Validation of Hysys Software for Design of a Sieve Tray Column

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Department of Chemical Engineering and Chemical Technology

Faculty of Engineering and Technology

November, 2016
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<td>Prof. Gurashi Abdalla Gasmeseed</td>
<td>chair person</td>
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Date : November/2016
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Date of Examination: 12/11/2016

Verses
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(26) واحلل عقدة من لساني (27) يفقهوا 

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صدق الله العظيم

"سورة طه"
Dedication

To My Mother

To

My Father

To

My Husband

To

My be loved children

To

My sister and Relatives

And my Colleagues
Acknowledgment

In the name of Alla who made all things possible.

I’d like to thank – heartily – my Supervisor Dr. Gurashi Abdalla Gasmelseed for his encouragement, guidance, and support from the early beginning to the end of this research.

Besides I offer my regards and blessings to all of those who stand beside me to complete this task.
Validation Of Hysys Software for Design Of a Sieve Tray Column
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Abstract

The separation of liquid mixtures into their various components is one of the major operations in the process industries. Distillation is the most widely used method of achieving separation of different component in side streams. The types of Distillation Process are Batch Processes and Continuous processes. Sieve trays are simply metal plates with holes in them. Vapour passes straight upward through the column through the plates. The arrangement, number and size of the holes are design parameters, which should be specified. The advantages of sieve trays are simpler and cheaper construction has led to their increasing use; The Objectives of this research is to design a sieve trays column and Comparing the results of using manual calculations with computer software Aspen Hysys. The mixture of acetone-water was separated in a sieve tray column. The equilibrium data were obtained and plotted in diagram to obtain the number of theoretical stages. The design parameters, the column efficiency, column height, tray space and column diameter were calculated. Using the same data, Hysys software was used to determine the above design parameters. These design data were compared with that calculated manually and they found to be satisfactory. number of tray 25,25tary diameter 0.8 m, 0.77 m, weir length 0.64m, 0.60m weir hight 0.045m, 0.05m tery space 0.55m, 0.05m in Hysys software and manunl calculation respectively. this study recommneded Control tuning and stability of the system have to be investigated. further work may be carried out for design of a sieve tray column for multicomponents systems. Flooding against column operation conditions needs to be investigated.
التحقق باستعمال برنامج hysys في تصميم برج المناخل

سارة وداعة الله محمد ابراهيم

ملخص الدراسة

يعتبر فصل الخليط السائل الى مكوناته من العمليات الأساسية في معظم الصناعات كما يعتبر التقطير أكثر العمليات استعمالاً في تحقيق الفصل لأنواع المكونات المختلفة. ينقسم التقطير إلى التقطير بالفصل بالدفعات والفصل المستمر وان أبراج المناخل هي ببساطة صواني من المعادن تحتوي على مناخلاً فيها حيث تتصاعد الأبخرة الى أعلى خلال البرج من خلال الصواني يتم ترتيب المناخل وعددها وعدد الأخرام كعوامل تصميمي ويتبنى تجديداً مما أدى إلى تزايد استعمال أبراج المناخل حيث أنها مبسطة التصميم ورخيصة التصنيع والتصنيع. أن الهدف من هذا البحث هو تصميم برج المناخل مع مقارنة نتائج التشغيل باستخدام التحسينات اليدوية والتحسينات باستخدام برنامج hysys.

تم فصل خليط الاستون والبخار في برج المناخل وتتم الحصول على بيانات الاتزان ورسمها في رسم بيانياً وذلك للحصول على عدد الصواني المتداخلة وقد تم حساب الكفاءة والارتفاع والمسافة بين الصواني والقطر باستخدام نفس البيانات.

وبعد استعمال برنامج hysys، وجد أن هذه البيانات متلبسة مع البيانات المحسوبة على برنامج hysys المحسوبة بدءاً مع البيانات المتكونة عليها من برنامج hysys تقريباً. عدد الصواني الحقيقية 25.25 قطر الصواني 0.8 م، 0.77 م، طول الحائط 46 م، 0.05 م، المسافة بين الصواني 0.55 م، 0.5 م في برنامج Hysys وارتفاع الحائط. وتحل الاعدادات اليدوية على التوالي. توصي هذه الدراسة بوضع نظام تحكم وتحليل الاستقرارية والضبط كما نوصي مستقبلاً بتصميم برج المناخل لأنظمة متعددة المكونات ولايد من أخذ الفيضان في الاعتبار عند ظروف تشغيل البرج المختلفة ودراسته والتحقق فيه.
**List of abbreviations**

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<tr>
<td>$\rho_l$</td>
<td>Liquid density</td>
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<tr>
<td>PPDS</td>
<td>Physical property data service available</td>
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<td>$FLV$</td>
<td>Liquid vapour flow factor</td>
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<td>$U_f$</td>
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<td>$U_v$</td>
<td>Maximum allowable $(v)$ velocity</td>
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<tr>
<td>$D_c$</td>
<td>Column diameter</td>
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<td>$A_c$</td>
<td>Column area</td>
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<td>$h_{ow}$</td>
<td>Weir crest</td>
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<td>$h_d$</td>
<td>Hole diameter</td>
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<tr>
<td>$h_r$</td>
<td>Residual head</td>
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<tr>
<td>$h_t$</td>
<td>Total plate pressure drop</td>
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<tr>
<td>$hap$</td>
<td>The high of the bottom edge above the plate</td>
</tr>
<tr>
<td>$l_w$</td>
<td>Weir length,$m$</td>
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<td>$U_v$</td>
<td>Maximum allowable $(v)$ velocity cross sectional area</td>
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<tr>
<td>$L_w$</td>
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<td>Clear Liquid back-up</td>
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Chapter One

Introduction
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Introduction

1.1 Distillation:
The separation of liquid mixtures into their various components is one of the major operations in the process industries and distillation. The most widely used method of achieving this end is the key operation in any oil refinery. In processing, the demand for purer products, coupled with the need for greater efficiency, has promoted continued research into the techniques of distillation.

