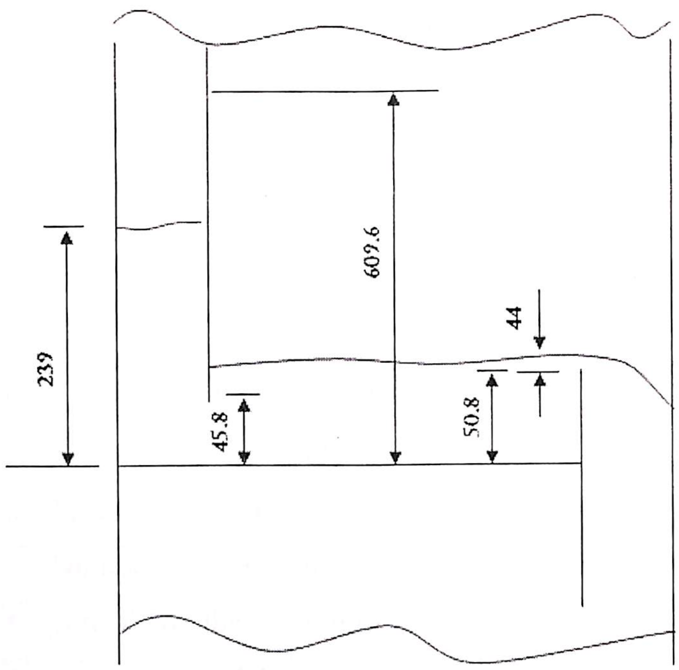
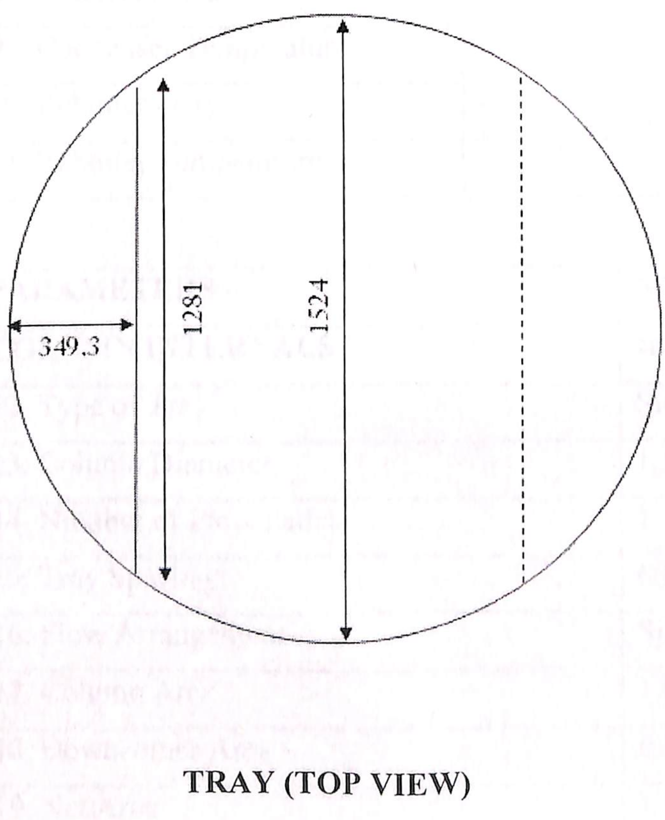


Table 2: Nozzles and Connections Details

Mark	No.	Size (in) / Rating	Service
M1	1	24 / #300	Manhole
M2	1	24 / #300	Manhole
N1	1	4 / #300	Feed
N2	1	8 / #300	Overhead
N3	1	10 / #300	To reboiler
N4A/N4B/N4C	3	2 / #300	DP guage
N5	1	12 / #300	Reboiler outlet
N6A/N6B	2	3 / #300	Level Transmitter



TRAY (FRONT VIEW)



TRAY (TOP VIEW)

RESULTS

Important Results are summed up as under:

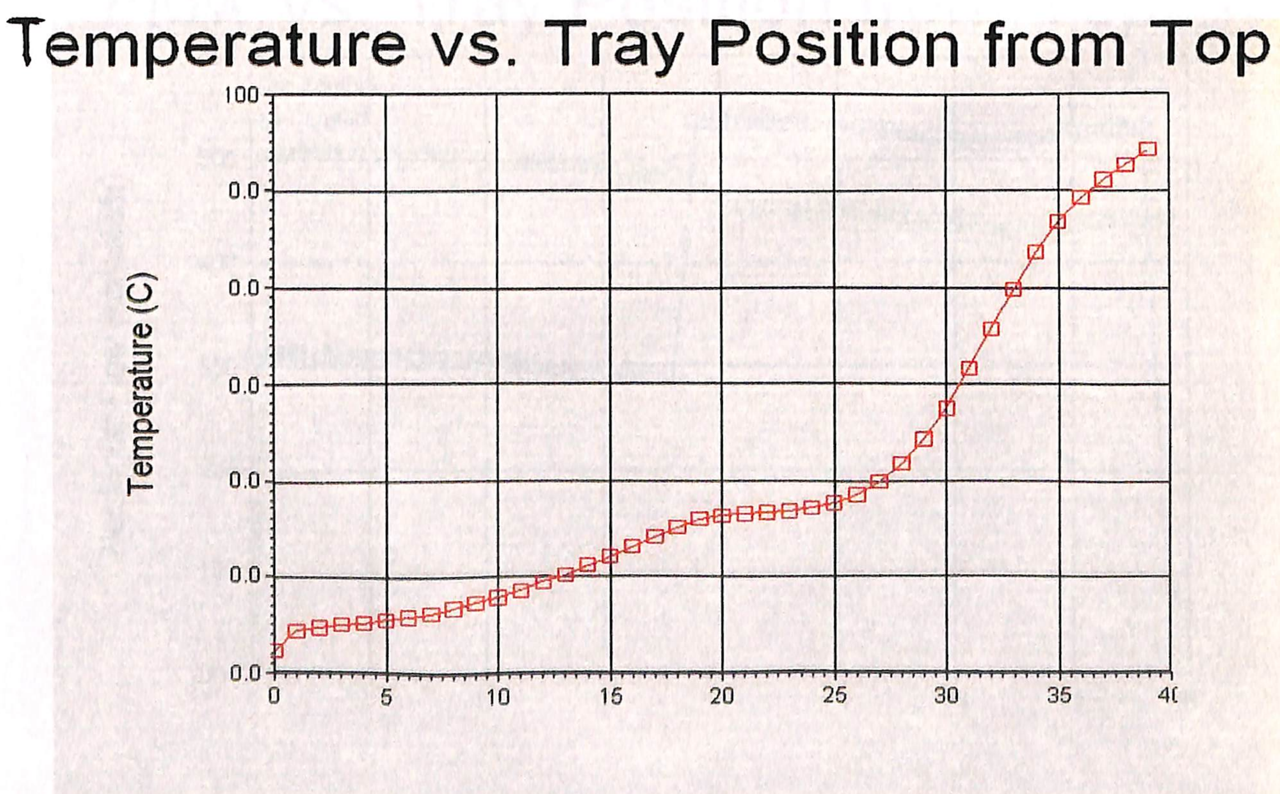
Table 3: Important Results of Column Design

