Heat Pump in Multicomponent Non-Ideal Distillation

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by

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CERTIFICATION

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ABSTRACT

This study deals with reducing energy requirement in multicomponent distillation processes using heat pump technique.

Heat pump with and without split tower technique was considered for two ternary (Methanol-Ethanol-Water and Acetone-methanol-Water), and one quaternary feed systems (n-Hexane – MCP – Ethanol and Benzene)and many variables have considered such as operating pressure, feed composition, and fractional recovery.

The separation was carried out using one configuration for each feed system according to their non-ideality.

In all cases the feed systems were assumed as a liquids at their boiling point and four degree of recoveries of 0.9, 0.925, 0.95, and 0.99 were studied, for feed systems a and b and 0.9 for feed system c.

The operating pressure of each column in configuration was assumed to be changed from 0.5-3.0 atm in order to minimize the total energy consumption.

With Split tower technique column 1 assumed to be operated at 0.5 atm and column 2 was operated at different operating pressure for six times from 0.5 to 3.0 atm, then the first column was assumed to be operated at 1.0 atm and the second changed from 0.5 to 3.0 until column one reached the operating pressure 3.0 atm this means that 36 possibilities were studied for feed system a and b and 216 possibilities for the quaternary feed system c.

Total reboiler load, percentage saving in total energy consumption and percentage saving in total annual cost have been considered.

Computer programmes written in FORTRAN 90 language were developed for the design of multicomponent distillation with and without heat pump system with and without split tower technique.

The results show that the percentage reduction in energy requirements using heat pump was ranging from 12.14% to 76.02% when the same operating pressure was considered in each column in the configuration. While with split tower technique the percent reduction ranging from 12.02 % to 78.00 %. Also this technique give a percentage saving in total annual cost ranging from 10.00% to 77.36% for all cases studied.

The present work was compared with previous work using heat integration [25] for the same systems and the same variables.

The results show that when using heat pump technique there is an average of 20% saving in total energy consumption over that with heat integration.

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Appendices

Appendix A

Nomenclature

Symbol	**Definition**	**Dimensions**
$[A_{BC}]$	Tridiagonal matrix	-
A, B, C	Antoine equation coefficient	
A_C	Area of condenser	m^2
A_j	Lower diagonal element of matrix $[A_{BC}]$	
A_R	Surface area of reboiler	m^2
A_S	Area of subcooler	m^2
A_{RC}	Area of reboiler condenser	m^2
B_j	Diagonal element of matrix $[A_{BC}]$	
b_1, b_2, b_3	Coefficients of vapor enthalpy coefficient	
C_A	Cross sectional area of the column	m^2
CAN	Total annual cost	ID
C_C	Cost of condenser	ID
C_{comp}	Cost of compressor	ID
C_{EH}	Equipment cost	ID
C_j	Upper diagonal element of matrix $[A_{BC}]$	
C_R	Cost of reboiler	ID
C_{RC}	Cost of reboiler condenser	ID
C_S	Cost of steam	ID
C_T	Cost of distillation column	ID
C_W	Cost of cooling water	ID
C_1	Cost of distillation column with dimension parameter 100	ID
C_2	Cost of tray within area 9.29 m^2	ID
C_3	Cost of heat exchanger with surface area 92.9 m^2	

c_1, c_2, c_3	Coefficients of the liquid enthalpy equation	
D	Distillate rate	kg mole / hr
{D}	Column vector	
E_j	Equilibrium function	
E_1	Slop of column diameter parameter versus cost curve	
E_2	Slope of tray area versus cost curve	
E_3	Slope of heat transfer area versus cost curve	
F	Feed rate	kg mole / hr
H	Enthalpy of vapor	kJ/ Kg mole
h	Enthalpy of liquid	kJ/Kg mole
i	Component number	
j	Stage number	
K	Equilibrium ratio y/x	
k	Iteration number	
L	Liquid flow rate	kg mole/hr
$M_{i, j}$	Material balance function	
m	Number of component	
N	Number of theoretical plates	
N_m	Minimum number of plates	
P_T	Total pressure	atm
p_i	Vapor pressure	atm
p_j	Auxiliary quantities in the tridiagonal matrix algorithm	
Q	Heat load	kJ/hr
q	Ratio between the heat required to vaporize the feed to the latent heat of vaporization	

R	Reflux ratio	
R_C	Fractional recovery	
R_m	Minimum reflux ratio	
S_j	Summation function	
T	Temperature	C°
U	Overall heat transfer coefficient	$kJ/hr.m^2.C°$
V	Vapor flow rate	kg mole / hr
W	Bottom product flow rate	kg mole / hr
x_i	Liquid mole fraction of component i	
y_i	Vapor mole fraction of component i	
Z_i	Mole fraction of component i in the feed	

Greek	**Definition**
α_{ij}	Relative volatility of component i with respect to component j
Δ	Difference in quantity
ϕ	Parameter in Underwood equation for calculating minimum reflux ratio
λ	Binary interaction forces
Λ	Wilson constant
Φ	Parameter in Underwood equation
ε	Compressor efficiency
γ_i	Activity coefficient

Subscripts	**Definition**
A, B, C, D	Components
b	Bottom product
c	Condenser

d	Distillate
f	Feed
i, j	Components
j	Stage number
L	Liquid
lm	Logarithmic mean
m, min	Minimum
R	Reboiler
T	Top
V	Vapor

Chapter One

Introduction

The original meaning of energy conservation is related to the first law of thermodynamics, which states that energy is always conserved, never destroyed, but changes from one form and level to another. In the chemical industry, the meaning of energy conservation includes conserving the temperature level of the energy and in consequent the availability of the energy to produce work.

Separation technologies include distillation, extraction, adsorption, crystallization and membrane-based technologies in addition to a few more. These processes play a variety of roles in industry: the removal of impurities from raw materials, the purification of products from by products and the removal of contaminants from air and water effluents.

Overall, these processes account for 40 to 70% of both the capital and operating costs of a broad range of industries. Separation operations significantly impact energy consumption, manufacturing profits and product costs. For example, of the 5.8 quads (1 quad = 10^{15} Btu, or 170 million bbl of oil) of energy consumed by the U.S. chemical process industries annually, about 43% goes for separation processes [59].

Of the various separation techniques, distillation is clearly the master, accounting for more applications than all the others combined. In fact, about 95% of all separations are made with distillation, and more than 95% of the energy consumed by separation processes goes for distillation.

Therefore distillation processes require large amounts of work and heat energy to perform the required separations, these processes are prime areas for better energy utilization.

The greatest energy reduction in the immediate future can be accomplished by operating existing distillation systems more efficiently. Over the years, there have been many searches for lower energy alternatives or improved efficiencies in existing separation and purification technology.

This study concerns with saving of energy using heat pump system, in this system a compressor is used to recycle the latent heat from overhead or flashed vapors and then recompress the vapor to conditions suitable for driving the reboiler at the bottom of the tower.

Split tower technique was considered with and without heat pump where this can offer significant energy saving over a conventional distillation column. A split tower arrangements consists of two columns operates at different pressures and by using this arrangements, we have cut the energy used almost in half.

To design a distillation column for separating non-ideal multicomponent mixtures two methods were used:

1. Short-cut method.
2. Plate-to-plate method (rigorous method).

The first method is an approximate method gives the first estimation of variables that can be used in rigorous method.

The design calculation requires the prediction of the vapor-liquid equilibrium in terms of stage temperature, pressure, and phase composition.

For an ideal solution, the equilibrium relationship for any component in an equilibrium stage is defined in terms of distribution coefficient K,

Where

$$K_i = \frac{y_i}{x_i} \quad \text{..} \quad (1.1)$$

For non-ideal solution, additional variable is added γ_i (activity coefficient) represent the degree of deviation from ideality liquid phase only

$$K_i = \frac{y_i}{x_i} = \frac{\gamma_i p_i}{P} \quad \text{……} \quad (1.2)$$

Many methods are presented in the literature that predict the activity coefficient, Wilson model [84] was the most popular model because Wilson equation has found wide application, both in the correlation of binary data and in the prediction of multicomponent vapor-liquid behavior. Its popularity is based on it's simplicity of expression coupled with accuracy comparable to that obtained by other new complex models.

The aim of this work:

The aim of this work is reducing the energy requirements in multicomponent distillation by applying heat pump with and without split tower technique. This was done by studying effects of operating pressure, feed composition and fractional recovery on two ternary and one quaternary non-ideal feed systems and three computer programmes were used in the calculation.

Chapter Two

Literature Survey

2.1 Introduction:

Distillation is used for many commercial processes, such as separation of row gasoline, demenaralize water, alcohol, paraffin, kerosene, and many other liquids. Therefore many techniques have been proposed for the reduction of the energy requirements of distillation processes. Heat pumping may be one of the most important schemes for reduction of energy requirements. The heat-pumped distillation column has been used in industrial application for many years. Many authors have investigated vapor-recompression schemes in the past few years, and that will considered in this chapter.

The design of non-ideal multicomponent distillation system requires two main parts; first is the prediction of deviation from ideality and second is the application of the two design methods short-cut and rigorous plate to plate calculation [26].

2.2 Definition of Energy of Separation:

Large amounts of heat are used every year by distillation plants. So it is important to see whether any of this heat can be saved. Energy supplied to separation process may be in the form of mechanical, chemical, electrical or thermal energy depending upon the type of the process [3].

Fundamentally, distillation consists of unmixing a mixture. In theory energy is required because of the decrease in entropy of the system which is occurred when it is separated into its components. At any rate the only work required is that equivalent to the entropy change in mixing. In practice, extra work is required owing to

a. Inefficiencies of the separation process.

b. Heat losses.

To determine the efficiency of a process the ideal or theoretical energy requirement must be found. The least work will be required by a process which is isothermal and reversible, although it is only the initial and final states of the substance involved which effect the energy requirement [49, 40]

2.3 Energy Conservation Schemes for Distillation Processes:

Distillation is the workhorse of the chemical and petroleum industries, for producing high purity chemicals and for recovering organic solvents from waste streams. Unfortunately, distillation is energy intensive and inefficient. It consumes large quantities of energy in the form of steam, cooling water, chilled water, or refrigerated brine. The greatest energy reduction could be accomplished by operating existing distillation systems more efficiently. Over the years, there have been many searches for lower energy alternatives or improved efficiencies in existing separation and purification technology and these can be categorized into three groupsfirst; improved operation with little or no capital expenditure, second: extensive modification of existing equipments, third: design of new distillation system[3].

2.3.1 Improvement the operation for Existing Conventional Columns:

Distillation is a column-type process that separates components of a liquid mixture by their different boiling points. In distillation column vapor and liquid phases flow counter currently within the mass transfer zones. The column is fitted with trays or packing to maximize contact between the phases. Some of the overhead streams are returned as a liquid reflux .A reboiler at the bottom provides a vapor stream to the bottom plate. To enhance performance and reducing the energy requirement we can lower the reflux ratio, and increase the number of trays [43].

2.3.1.1 Reflux ratio:

Some of the overhead stream has to be returned to the column as a liquid reflux to enhance the performance. For minimum energy consumption, a distillation column must be operated at a calculated reflux ratio with overheads and bottoms at minimum quality requirements. (Quite often, however, operators in the interest of smooth safe operation will carry excess reflux rates at any feed rate). Provide tools and incentives to achieve minimum reflux ratio and to minimize energy consumption. Tools are good instrumentation, good analytical measurements, and good operating instructions.

In distillation, most of the reboiler energy is transferred to the cooling fluid at the condenser and is lost from the process, in conventional distillation; the energy produced at the reboiler is approximately equal to that lost at the condenser.

Thus distillation energy may be saved by implementing technologies that reduce the reflux ratio. In practice the optimum reflux ratio, R_D (optimum) is 1.1 to 1.5 R_D (min). The number of theoretical stages and reflux ratios has opposite effects: At total reflux (where there is no flow of feed as well as withdrawal of product streams), the numbers of stages are minimum. And infinite numbers of stages require only minimum reflux. Hence optimum values have to be used in the design, therefore the optimization of the reflux ratio of the distillation column can produce significant energy saving [56].

2.3.1.2 Quality Specifications:

Overhead and bottom purity specifications should always be challenged. Small changes in these concentrations can result in considerable changes in energy requirement. For example a 70-tray distillation column for a system of relative volatility of 1.4 is designed for an overhead concentration of 98% and a bottoms concentration of 0.4% when operating at 99% overhead 0.3% bottoms will result in an 8% increase in energy consumption. Conversely, a decreased overhead and or bottoms concentration can result in significant energy savings [19].One should be more analyzers to produce only the purities really needed instead of wasting steam to play it safe and over purifying the chemicals [55, 29].

2.3.1.3 Incorrect feed plate location/proper feed tray:

With wrong feed-tray, mixing of liquids of different composition will reduce column efficiency and increase steam consumption [19, 34, 54].

If the original design feed temperature or composition of the products has changed, it may be desirable to recalculate the optimum feed points. Most continuous columns have multiple feed points that can be readily changed. For four component feed stock of composition 0.25, 0.25, 0.25, 0.25, feed temperature of 76.5C°and recovery of light and heavy key components of the feed is 0.995, 0.995, respectively. The optimal feed plate location was 10; reboiler load was 8.47 million kJ/hr and steam consumption 0.2003 kg/hr.

Changing the recovery to 0.95 for both light and heavy key component the optimum feed location was 6 the reboiler load was 7.9 million kJ/hr and steam consumption was 0.188 kJ/hr.

2.3.1.4 Feed optimization:

A distillation column feed can vary from sub cooled liquid to superheated vapour. The thermal condition of the feed is an important parameter in the design of a distillation column because changes in the condition can affect both the capital and operating costs for a given system.

Giving examples, Patterson and Wells [47] showed how operating costs can change significantly with changes in the feed condition indicating that, in the optimization of a distillation system design, the feed condition cannot be ignored. Heating feed can reduce load on reboiler, but economics strongly dependent on split between overhead and bottoms. For 80% overhead, 20% bottoms split; a vapour feed gives greatest savings. For 20% overhead, 80% bottoms split, negligible savings result, in vaporizing portion of feed [54].

2.3.1.5 Column auxiliary:

Improperly operated distillation column auxiliaries, such as steams traps and vacuum steam jets, can result in a sizable energy waste. Therefore the most energy waste is steam leakage from bad steam traps and leaking fitting .steam traps are blamed for inefficient and causing as much as 10% of the generated heat from steam to be lost [64].

2.3.2 Extensive modification of existing equipment

This is particularly useful for columns operated at rate significantly below design feed rates such as columns with large safety factors, due to the lack of good relative volatility data during the original design calculation.

The solution is to repair the column internals to the newer and more efficient trays [64].

2.3.3 Design of new systems:

The design of a distillation system should have an opportunity to look at the whole process, since the design of reactors, extraction equipment, and other non distillation equipments, often has a large effect on energy requirements for the actual distillation steps, for instance, by altering reactor conditions it is sometimes possible to recover exothermic reaction heat at a temperature level high enough for it to be used in distillation column reboilers and feed preheaters.

The designer should start with premise that exothermic processes will supply all heat energy requirements. Then, he should design a process to accomplish this and only add outside energy to eliminate the capital that cannot be justified by the energy saved [64].

2.3.3.1Vapor-Liquid Equilibrium data:

It is necessary to have reliable vapor-liquid equilibrium data, such data is expensive but it is even more expensive to design large safety factors into equipments.

Better vapor-liquid equilibrium data are needed and essential for a successful design of a distillation column [64].

2.3.3.2 Optimum control:

As man-power, energy and feed stock costs increase, the designer can spend more money to keep the columns under a tight optimum control conditions.

2.3.3.3 Designing column internal:

Since energy costs increase continuously every year, column internals should be designed to operate over greater ranges of possible rates at minimum steam usage:

1. Packed towers and valve trays should be used where dual flow is encountered where sieve trays have insufficient turn down.
2. When very large single train columns are being considered it is some times more advantageous to install two columns each holding half of the load.
3. When multiple distillation trains are necessary, it is sometimes best to design some to run at maximum capacity and the other to have extreme flexibility for handling capacity swings [64].

2.4 Energy Conservation:

2.4.1 Interreboiler and Intercondensers:

The generally accepted approach of applying heat only at the bottom of the tower, and withdrawing heat only at the top, is most often directed by the economic and operability requirements imposed on the design; it is a conventional practice. In situations where energy costs are low, the thermodynamic inefficiencies inherent with this approach are usually not worth reducing. However, in multistage distillation, it is possible to add and remove heat at numerous locations in the distillation column. It is theoretically possible- but seldom practical- to apply this concept to each equilibrium stage in the column by adding finite quantities of heat to every stripping stage, and removing finite quantities of heat from every rectification stage.

When the same amount of energy is divided up and added to several intermediate points between the feed tray and the bottom tray the temperature levels of the energy can be progressively lower as the feed tray is approached. Interreboilers and intercondensers when applied in accordance with typical economic and operating criteria can produce significant reductions in the operation cost of a distillation column. Using multiple condensers and multiple reboilers can have significant effects on the design of a distillation column itself [57, 71].

This system was studied by many workers: Benedict [6], Timmers [63], Kayihan [33], and Sabarathinam [56].

2.4.2Heat integration method:

Heat integration means using the heat removed from the top of one column as the heating source for another column, under certain temperature difference between the heat source and heat sink. This must be done while the column is operating under the same pressure or different pressures. The aim of this method is reducing the external heating and cooling source by interchanging the process streams, this was studied by Freshwater [17], King [34], William and Thomas [70], Nakkash [40] Westerberg [69], Saxena [58].