When distillation operated at a continuous steady state, new feed is always being introduced unless the process is disturbed due to changes in feed, heating, temperature, or condensing, the rate of feed being introduced and the amount of top and bottom products being removed are normally balanced. This is known as continuous steady-state fractional distillation. [1]

1.2 Types of Distillation Processes:

1.2.1 Batch Processes:

In batch distillation, the more volatile component is evaporated from the still which therefore becomes progressively richer in the less volatile constituent. Distillation is continued, either until the residue of the still contains a material with an acceptably low content of the volatile material, or until the distillate is no longer sufficiently pure with respect of the volatile component. In batch operation, the feed to the column is introduced batch-wise. That is, the column is charged with a 'batch' and then the distillation process is carried out. When the desired task is achieved, a next batch of feed is introduced. [4]
1.2.2 Continuous Columns:

In contrast, continuous columns process a continuous feed stream. No interruptions occur unless there is a problem with the column or surrounding process units. They are capable of handling high throughputs and are the most common of the two types. It shall concentrate only on this class of columns.

Continuous distillation is an ongoing distillation in which a liquid mixture is continuously (without interruption) fed into the process and separated fractions are removed continuously as output streams as time passes during the operation. Continuous distillation produces at least two output fractions, including at least one volatile distillate fraction, which has boiled and been separately captured as a vapor condensed to a liquid. There is always a bottoms (or residue) fraction, which is the least volatile residue that has not been separately captured as a condensed vapor.

Continuous distillation differs from batch distillation with the respect to that concentrations should not change over time. Continuous distillation can be run at a steady state for an arbitrary amount of time. For any source material of specific composition, the main variables that affect the purity of products in continuous distillation are the reflux ratio and the number of theoretical equilibrium stages (practically, the number of trays or the height of packing). Reflux is a flow from the condenser back to the column, which generates a recycle that allows a better separation with a given number of trays. Equilibrium stages are ideal steps where compositions achieve vapor-liquid equilibrium, repeating the separation process and allowing better separation given a reflux ratio. A column with a high reflux ratio may have fewer stages, but it refluxes a large amount of liquid, giving a wide column with a large holdup.
Conversely, a column with a low reflux ratio must have a large number of stages, thus requiring a taller column.

1.3 Types Of Continuous Process:

Continuous columns can be in future classified according to:

- The nature of the feed that they are processing.
- The number of products steam they have.
- The number of feed plates introduction. [4]

Fig. (1.1) Continuous distillation column
1.4 Fractional distillations:

For many cases, the boiling points of the components in the mixture will be sufficiently close that Raoult's law must be taken into consideration. Therefore, fractional distillation must be used in order to separate the components well by repeated vaporization-condensation cycles within a packed fractionating column. This separation, by successive distillations, is also referred to as rectification. [4]

As the solution to be purified is heated, its vapors rise to the fractionating column. As it rises, it cools, condensing on the condenser walls and the surfaces of the packing material. Here, the condensate continues to be heated by the rising hot vapors; it vaporizes once more. However, the composition of the fresh vapors are determined once again by Raoult's law. Each vaporization-condensation cycle (called a theoretical plate) will yield a purer solution of the more volatile component. In reality, each cycle at a given temperature does not occur at exactly the same position in the fractionating column; theoretical plate is thus a concept rather than an accurate description. [4]

More theoretical plates lead to better separations. A spinning band distillation system uses a spinning band of Teflon or metal to force the rising vapors into close contact with the descending condensate, increasing the number of theoretical plates. [4]

1.5 Flash or Equilibrium Distillation:

Flash or equilibrium distillation, frequently carried out as a continuous process, consists of vaporising a definite fraction of the liquid feed in such a way that the vapour evolved is in equilibrium with the residual liquid. The feed is usually pumped through a fired heater and enters the still through a valve where the pressure is reduced. The still is essentially
a separator in which the liquid and vapour produced by the reduction in pressure have sufficient time to reach equilibrium. The vapour is removed from the top of the separator and is then usually condensed, while the liquid leaves from the bottom. [1]

**1.6 Steam Distillation:**

Like vacuum distillation, steam distillation is a method for distilling compounds which are heat-sensitive. The temperature of the steam is easier to control than the surface of a heating element, and allows a high rate of heat transfer without heating at a very high temperature. This process involves bubbling steam through a heated mixture of the raw material. By Raoult's law, some of the target compound will vaporize (in accordance with its partial pressure). The vapor mixture is cooled and condensed, usually yielding a layer of oil and a layer of water. [4]

**1-7 Objectives:**

1. Design of a distillation sieve tray, column.

2. Comparing the results using manual calculation with computer software Aspen Hysys.
Chapter Two

Literature Review
2. Literature Review:

2.1 Trays and Plates:

The terms "trays" and "plates" are used interchangeably. There are many types of tray designs, but the most common ones are:

2.1.1 Sieve trays:

Sieve trays are simply metal plates with holes in them. Vapour passes straight upward through the liquid on the plate. The arrangement, number and size of the holes are design parameters.

The advantages of sieve trays are simpler and cheaper construction has led to their increasing use. [2]

Fig (2.1): Sieve tray
2.1.2 Bubble cap trays

A bubble cap tray has riser or chimney fitted over each hole, and a cap that covers the riser. The cap is mounted so that there is a space between riser and cap to allow the passage of vapour. Vapour rises through the chimney and is directed downward by the cap, finally discharging through slots in the cap, and finally bubbling through the liquid on the tray. [2]

Bubble-cap trays are rarely used for new installations on account of their high cost and their high pressure drop. In addition, difficulties arise in large columns because of the large hydraulic gradients which are set up across the trays. Bubble cap trays are capable of dealing with very low liquid rates and are therefore useful for operation at low reflux ratios. [4]

Fig (2.2): Bubble cap tray
2.1.3 Valve trays

The valve tray, which may be regarded as intermediate between the bubble cap and the sieve tray, offers advantages over both. The important feature of the tray is that liftable caps act as variable orifices which adjust themselves to changes in vapour flow. The valves are either metal discs of up to about 38 mm diameter, or metal strips which are raised above the openings in the tray deck as vapour passes through the trays. The caps are restrained by legs or spiders which limit the vertical movement and some types are capable of forming a total liquid seal when the vapour flow is insufficient to lift the cap. [1]

2.1.3.1 Advantages claimed for valve trays include:

(a) Operation at the same capacity and efficiency as sieve trays.