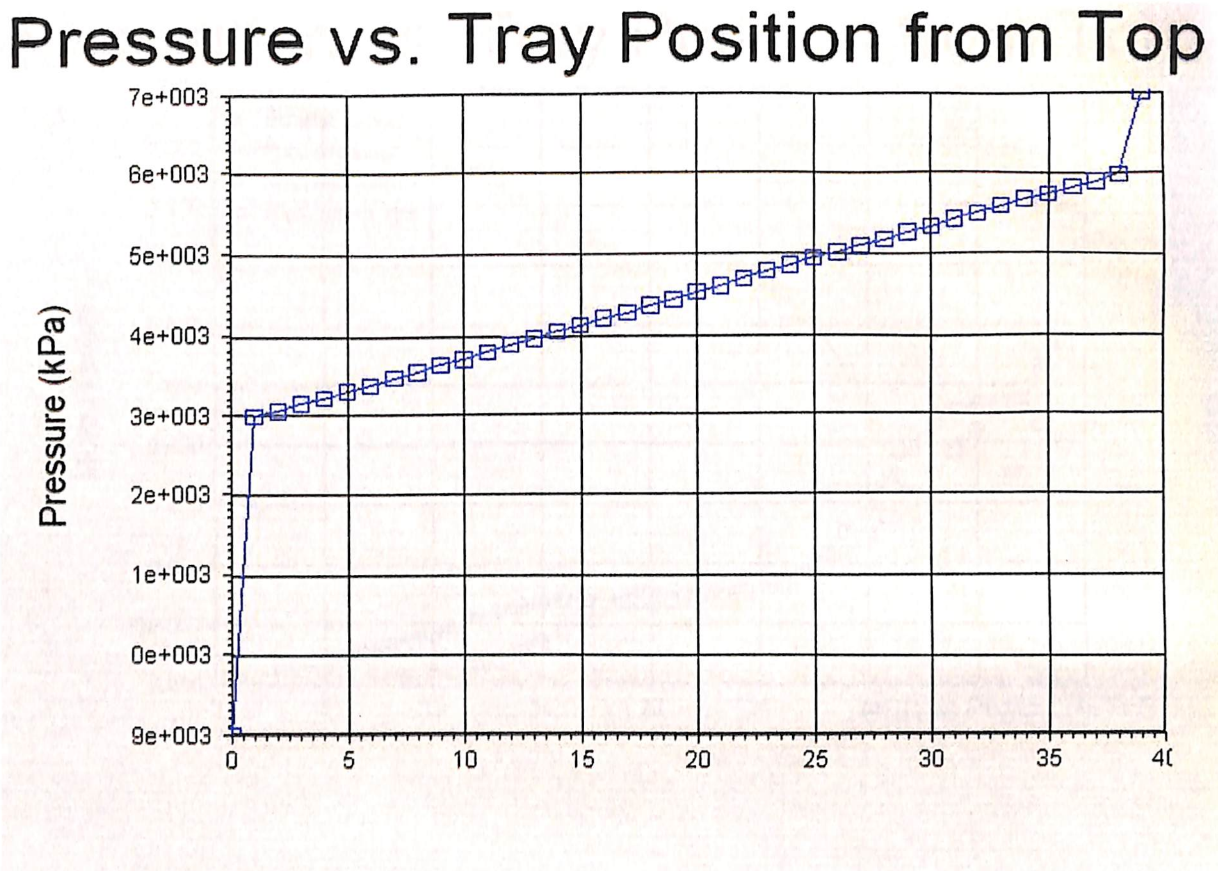
PARAMETERS	THEORETICAL VALUE	VALUE FROM ASPEN HYSYS
COLUMN PARAMETERS		
1. Minimum Number of Stages	13	15
2. Minimum Reflux	1.53	1.53
3. Number of Theoretical Stages	26	27
4. Actual Number of Stages	38	38
5. Stripping Feed Tray Location	19	19
6. Rectifying Feed Tray Location	19	19
7. Reflux Ratio	2.2935	2.293
8. Condenser Duty	-	$8.562 * 10^6$ KJ / hr
9. Condenser Temperature	-	42.53 °C
10. Reboiler Duty	-	$8.359 * 10^6$ KJ / hr
11. Reboiler Temperature	-	94.32 °C