2.4.3Vapor recompression or heat pump method:

Vapor recompression consists of taking the overhead vapors of a column, condensing the vapor to liquid, and using the heat liberated by the condensation to reboil the bottoms liquid from the same column. The temperature driving force needed to force heat to flow from the cooler overhead vapors to the hotter bottoms product liquid is set up by either indirect vapor recompression, where the heat at top of column is transferred to an auxiliary medium, which takes it to the bottom and release it there, or direct vapor recompression, either by using the top vapor as working fluid through (vapor recompression) thus exchanging the heat between the top and the bottom of the tower, or using the bottom vapors as a working fluid by heating and flashing the bottom against the condensing top vapor , and injecting them directly into the column bottom . This was studied by Robinson and Gilliland [54], Freshwater [17, 18],Flower and Jackson [14], Pratt [49], King [34], Wolf and Weiber [73], Null [46], Danziger [9], Wisz et al [72], Frederic and Alexandra [16], Neill et al [44], Lynd and Grethlein [37], Meilli and Guxens [38]Nakkash and al Ramadhani [41].

2.4.4Multieffect method:

The condensing overhead vapors of one distillation column can be used to provide the reboiling duty of another column, where the condensing temperature is higher than the reboiling temperature. This creates in effect the equivalent of a multieffect evaporator system, except that the distillation columns are used, rather than the direct evaporation.

This may be used for multicomponent separation where each column is used to separate one of the components, or for binary separation in which a single feed is divided to a number of parallel columns.

Freshwater [17] studied a system of ethanol/water mixture and reported the limiting effect of temperature drop across the columns using multieffect method by starting with feed to the first column at atmospheric pressure and fixing the driving force (ΔT)in each heat exchanger equal to $5C^{\circ}$. Each column operates under different pressure and provides heat for the next column, the first column receive heat from an external source.

If the feed entering is liquid at it is boiling point for each column then the heat required in each case is exactly one quarter of that was used previously and the efficiency four times increased.

Nakkash [40] studied multieffect principle using ethanol/water/and benzene system, there was 20% saving of the energy used in this technique than for conventional process arrangement.

2.5 Energy conservation in multicomponent distillation for non-ideal systems:

There are many methods presented deal with the problem of energy saving in multicomponent distillation system .In principle the same methods for ideal system may be applied for non-ideal systems.

2.5.1 Energy integration through heat stream matching:

The concept of energy integration involves the matching or sharing of the heat streams where ever possible within a configuration.

Thus, heat stream matching may be possible between the sensible heat of the feed stream, with the heat load of the overhead condenser and reboiler and with the heat load at the intermediate heaters and coolers. The feasibility of energy matching depends on the amount of energy available at each source and the amount required for each sink, therefore imbalance heat loads between streams to be matched is not less than $5C^{\circ}$.

Nakkash [40] studied the separation of non-ideal mixtures of (acetone, cumene, and phenol), according to Heaven [24] equation, two columns are required for the separation and two configuration are possible, but it was found that the previous rule considered by Heaven [24] can not be applied to highly deviated non-ideal mixture. This because the separation needed a very high reflux ratio and high number of theoretical stages. This will increase the reboiler load for the two configurations. The high reflux is due to the non-ideality.

Nakkash and Hababa [42] studied the heat synthesis of heat integration between reboilers and condensers into ternary and quaternary non-ideal feed systems and the results was also agree with the previous works .

Energy integration was also studied by several workers for ideal systems (Hwa [30], Nishida [45], Rathore [52], Nakkash [40], Pibouleau [48], Isla [31], and Naka [39]) and a few workers have been studied energy integration for non-ideal systems Nakkash [40], Hababa [25].

2.5.2 Multieffect method:

It is possible to employ multieffect operation distillation as practiced with evaporator, using the heat rejected from the condenser to the reboiler of a subsequent column figure (2.3). This system may be used for multicomponent separation where each column is used to separate one of the components in which a single feed is divided to a number of parallel column.

Each column must be operated at a lower pressure than the previous one, to enable a positive temperature difference to be maintained in each

reboiler –condenser. This process is applicable to low strength feed, especially with non-ideal system with positive deviation from Raoult's law taking advantage of the fact that minimum reflux ratio decreases with increasing the feed concentration.

Consequently, the feed is passes into a stripping column operating at a relatively high reflux ratio and the overhead vapor is used to provide the boil up in a center feed column operating at lower pressure. This method was studied by several workers: Freshwater [17], King [34].

2.5.3 Split Tower:

A split tower arrangement consists of splitting the feed into two equivalent streams and distilling in two smaller columns. The two columns operate at different pressures, one higher than the other, resulting in its overhead vapor having a condensing temperature high enough to be able to use the condensing vapor to provide the reboiling duty in the lower pressure column. The bubble point temperature of the overhead vapor must be high enough over the bubble point of the lower pressure reboiling bottoms to provide a sufficient ΔT for the condenser-reboiler. The feed stream will be split so that the condenser duties of the high pressure column approximately match the required reboiler duty of the low pressure column. The heat input to the reboiler, of the high pressure column rises to the condenser where it then provides the reboiling duty of the other column. By use of the split tower arrangement, we have cut our energy use almost in half; note that instead of two columns, any number of columns can be used in the split tower fashion.

However, for each additional tower, an extra ΔT must be supplied, plus the temperature drop across the column. In addition, the energy savings drops as each column is added. The two-tower system saves 50% of the energy [59].

Abu – Eishan [2] considered the reduction of energy consumption using split tower arrangement figure (2.4). This arrangement can be afford significant energy savings over a conventional distillation column. It consists of splitting the feed into two equivalent streams and distilling in to smaller columns. The two columns operated at different pressures one higher than the other, resulting in its overhead vapor having a condensing temperature high enough to be able to use the condensing vapor to provide the reboiler duty in the lower pressure column.

Hababa [25]studied this technique for two ternary and one quaternary non-ideal system. His results show that when operating the towers at the same pressure (high or low), there is no energy integration thus very low percentage saving in total energy consumption was obtained. While when operating at different pressures energy integration takes place, which give high percentage saving in total energy consumption.

2.5.4 Thermal coupling system:

A distillation system contains a thermal coupling when a heat flux is utilized for more than one fractionation and the heat transfer between fractionation sections occurs by a direct contact of vapor and liquid, compared with a conventional system.

Thermally coupled distillation systems can separate close boiling components with considerable saving of heating and cooling costs.

Tobyia [64], studied thermal coupling system dealing with four components feed stocks ideal mixtures , many variables have been discussed , four feed types , five different compositions , two fractional recoveries 99% and 95% , three distribution factors , all these variables were studied on twenty seven configuration. The results showed that the percentage saving in cost and energy saving varying from 1 to 86%.

2.5.5 Heat pump or vapor recompression process:

This process consists of putting heat into the bottom of a column and taking it out again at the top. The vapor recompression cycle has a set ratio between available condenser sides to reboiler duty. The reboiler heat flow obtained will be equal to the sum of condenser duty plus the work added to the gas stream and its inefficiencies. In all cases where the column reboiler and condenser duties do not match in the manner stated an auxiliary system would be needed to supply the excess column condenser or reboiler duty.

The advantage of vapor recompression lies in its ability to move large quantities of heat between the condenser and reboiler of the column with a small work input. This results from cases where there is only a small difference between the overhead and bottoms temperature. A conventional column with steam heating and water cooling may use ten times the Btu's of a column running with vapor recompression.

Heat pumping systems for distillation column may operate between the condenser as a source and the reboiler as a sink, either directly by compression of the overhead vapor or indirectly by using a secondary heat transfer medium.

Therefore vapor recompression process can be divided into two processes, these processes were studied by many workers, Robinson and Gilliland [54], Freshwater [17, 18], Flower and Jackson [14], Pratt [49], King [34], Null [46], Danziger [9], Wisz et al [72], Frederic and Alexandra [16], Neill et al [44], Lynd and Grethlein [37], Guxens [22], Nakkash and Al-Ramadany [41].

2.5.5.1 Direct vapor recompression process:

In this process was the vapor from top of the column are compressed to a temperature above the temperature of the bottom, by pumping and condensing in the combined condenser and reboiler Figure (2.1) thus providing the necessary heat for distillation. This type of process is used when distillation involves a relatively close - boiling mixture.

The thermodynamic efficiency was calculated using direct vapor recompression for a column separating phenol / cresol at minimum reflux without heat recovery and assuming the compressor to be hundred percent efficient, the value of this efficiency was 36%, this represent a considerable improvement while the reported thermodynamic efficiency without heat recovery by direct vapor recompression was 11.5% [3].

2.5.5.2 Indirect vapor recompression process:

In this process the vapor from the top of the column condensed in the evaporative condenser, the vapors transfer their latent heat during condensation, by reboiling the water on the water side of the condenser, the steam produced is compressed by the use of high pressure to result high temperature where it can be used to heat the bottom fraction of the column figure (2.2). The same above example given above for separating phenol /cresol to produce 94% phenol, at minimum reflux, gave a thermodynamic efficiency of 19% while without heat recovery was 11.5% [3].

Direct vapor recompression is used because of that the indirect vapor recompression requires some extra compression, since the temperature difference between vaporization and condensation of the fluid must be enough to overcome the temperature difference of the distillation and provide temperature difference driving forces for two heat exchangers. While for direct vapor recompression process only the temperature difference driving force for one exchanger need be provided in addition to overcoming the temperature difference of the distillation.

2.5.5.3 Literature review for heat pump or vapor recompression process in distillation:

One alternative to recover the waste heat in the condenser is the use of heat pump. The idea is not new. Null [46] and Freshwater [18] appear to be first to introduce it in 1950's. Later Freshwater [18] proposed several arrangements where the heat pump would extract heat from somewhere between the feed entrance and the condenser at the top.

Wolf and Weiber [73] showed that heat pumps optimized with the union carbide multiple downcomer (MD) trays and high flux tubing (high efficiency components) could reduce equipment costs by 30% and energy cost by half , when compared to heat pumps having conventional multipass valve trays and bare tube reboilers.

Null [46] proposed a simple system where the heat pump could be used to transfer heat from the vapor at the top of an adiabatic column to the reboiler at the bottom i.e. between the extreme points of the column. This was to be carried out by an external working fluid, where direct refrigeration or chiller water was needed to condense the overhead.

Patterson and wells [47] proposed an arrangement where the vapor at the top is compressed and later condensed in the reboiler and the bottom product is flashed through the expansion valve and vaporized and vaporized in the condenser. The first arrangement eliminates the condenser and has the advantage of not being restricted by the cooling medium. The second eliminates the reboiler.

Danziger [9], described process of vapor recompression in a pilot plant to measure the energy saving. The plant was running at total reflux with the mixture cis-/trans – declain, the boil up rate was 360 kg/hr, at a top pressure of 175mbar. The turbo –blower have a speed of 3600 rev/min and it compressed vapor from 175mbar to 314mbar.

This pressure corresponds to a saturated vapor temperature of 143C°, the bottom temperature was 135.5C°, and consequently the temperature difference in the reboiler /condenser was 7.5C °. The energy saved through the application of heat pump for an adiabatic column is over 80% compared with that of conventional distillation apparatus.

Quadri [51] analyzed the propane – propylene (p-p) splitter heat pump using generalized equations for the process modeling of the complete system, stresses the importance of proper selection of the compressor and reboiler–condenser. He also identified the interrelationship among power, compressor operating curve, reboiler temperature difference, and column load.

Wisz et al [72] discussed the use of these high efficiency components in the (p-p) separation and the design and operation of heat pumps with both single and dual – stage compressors. Systems with two stages of overhead compression, although somewhat more complex, offer an additional degree of energy saving over the single stage since only a small fraction 10% or less of the compressed vapor need to be elevated to a pressure high enough to reject the heat of compression for cooling water or air.

Frederic and Alexandra [16] considered recovering energy by mechanical vapor recompression. An example was given for extraction of xylene in a refinery plant using (MVRS) give saving nearly 47.3% in total annual cost.

Neill et al [44] discussed the use of high efficiency components with the operation of heat pumps in separations that involved close boiling components (mixtures with closed relative volatilities), like (p-p) separation or iso –n-butane or mixed butanes. The result of using heat pump system gives a large reduction in cooling water or air and elimination of steam or other heating media for the reboiler.The heat of compression and other sensible heat affected in the system must be balanced by trim cooler. Trim cooler heat load is usually a small fraction of the reboiler load. The energy of compression may be minimized by design for a small temperatures difference across the reboiler (thus reducing the compressor discharge pressure), and sub-cooling the condensate, which in turn reduces flash vapor recycle when the reflux is returned to the column.

An example of operation rate of 120,000 Mt/yr of propylene at a purity of 0.995% in (p-p) splitter, the feed basis was 72.3% propylene, 26.7% propane, and 1% C4. The total equipment cost for conventional system was $3,210,000. A system with heat pump was $2,630,000, so the saving in cost was 30%.

Lynd and Grethlein [37] presented a new approach for using heat pump; distillation with intermediate heat pumps and optimal side stream return (IHOSR distillation). The (IHOSR) strategy appears to be most attractive in cases with either a dilute feed or a large column temperature drop. Three examples considered using (IHOSR method), the first involve separating ethanol-water, heat was moved from the rectifying section to the reboiler, and vapor is withdrawn from the distillation column, compressed, condensed in the reboiler, and returned at a level above the point of withdrawal. The second example involves separating 10 wt% ethanol – water feed to distillate and bottoms of 95 wt%. And 0.1 wt% respectively , by moving heat from the overhead vapor to the rectifying section , liquid removed , vaporized , and returned to the column as vapor below the point of side stream withdrawal. The third example involves separating of n – hexane /n – octane mixture by removing both vapor and liquid intermediate column sections and return them according to the (IHOSR method).

These examples when compared with overhead to reboiler (OTR) heat pump method and conventional distillation, the results of comparison was the first example, both heat pumps methods are preferred over conventional distillation.

Meilli and Guxens [38] showed that vapor recompression technique was extremely economical solution to save energy, this was especially true whenever the distillation column were equipped with regular packing which have a low pressure drop. They considered three examples, C3 – splitter, ethylene styrene separation and EDC – purification, and they have seen that the percentage saving in energy for the three examples were 86%, 83%, 81% respectively.

Guxens [22] presented a simulation program for distillation columns with vapor recompression together with an optimization algorithm based on the box's complex algorithm, and using this program for both high and low volatility systems. They proposed an optimal deign configuration for each case. Four design parameters and four operation variable have been optimized to take into account both investment and operating cost in the objective function.

Nakkash and Al-Ramadhani [41] studied heat pumps in multicomponent distillation of nearly ideal hydrocarbon systems. They found that heat pump system gave a saving in total annual cost from 5% to 77% over the conventional system and the percentage reduction in exergy losses with heat pump was 27% to 97% over the conventional system.

2.6 Thermodynamic properties and models:

Thermodynamic properties and models play a major role in separation operations, particularly with energy requirements, phase equilibria, and sizing equipments.

Vapor liquid equilibrium calculations are usually carried out for separation processes, with several versions of the EOS. The prediction of mixture vapor-liquid equilibrium is more complicated than the prediction of pure components.

Phase equilibrium relation is one of the fundamental properties which are necessary for the calculation of the separation processes, and useful equations have been proposed for expressing these relations.

2.6.1 Fundamental equations for the VLE relation:

To describe the phase equilibrium of a system of N components at a temperature T, pressure P, and at equilibrium, the vapor phase fugasity is equal to the liquid phase fugasity for every component:

$$f_i^L = f_i^V \qquad \text{for} \qquad i=1,2,3,\ldots\ldots\ldots N \tag{2.1}$$

Equation of state in both phases

$$\hat{\Phi}_{iL} = \frac{\hat{f}_{iL}}{x_i P} \quad , \quad \hat{\Phi}_{iV} = \frac{\hat{f}_{iV}}{y_i P} \tag{2.2}$$

$$\hat{\Phi}_{iL}\, x_i = \hat{\Phi}_{iV}\, y_i \tag{2.3}$$

$$K = \frac{\hat{\Phi}_{iL}}{\hat{\Phi}_{iV}} \tag{2.4}$$

Equation of state in vapor phase and activity coefficient in liquid phase

$$\hat{f}_{iV} = \hat{f}_{iL} \tag{2.5}$$

The vapor phase fugasity can be written in terms of the vapor phase fugasity coefficient Φ_i^V vapor mole fraction y_i and total pressure P as following

$$\hat{f}_{iV} = \hat{\Phi}_{iV}\, y_i P \tag{2.6}$$

$$\hat{a}_{iL} = \frac{\hat{f}_{iL}}{\hat{f}_{iV}} \tag{2.7}$$

Also the liquid phase fugasity can be written in terms of liquid phase activity coefficient γ_i and liquid mole fraction x_i as following:

$$\hat{f}_{iL} = \gamma_i\, x_i\, f^{\circ}_{iL} \tag{2.8}$$

$$\ln \frac{f^{\circ}_{iL}}{f^{\circ}_{isat}} = \frac{V}{RT}(p - p_i^{sat}) \tag{2.9}$$

$$\hat{f}_{iL} = f^{\circ}_{sat} \tag{2.10}$$

$$\Phi_i^{\circ sat} = \frac{f_i^{\circ sat}}{p_i^{sat}} \tag{2.11}$$

$$\hat{f}_{iL} = \Phi_i^{\circ sat}\, p_i^{sat} \exp^{\frac{V}{RT}(p - p_i^{sat})} \tag{2.12}$$

$$\hat{\Phi}_{iV}\, y_i\, p = \gamma_i\, x_i\, \Phi_i^{sat}\, p_i^{sat} \exp^{\frac{V}{RT}(p - p_i^{sat})} \tag{2.13}$$

2.6.2 Ideal vapor liquid equilibrium:

A mixture is called ideal if both liquid and vapor are ideal mixtures of ideal components, thus the partial pressure of component i, p_i in the vapor phase is proportional to its mole fraction in the vapor phase according to

$$P_i^{\circ} = y_i P \tag{2.14}$$

For an ideal mixture, the equilibrium relationship for any component is defined as:

$$K_j = \frac{y_i}{x_i} \tag{2.15}$$

The K_j value can be predicted from Raoult's law, where:

$$K_j = \frac{p_i^{\circ}}{P} \tag{2.16}$$

Ideal solutions occur when molecular diameter is equal, chemical interactions are present, intermolecular forces between like and unlike molecules are equal, and in which all the activity coefficients are unity (i.e. $y_i = 1$ for all i).