(b) A low pressure drop which is fairly constant over a large portion of the operating range.

(c) A high turndown ratio, that is it can be operated at a small fraction of design capacity.

(d) A relatively simple construction which leads to a cost of only 20 per cent higher than that of a comparable sieve tray. Valve trays, because of their proprietary nature, are usually designed by manufacturers, although it is possible to obtain an estimate of design and performance from published.

In valve trays, perforations are covered by liftable caps. Vapour flows lifts the caps, thus self creating a flow area for the passage of vapour. The lifting cap directs the vapour to flow horizontally into the liquid, thus providing better mixing than is possible in sieve trays. [2]
2.2 Liquid and Vapour Flows in a Tray Column:

The next few figures show the direction of vapour and liquid flow across a tray, and across a column. [4]

Fig (2.3): valve tray

Fig (2.4): direction of vapour and liquid flow across a tray, and across a column.
Each tray has 2 conduits, one on each side, called ‘downcomers’. Liquid falls through the downcomers by gravity from one tray to the one below it. A weir on the tray ensures that there is always some liquid holdup on the tray and is designed such that the holdup is at a suitable height. The area allowed for the passage of vapour on each tray is called the active tray area. As the hotter vapour passes through the liquid on the tray above, it transfers heat to the liquid. In doing so, some of the vapour condenses adding to the liquid on the tray. The condensate, however, is richer in the less volatile components than is in the vapour. Additionally, because of the heat input from the vapour, the liquid on the tray boils, generating more vapour. This vapour, which moves up to the next tray in the column, is richer in the more volatile components. This continuous contacting between vapour and liquid occurs on each tray in the column and brings about the separation between low boiling point components and those with higher boiling points.

2.3 Main Components of Distillation Columns:

2.3.1 Basic distillation equipments and operation:

Distillation columns are made up of several components, each of which is used either to transfer heat energy or enhance material transfer. A typical distillation contains several major components:

- a vertical shell where the separation of liquid components is carried out column internals such as trays/plates and/or packings which are used to enhance component separations
- a reboiler to provide the necessary vaporisation for the distillation process
- a condenser to cool and condense the vapour leaving the top of the column
• a reflux drum to hold the condensed vapour from the top of the column so that liquid (reflux) can be recycled back to the column.[4]

2.4 Basic Operation and Terminology:

The liquid mixture that is to be processed is known as the feed and this is 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 vapour. 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 vapour 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, bottoms. [4]

Fig (2.5) bottoms product
The vapour moves 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.

Thus, there are internal flows of vapour and liquid within the column as well as external flows of feeds and product streams, into and out of the column.

![Diagram of a distillation column with a condenser, reflux drum, and top product](image)

**Fig (2.6) Top product**

### 2.5 Distillation Principles:

Separation of components from a liquid mixture via distillation depends on the differences in boiling points of the individual components. Also, depending on the concentrations of the components present, the liquid mixture will have different boiling point characteristics. Therefore, distillation processes depends on the vapour pressure characteristics of liquid mixtures. [4]
2.5.1 Vapour Pressure and Boiling:

The vapour pressure of a liquid at a particular temperature is the equilibrium pressure exerted by molecules leaving and entering the liquid surface. Here are some important points regarding vapour pressure:

- Energy input raises vapour pressure.
- Vapour pressure is related to boiling.
- A liquid is said to ‘boil’ when its vapour pressure equals the surrounding pressure. The ease with which a liquid boils depends on its volatility.
- Liquids with high vapour pressures (volatile liquids) will boil at lower temperatures.
- The vapour pressure and hence the boiling point of a liquid mixture depends on the relative amounts of the components in the mixture.
- Distillation occurs because of the differences in the volatility of the components in the liquid mixture.

2.5.2 Boiling Point Diagram:

The boiling point diagram shows how the equilibrium compositions of the components in a liquid mixture vary with temperature at a fixed pressure. Consider an example of a liquid mixture containing 2 components (A and B) - a binary mixture. This has the following boiling point diagram.

The boiling point of A is that at which the mole fraction of A is 1. The boiling point of B is that at which the mole fraction of A is 0. In this example, A is the more volatile component and therefore has a lower boiling point than B. The upper curve in the diagram is called the dew-point curve while the lower one is called the bubble-point curve.
The dew-point is the temperature at which the saturated vapour starts to condense.

The bubble-point is the temperature at which the liquid starts to boil.

The region above the dew-point curve shows the equilibrium composition of the superheated vapour while the region below the bubble-point curve shows the equilibrium composition of the subcooled liquid.

This difference between liquid and vapour compositions is the basis for distillation operations.

**Fig (2.7) Boiling Point Diagram**

### 2.6 Relative Volatility:

Relative volatility is a measure of the differences in volatility between 2 components, and hence their boiling points. It indicates how easy or difficult a particular separation will be. The relative volatility of component ‘i’ with respect to component ‘j’ is defined as

\[
\alpha_j = \frac{y_j / x_j}{y_j / x_j} 
\]

\( y_j \) = mole fraction of component ‘i’ in the vapour

\( x_j \) = mole fraction of component ‘i’ in the liquid

Thus if the relative volatility between 2 components is very close to one, it is an indication that they have very similar vapour pressure characteristics. This means that they have very
similar boiling points and therefore, it will be difficult to separate the two components via distillation.

2.7 Factors Affecting Distillation Column Operation:

2.7.1 Feed Conditions:

The state of the feed mixture and feed composition affects the operating lines and hence the number of stages required for separation. It also affects the location of feed tray. During operation, if the deviations from design specifications are excessive, then the column may no longer be able handle the separation task. To overcome the problems associated with the feed, some column are designed to have multiple feed points when the feed is expected to containing varying amounts of components.