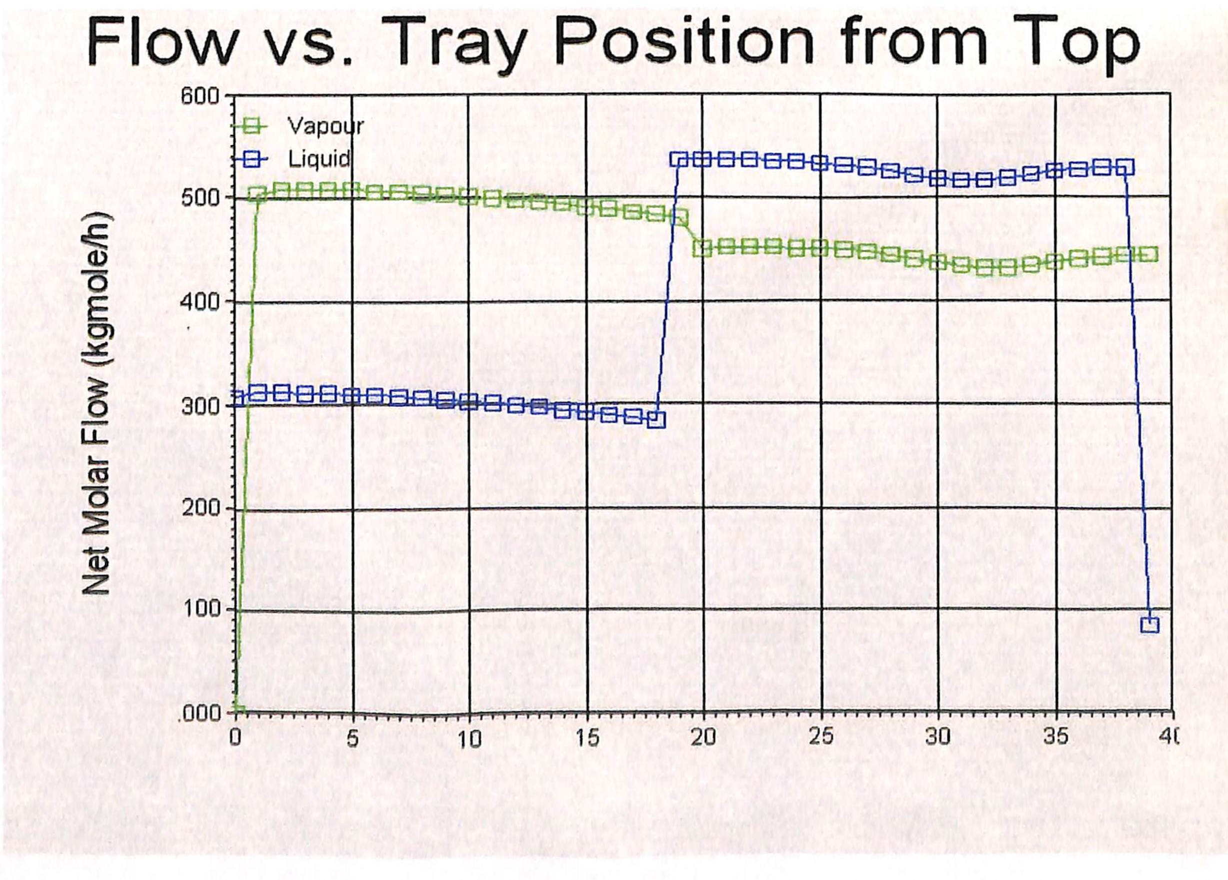
PARAMETERS	NOTE: The Theoretical Values are almost same as the Values taken from Aspen Hysys
COLUMN INTERNALS	
12. Type of Tray	Sieve Tray
13. Column Diameter	1.524 m
14. Number of Flow Paths	1
15. Tray Spacing	609.6 mm
16. Flow Arrangement	Single Pass Cross Flow
17. Column Area	1.824 m^2
18. Downcomer Area	0.3153 m^2
19. Net Area	1.5087 m^2
20. Active Area	1.194 m^2
21. Hole Area	0.1678 m^2

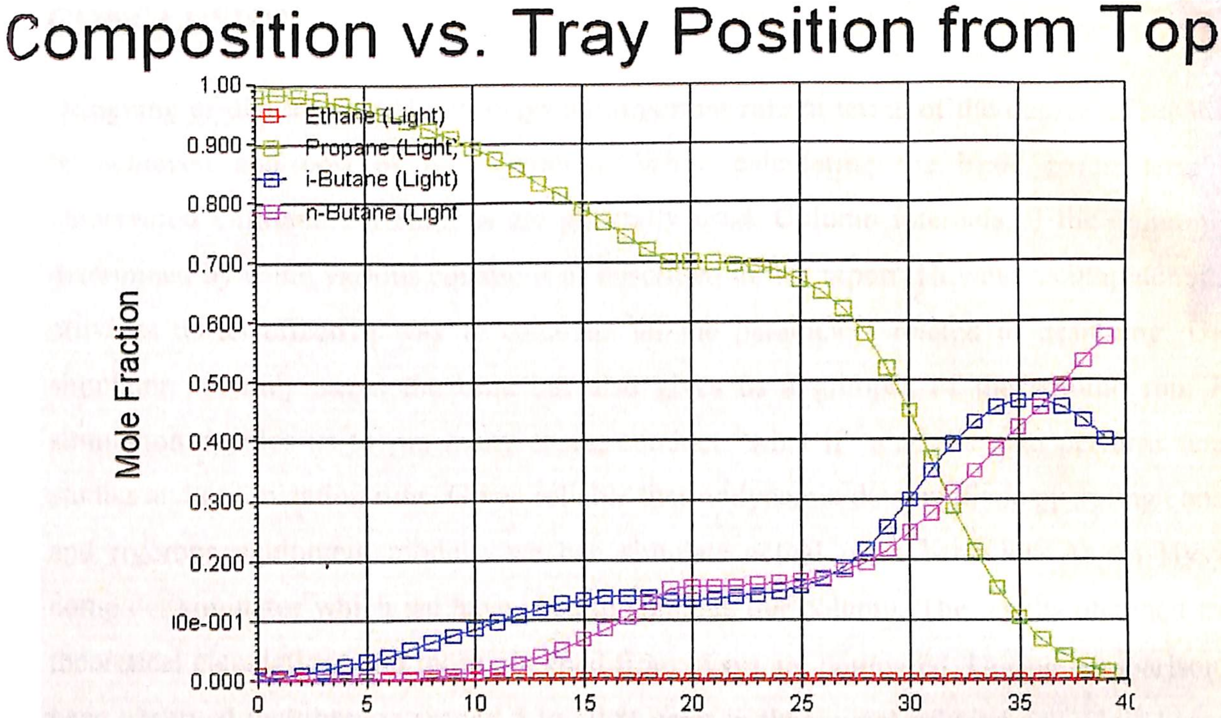
22. Weir Type	STRAIGHT
23. Weir Height	50.8 mm
24. Weir Length	1281 mm
25. Hole Diameter	5 mm
26. Maximum Weir Loading	89.42 m^3 / h – m
27. Hole Pitch	12.7 mm
28. Plate Thickness	5 mm
29. Total Plate Pressure Drop	580.8776 Pa
30. Downcomer Type	Vertical
31. Downcomer Clearance	45.8 mm
32. Downcomer Clearance Area	0.0587 m^2
33. Downcomer Backup	239 mm
34. Downcomer Residence Time	4.1068 sec
35. Number Of Holes	8545