Ideal gas mixture is an ideal solution, and any equation applying to ideal solution can be also applied to ideal gas mixture. The converse is not true; there are many ideal solutions that are not ideal gases. The ideal gas mixture is an imaginary gas mixture and every component of the ideal gas mixture obeys the ideal gas law equation (2.17) in pure state as well as in mixture environment.

However, the real gases do not obey the ideal gas law expect at very low pressures.

$$Z = \frac{PV}{RT} = 1 \tag{2.17}$$

2.6.3 Non ideal vapor liquid equilibrium:

Deviation from ideality may occur in the liquid and in the vapor or in both. Such changes as usually more prominent in liquids because molecules are much closer together in liquid than in vapor.

Vapor phase in non-ideal solutions usually behave approximately as perfect gases at pressure below 3 atm.

For non ideal systems addition variables γ_i (activity coefficient) appears in vapor-liquid equilibrium

$$y_i = \frac{\gamma_i \overset{\circ}{p_i}}{P} \cdot x_i \qquad (2.18)$$

Where γ_i represent degree of deviation from reality. When $\gamma_i = 1$, the mixture is said to ideal simplifies to Raoult's law. For non-ideal mixtures $\gamma_i \neq 1$, exhibit either positive deviation from Raoult's law ($\gamma_i > 1$), or negative deviation from Raoult's law ($\gamma_i < 1$).

Activity and activity are closely linked with the theory of excess functions. Excess functions are thermodynamic properties of solution which are in excess of an ideal solution at same temperature, pressure and composition. For an ideal solution all excess function are zero.

2.6.4 Calculation of activity coefficients:

The prediction of liquid phase activity coefficient is most important for non-ideal distillation design calculation.

There are several excess energy g^E models to calculate the activity coefficient in multicomponent systems. The most important models are (Wilson, NTRL, UNIQUAC, and UNIFAC).In all these models, the model parameters are determined by fitting the experimental data of binary mixtures. Using binary interaction data only can make the prediction of vapor-liquid equilibrium of multicomponent mixture possible.

Each one of these models has advantage and disadvantage. The selection of appropriate model for a given mixture is based on the three characteristics, which are temperature, pressure, and composition.

2.6.4.a NTRL model:

The NRTL (non-random, two liquid model) developed by Renon and Prausnitz [53].This model has the advantage of containing an adequate number of parameters (three at a given temperature) to give a good representation of strong deviations from ideality, including liquid-liquid equilibria, for all types of non electrolytic systems. Its extension to multicomponent mixtures does not require additional parameters, only parameters for binary interactions need to be known to calculate the property of any mixture of non electrolytes.

The following equations represent NTRL model:

$$G_{ij} = \exp(-\alpha_{ij}\tau_{ij}) \qquad (2.19)$$

$$\ln\gamma_i = \frac{\sum_{j=1}^{n_c}\tau_{ji}G_{ji}x_j}{\sum_{j=i}^{n_c}G_{ji}x_j} + \sum_{j=i}^{n_c}\frac{x_j}{\sum G_{Kj}x_K}\cdot\left(\tau_{ij} - \frac{\sum_{K=1}^{n_c}x_K\tau_{Kj}G_{Kj}}{\sum_{K=1}^{n_c}x_K G_{Kj}}\right) \qquad (2.20)$$

The NTRL group interaction parameters $\tau_{12} = b_{12}/RT, a, b_{12}, and\, b_{21}$ are NTRL constant.

2.6.4.b UNIQUAC model:`

Abrams and Prausnitz [1] develop the UNIQUAC (universal quasi chemical) activity coefficient model. This model distinguishes two contributions (C) and (R).

$$\ln\gamma_i = \ln\gamma_i^c(combinational) + \ln\gamma_i^R(residual) \qquad (2.21)$$

The combinational part basically accounts for non-ideality of a mixture arising from differences in size and shape of constitute molecular species; whereas the residual part considers the difference between inter-molecular and interaction energies.

The two parameter in UNIQUAC equation gives a good representation of the vapor-liquid equilibria for binary and multicomponent mixture.

$$\tau_{ij} = \exp(A_{ij}) \qquad (2.22)$$

$$\Phi = \frac{r_i x_i}{\sum_{j=1}^{n_c} r_i x_i} \qquad (2.23)$$

$$\theta = \frac{q_i x_i}{\sum_{j=i}^{n_c} q_i x_i} \qquad (2.24)$$

$$L_j = Z(r_i - q) - (r \; -1) \qquad \text{Where Z=10} \qquad (2.25)$$

$$\ln \gamma_i^c = \ln \frac{\Phi_i}{x_i} + \frac{Z}{2} q_i \ln \frac{\theta_i}{\Phi_i} + L_i - \frac{\Phi_i}{x_i} \sum_{j=1}^{n_c} x_i L_i \qquad (2.26)$$

$$\ln \gamma_i^r = -q_i \ln\left(\sum_{j=i}^{n_c} \theta_i \tau_{ij} \right) + q_i - q_i \sum_{j=i}^{n_c} \frac{\theta_j \tau_{ij}}{\sum_{K=1}^{n_c} \theta_K \tau_{Kj}} \qquad (2.30)$$

$$\ln \gamma_i = \ln \gamma_i^c + \ln \gamma_i^r \qquad (2.31)$$

The molecular vendor walls volume r_i and the molecular surface qi are usually calculated as the sum of the group volume and area parameters.

2.6.4.c UNIFAC model:

The UNIFAC (UNIQUAC functional group activity coefficients) group contribution method, first presented by Fredenslund [15], and Prausnitz [50].In UNIFAC model each molecule is taken as a composite of subgroups; for example t-butanol is composed of 3 "CH_3" groups 1"C" group and 1 "OH" group and Ethane, which contain two "CH_3" groups. The interaction parameters between different molecules are defined in literature.

This method is based theoretically on UNIQUAC equation (2.30). The activity coefficient consists of two parts, combinational and residual contribution.

$$\ln\gamma_i = \ln\gamma_i^c + \ln\gamma_i^r \tag{2.32}$$

Combinational contribution γ_i^c taking into account effects arising from difference in molecular size and shape while residual contribution γ_i^r taking into account energies interactions between the functional group in the mixture, th combinational parts is given in equation.

$$\ln\gamma_i^c = \ln\frac{\Phi_i}{x_i} + 5q_i\ln\frac{\theta_i}{\Phi_i} + L_i - \frac{\Phi_i}{x_i}\sum_j x_j L_i \tag{2.33}$$

The residual contribution is given by:

$$\ln\gamma_i^r = \sum_K v_K^i\left(\ln\Gamma_K - \ln\Gamma_K^i\right) \tag{2.34}$$

UNIFAC model is extensively used to describe thermodynamic in chemical engineering literature and is widely used in process simulation.

2.6.4.d Wilson model:

Wilson [71] predicted his equation to calculate the liquid phase activity coefficient.

$$\gamma_K = \ln\left[\sum_{j=1}^{n} x\right] + \sum_{K=1}^{n}\frac{x_K\Lambda_{Ki}}{\sum_{K=1}^{n}\Lambda_{Ki}} \tag{2.35}$$

Where

$$\Lambda_{ij} = \frac{v_j^L}{v_i^L}\cdot\exp\left[-{(\lambda_{ij}-\lambda_{ii})}\big/{RT}\right] \tag{2.36}$$

$$\Lambda_{ij} = \lambda_{ji} \tag{2.37}$$

$$\lambda_{ij} = \lambda_{ji} \tag{2.38}$$

$$\Lambda_{ii} = \Lambda_{jj} = 1 \tag{2.39}$$

Wilson's equation has many features:

1. System must be miscible.
2. Able to describe non-ideal systems.
3. Based on local volume fractions, which were related to local molecule segregations caused by differing energies of interaction between pairs of molecules.

4. Two interaction parameters for each component pair.

The limits of application of Wilson's equation are:

1. This equation cannot be used for a solution forming two liquid phases.
2. This equation can be applied to solutions in which the activity coefficient shows the maximum value.
3. Unable to predict immiscibility.

2.7 Design models of multicomponent non-ideal distillation

To model multicomponent non-ideal distillation columns, there are two main methods:

1. Short-cut approximate method.
2. Rigorous plate-to-plate method.

2.7.1Short-cut approximate model:

Short-cut method is used in the design of multicomponent distillation column. This method is used to get a first estimation of the number of theoretical stages; reflux ratio, top and bottom temperature, reboiler and condenser load and feed plate location. This method gives not accurate results when it is used for the design of non-ideal systems because they are based on the assumption of ideality and constant relative volatility. Therefore distillation columns for non-ideal systems must be designed by plate to plate calculation procedure because it is more accurate and convenient to give temperature composition, steam flowrates and heat load at each stage.

Smith and Brinkley [60] developed a method for determining the distribution of components in ideal multicomponent separation processes.

Their method is based on the solution of finite difference equation that can be written for multistage separation processes and can be used for extraction and absorption as well as distillation processes.

Eckert and Hlavelek [10] modified the Smith and Brinkley [60] method for use in computing non-ideal multicomponent distillation, they tested their procedure for number of strongly non-ideal systems.

They were found that the storage requirement and computer time expenditure are by a factor 100 lower in comparison with plate to plate calculation.

A developed method called "Empirical correlation" which is based on the method of Fenske [12], Colburn [7] and Gilliland [20] and Underwood [66] and Erbar and Maddox [11] to calculate the actual stage requirements for multicomponent distillation.

2.7.2 Rigorous methods for multicomponent non-ideal systems

Final design of multistage equipment for multicomponent separation requires rigorous determination of temperature, pressure, stream flow rates, stream composition and heat load at each stage.

This determination is made by solving material balance, enthalpy balance and equilibrium relation for each stage.

Lewis and Matheson [36] developed an equation tearing procedure, it was formulated to determine the stage requirements, for specifications of the separation of two key components, a reflux ratio and feed stream location.

Thiele and Geddes [62] gave a multicomponent distillation method, its calculations made in terms of mole ratio of vapour or liquid to product, it is necessary to specify the number of ideal (theoretical) stages and the reflux ratio and then assume the first temperature for each plate. This method was found to be numerically unstable when attempts were made to program it for a digital computer. So Holland [28] and co-workers developed an improved Thiele-Geddes procedure called the theta method, which has been apply with considerable success.

Amundson and Pontinen [4] show that the equations of material balance, enthalpy balance and vapour-liquid equilibrium relation equation, (MESH equations) could be combined and solve component by component rather than using a stage by stage solution procedure.

Ferraries and Donati [13] present a top-bottom Newton – Raphson procedures method for the solution of the material and enthalpy balance, and equilibrium equations in multicomponent non-ideal distillation. Their method based on a linearization of the equations and on a procedure for the solution of linearized system that operates through a sequence of computations from top to the bottom of the column.

Saito and Sugie [57] modified a successive iteration method proposed for calculating of non-ideal multicomponent distillation column with one feed.

Tsouboka and Katayma [65] developed a new procedure or the matrix method by using a method similar to a second-order Rung-Kutta procedure.

This method is more flexible for strongly non-ideal systems than the relaxation method and is as simple as the original matrix method.

Yorizane and Yoshida [74] made their calculation on relaxation method or extractive distillation for a system of Acetone – Methanol – Water using normal two column method.

Kinoshita and Hashimoto [35] developed a powerful new simulation procedure for multicomponent distillation column processing non-ideal solutions or reactive solutions.

The main calculation loop is the Newton-Raphson procedures, in which liquid mole fractions are chosen for the independent variables and the functions to be zeroed at originally defined.

This procedure presents much greater stability in finding a solution. The liquid mole fraction can be considered as key variables to be chosen for the independent variables, in other words, the liquid mole fractions initially assumed, and they are repeatedly modified at each iterative step until convergence criterion is satisfied. The most popular method using this idea is the tridiagonal matrix method developed by Amundson [4] and extended by Wang [68].

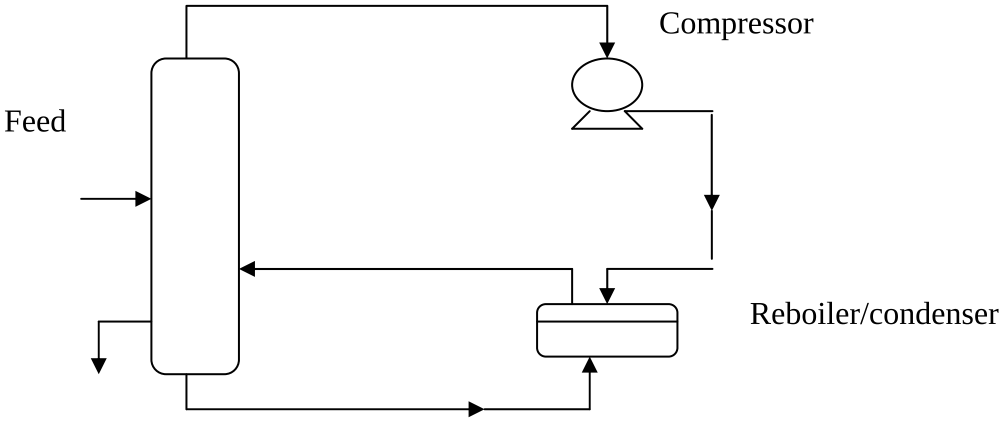

Figure (2.1) Direct vapor recompression

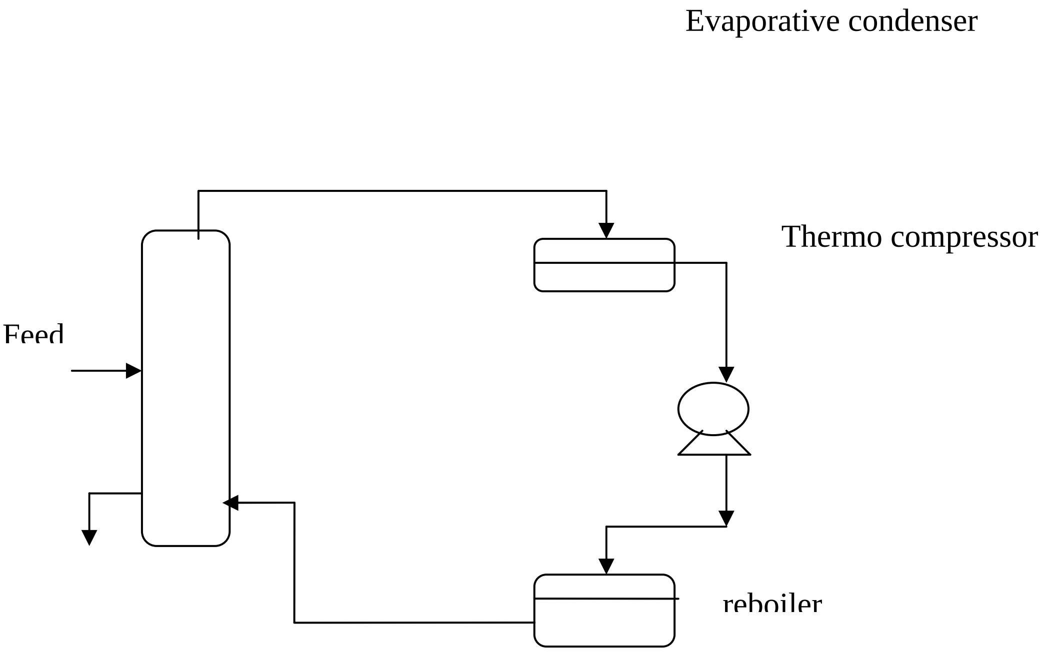

Figure (2.2) Indirect vapor recompression

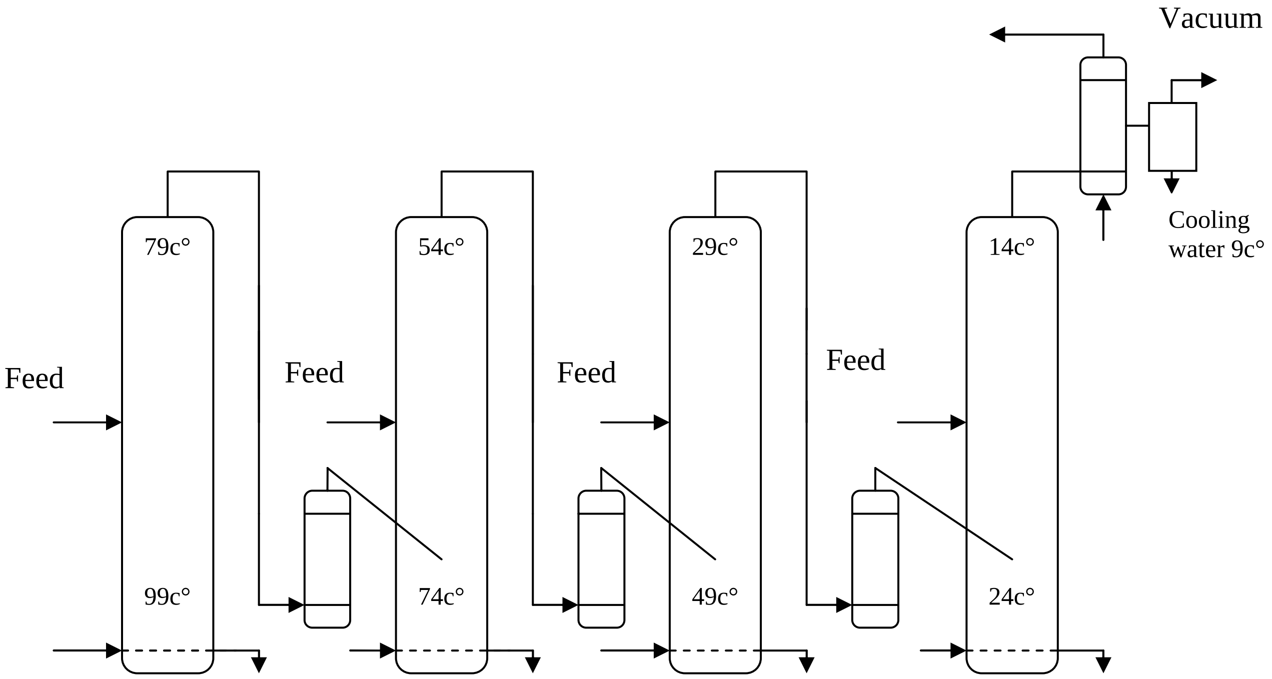

Figure (2.3) Multiple effect distillation showing limiting effect of temperature drop across column