2.7.2 Reflux Conditions:

As the reflux ratio is increased, the gradient of operating line for the rectification section moves towards a maximum value of 1. Physically, what this means is that more and more liquid that is rich in the more volatile components are being recycled back into the column. Separation then becomes better and thus less trays are needed to achieve the same degree of separation. Minimum trays are required under total reflux conditions.

On the other hand, as reflux is decreased, the operating line for the rectification section moves towards the equilibrium line. The ‘pinch’ between operating and equilibrium lines becomes more pronounced and more and more trays are required. This is easy to verify using the McCabe-Thiele method.
The limiting condition occurs at minimum reflux ratio, when an infinite number of trays will be required to effect separation. Most columns are designed to operate between 1.2 to 1.5 times the minimum reflux ratio because this is approximately the region of minimum operating costs (more reflux means higher reboiler duty).

2.7.3 Vapour Flow Conditions:

Adverse vapour flow conditions can cause

- foaming
- entrainment
- weeping/dumping
- flooding

2.7.3.1 Foaming:

Foaming refers to the expansion of liquid due to passage of vapour or gas. Although it provides high interfacial liquid-vapour contact, excessive foaming often leads to liquid buildup on trays. In some cases, foaming may be so bad that the foam mixes with liquid on the tray above. Whether foaming will occur depends primarily on physical properties of the liquid mixtures, but is sometimes due to tray designs and condition. Whatever the cause, separation efficiency is always reduced.

2.7.3.2 Entrainment:

Entrainment refers to the liquid carried by vapour up to the tray above and is again caused by high vapour flow rates. It is detrimental because tray efficiency is reduced: lower volatile material is carried to a plate holding liquid of higher volatility. It could also contaminate high purity distillate. Excessive entrainment can lead to flooding.
2.7.3.3 Weeping/Dumping:

This phenomenon is caused by low vapour flow. The pressure exerted by the vapour is insufficient to hold up the liquid on the tray. Therefore, liquid starts to leak through perforations. Excessive weeping will lead to dumping. That is the liquid on all trays will crash (dump) through to the base of the column (via a domino effect) and the column will have to be re-started. Weeping is indicated by a sharp pressure drop in the column and reduced separation efficiency.

3.7.3.4 Flooding:

Flooding is brought about by excessive vapour flow, causing liquid to be entrained in the vapour up the column. The increased pressure from excessive vapour also backs up the liquid in the downcomer, causing an increase in liquid holdup on the plate above. Depending on the degree of flooding, the maximum capacity of the column may be severely reduced. Flooding is detected by sharp increases in column differential pressure and significant decrease in separation efficiency.

2.8 Column Diameter:

Most of the above factors that affect column operation is due to vapour flow conditions: either excessive or too low. Vapour flow velocity is dependent on column diameter. Weeping determines the minimum vapour flow required while flooding determines the maximum vapour flow allowed, hence column capacity. Thus, if the column diameter is not sized properly, the column will not perform well. Not only will operational problems occur, the desired separation duties may not be achieved.
2.9 State of Trays:

number of trays required for a particular separation duty is determined by the efficiency of the plate. Thus, any factors that cause a decrease in tray efficiency will also change the performance of the column. Tray efficiencies are affected by fouling, wear and tear and corrosion, and the rates at which these occur depends on the properties of the liquids being processed. Thus appropriate materials should be specified for tray construction.

2.10 Weather Conditions:

Most distillation columns are open to the atmosphere. Although many of the columns are insulated, changing weather conditions can still affect column operation. Thus the reboiler must be appropriately sized to ensure that enough vapour can be generated during cold and windy spells and that it can be turned down sufficiently during hot seasons. The same applies to condensors.

These are some of the more important factors that can cause poor distillation column performance. Other factors include changing operating conditions and throughputs, brought about by changes in upstream conditions and changes in the demand for the products. All these factors, including the associated control system, should be considered at the design stages because once a column is built and installed, nothing much can be done to rectify the situation without incurring significant costs. The control of distillation columns is a field in its own right,
Chapter Three

Materials and Methods
Chapter Three

Material and Methods

3.1 Selection of plate type:

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.

3.1.1 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, bubble-cap. [2]

3.1.2 Operating range:

This is the most significant factor. By operating range is meant the range of vapour and liquid rates over which the plate will operate satisfactorily. 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 vapour rates.

Sieve plates rely on the flow of vapour through the holes to hold the liquid on the plate,
and cannot operate at very low vapour 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.

3.1.3 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.

3.1.4 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. [2]

3.2 Number of plates required in a distillation column:

In order to develop a method for the design of distillation units to give the desired fractionation, it is necessary, in the first instance, to develop an analytical approach which enables the necessary number of trays to be calculated. First the heat and material flows over the trays, the condenser, and the reboiler must be established. Thermodynamic data are required to establish how much mass transfer is needed to establish equilibrium between the streams leaving each tray. The required diameter of the column will be dictated by the necessity to accommodate the desired flowrates, to operate within the available drop in pressure, while at the same time effecting the desired degree of mixing of the streams on each
Four streams are involved in the transfer of heat and material across a plate, as shown in Figure (2-9) in which plate $n$ receives liquid $L_{n+1}$ from plate $n + 1$ above,

and vapour $V_{n-1}$ from plate $n - 1$ below.

Plate $n$ supplies liquid $L_n$ to plate $n - 1$,

and vapour $V_n$ to plate $n + 1$.