GRAPHS




Pressure vs. Tray Position from Top
Pressure (kPa)
7e+003
6e+003
5e+003
4e+003
3e+003
2e+003
1e+003
0e+003
9e+003
0
5
10
15
20
25
30
35


Flow vs. Tray Position from Top
Vapour
Liquid
Net Molar Flow (kgmole/h)
600
500
400
300
200
100
.000
0
5
10
15
20
25
30
35


Composition vs. Tray Position from Top
Ethane (Light)
Propane (Light)
i-Butane (Light)
n-Butane (Light
Mole Fraction
1.00
0.900
0.800
0.700
0.600
0.500
0.400
0.300
0.200
l0e-001
0.000
0
5
10
15
20
25
30
35
40

CONCLUSION

Designing of distillation column plays an important role in terms of the degree of separation to be achieved and cost of the operation. While calculating the basic parameters, Fenske Underwood Gilliland correlations are generally used. Column internals of the column can be determined by using various equations as described in this report. However, computer simulator provides us an effective way to calculate all the parameters related to designing. Use of a simulator not only saves the time but also gives us a glimpse of the column run. Process simulation enables us to run many cases, conduct "what if" analysis, and perform sensitivity studies and optimization runs. Given reliable thermodynamic data, realistic operating conditions, and rigorous equipment models, we can simulate actual plant behavior. Aspen Hysys is a computer simulator which we have used to simulate our column. The results obtained from the theoretical calculations and those obtained from hysys are compared. During comparison, it has been observed that there is around 5 to 10 % error in theoretical calculations while keeping the hysys results as reference. Thus, designing of depropanizer column is done and simulation results are compared.

REFERENCES

1. Walas, Stanley M., *Chemical Process Equipment Selection and Design*, Butterworth – Heinemann series in Chemical Engineering, 1990.
2. Crowl, D.A., *Distillation Column – Design Case*, October, 2007.
3. Kister, Henry Z., *Distillation Design*, Mc Graw Hill Inc., 1992.
4. Kooijman, Hendrik Adriaan, *Dissertation on Dynamic Non equilibrium Column Simulation*, Clarkson University, February 1995.
5. Strigle, Ralph F., Jr., *Packed Tower Design and Application Random and Structured Packings*, Second Edition, Gulf Publishing Company, 1994.
6. Sinott, R.K., *Coulson and Richardson's, Chemical Engineering*, Third Edition, Volume 6, Butterworth – Heinemann series in Chemical Engineering, 2003.
7. *Aspen Hysys Reference Guide*, Hyprotech, 2002.
8. Perry, Robert H. with Green, Don W. and Maloney, James O., *Perry's Chemical Engineers Handbook*, Seventh Edition, Mc Graw Hill Inc., 1997.
9. Treybal, Robert E., *Mass Transfer Operation*, Third Edition, Mc Graw Hill International Editions, 1981.
10. Kister, Henry Z., *Distillation Operation*, Mc Graw Hill Inc., 1992.