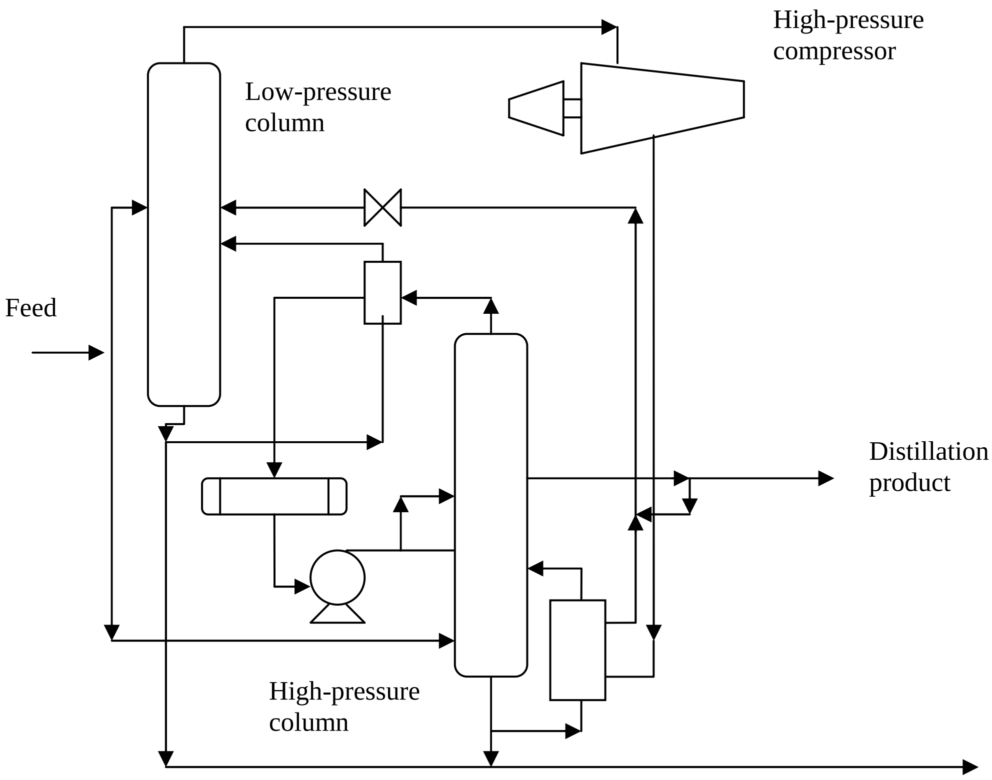

Figure (2.4) Split tower technique.

Chapter Three

Theoretical Aspects

3.1 Introduction

Distillation is a widely used method for separating mixtures based on differences in the conditions required to change the phase of components of the mixture. To separate a mixture of liquids, the liquid can be heated to force components, which have different boiling points, into the gas phase. The gas is then condensed back into liquid form and collected

In this work, non-ideal mixtures were studied where the non-ideal systems often occur and are important because they include industrially significant mixtures like ethanol / water, acetone / water and similar polar systems.

The behavior of non-ideal systems can be explained by reference to the inter-molecular forces at the surface of the liquid.

Non-ideal solutions are usually formed by components whose native environments are not the same, i.e. components that are not form the same molecular species, e.g. ethanol /water, and chloroform / acetone.

Deviation from ideality may occur in the liquid, in the vapor, or in both. In this work, the deviation was considered in liquid phase only, Wilson model is considered to calculate the degree of deviation in order to make a design analysis for non-ideal multicomponent system and the energy consumptions.

3.2 Method of Analysis:

The analysis emphasizes to investigate the possibility of energy recovery in non-ideal multicomponent distillation using direct heat pump technique to separate two ternary and one quaternary system feed stock mixtures into a relatively pure products, and how these possibilities vary when the feed composition , degrees of recovery and operating pressure vary.

The separation was carried out using different configuration for each feed system depending upon its non-ideality. The material and energy balances were carried out for each configuration using a modified computer programmes for short-cut method (empirical method) and rigorous plate-to-plate calculation.

In all cases the feed systems are assumed as a liquids at there boiling points and four degree of recoveries of 0.9, 0.925, 0.95, and 0.99 in which the flow rates, composition and temperature of all process streams within the configuration are determined together with the process design of all columns, heat load for the condensers and reboilers, number of plates, actual reflux ratio and feed plate location.

The compositions, flow rate and temperature profile for each stage were obtained using a modified rigorous plate-to-plate calculation.

Short-cut method was used to design the distillation column for non-ideal multicomponent system, to get the first estimation of the process variables, then the modified rigorous plate-to-plate program is used to design column for non-ideal multicomponent system, where each column in the configuration was designed separately.

Split tower and heat pump techniques were used as methods for analysis energy recovery in multicomponent non-ideal distillation. Modified programmes were used in order to calculate percentage saving in total energy consumption and total annual cost when using heat pump technique.

3.3 Variable Specifications:

The variables considered for the process system are:

1. Feed composition.

2. Degree of recovery.
3. Operating pressure.

Two ternary and one quaternary non-ideal feed systems table (3.1) have chosen because the physical properties and the binary interaction forces are available in the literature. Also they were studied by Nakkash and Hababa [42] and are given in appendix (C).

These systems have been studied with different feed compositions of ternary and quaternary feed systems table (3.2) and four different recoveries of 0.9, 0.925, 0.95, and 0.99, for feed systems a and b and 0.9 for feed system c for their conventional configuration, (due to their non-ideality).

All feed systems were assumed to be liquids at their boiling point, total condensation to give a maximum heat removed at the top of the column.

The operating pressure of each column in configuration was assumed to be changed from 0.5 – 3.0 atm in order to minimize the total energy consumption.

The ratio of actual to minimum reflux ratio was taken constant as a value of 1.25 as it was considered in previous studies [29].

The number of possibilities studied for heat pump system using ternary mixtures were (1 configuration * 4 feed composition * 4 degrees of recovery *36 possible operating pressure), and for quaternary system were (1 configuration * 5 feed composition * 1 degree of recovery * 27 possibility of operating pressure).

component	**Ternary feed system**	
	a	**b**
A	methanol	Acetone
B	Ethanol	Methanol
C	water	water

component	**Quaternary feed system**
	C
D	n-hexane
E	MCP
F	Ethanol
G	Benzene

Table (3.1) feed system components

component	Ternary feed composition			
	1	2	3	4
A	0.333	0.8	0.1	0.1
B	0.333	0.1	0.8	0.1
C	0.333	0.1	0.1	0.8

component	Quaternary feed composition				
	1	2	3	4	5
D	0.25	0.7	0.1	0.1	0.1
E	0.25	0.1	0.7	0.1	0.1
F	0.25	0.1	0.1	0.7	0.1
G	0.25	0.1	0.1	0.1	0.7

Table (3.2) ternary and quaternary feed system composition

3.4 Predication of vapor– liquid equilibrium for non–ideal systems:

Most experimental work has dealt with binary systems; therefore, several expressions have been presented through the years that attempt to utilize the available binary experimental vapor-liquid equilibrium data to predict vapor-liquid behavior, greatest success has been achieved with the more important expressions; Wilson, NTRL, UNIQUAC, and UNIFAC.

Among the models, the Wilson equation [71]has found wide application, both in the correlation of binary data and in the prediction of multicomponent vapor-liquid equilibrium behavior.

In this work, the Wilson model was chosen because it provides that Wilson model is a simplest model coupled with accuracy comparable to that obtained by the other more complex models.

3.4.a Wilson model:

Wilson equation [71] derived to solve the equation of excess free energy,

$$\ln\gamma_k = -\ln\left[\sum_{j=1}^{n} x_j \Lambda_{ij}\right] + 1 - \sum_{k+1}^{n}\left[\frac{x_k \Lambda_{ki}}{\sum_{j=1}^{n} \Lambda_{kj}}\right] \qquad \ldots (3.1)$$

Where

$$\Lambda_{ij} = \frac{v_j^L}{v_L^l} \exp\left[-\left(\lambda_{ij} - \lambda_{ii}\right)/RT\right] \qquad \ldots (3.2)$$

And

$$\Lambda_{ji} = \frac{v_i^L}{v_j^L} \exp\left[-\left(\lambda_{ji} - \lambda_{jj}\right)/RT\right] \qquad \ldots (3.3)$$

Although $\lambda_{ij} = \lambda_{ji}$ … (3.4)

$\Lambda_{ij} = \Lambda_{ji}$ … (3.5)

and $\Lambda_{ii} = \Lambda_{jj} = 1.0$ … (3.6)

Λ_{ij} and Λ_{ji} are Wilson's constants.

3.4.b Enthalpy of non-ideal systems:

The prediction of enthalpy of non-ideal system is important because of the excess free energy, where for ideal system the excess free energy equal zero.

$$\Delta G^E = \Delta G - \Delta G^i \qquad \dots (3.7)$$

The relationship between excess molal free energy excess molal enthalpy and entropy is:

$$\Delta G^E = \Delta H^E - T \ \Delta S^E \qquad \dots (3.8)$$

Wilson [71] gives an expression to determine excess enthalpy of non-ideal solutions

$$H^E = \sum_{i=1}^{C} x_i \overline{H}_i^E = \sum_{i=1}^{C} \left\{ x_i \left[\frac{\sum_{i=1}^{C} x_i \left(\delta \Lambda_{ij} / \delta T \right)_R T^2}{x_i + \sum_{i,j\neq 1} x_j \Lambda_{ij}} \right] \right\} \qquad \dots (3.9)$$

Where

$$\left(\delta \Lambda_{ij} / \delta T \right)_R T^2 = \left(\lambda_{ij} - \lambda_{ij} \right) \Lambda_{ij} \qquad \dots (3.10)$$

For binary systems, H^E become:

$$H^E = x_1 \left(\frac{x_2 \Lambda_{12}}{x_1 + \Lambda_{12} x_2} \right) \left(\lambda_{12} - \lambda_{11} \right) + x_2 \left(\frac{x_1 \Lambda_{21}}{x_2 + \Lambda_{21} x_1} \right) \left(\lambda_{12} - \lambda_{22} \right) \qquad \dots (3.11)$$

For more than two component equation, H^E become:

$$H^E = \sum_{i=1}^{C} \left\{ x_i \left[\frac{\sum_{i=1}^{C} x_i \left(\lambda_{ij} - \lambda_{ii} \right) \Lambda_{ij}}{x_i + \sum_{i,j\neq 1} X_j \Lambda_{ii}} \right] \right\} \qquad \dots (3.12)$$

It is therefore convenient to calculate the excess enthalpy for any solution and then adding this value to the enthalpy value of ideal system to get the actual value of enthalpy for the solution. Therefore, equation (3.12) was used in this work to calculate the excess enthalpy of non-ideal multicomponent system in liquid phase only.

3.5 Design models:

The calculations were made by using two main modified computer programmes for Short-cut model and Rigorous plate-to-plate model for non-ideal multicomponent distillation system.

3.5.1 Mathematical model for the short-cut design calculation:

This model based on the "empirical correlation method" which calculates:

1. Minimum reflux ratio.
2. Minimum number of plates.
3. Actual number of plates.
4. Actual reflux ratio.
5. Top and bottom temperature.
6. Heat load at top and bottom of the column.
7. Feed plate location.

Underwood [66] equation was used to determine the minimum reflux ratio:

$$\sum_{i=1}^{N} \frac{\alpha_i x_{i,d}}{\alpha_i - \phi} = R_m + 1 \qquad \text{… (3.13)}$$

ϕ was calculated from Fenske [12] relations :

$$\sum_{i=1}^{N} \left(\frac{\alpha_i x_{if}}{\alpha_i - \phi} \right) = 1 - q \qquad \text{… (3.14)}$$

And $\alpha_{LK} \le \phi \le \alpha_{HK}$

Where:

$$q = \frac{\text{Heat required to vaporize the feed}}{\text{Latent heat of vaporization}}$$

The value of ϕ is calculated from previous equations by trial and error, then R_m is calculated from equation (3.13).

Operating reflux ratio $= R = R_m * 1.25$ … (3.15)

Gilliand [20], Van Winkle and Todd [67] relationships were used to calculate the theoretical number of plates:

a. for $(R - R_m)/(R + 1) < 0.125$

$$\frac{N - N_m}{N + 1} = 0.5039 + 0.5068\left(\frac{R - R_m}{R + 1}\right) - 0.0908\log\left(\frac{R - R_m}{R + 1}\right) \quad \ldots (3.16)$$

b. for $(R - R_m)/(R + 1) > 0.125$

$$\frac{N - N_m}{N + 1} = 0.6257 - 0.9868\left(\frac{R - R_m}{R + 1}\right) + 0.516\left(\frac{R - R_m}{R + 1}\right)^2 - 0.1738\left(\frac{R - R_m}{R + 1}\right)^3$$

… (3.17)

Vapor – liquid equilibrium were calculated using both Wilson and Antoine equations

$$K_i = \frac{\gamma_i p_i}{p} \quad \ldots (3.18)$$

$$\ln p_i = A + \frac{B}{C + T} \quad \ldots (3.19)$$

The Antoine coefficients were given in appendix (C).

The ideal enthalpy of vapor and liquid were used in the form of third degree polynomials as a function of temperature. The coefficients of these polynomials were given in appendix (C).

The sum of ideal enthalpy and excess is the actual enthalpy of the components in the non-ideal systems.

The details of shortcut method were explained in appendix (A) and the modified program flow charts were given in appendix (B).

3.5.2 Mathematical models for plate-to-plate calculations:

Figure (3.4) considered a model column; this column has n-equilibrium stages including a condenser and reboiler, the stages numbered from top to bottom with the condenser as first stage and the reboiler as the N-th stage it is assumed that one feed stream F_j, one vapor side stream W_j, one liquid side stream U_j, and one intercooler or interheater Q_j exist at each stage except for the condenser and reboiler. This model column could be reduced to any simpler one by setting the undesired quantities to zero.

For conventional system column, all the quantities of the external streams except F_F, Q_A, Q_B and Q_D are zero each stage in the model column was assumed an equilibrium stage.

The composition, temperature and flow rate profiles were determined using plate-to-plate calculations.

The input data needed to obtain the solution was taken from the short-cut method, which is used to get a first estimation of the solution for plate – to – plate calculation:

1. Distillate rate.
2. Temperature of all feed streams.
3. Operating pressure.
4. Number of theoretical trays.
5. Antoine constants.
6. Wilson binary interaction energies.
7. Liquid and vapor enthalpy data.
8. Feed plate location.
9. Feed temperature, feed pressure and feed composition.

The starting point of plate-to-plate program is the first assumption of feed mole fractions; it was postulate that the number of total theoretical stages, flow rates of output streams, feed conditions, reflux ratio and operating pressure specified from short-cut program.

The calculation procedure can be summarizing as follows:

1. Assume $x_{i,j}$'s (i=1,………,m-1;j=1,………..,N).
2. The variables, $x_{m,j}$'s are determined from

$$x_{m,j}=1-\sum_{K=1}^{m-1} x_{k,j} \qquad \text{… (3.20)}$$

3. The temperature and vapor mole fraction can be calculated by solving

$$\sum K_{i,j}\, x_{i,j}=1.0 \qquad \text{… (3.21)}$$

4. Step 3 allow calculating the vapor and liquid enthalpies H_j's and h_j's.
5. From material balance, the vapor and liquid flow rates, V_j's and L_j's can be calculated.
6. The tridiagonal matrix equations are solved m times and the liquid mole fractions were calculated.

After normalization of these mole fractions, the liquid mole fractions, $X_{i,j}$'s are determined.

The values of $X_{i,j}$'s must equal the value of $x_{i,j}$'s assumed in step 1 and 2 respectively. In other words the variables calculated in step (1) through (6) satisfy all the basic equations derived only if $X_{i,j}$'s is exactly equal to $x_{i,j}$'s respectively.

Therefore, the following functions were defined:

$$F_{i,j} = x_{i,j} - X_{i,j}\ (i=1,\ldots,m;j=1,\ldots,N) \qquad \text{…(3.22)}$$

For column has n equilibrium stages including the condenser and reboiler which are assumed to be ideal stages, the stages are numbered from top to bottom with the condenser as the first stage and reboiler as the nth stage figure(3.4).

Mass and heat balances and equilibrium relations comprise the set of equations to be solved and were written as:

1- Material balance equation (m – equation)

For 1 ≤ i ≤ m and 1 ≤ j ≤ n

$$M_{i,j}\left(X_{i,j},V_j,T_j\right)=L_{j-1}X_{i,j-1}-\left(V_j+W_j\right)y_{i,j}-\left(L_j+U_j\right)X_{i,j}+V_{j+1}y_{i,j}+F_jZ_{i,j}=0$$

… (3.23)

2- Vapor – liquid equilibrium (E – equation)

$$E_j\left(X_{i,j},V_j,T_j\right)=y_{i,j}-K_{i,j}X_{i,j}=0 \qquad \ldots (3.24)$$

Where

$$K_{i,j}=\gamma_{i,j}\left(x_jT_j\right)p_{i,j}\Big/p \qquad \ldots (3.25)$$

3- Summation equation (S – equation)

$$S_j\left(X_{i,j},V_j,T_j\right)=\sum_{i=1}^{m}y_{i,j}-1.0\ =0 \qquad \ldots (3.26)$$

$$S_j\left(X_{i,j},V_j,T_j\right)==\sum_{i=1}^{m}x_{i,j}-1.0\ =0 \qquad \ldots (3.27)$$

4- Heat balance equation (H – equation)

$$H_j\left(X_{i,j},V_j,T_j\right)=L_{j-1}h_{j-1}-\left(V_j+W_j\right)H_j-\left(L_j+U_j\right)h_j+V_{j+1}H_{j+1}+F_jH_{Fj}-Q_j=0$$

… (3.28)

Equation (3.23) and (3.24) were combined and the liquid low rate (L's) is expressed in terms of vapor flow rate (V's) by an overall material balance of all stages from the condenser through the j-th stage.

$$L_j=V_{j+1}+\sum_{k=2}^{j}\left(F_k-W_k-V_k\right)-D\ ,\ 2\le j\le n-1 \qquad \ldots (3.29)$$

Where

$$D=V_1+U_1 \qquad \ldots (3.30)$$

There are n (2m+2) equations and n (2m+3) unknown variables are vapor and liquid flow rates, vapor and liquid mole fractions and stage temperatures.