The action of the plate is to bring about mixing so that the vapour $V_n$, of composition $y_n$, approaches equilibrium with the liquid $L_n$, of composition $x_n$. The streams $L_{n+1}$ and $V_{n-1}$ cannot be in equilibrium and, during the interchange process on the plate, some of the more volatile component is vaporised from the liquid $L_{n+1}$, decreasing its concentration to $x_n$, and some of the less volatile component is condensed from $V_{n-1}$, increasing the vapour concentration to $y_n$. The heat required to vaporise the more volatile component from the liquid is supplied by partial condensation of the vapour $V_{n-1}$. Thus the resulting effect is that the more volatile component is passed from the liquid running down the column to the vapour rising up, whilst the less volatile component is transferred in the opposite direction.
Figure (3.1). Material balance
3.3 Calculation of the number of plates using the Lewis–Sorel method:

If a unit is operating, so that a binary feed $F$ is distilled to give a top product $D$ and a bottom product $W$, with $x_f$, $x_d$, and $x_w$ as the corresponding mole fractions of the more volatile component, and the vapour $V_t$ rising from the top plate is condensed, and part is run back as liquid at its boiling point to the column as reflux, the remainder being withdrawn as product, then a material balance above plate $n$, indicated by the loop I in Figure (3.9) gives:

$$V_n = L_{n+1} + D$$ \hspace{1cm} (3.1)

Figure(2-11). Material balances at top and bottom of column

Expressing this balance for the more volatile component gives:

$$y_n V_n = L_{n+1} X_{n+1} + D x_d$$ \hspace{1cm} (3.2)

Thus: $y_n V_n = L_{n+1} X_{n+1} + D x_d$ \hspace{1cm} (3.3)

$$Y_n = \frac{L_{n+1} X_{n+1} + D x_d}{V_n}$$ \hspace{1cm} (3.4)

This equation relates the composition of the vapour rising to the plate to the composition of the liquid on any plate above the feed plate. Since the molar liquid overflow is constant, $L_n$ may be replaced by $L_{n+1}$ and:

$$y_n = \frac{L_{n} X_{n+1} + D x_d}{V_n}$$ \hspace{1cm} (3.5)

Similarly, taking a material balance for the total streams and for the more volatile component from the bottom to above plate $m$, as indicated by the loop II in Figure (3.10)

$$L_m = L_{m+1}$$ \hspace{1cm} (3.6)
This equation, which is similar to equation (3.5) gives the corresponding relation between the compositions of the vapour rising to a plate and the liquid on the plate, for the section below the feed plate. These two equations are the equations of the operating lines. In order to calculate the change in composition from one plate to the next, the equilibrium data are used to find the composition of the vapour above the liquid, and the enrichment line to calculate the composition of the liquid on the next plate. This method may then be repeated up the column, using equation (3.9) for sections below the feed point, and equation (3.5) for sections above the feed point. [1]
Figure (3.2) Material balances at top and bottom of column
3.4 Design of industrial distillation columns:

Design and operation of a distillation column depends on the feed and desired products. Given a simple, binary component feed, analytical methods such as the McCabe-Thiele method or the Fenske equation can be used. For a multi-component feed, simulation models are used both for design and operation.

Moreover, the efficiencies of the vapor-liquid contact devices (referred to as plates or trays) used in distillation columns, a distillation column needs more plates than the number of theoretical vapor-liquid equilibrium stages.

An indication of numbers: the separation of two compounds with relative volatility of 1.1 requires a minimum of 130 theoretical plates with a minimum reflux ratio of 20. With a relative volatility of 4, the required number of theoretical plates decreased to 9 with a reflux ratio of 0.66. In another source, a boiling point difference of 30 °C requires 12 theoretical plates and, for a difference of 3 °C, the number of plates increased to 1000.

The reflux ratio is the ratio of the amount of moles returned as refluxed liquid to the fractionating column and the amount of moles of final product, both per unit. [4]
3.5 Aspen Hysys:

The goal of programs like Aspen Hysys and Aspenone of course, to provide with the capability to design an entire process as completely and accurately as possible. Aspenhysys has a much better and intuitive interface than Aspenone.

Aspen Hysys does not wait until you've entered everything before beginning calculations. It always calculates as much as it can at all times and results are always available, even during calculations. Any changes that is make to the data are automatically propagated throughout the program to anywhere that entry appears and all necessary recalculations are instantly carried out. It tends to be a lot easier to catch errors this way as you build your simulation. when you will not want HYSYS calculating the entire flowsheet over again every time make a small change..

There are two files that HYSYS reads whenever you start Hysys. One of these is the Preferences File, the other is the workbook format file. The preferences file tells Hysys many important things like which units you want to use, where to look first for your stored files . multiple preference files saved and build up various unit sets, in one file you alter upon need, but the one that is loaded upon startup is the file named hysys If launch Hysys from start menu, user's profile has been configured with Y:\Hysys as the start directory The workbook shows the contents of every stream in the process and the connections of every unit entire simulation using the Workbook.

If start file by double clicking it simulations should all be stored under Y:\Hysys\Cases[3].
Chapter Four

Result and Discussion
Chapter Four
Result and Discussion

4.1: Case study (1)

Acetone is to be recovered from an aqueous waste stream by continuous distillation. The feed contains 10 per cent w/w acetone of 98 per cent purity is wanted and the aqueous effluent does not contain 50 ppo acetone. The feed is at 20 °C. the number of ideal stages are required to calculated using manual.

4.1.1 Procedure

The column is operated of at pressure

<p>| | | | | | | |</p>
<table>
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<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>(x)</td>
<td>0.35</td>
<td>0.40</td>
<td>0.45</td>
<td>0.50</td>
<td>0.55</td>
<td>0.60</td>
</tr>
<tr>
<td>(y)</td>
<td>0.8201</td>
<td>0.8269</td>
<td>0.8376</td>
<td>0.8387</td>
<td>0.8455</td>
<td>0.8532</td>
</tr>
<tr>
<td>°C</td>
<td>61.26</td>
<td>60.75</td>
<td>60.35</td>
<td>59.95</td>
<td>59.54</td>
<td>59.12</td>
</tr>
</tbody>
</table>