The solution of these equations is by transforming the material balance equation (3.29) to a tridiagonal matrix. The details were given in appendix (A) and the modified program flow charts were given in appendix (B).

3.6 Application of heat pump system to multicomponent distillation:

The high-energy requirements for distillation can often be reduced by using heat pump or vapor-recompression, to pump heat from the condenser to the reboiler .this is accomplish by using compressor to raise the temperature level of the available heat from that of the condenser to that of the reboiler.

In this work, the study of energy recovery in multicomponent distillation using heat-pump system was dealt thoroughly using a developed computer programme.

To separate non-ideal mixtures of M-component feed into relatively pure products requires (M-1) distillation columns.

The heating medium is steam, its temperature depend on the boiling temperature of mixtures. The coolant medium is cooling water or air at ambient temperature.

This technique made a large reduction in cooling water or airflow and elimination of steam or other heating media for the reboiler. The heat of compression and other sensible heat effects in the sequence must be balanced by a trim cooler. Trim cooler heat load is usually a small fraction of the reboiler load so this load can be neglected, Neill et al [44].

After the trim cooling the overhead vapors are totally condensed in the reboiler and the resulting stream split into reflux and product. Sub-cooling the condensate in sub-cooler, this in turn reduces flash vapor recycle when the reflux was returned to the column.

The only change from conventional system is the elimination of the condenser, figure (3.1), (3.2) and (3.3).

The ratio of heat pumped to the reboiler Q_R to compression work (W) required, can be:

$$\frac{Q_R}{W} = (\varepsilon T_2)/(T_2 - T_1) \qquad \text{… (3.31)}$$

Where ε is compressor efficiency, T_2 is the temperature of compression vapor and it must be equal to the reboiler temperature plus the temperature difference in the reboiler (i.e. $T_2 = T_{reb} + \Delta T_{reb}$), and T_1 is the overhead vapors temperature.

For any heat pump system there is a limited range of ΔT_{reb} which economics will be favorable. In this study, the value of ε was set to be 0.7.

In heat pump system, the important equipment is the compressor and its cost.

3.6.1 Mathematical models for heat pump calculation:

In order to investigate the effect of using heat pump system in multicomponent distillation many variables must be calculated for the condenser and reboiler:

$$Q_C = V_{ac}(H_{VI} - h_{LI}) \qquad \text{… (3.32)}$$

$$Q_R = D \cdot h_D + W \cdot h_w + Q_c - F \cdot H_{iF} \qquad \text{… (3.33)}$$

Where Q_c and Q_R are the heat loads for condenser and reboiler in Kcal/hr.

$$W_{comp} = V_{ac}(H_{VI} - H_{VO}) \qquad \text{… (3.34)}$$

Where, W_{comp} is the work of the compressor in Kcal/hr .and H_{VO} is the enthalpy of vapor stream out of the compressor

The heat exchanged by the subcooler was given by:

$$Q_{SC} = V_{AC}(h_{LO} - h_D) \qquad \text{… (3.35)}$$

The details were given in appendix (A).

3.6.2 Cost estimation:

Equipment cost, operating cost were calculated depend on the equations of Henry [27] method.

Equipment cost = C_E = cost of column + cost of reboiler + cost of condenser

Operating cost = 28% (total equipment cost) + cost of utilities

Cost of utilities = cost of steam + cost of cooling water

Total annual cost = depreciation + annual operating cost

Total annual cost= can = (equipment cost / service life) +annual operating cost. The details were given in appendix (A).

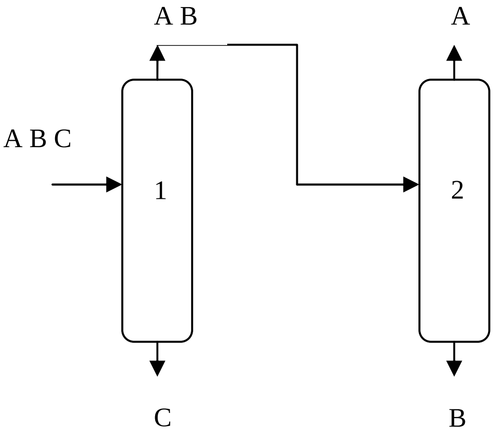

Figure (3.1) Feed system b configuration

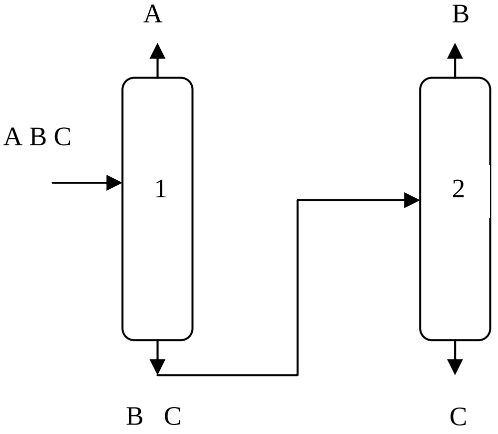

Figure (3.2) Feed system b configuration

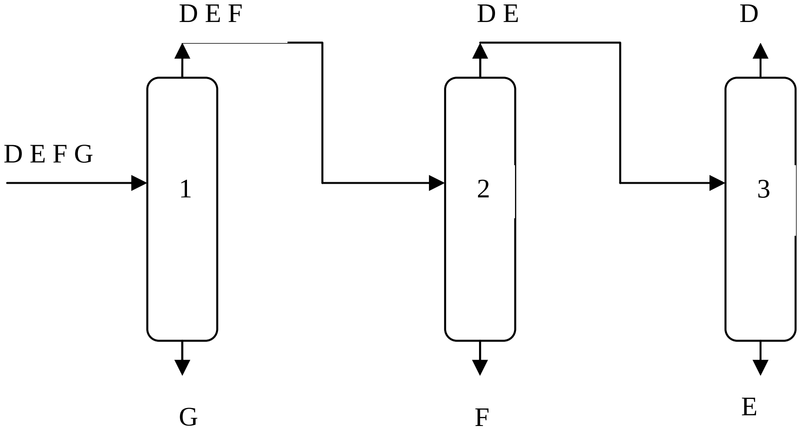

Figure (3.3) Feed system c configuration.

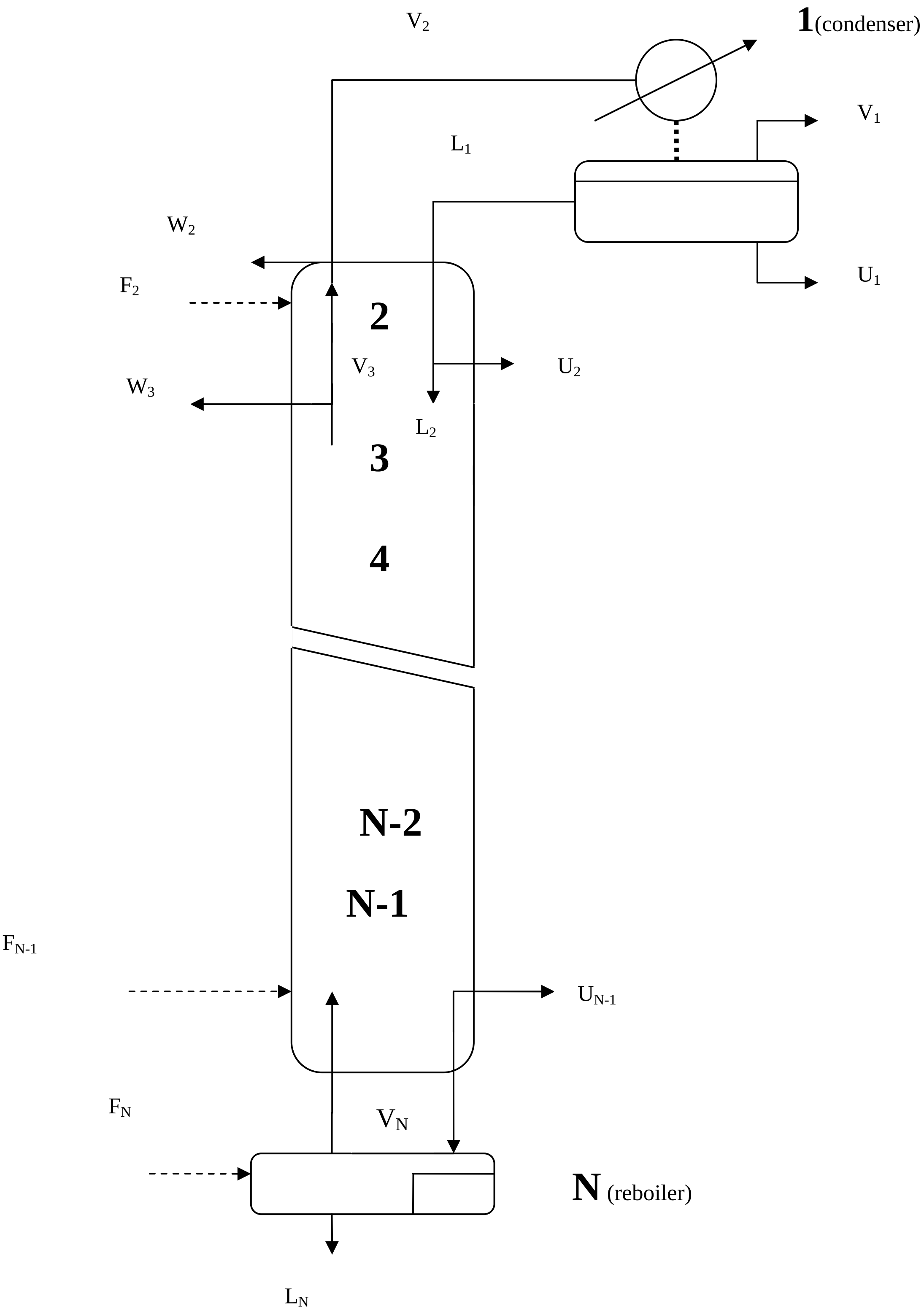

Figure (3.4) model column for mathematical calculation.

Chapter Four

Results and Discussion

4.1 Introduction:

In this chapter, a number of non-ideal feed systems were studied by applying different feed compositions, fractional recovery and operating pressure, by using short-cut model and rigorous plate-to-plate model.

All results presented in this chapter were taken from plate-to-plate calculation.

Total reboiler load for all feed systems at different feed compositions, fractional recoveries and operating pressure were reported in tables, and/or represented graphically to ease the comparison.

Percentage saving in total energy consumption using heat pump were reported in tables and/or presented graphically, and finally percentage saving in total annual cost for all feed systems were reported too.

4.2 The analysis of the feed systems:

Two ternary and one quaternary feed system were chosen, where feed system a [methanol, ethanol, water]: anhydrous ethanol is manufactured by the fractionation of an aqueous feed in a column to which dehydration liquid is introduced near the top.

Benzene, ethyl-ether are commercial dehydration agents, each of these has the property of forming an aqueous azeotrope which boils at a lower temperature than does 95% alcohol; thus the water is separated from the alcohol and the latter is withdrawn from the bottom of the columns as an anhydrous product.

In this work, the method of Griswold and Dinwidde [21 has been used for the separation of ethanol-water by methanol (dehydration alcohol by a non-azeotropic distillation).

Where for feed system a, the third component forms a regular binary solution with ethanol and with boiling point lower than that of ethanol-water azeotrope.

Therefore, it should be possible to dehydrate the ethanol by fractionation the ternary mixture into a dry binary overhead product and a water bottom. Methanol is an obvious agent does not form an azeotrope with water as in figure (4.1).

For feed system b [acetone, methanol, water]: the separation of acetone and methanol (binary azeotrope) by adding water as a solvent to break the azeotrope between acetone and methanol. The water lowers the volatility of the methanol, so the acetone can be separated from the top of the column and methanol and water from the bottom where it is fed to the second column to separate methanol from water figure (4.2).

For the feed system c [n-hexane, methylcyclopentane, ethanol, benzene], Belknap [5] and Kaes [32] show that the ternary systems ethanol-benzene-methylcyclopentane and n-hexane-ethanol-methylcyclopentane are deviated greatly from ideal liquid phase behavior, with no ternary azeotrope, also the quaternary system exhibition non-ideal liquid phase behavior as in figure (4.3).

This work studied the effect of the several parameters for two ternary and one quaternary feed system on total reboiler load, percentage saving in total energy consumption and percentage saving in total annual cost.

1. Effect of pressure.
2. Effect of feed composition.
3. Effect of fractional recovery.

In order to study the optimum operating conditions, a complete design of each feed system was done using short-cut method first, to get the initial estimation of the design calculation to plate-to-plate calculation method.

Heaven equation [24], calculates the number of sequence of separation of n-components, cannot be applied to this work because a special configuration due to non-ideality is needed.

Heat pump was studied for all feed systems in order to get the best configuration for each feed system with heat pump and with maximum saving energy consumption and in total annual cost.

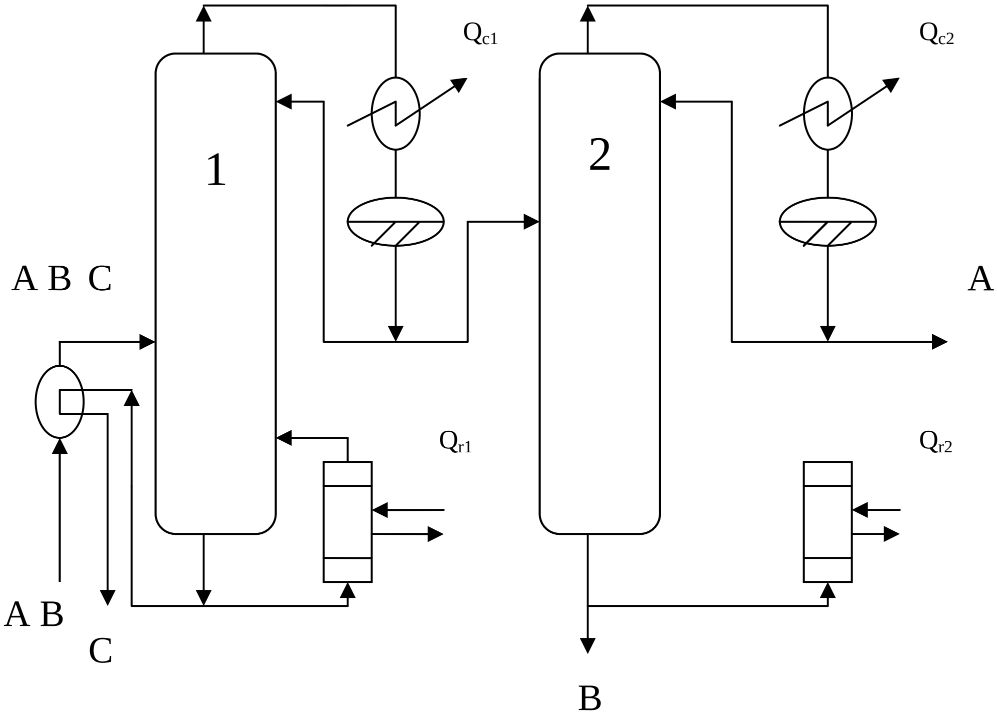

Figure (4.1) Configuration of feed system a

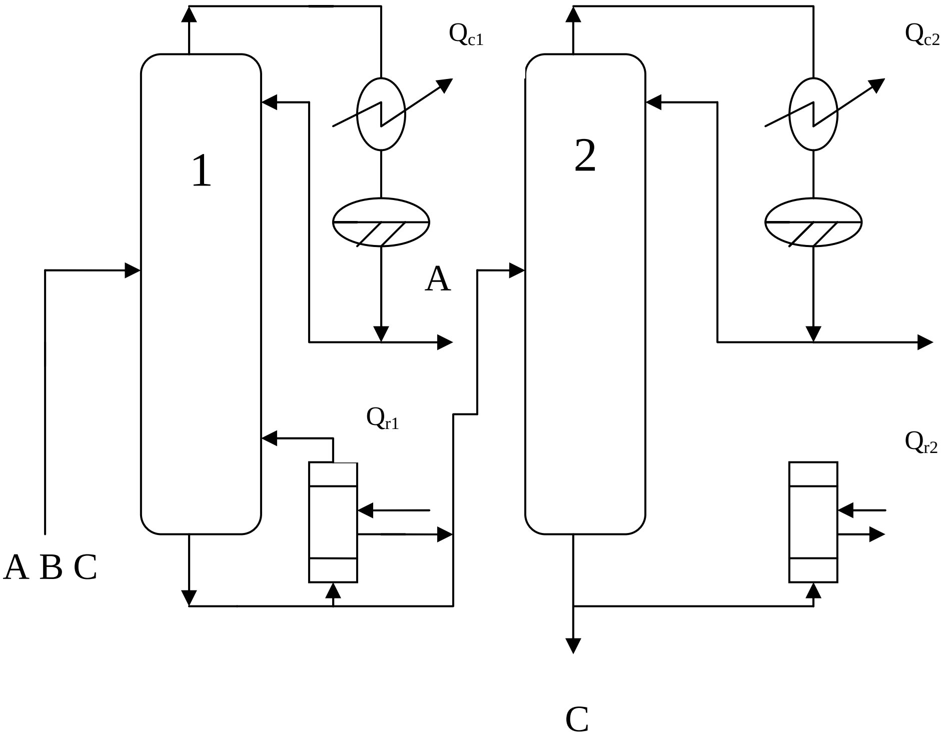

Figure (4.2) Configuration of feed system b.

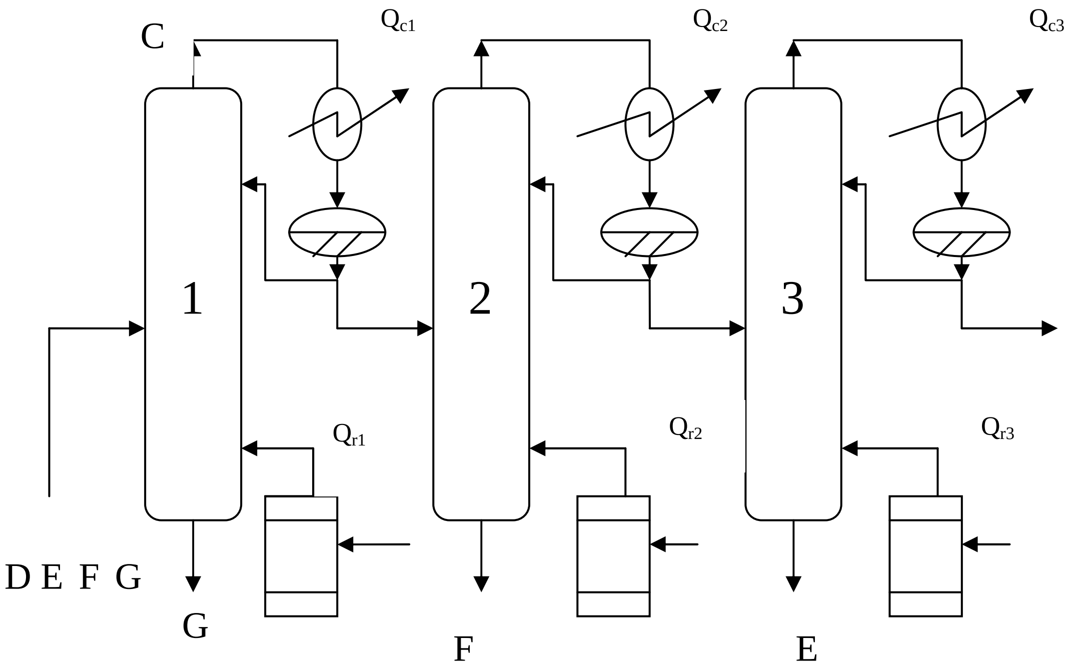

Figure (4.3) Configuration of feed system c.

4.3 Effect of process variables on total reboiler load using split tower technique without heat pump:

4.3.1 Effect of pressure:

Split tower technique was used in this work, where each column in the configuration was assumed to be operated at different operating pressure. Six operating pressure were taken 0.5, 1.0, 1.5, 2.0, 2.5, and 3.0 atm for feed system a and b. Column 1 assumed to be operated at 0.5 atm and column 2 was operated at different operating pressure for six times from 0.5 to 3.0 atm. Then the first column was assumed to be operated at 1.0 atm and the second changed from 0.5 to 3.0 until column one reached the operating pressure 3.0 atm, this means that there are 36 possibilities for the ternary feed systems a and b at different feed compositions and recovery and all these possibilities were reported in tables (4.1)-(4.4) and (4.5)-(4.8) for feed systems a and b respectively. Figure (4.4) and (4.5) show the change of the total reboiler load as a function of changing the operating pressure of column one and column two for feed systems a and b, respectively.

For quaternary feed system c, there are 216 possibilities if the three columns in figure (4.3) assumed to be operated at different pressure for six times. Therefore, in this work only three different operating pressures of 1, 2, 3 was assumed and the possibilities of changing pressure was given in table (4.9). Table (4.10) gives the effect of changing pressure on total reboiler load for five feed composition.

Split tower technique was used to reduce the number of reboilers in the configuration and to reuse the heat from the condenser in the reboiler.

The results show that as the operating pressure increases the total reboiler load increases too, this is due to the increasing in the bottom temperature of the column.

For feed system a as the operating pressure increases the total reboiler load increases very slightly, this is due to the ease separation of Methanol – Ethanol from water, while for feed system b there was a large increases in total reboiler load with increasing the pressure that is due to the difficult separation of acetone from methanol – water.

4.3.2 Effect of feed composition:

Four feed compositions as in figure(3.1) four fractional recoveries and six operating pressure for ternary feed systems a and b, and five feed compositions one fractional recovery and three operating pressure for feed system c were studied. The results obtained were tabulated in tables (4.1) – (4.4) for feed systems a, tables (4.5) – (4.8) for feed system b and table (4.10) for feed system c.

Total reboiler load as a function of feed composition and fractional recovery has been shown in form of histograms for feed systems a and b figures (4.6, 4.7) and for feed system c figure (4.8).

From the results it was found that for system a and b there is maximum total reboiler load in feed composition 3, this was due to the high reflux ratio in the which will increase the reboiler load and it was decreased as the feed composition changed to 2, 1 and 4, respectively. For feed system c, table (4.9) show 27 possibilities of changing the operating pressure for the three columns, table (4.10) shows that there is a maximum total reboiler load for feed composition 2 this was decreased as the feed composition changed to 3, 4, 1 and 5, respectively.

4.3.3 Effect of fractional recovery:

The fractional recovery was defined as the fraction of the amount of component, which is produced in the product divided by the amount of the component in the feed.

In this work four fractional recoveries of 0.9, 0.925, 0.95, 0.99 were considered, these recoveries was applied to short-cut method in order to get the initial estimation of reflux ratio, distillate rate, feed temperature, appendix (C) shows the composition profile results from plate – to plate calculation for feed composition (1) of feed system a to get the accurate composition profile at each stage.

All results were given in tables (4.1- 4.4) for feed system a, and tables (4.5 – 4.8) for feed composition b and were shown in figures (4.6) for feed system a and figure (4.7) for feed system b. For feed system c, only one recovery of 0.9 was studied and the results were given in table (4.10) and figure (4.8). These results show that as the fractional recovery increase the total reboiler load will increase too for all systems.

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.49	1.63	1.76	1.87	2.00	2.11
1.0	1.50	1.65	1.77	1.89	2.01	2.13
1.5	1.53	1.67	1.78	1.91	2.03	2.15
2.0	1.54	1.68	1.80	1.92	2.04	2.17
2.5	1.55	1.69	1.81	1.93	2.06	2.18
3.0	1.57	1.71	1.83	1.94	2.07	2.20

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.57	1.76	1.92	2.02	2.14	2.25
1.0	1.58	1.80	1.95	2.05	2.17	2.30
1.5	1.61	1.81	1.96	2.06	2.18	2.31
2.0	1.62	1.82	1.98	2.08	2.20	2.32
2.5	1.64	1.84	2.00	2.10	2.21	2.35
3.0	1.65	1.85	2.01	2.11	2.23	2.36

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.68	1.85	2.03	2.14	2.30	2.43
1.0	1.69	1.86	2.05	2.15	2.31	2.45
1.5	1.70	1.87	2.06	2.16	2.32	2.46
2.0	1.72	1.90	2.08	2.17	2.34	2.47
2.5	1.74	1.91	2.10	2.19	2.35	2.49
3.0	1.76	1.93	2.11	2.20	2.37	2.51

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.91	2.03	2.10	2.26	2.39	2.52
1.0	1.91	2.05	2.12	2.29	2.40	2.56
1.5	1.93	2.06	2.12	2.32	2.45	2.57
2.0	1.94	2.08	2.13	2.35	2.47	2.59
2.5	1.98	2.08	2.17	2.38	2.49	2.63
3.0	1.98	2.09	2.30	2.40	2.52	2.68

Table (4.1) The effect of pressure change (atm.) of column 1 and 2 on total reboiler load (kcal. /hr*10^-6) for feed system a (feed composition 1).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.52	1.65	1.76	1.90	2.05	2.10
1.0	1.54	1.67	1.79	1.92	2.08	2.12
1.5	1.55	1.69	1.80	1.93	2.09	2.13
2.0	1.56	1.69	1.80	1.93	2.10	2.13
2.5	1.57	1.70	1.81	1.94	2.10	2.14
3.0	1.58	1.71	1.82	1.95	2.11	2.15

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.66	1.82	1.94	2.07	2.37	2.49
1.0	1.67	1.83	1.96	2.09	2.39	2.50
1.5	1.68	1.83	1.96	2.09	2.39	2.51
2.0	1.69	1.84	1.97	2.10	2.40	2.52
2.5	1.70	1.85	1.97	2.11	2.41	2.53
3.0	1.70	1.85	1.98	2.11	2.41	2.53

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.75	1.95	2.05	2.25	2.55	2.87
1.0	1.77	1.97	2.08	2.27	2.57	2.90
1.5	1.77	1.97	2.08	2.28	2.58	2.90
2.0	1.78	1.98	2.10	2.29	2.59	2.91
2.5	1.79	1.99	2.10	2.30	2.60	2.92
3.0	1.80	2.00	2.11	2.31	2.61	2.93

R_C = 0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	2.01	2.13	2.40	2.70	2.90	3.10
1.0	2.03	2.15	2.41	2.72	2.92	3.13
1.5	2.03	2.17	2.43	2.73	2.93	3.15
2.0	2.05	2.19	2.45	2.74	2.95	3.16
2.5	2.06	2.20	2.47	2.75	2.96	3.20
3.0	2.08	2.22	2.48	2.77	2.98	3.25

Table (4.2) The effect of pressure change of column 1 and 2 on total reboiler load (kcal. /hr*10^-6) for feed system a (feed composition 2).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	2.67	2.92	3.20	3.41	3.64	3.80
1.0	2.70	2.95	3.22	3.45	3.67	3.82
1.5	2.71	2.98	3.24	3.47	3.69	3.85
2.0	2.75	3.02	3.27	3.52	3.74	3.88
2.5	2.85	3.11	3.36	3.60	3.81	3.95
3.0	2.95	3.20	3.47	3.70	3.91	4.05

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	3.02	3.30	3.65	3.86	4.09	4.20
1.0	3.08	3.34	3.70	3.90	4.12	4.25
1.5	3.10	3.36	3.72	3.94	4.15	4.27
2.0	3.20	3.45	3.82	4.02	4.24	4.36
2.5	3.21	3.49	3.85	4.05	4.28	4.40
3.0	3.25	3.52	3.90	4.11	4.31	4.45

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	2.48	2.68	2.95	3.20	3.40	3.65
1.0	2.70	2.90	3.10	3.44	3.63	3.88
1.5	2.80	3.00	3.27	3.53	3.72	3.97
2.0	2.98	3.28	3.54	3.80	4.00	4.25
2.5	3.30	3.50	3.78	4.03	4.23	4.48
3.0	3.58	3.78	4.00	4.31	4.50	4.75

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	3.55	4.23	4.11	4.22	4.41	4.51
1.0	3.80	4.25	4.43	4.63	4.75	5.30
1.5	3.99	4.36	4.63	4.75	4.95	5.60
2.0	4.15	4.56	4.65	4.98	5.20	5.71
2.5	4.58	4.70	4.95	5.29	5.40	5.92
3.0	4.59	4.82	5.18	5.30	5.50	6.23

Table (4.3) The effect of pressure change of column 1 and 2 on total reboiler load (kcal. /hr*10^-6) for feed system a (feed composition 3).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.87	1.13	1.28	1.48	1.58	2.00
1.0	1.00	1.14	1.30	1.50	1.67	2.01
1.5	1.03	1.16	1.33	1.53	1.70	2.03
2.0	1.04	1.17	1.34	1.54	1.71	2.05
2.5	1.06	1.20	1.37	1.57	1.75	2.08
3.0	1.09	1.22	1.40	1.60	1.78	2.10

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.00	1.15	1.32	1.52	1.70	1.95
1.0	1.02	1.17	1.35	1.54	1.73	2.00
1.5	1.03	1.20	1.38	1.57	1.75	2.03
2.0	1.05	1.22	1.40	1.59	1.78	2.05
2.5	1.08	1.27	1.43	1.59	1.82	2.09
3.0	1.13	1.30	1.47	1.63	1.86	2.12

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.03	1.20	1.35	1.50	1.70	1.87
1.0	1.05	1.22	1.38	1.52	1.72	1.90
1.5	1.08	1.26	1.42	1.56	1.73	1.93
2.0	1.10	1.29	1.43	1.58	1.77	1.95
2.5	1.13	1.30	1.46	1.60	1.80	1.97
3.0	1.15	1.32	1.48	1.63	1.82	2.00

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.10	1.31	1.47	1.62	1.83	2.12
1.0	1.12	1.32	1.50	1.64	1.85	2.12
1.5	1.15	1.33	1.55	1.66	1.89	2.22
2.0	1.19	1.35	1.64	1.68	1.92	2.25
2.5	1.28	1.43	1.66	1.80	2.12	2.28
3.0	1.40	1.56	1.85	1.90	2.19	2.43

Table (4.4)The effect of pressure change of column 1 and 2 on total reboiler load(kcal. /hr*10^-6) for feed system a (feed composition 4).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.90	0.91	0.92	0.95	0.97	1.00
1.0	0.95	0.97	0.98	1.01	1.03	1.05
1.5	1.02	1.05	1.06	1.07	1.08	1.13
2.0	1.12	1.13	1.15	1.18	1.20	1.22
2.5	1.18	1.20	1.20	1.23	1.23	1.28
3.0	1.27	1.29	1.30	1.32	1.33	1.36

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.93	0.95	0.97	0.99	1.01	1.05
1.0	0.99	1.02	1.03	1.06	1.09	1.12
1.5	1.06	1.08	1.10	1.12	1.09	1.19
2.0	0.16	1.18	1.20	1.23	1.24	1.28
2.5	1.25	1.26	1.28	1.31	1.35	1.36
3.0	1.44	1.45	1.47	1.50	1.51	1.55

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.99	1.00	1.02	1.04	1.07	1.10
1.0	1.03	1.05	1.07	1.10	1.10	1.14
1.5	1.10	1.13	1.15	1.17	1.19	1.23
2.0	1.18	1.20	1.22	1.24	1.26	1.30
2.5	1.26	1.29	1.30	1.33	1.35	1.38
3.0	1.35	1.35	1.39	1.40	1.42	1.45

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.09	1.10	1.13	1.15	1.17	1.20
1.0	1.10	1.12	1.15	1.17	1.18	1.21
1.5	1.15	1.17	1.18	1.20	1.22	1.25
2.0	1.22	1.25	1.27	1.28	1.30	1.34
2.5	1.30	1.31	1.33	1.35	1.37	1.41
3.0	1.35	1.37	1.40	1.41	1.44	1.46

Table (4.5) The effect of pressure change of column 1 and 2 on total reboiler load (kcal. /hr*10^-6) for feed system b (feed composition 1)

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.30	1.36	1.42	1.46	1.56	1.75
1.0	1.48	1.55	1.60	1.65	1.75	1.95
1.5	1.68	1.73	1.80	1.83	1.94	2.12
2.0	1.83	1.87	1.92	1.97	2.08	2.25
2.5	1.97	2.04	2.05	2.10	2.21	2.40
3.0	2.18	2.23	2.30	2.33	2.44	2.59

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.40	1.41	1.55	1.57	1.58	1.80
1.0	1.60	1.61	1.70	1.75	1.80	2.02
1.5	1.82	1.92	1.99	2.03	2.07	2.27
2.0	1.96	2.03	2.06	2.12	2.17	2.36
2.5	2.10	2.16	2.20	2.27	2.30	2.50
3.0	2.33	2.40	2.43	2.48	2.53	2.75

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.50	1.55	1.60	1.65	1.69	1.88
1.0	1.70	1.75	1.79	1.85	1.88	2.00
1.5	1.90	1.94	1.98	2.03	2.07	2.27
2.0	2.10	2.15	2.18	2.24	2.27	2.47
2.5	2.33	2.38	2.42	2.47	2.50	2.70
3.0	2.57	2.62	2.65	2.70	2.75	2.92

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.70	1.72	1.74	1.76	1.79	1.99
1.0	1.92	1.93	1.95	1.98	2.00	2.20
1.5	2.08	2.10	2.11	2.19	2.20	2.42
2.0	2.30	2.32	2.36	2.38	2.41	2.62
2.5	2.50	2.53	2.57	2.60	2.66	2.88
3.0	2.78	2.79	2.80	2.84	2.89	3.09

Table (4.6) The effect of pressure change of column 1 and 2 on total reboiler load (kcal. /hr*10^-6) for feed system b (feed composition 2).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.31	1.32	1.32	1.35	1.37	1.40
1.0	1.35	1.36	1.37	1.40	1.40	1.43
1.5	1.52	1.55	1.55	1.57	1.58	1.62
2.0	1.67	1.68	1.70	1.73	1.75	1.77
2.5	1.91	1.92	1.93	1.96	1.97	2.00
3.0	2.10	2.11	2.12	2.15	2.17	2.18

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.35	1.36	1.36	1.38	1.40	1.42
1.0	1.57	1.58	1.58	1.60	1.62	1.65
1.5	1.85	1.85	1.85	1.89	1.90	1.93
2.0	2.21	2.23	2.22	2.25	2.26	2.30
2.5	2.50	2.52	2.53	2.55	2.57	2.60
3.0	2.80	2.80	2.81	2.84	2.86	2.88

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.37	1.38	1.39	1.40	1.42	1.43
1.0	1.72	1.73	1.74	1.75	1.76	1.78
1.5	2.24	2.25	2.25	2.26	2.27	2.30
2.0	2.64	2.65	2.66	2.66	2.68	2.70
2.5	2.75	2.76	2.78	2.86	2.79	2.80
3.0	3.38	3.40	3.41	3.42	3.43	3.44

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	1.90	1.92	1.95	1.98	2.05	2.08
1.0	2.50	2.53	2.54	2.56	2.58	2.59
1.5	2.60	2.63	2.64	2.66	3.67	3.68
2.0	3.33	2.33	3.35	3.37	3.38	3.39
2.5	3.50	3.52	3.53	3.54	3.56	3.58
3.0	4.82	4.85	4.86	4.88	4.88	4.90

Table (4.7) The effect of pressure change of column 1 and 2 on total reboiler load (kcal. /hr*10^-6)for feed system b (feed composition 3).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.42	0.44	0.50	0.55	0.59	0.65
1.0	0.47	0.50	0.54	0.60	0.62	0.70
1.5	0.50	0.55	0.58	0.63	0.67	0.73
2.0	0.53	0.59	0.63	0.67	0.70	0.77
2.5	0.60	0.64	0.68	0.73	0.77	0.84
3.0	0.64	0.68	0.72	0.77	0.82	0.87

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.45	0.50	0.55	0.59	0.60	0.65
1.0	0.50	0.53	0.58	0.62	0.64	0.70
1.5	0.53	0.58	0.62	0.67	0.68	0.75
2.0	0.57	0.60	0.65	0.70	0.72	0.77
2.5	0.63	0.66	0.71	0.75	0.77	0.85
3.0	0.67	0.70	0.76	0.80	0.82	0.88

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.49	0.53	0.57	0.60	0.65	0.72
1.0	0.50	0.57	0.60	0.64	0.68	0.75
1.5	0.55	0.60	0.64	0.69	0.73	0.79
2.0	0.59	0.63	0.67	0.72	0.76	0.82
2.5	0.62	0.68	0.70	0.75	0.80	0.86
3.0	0.67	0.70	0.75	0.78	0.82	0.89

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	0.55	0.58	0.64	0.65	0.70	0.75
1.0	0.57	0.60	0.66	0.67	0.73	0.78
1.5	0.59	0.64	0.68	0.70	0.74	0.79
2.0	0.61	0.69	0.70	0.72	0.76	0.81
2.5	0.67	0.74	0.78	0.79	0.85	0.90
3.0	0.70	0.75	0.79	0.81	0.87	0.92

Table (4.8) The effect of pressure change of column 1 and 2 on total reboiler load (kcal. /hr*10^-6) for feed system b (feed composition 4).