The equilibrium curve is drawn with sufficient accuracy to determine the stages above the feed by plotting the concentrations at increments of 0.1 as shown in fig. (4.1)
Molecular weights, acetone 58, water 18

\[
\text{Mol fractions acetone feed} = \frac{10}{\frac{58}{10} + \frac{90}{18}} = 0.033 \quad \text{(4.1)}
\]

\[
\text{top product} = \frac{98}{\frac{58}{98} + \frac{2}{18}} = 0.94 \quad \text{(4.2)}
\]

\[
\text{bottom product} = 50\times10^{-6} \times \frac{18}{58} = 15.5\times10^{-6} \quad \text{(4.3)}
\]

**4.1.2 Feed condition (q-line)**

Bubble point of feed (interpolated) 83° C
Latent heats, water 41,360, acetone 28,410 J/mol

Mean specific heats, water 75.3, acetone 128 J/mol° C

Latent heat of feed D 28,410 O 0.033 C (1 – 0.033) 41,360 D 40,933 J/mol

Specific heat of feed D (0.033 O 128) C (1 – 0.033) 75.3 D 77.0 J/mol

Heat to vaporise 1 mol of feed D (83 – 20) 77.0 C 40,933 D 45,784 J

\[ q = \frac{45.784}{40.933} = 1.12 \] ........................... (4.5)

Slope of q line \[ \frac{1.12}{1.12 - 1} = 9.32 \] ........................... (4.6)

For this case the condition of minimum reflux occurs where the top operating line just touches the equilibrium curve at the point where the q line cuts the curve. From the Figure (4.1)

\( \phi \) for the operating line at minimum reflux = 0.65............................. (4.7)

\[ R_{min} = \frac{0.94}{0.65} - 1 = 0.45 \]

Take \( R = R_{min} \times 3 \)

As the flows above the feed point will be small, a high reflux ratio is justified; the condenser duty will be small.

\[ At\ R = 3 \times 0.45 = 1.35, \ \phi = \frac{0.94}{1 + 1.35} = 0.4 \] ........................... (4.8)

For this Case it is convenient to step the stages off starting at the intersection of the operating lines. This gives three stages above the feed
up to \( y = 0.8 \). The top section is drawn to a larger scale, Figure (4.2), to determine the stages above \( y = 0.8 \): three to four stages required; total stages above the feed 7.

![Figure (4.2): top section enlarged](image)

Below the feed, one stage is required down to \( x = 0.04 \). A log-log plot is used to determine the stages below this concentration. Data for log-log plot:

operating line slope, from Figure (4.1) = \( 0.45/0.09 = 5.0 \) ................ (4.9)

operating line equation, \( y = 4.63 (x - x_b) + x_b \)

\[
= 5.0x - 62.0 \times 10^{-6}
\]

equilibrium line slope, from \( v - 1 - e \) data = \( 0.6381/0.05 = 12.8 \)

table (4.2) vapor-liquid Equilibrium

<table>
<thead>
<tr>
<th>( x )</th>
<th>( 4 \times 10^{-2} )</th>
<th>( 10^{-3} )</th>
<th>( 10^{-4} )</th>
<th>( 4 \times 10^{-5} )</th>
<th>( 2 \times 10^{-5} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Equilibrium line ( y )</td>
<td>0.51 ( 1.3 \times 10^{-2} )</td>
<td>( 1.3 \times 10^{-3} )</td>
<td>( 5.1 \times 10^{-4} )</td>
<td>( 2.6 \times 10^{-5} )</td>
<td></td>
</tr>
<tr>
<td>Operating line ( y )</td>
<td>0.20 ( 4.9 \times 10^{-3} )</td>
<td>( 4.4 \times 10^{-4} )</td>
<td>( 1.4 \times 10^{-2} )</td>
<td>( 3.8 \times 10^{-5} )</td>
<td></td>
</tr>
</tbody>
</table>
From Figure (4.3), number of stages required for this Rectifying section including the reboiler = 8

Figure (4.3) log-log plot of McCabe-Thiele diagram

Total number of stages below feed = 9

Total stages = 7 + 9 = 16 [2]
4.2 Case study (2)

Statement of the design problem

Design of the plates for the column specified in case study (1). The minimum feed rate is taken as 70 per cent of the maximum (maximum feed 10,000 kg/h) for sieve plates.

4.2.1 Procedure:

As the liquid and vapour flow-rates and compositions is varied up the column, plate designs should be made above and below the feed point. Only the bottom plate will be designed in detail in this case.

Taking:

Number of stages = 16

Slope of the stripping operating line = 5.0

Slope of top operating line = 0.57

Top composition 94 per cent mol. 98 per cent w/w.

Bottom composition — essentially water.

Reflux ratio = 1.35

4.2.2 Flow-rates

Mol. weight of feed = 0.033 x 58 + (1 - 0.033)18 = 19.32 ................. (4.10)

Feed rate = 13,000/19.32 = 672.9 kmol/h ......................... (4.11)
A mass balance on acetone gives:

Top product, \( D = 672.9 \times 0.033/0.94 \) \( D \) \( 23.6 \text{ kmol/h} \)

Vapour rate, \( V = D (1 + R) \) \( D \) \( 23.6 (1 + 1.35) = 55.5 \text{ kmol/h} \)

An overall mass balance gives:

Bottom product, \( B = 672.9 - 23.6 = 649.3 \text{ kmol/h} \)

Slope of the bottom operating line \( L_{\text{m}} / V_{\text{m}} = 5.0 \)

and \( L_{\text{m}} = V_{\text{m}} - B \) from which:

vapour flow below feed, \( V_{\text{m}} = 162.3 \text{ kmol/h} \)

liquid flow below feed, \( L_{\text{m}} = 811.6 \text{ kmol/h} \)

4.2.3 Physical properties

The base pressure is estimated column efficiency is 60 per cent, the reboiler is equivalent to one stage

\[
\text{Number of real stages} = \frac{16 - 1}{0.6} = 25
\]

Assume 100 mm water, pressure drop per plate.

\[
\text{Column pressure drop} = 100 \times 10^{-3} \times 1000 + 9.81 \times 25 = 24,525 \text{ Pa}
\]

Top pressure, 1 atm (14.7 lb/in\(^2\)) = \( 101.4 \times 10^3 \text{ Pa} \)