Possibility No.	Operating pressure		
	Column one	Column two	Column three
1	1	1	1
2	1	1	2
3	1	1	3
4	1	2	1
5	1	2	2
6	1	2	3
7	1	3	1
8	1	3	2
9	1	3	3
10	2	1	1
11	2	1	2
12	2	1	3
13	2	2	1
14	2	2	2
15	2	2	3
16	2	3	1
17	2	3	2
18	2	3	3
19	3	1	1
20	3	1	2
21	3	1	3
22	3	2	1
23	3	2	2
24	3	2	3
25	3	3	1
26	3	3	2
27	3	3	3

Table (4.9) Possibilities of pressure change in the three columns of feed system c.

Possibility No.	Feed composition				
	1	2	3	4	5
1	6.00	11.00	8.55	8.50	4.55
2	7.01	12.99	9.02	8.75	4.76
3	7.51	14.00	10.45	8.93	5.00
4	6.50	12.23	9.21	8.70	5.50
5	7.00	13.60	10.45	8.88	5.93
6	7.31	15.01	11.20	9.00	6.00
7	6.62	12.00	9.70	8.74	8.05
8	7.30	14.21	10.52	8.95	8.42
9	7.90	15.12	11.50	9.08	8.66
10	7.00	12.60	9.25	9.11	4.82
11	7.50	13.85	10.00	9.20	5.10
12	8.01	15.42	10.74	9.46	5.4
13	7.15	12.30	9.50	9.21	5.92
14	7.52	14.47	10.45	9.32	6.13
15	8.01	15.44	11.50	9.50	6.44
16	7.20	12.6	10.25	9.29	8.39
17	7.77	14.23	11.00	9.43	8.45
18	8.25	15.42	11.93	9.55	8.99
19	7.33	12.63	9.48	9.98	5.00
20	7.88	14.22	10.45	10.07	5.46
21	8.39	15.50	11.25	10.20	5.60
22	7.57	12.78	10.00	10.00	6.20
23	7.92	14.20	10.85	10.19	6.45
24	8.21	15.31	11.80	10.33	6.75
25	7.60	13.20	10.25	10.02	8.50
26	8.12	14.75	11.36	10.22	8.93
27	8.52	16.52	12.00	10.35	9.00

Table (4.10) Effect of reboiler load (kcal/hr*10^-6) with change in feed composition for (27 possibilities) of feed system c.

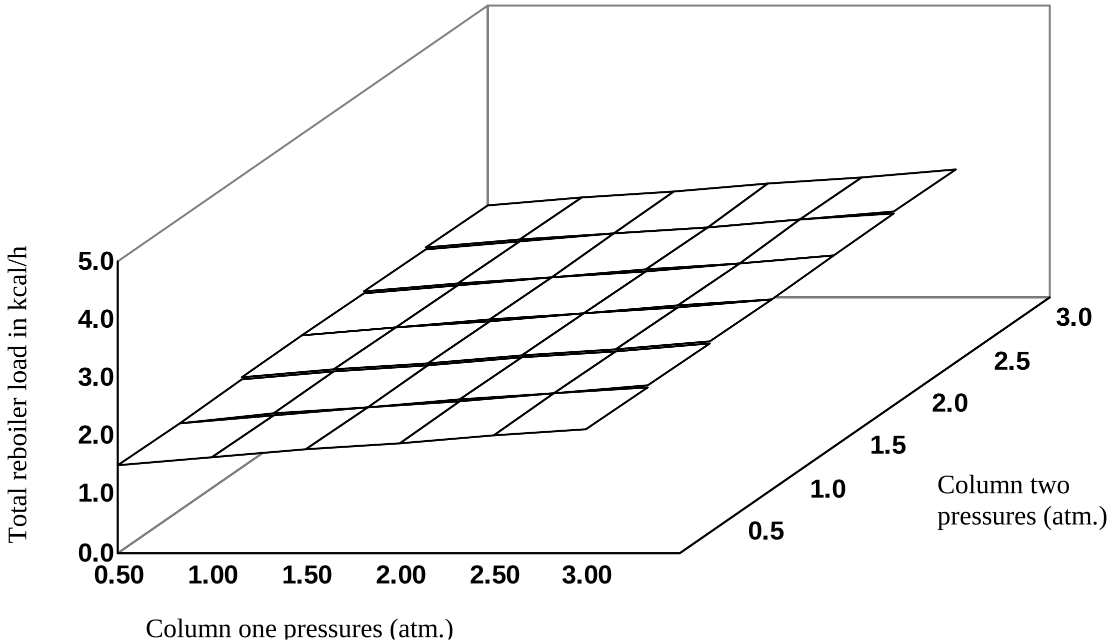

Figure (4.4) the effect of pressure change of column 1 and 2 on total reboiler load for feed system a (feed composition 1 and recovery = 0.9).

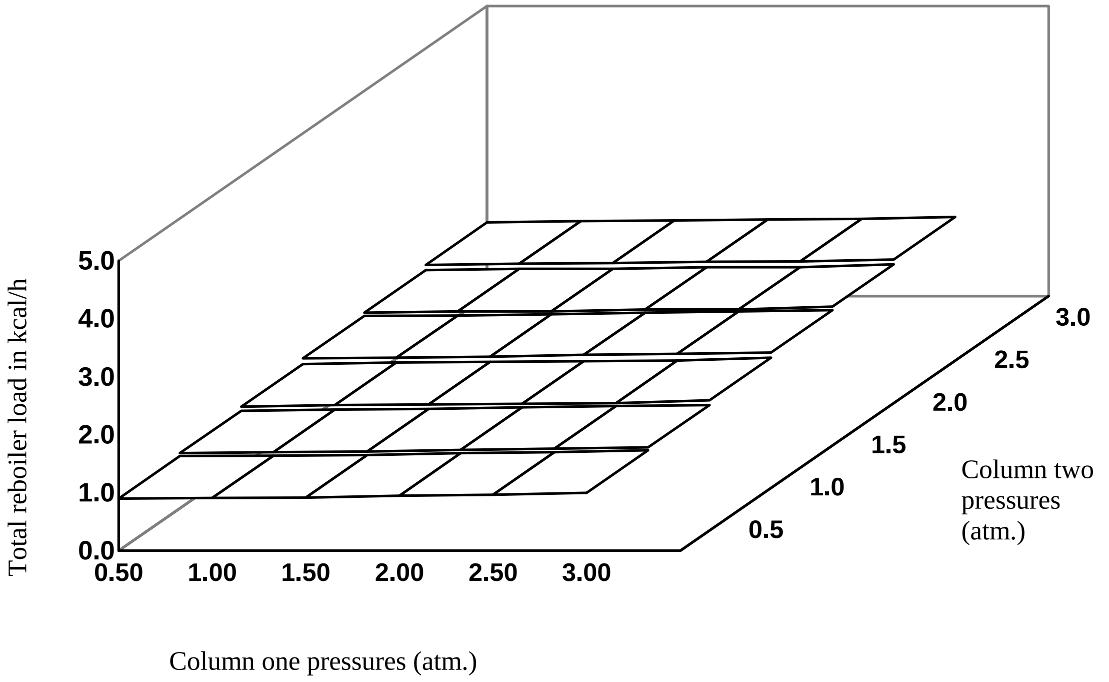

Figure (4.5) the effect of pressure change of column 1 and 2 on total reboiler load for feed system b (feed composition 1 and recovery = 0.9).

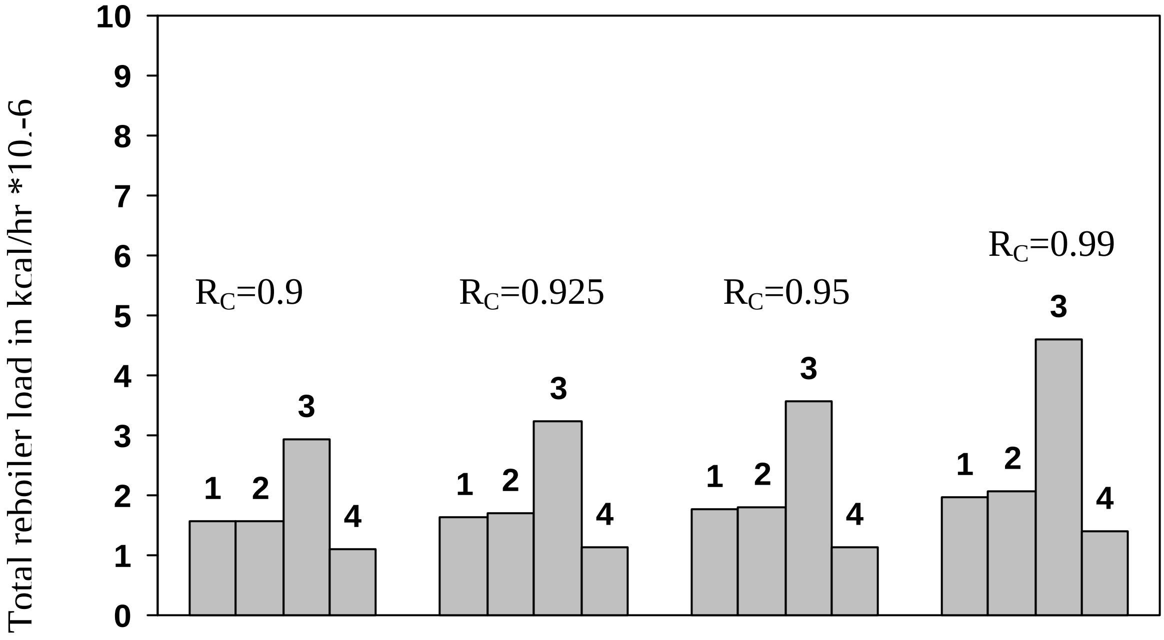

Figure (4.6) the effect of feed composition change on total reboiler load feed system a (column one at 3 atm and column two at 0.5 atm.).

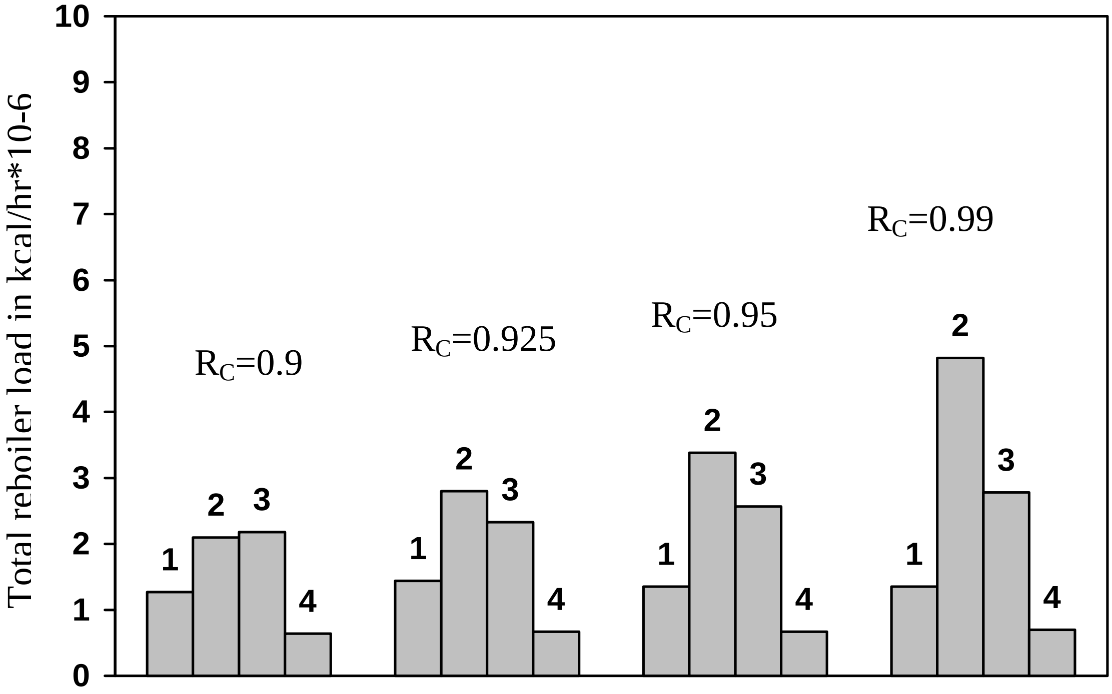

Figure (4.7) the effect of composition change on total reboiler load for feed system b (column one at 3 atm. and column two at 0.5 atm.).

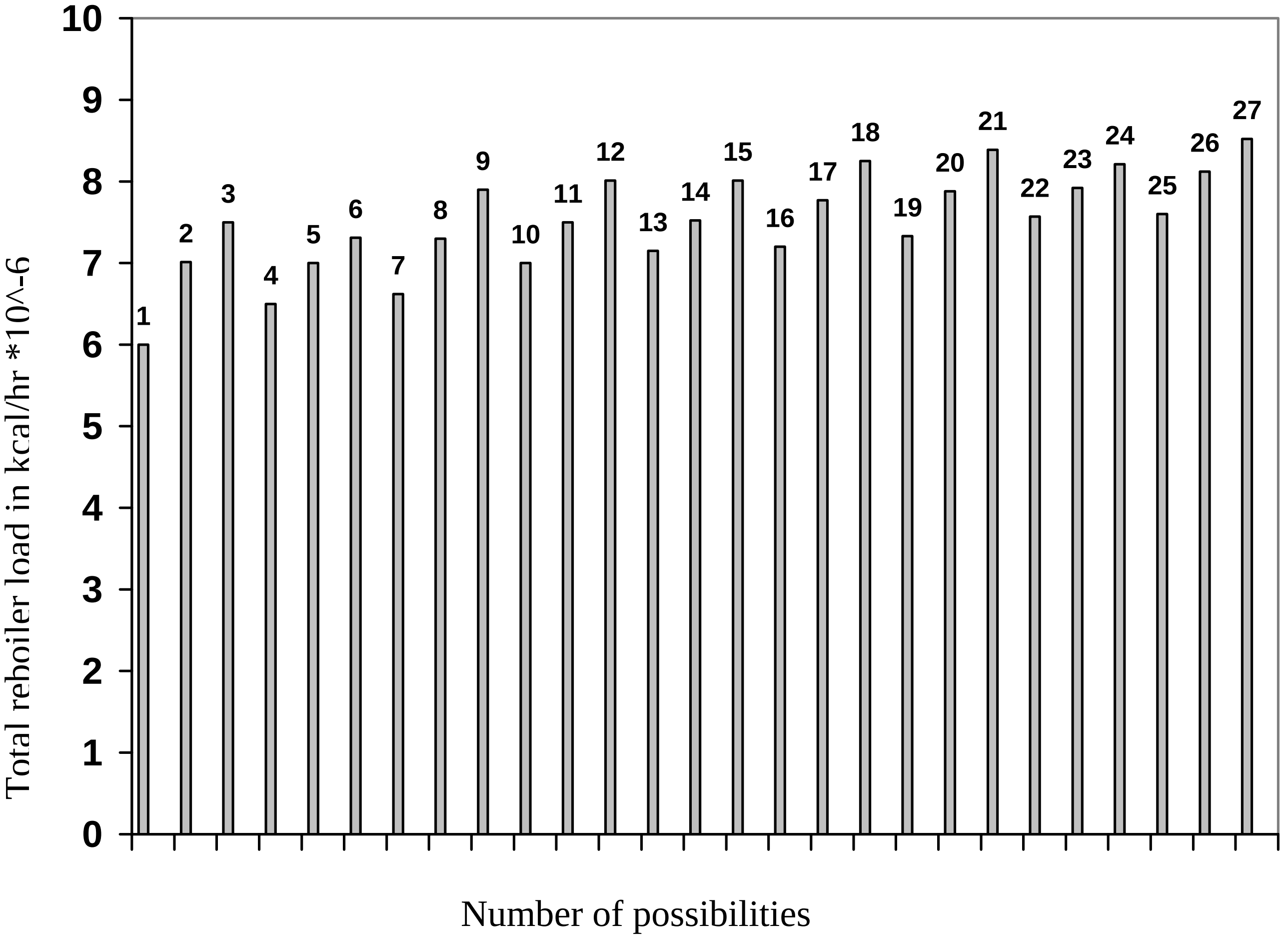

Figure (4.8) the effect of all possibilities of the pressure change for the three columns on total reboiler load for feed system c (feed composition one).