Estimated bottom pressure = \( 101.4 \times 10^3 + 24,525 \)

\[= 125,925 \text{ Pa} = 1.26 \text{ bar} \]

From steam tables, base temperature 106\(^\circ\) C

\[\rho_u = 0.72 \text{ kg/m}^3\]
\[ \rho_u = 954 \text{ kg/m}^3 \]

Surface tension \(57 \times 10^{-3} \text{ N/m}\)

Top, 98% w/w acetone, top temperature 57\(^\circ\) C

From PPDS

\[ \rho_u = 2.05 \text{ kg/m}^3, \rho_u = 753 \text{ kg/m}^3 \]

Molecular weight 55.6

Surface tension \(23 \times 10^{-3} \text{ N/m}\)

**4.2.4 Column diameter:**

\[
F_{LV} \text{ bottom} = 5.0 \sqrt{\frac{0.72}{954}} = 0.14 \quad \text{......................... (4.14)}
\]

\[
F_{LV} \text{ top} = 0.57 \sqrt{\frac{0.72}{753}} = 0.03
\]

Tray spacing as 0.5m

From fig (4.4)

Base \(K_1 = 7.5 \times 10^{-2}\)

Top \(K_1 = 9.0 \times 10^{-2}\)

Correction for surface tensions

Base \(K_1 = \left( \frac{57}{20} \right)^{0.2} \times 7.5 \times 10^{-2} = 9.3 \times 10^{-2}\)

Base \(K_2 = \left( \frac{23}{20} \right)^{0.2} \times 9.0 \times 10^{-2} = 9.3 \times 10^{-2}\)
Base \( u_f = 9.3 \times 10^{-2} \sqrt{\frac{954 - 0.72}{0.72}} = 3.38 \text{ m/s} \) .......................... (4.15)

Top \( u_f = 9.3 \times 10^{-2} \sqrt{\frac{953 - 2.05}{2.05}} = 1.78 \text{ m/s} \) .......................... (4.16)

Design for 85 per cent flooding at maximum flow rate:

Base \( \hat{u}_v = 3.38 \times 0.85 = 2.87 \text{ m/s} \)

Top \( \hat{u}_v = 1.78 \times 0.85 = 1.51 \text{ m/s} \)

Maximum volumetric flow-rate

Base = \( \frac{162.3 \times 18}{0.72 \times 3600} = 1.13 \text{ m}^3/\text{s} \) .......................... (4.17)

Top = \( \frac{55.5 \times 55.6}{2.05 \times 3600} = 0.42 \text{ m}^3/\text{s} \) .......................... (4.18)

Net area required

Bottom = \( \frac{1.13}{2.87} = 0.40 \text{ m}^2 \) .......................... (4.19)

Top = \( \frac{0.42}{1.51} = 0.28 \text{ m}^2 \) .......................... (4.20)

As first take downcomer area as 12 per cent of total.

Column cross-sectioned area

base = \( \frac{0.40}{0.88} = 0.46 \text{ m}^2 \) .......................... (4.21)

Top = \( \frac{0.28}{0.88} = 0.32 \text{ m}^2 \) .......................... (4.22)
Column diameter

\[ \text{base} = \sqrt{\frac{0.46 \times 4}{\pi}} = 0.77 \text{m} \] ........................ (4.23)

\[ \text{Top} = \sqrt{\frac{0.34 \times 4}{\pi}} = 0.64 \text{m} \] .......................... (4.24)

Use same diameter above and below feed, reducing the perforated area for plates above the feed.

Nearest standard pipe size (BS 1600 Pt.2); outside diameter 812.8 mm (32 in); standard wall thickness 9.52 mm; inside diameter 794 mm.

4.2.5 Liquid flow volumetric rate

Maximum volumetric liquid rate = \(\frac{811.6 \times 18}{3600 \times 954} = 4.3 \times 10^{-3} \text{ m}^3/\text{s} \) g ...... (4.25)

The plate diameter is outside the range of Figure (4.5), but it is clear that a single pass plate can be used.

4.2.6 Provisional plate design

Column diameter \( D_c = 0.79 \text{ m} \)

Column area \( A_c = 0.50 \text{ m}^2 \)

Downcomer area \( A_d = 0.12 \times 0.50 = 0.06 \text{ m}^2 \), at 12 per cent

Net area \( A_n = A_c - A_d = 0.50 - 0.06 = 0.44 \text{ m}^2 \)

Active area \( A_a = A_c - 2A_d = 0.50 - 0.12 = 0.38 \text{ m}^2 \)

Hole area \( A_h \) take 10 per cent \( A_a \) as first trial \( = 0.038 \text{ m}^2 \)
Weir length from Figure (4.8) = $0.76 \times 0.79 = 0.60$ m

Take weir height 50 mm

Hole diameter 5 mm

Plate thickness 5 mm

4.2.7 Cheek weeping

Maximum liquid rate $= \left( \frac{811.6 \times 18}{3600} \right) = 4.06$ kg/s ......................... (4.26)

Minimum liquid rate, at 70 per cent turn-down $= 0.7 \times 4.06 = 2.84$ kg/s

Maximum $h_{ow} = 750 \left( \frac{4.06}{954 \times 0.06} \right)^{2/3} = 27$ mm liquid ......................... (4.27)

Maximum $h_{ow} = 750 \left( \frac{2.85}{954 \times 0.60} \right)^{2/3} = 22$ mm liquid ......................... (4.28)

At minimum rate $h_w + h_{ow} = 50 + 22 = 72$ mm ......................... (4.29)

From fig (4.7)

$K_2 = 30.6$

$u_h (\text{min}) = \frac{30.6 - 0.90(25.4 - 5)}{(0.72)^{1/2}} = 14$ m/s ......................... (4.30)

Actual minimum vapour velocity $\frac{\text{min \ internal \ rate}}{A_h}$

$= \frac{0.7 \times 1.13}{0.038} = 20.8$ m/s ......................... (4.31)

So minimum operating rate will be well above weep point.
4.2.8 Plate pressure drop

Dry plate drop

Maximum vapour velocity through holes

\[ \dot{u}_h = \frac{1.13}{0.038} = 29.7 \text{ m/s} \] .......................... (4.32)

for plate thickness/hole dia. = 1, and \( A_h/A_p \approx A_h/A_a = 0.1, \)

\( C_0 = 0.84 \)

\[ h_y = 51 \left( \frac{29.7}{0.84} \right)^2 \frac{0.72}{954} = 48 \text{ mm liquid} \] .......................... (4.33)

Residual head

\[ h_r = \frac{12.5 \times 10^3}{954} = 13.1 \text{ mm liquid} \] .......................... (4.34)

Total plate pressure drop

\[ h_t = 48 + (50 + 27) + 13 = 138 \text{ mm liquid} \] .......................... (4.35)

4.2.9 Downcomer liquid back-up

Downcomer pressure loss

Take \( h_{w \text{p}} = h_w - 10 = 40 \text{ mm} \)

Area under apron, \( h_{w \text{p}} = 0.60 \times 40 \times 10^{-3} = 0.024 \text{ m}^2 \) .......................... (4.36)

As this is less than \( h_r = 0.06 \text{ m}^2 \) use \( A_{w \text{p}} \) in equation 11.92

\[ h_{dc} = 166 \left( \frac{4.06}{954 \times 0.024} \right)^2 = 5.2 \text{ mm} \] .......................... (4.37)

say 6 mm
Back – up in downcomer

\[ h_p = (50 + 27) + 138 + 6 = 221\,mm \] ........................................ (4.38)

\[ 0.22\,m \]

\[ 0.00 < \frac{1}{2} \text{ (plate spacing + weir height)} \]

So tray spacing is acceptable

Cheek residence time:

\[ t_r = \frac{0.06 \times 0.22 \times 954}{4.06} = 3.1\,s \] ........................................ (4.39)

\[ > 3\,s, \text{ satisfactory.} \]

4.2.10 Check entrainment

\[ u_v = \frac{1.13}{0.44} = 2.57\,m/s \] ........................................ (4.40)

Per cent flooding \[ = \frac{2.57}{3.38} = 76 \]

\[ F_{LV} = 0.14, \text{ from Figure (4.6), } \Psi = 0.018 \text{ well below } 0.1. \]

As the per cent flooding is well below the design figure of 85, the column diameter could be reduced, but this would increase the pressure drop.

4.2.11 perforated Area

From fig (4.9) at \[ \frac{l_w}{D_c} = 0.76 \]

\[ \theta_c = 99^\circ \]

Angle subtended at plate edge by unperforated strip = 180 – 99 = 81°(4.41)

mean length, unperforated edge strips = (0.79 – 50 \times 10^{-3}) \pi \times 81/180 = 1.05\,m (4.42)
area of unperforated edge strips = $50 \times 10^{-3} \times 1.05 = 0.053 \text{ m}^2$ ........ (4.43)

Mean length of calming zone = $(0.79 - 50 \times 10^{-3}) \sin \left(\frac{99}{2}\right) = 0.563 \text{ m} \ldots (4.44)$

Area of calming zone = $2(0.563 \times 50 \times 10^{-3}) = 0.056 \text{ m}^2$ .................. (4.45)

Total area for perforations, $A_p = 0.38 - 0.053 - 0.056 = 0.271 \text{ m}^2$ ..(4.46)

$A_h / A_p = 0.038 / 0.271 = 0.14$ ...........................................(4.47)

From figure (4.10) $l_p / d_n = 2.6$, satisfactory, within 2.5 to 4.0.

4.2.12 Number of holes

Area of one hole = $1.964 \times 10^{-5} \text{ m}^2$

Number of holes = $\frac{0.038}{1.964 \times 10^{-5}} = 1935 \ [2]$ .............................................(4.48)
Figure (4.4): Flooding velocity, Sieve plates
Figure (4.5) Selection of liquid-flow arrangement
Figure (4.6): Entrainment correlation for sieve plates
Figure (4.7): Weep-point correlation
Figure (4.8): Relation between downcomer area and weir length
Figure (4.9): Relation between angle subtended by chord, chord height and chord length
Figure (4.10): Relation between hole area and pitch
Figure (4.11): discharge coefficient, sieve plates
## Summary of data design

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Chapter Five

Conclusion & Recommendation
Chapter five

Conclusion and Recommendation

5.1 Conclusion:

Manual calculations of design parameters are found to be in agreement with those obtained from Hysys software. Hysys software can be used for research and development (R and D) to investigate the column hydrodynamics.

5-2 Recommendation:

1- Control tuning and stability of the system have to be investigated.

2- Further work may be carried out for design of a sieve tray for multicomponents systems.

3- Flooding against column operation conditions needs to be investigated.

4- It needs some modification to be comprehensive with higher precision.
References


3. www.owlnet.rice.edu/~ceng403/hysys/intro.html

4. http: lorien. ncl.ac.uk.distill.distiltyp.htm


8. (www.digitalengineeringlibrary.com)


10. copyright@1972 G.W.Gover,Sc.D and K.Aziz ,PHD. The flow of Complex Mixtures in pipes.


Appendices
Hysys Operation

Date input:

Setting Session Preferences

Defining a Fluid Package
Specifying the Feed Conditions

![Workbook - Case (Main)]

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Include Sub-Flowsheets
Show Name Only
Number of Hidden Objects: 0

Installing the Column

![Reflected Absorber Column Input Expert]
Entering the Initial Estimate Values

Entering inlet, outlet Streams and Number of Trays
Temperature Profile Page

Estimated Vapor Rate & Reflux Ratio
FEED = 672.9 kmol/hr
0.96 H2O
0.033 Acetone

Top, T 57°C
P, 1 atm

No of stage 16-25

Bottm T, 106°C
P, 1.26 bar

D = 23.60 kmol/h

98 w/w Acetone.

R = 1.35

B = F - D