4.4 Heat pump with split tower technique:

In calculating the condenser and reboiler load at the top and bottom of each column it was found that a considerable amount of energy would be possible to be recovered.

Heat pump technique was considered for the three non-ideal feed systems a, b and c figures (4.9, 4.10, and 4.11) with and without split tower technique or when the column were operated at the same pressure and/or at different pressure.

After applying heat pump system the energy requirements in distillation column was reduced and the percentage saving in total energy consumption is calculating using:

Percentage saving in total energy consumption= $\frac{Q_R - Q_{RC}}{Q_R}$ *100… (4.1)

Where:

Q_R = the total reboiler load for conventional system.

Q_R = the total reboiler load using heat pump system.

4.5 Effect of process variables on the percentage saving in total energy consumption with heat pump technique:

Many variables have been studied such as operating pressure, feed composition, and fractional recovery.

4.5.1 Effect of pressure:

The effect of changing the pressure on percentage saving in total energy consumption was shown in tables (4.11-4.14) (4.15 – 4.18) and (4.19) for feed system a, b and c, respectively. Figure (4.12) and (4.13) represent the change in percentage saving in total energy consumption with heat pump and with changing the pressure in column one for feed system a and b.

4.5.2 Effect of feed composition:

The effect of changing feed composition was studied for all feed composition (table (3.2)) and for all feed systems and the results were given in tables (4.11-4.14) for feed system a, (4.15-4.18) for feed system b and table (4.19) for feed system c.

Percentage saving in total energy consumption with heat pump as a function of feed composition and fractional recovery has been shown in form of histograms for feed systems a and b figures (4.14, 4.15) and for feed system c figure (4.16).

From these results the following was obtained

1. For feed system a, there was a maximum percentage saving in total energy consumption for feed system 2 this was decrease if the feed composition changed to 1, 3, and 4.
2. For feed system b, there was maximum percentage saving in total energy consumption for feed composition 3 and this was decreased if the feed changed to 4, 2, and 1.
3. For feed system c, table (4.19) shows the percentage saving in total energy consumption with different feed composition. that for possibilities number 1, 3,7,8,10,13,18,24 and 27 feed composition 5 show maximum percentage saving in total energy consumption, while for possibilities number 2,5,6,14,19,21,25 and 4,9,11,12,15,16,17,20 feed composition 4 and 3 respectively, represent a maximum percentage saving in total energy consumption, also feed composition 1 represent maximum percentage saving in total energy consumption for possibilities 22 and 23 and feed composition 2 represent maximum percentage saving in total energy consumption for possibility 26.

4.5.3 Effect of fractional recovery:

the effect of changing fractional recovery for all feed systems and compositions are given in tables (4.11-4.19) and were shown in figures (4.14) for feed system a and figure (4.15) for feed system b and figure (4.16) for system c, all the results show that as the fractional recovery increases the percentage saving in total energy consumption decreases, this is due to the increase in the temperature of the streams this require more energy input to the system in order to get high purity.

4.5.4 Regions of optimality:

The regions of optimality with respect to energy requirements were given in tables (4.20) and (4.21) for feed system a and b depending on operating pressure, composition and purity changes.

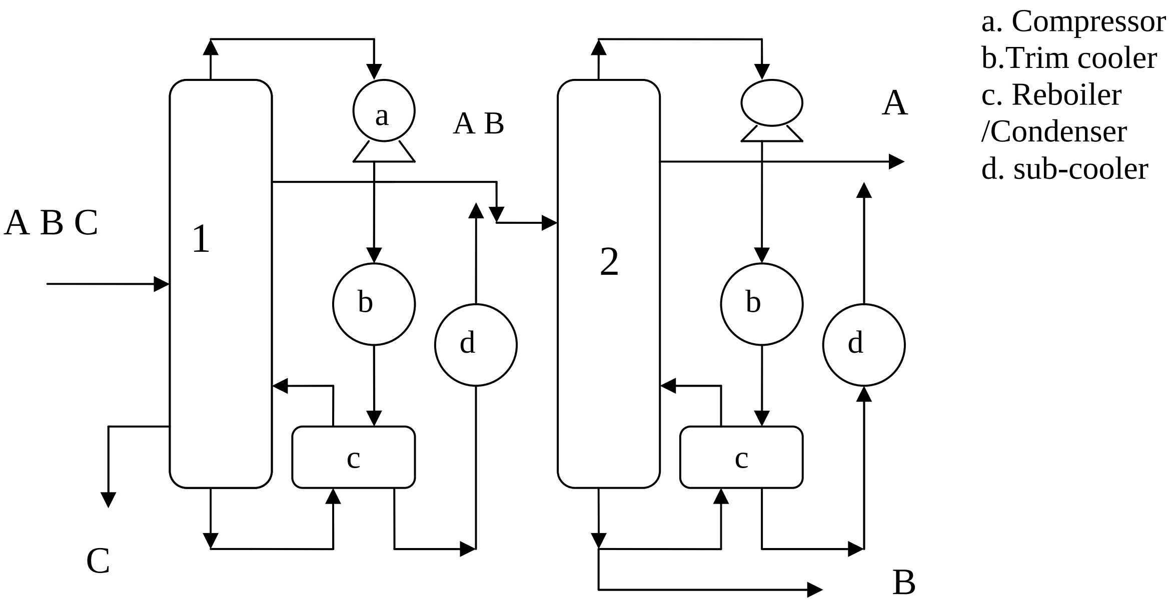

Figure (4.9) heat pump configuration for feed system a.

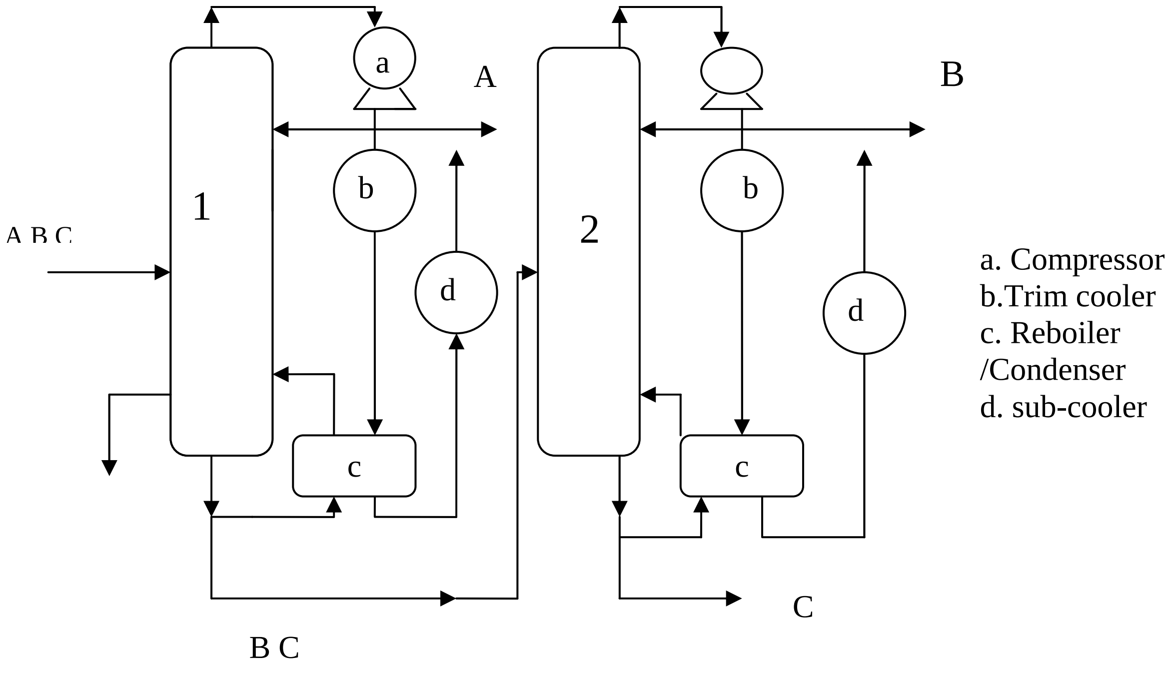

Figure (4.10) heat pump configuration for feed system b.

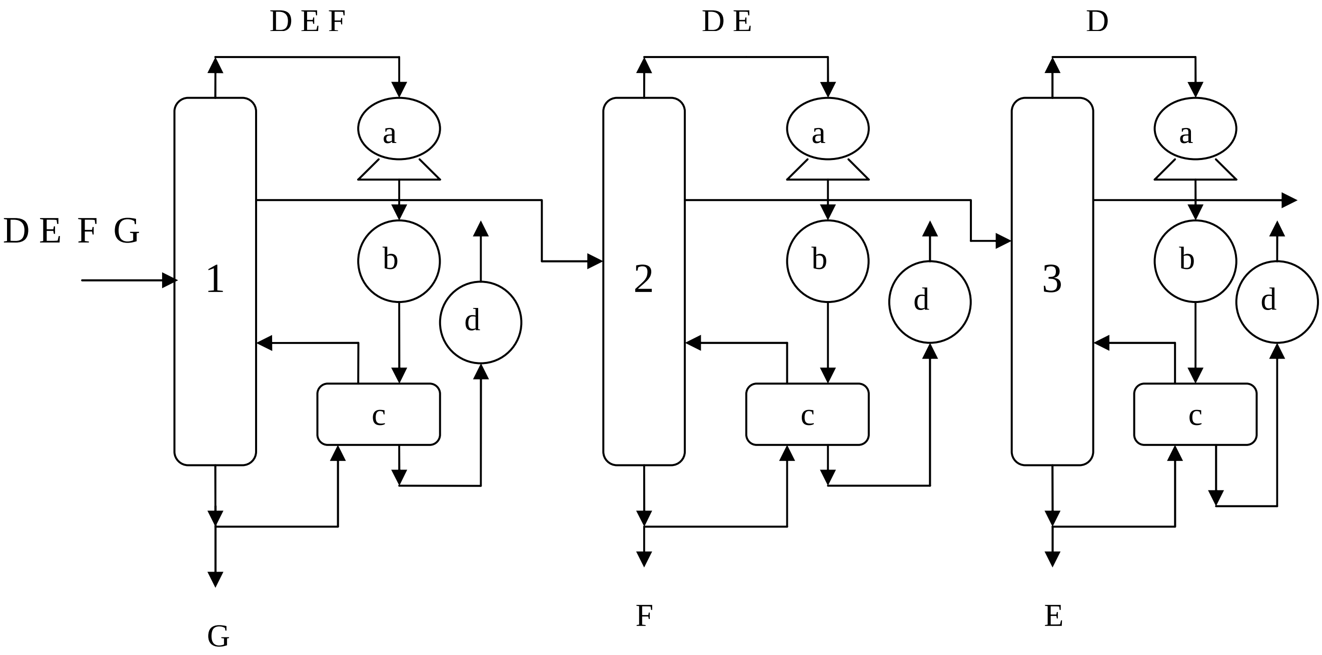

Figure (4.11) heat pump configuration for feed system c.

a. Compressor
b.Trim cooler
c. Reboiler
/Condenser
d. sub-cooler

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	25.59	28.64	30.46	30.00	29.98	29.99
1.0	28.65	33.69	36.65	33.61	32.65	33.50
1.5	28.36	33.51	34.05	33.60	33.00	33.21
2.0	22.65	40.25	41.56	46.31	26.31	46.00
2.5	53.65	55.69	59.64	57.89	57.02	61.01
3.0	63.21	63.53	65.52	65.81	64.30	65.23

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	24.66	25.69	28.98	28.65	28.63	28.65
1.0	25.31	32.36	30.00	30.24	30.00	30.02
1.5	25.06	31.56	29.36	30.00	29.05	30.04
2.0	20.56	39.25	18.56	45.01	24.69	44.32
2.5	50.36	53.12	55.36	56.69	55.36	56.23
3.0	60.64	60.98	61.52	62.36	62.06	62.30

R_C= 0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.88	24.33	25.36	26.98	25.12	26.02
1.0	23.90	30.36	26.36	28.95	27.93	28.05
1.5	23.00	30.32	34.21	28.84	27.90	28.04
2.0	20.36	35.69	37.89	41.25	41.06	42.35
2.5	48.25	51.24	22.65	55.36	54.85	55.45
3.0	56.33	59.69	59.69	59.92	58.00	59.21

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.87	24.03	24.53	25.00	24.56	24.99
1.0	23.00	24.55	25.33	27.36	27.22	27.56
1.5	23.64	24.35	32.28	27.21	27.22	27.44
2.0	16.94	33.65	35.21	40.20	39.15	40.18
2.5	46.22	50.00	19.33	53.11	48.65	52.66
3.0	44.36	56.09	56.65	57.00	56.13	56.55

Table (4.11) The effect of pressure change of column 1 and 2 on %saving in total energy consumption for feed system a (feed composition 1).

R_C=0.9

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	45.63	50.21	52.36	55.21	52.36	54.06
1.0	50.23	53.02	55.69	56.69	55.63	56.36
1.5	50.21	53.00	55.36	56.56	55.36	56.32
2.0	58.96	59.62	63.66	55.38	65.30	65.31
2.5	60.32	64.35	65.36	69.36	69.25	69.29
3.0	66.36	68.36	68.01	71.26	70.06	70.36

R_C= 0.925

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	44.32	45.36	49.63	50.25	50.13	50.15
1.0	46.23	47.85	54.45	52.36	51.22	52.15
1.5	46.01	40.69	54.00	52.31	52.00	52.09
2.0	55.36	57.58	57.66	63.85	57.00	57.25
2.5	59.01	60.56	62.35	61.82	71.31	63.55
3.0	62.01	63.58	65.36	66.25	65.56	76.02

R_C=0.95

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	41.02	43.62	46.40	48.90	48.40	48.80
1.0	44.22	45.69	51.00	48.53	48.16	48.22
1.5	19.21	33.21	50.03	48.41	48.00	48.03
2.0	51.00	53.00	55.00	56.84	56.00	56.22
2.5	54.05	56.02	57.32	59.00	58.49	58.91
3.0	55.00	57.00	58.01	63.99	62.02	63.66

R_C=0.99

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	40.30	43.36	45.36	47.36	47.01	47.22
1.0	43.05	45.63	48.31	48.62	48.44	48.55
1.5	17.96	45.75	48.05	48.40	48.00	48.12
2.0	50.63	52.30	54.36	55.36	55.20	55.23
2.5	53.01	55.36	58.65	58.32	57.21	58.24
3.0	43.69	56.11	59.00	60.50	60.12	60.40

Table (4.12) The effect of pressure change of column 1 and 2 on %saving in total energy consumption for feed system a (feed composition 2).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	24.25	25.54	26.46	27.13	26.13	26.79
1.0	25.02	25.74	27.88	29.20	29.20	29.82
1.5	24.55	25.21	28.43	30.07	28.66	30.46
2.0	17.25	34.28	35.00	38.35	36.46	37.85
2.5	42.50	47.40	45.99	46.64	42.10	48.10
3.0	47.95	47.98	48.25	49.40	46.00	46.71

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.69	24.36	25.00	26.36	26.00	26.20
1.0	24.02	25.36	27.03	28.90	28.02	28.56
1.5	24.00	25.25	26.56	28.80	28.00	28.51
2.0	15.99	33.69	34.21	37.36	35.00	35.36
2.5	42.00	46.32	44.56	45.36	40.06	46.22
3.0	44.69	47.00	47.36	48.99	45.00	46.11

R_C= 0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.65	23.62	24.25	25.54	25.00	25.22
1.0	23.90	23.70	25.36	27.32	26.00	26.30
1.5	23.36	23.36	25.00	27.12	26.36	27.00
2.0	24.56	32.36	33.65	36.23	36.02	36.06
2.5	41.00	42.00	43.36	44.21	44.02	44.09
3.0	42.32	46.36	45.38	46.36	45.00	45.92

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.00	23.03	23.53	24.90	23.52	24.11
1.0	23.23	23.55	24.53	26.26	24.12	25.06
1.5	23.02	23.35	24.50	25.99	25.00	25.30
2.0	24.35	30.15	32.51	33.00	32.15	32.36
2.5	40.00	41.00	45.36	46.13	46.04	46.08
3.0	40.02	45.29	46.35	47.90	46.36	47.06

Table (4.13) The effect of pressure change of column 1 and 2 on %saving in total energy consumption for feed system a (feed composition 3).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	24.00	25.00	25.46	27.00	26.45	26.99
1.0	24.56	25.99	28.32	28.25	28.05	28.10
1.5	24.00	25.01	28.30	28.06	28.00	28.02
2.0	35.45	35.12	35.68	38.65	37.69	37.19
2.5	43.00	46.50	22.99	45.69	45.20	45.50
3.0	40.25	47.02	47.25	47.90	47.00	47.63

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.09	24.36	25.00	26.36	26.00	26.00
1.0	23.22	25.36	27.83	27.88	27.02	27.25
1.5	23.13	25.25	27.06	27.00	26.52	26.55
2.0	25.99	30.69	34.21	37.36	35.00	36.36
2.5	37.00	46.32	18.56	45.36	45.00	45.23
3.0	42.31	47.00	47.36	46.00	45.38	45.99

R_C= 0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.01	23.00	23.99	25.80	24.56	25.44
1.0	23.90	23.89	24.34	25.88	24.90	25.36
1.5	23.36	23.36	24.00	25.19	25.01	25.15
2.0	24.62	30.36	32.55	33.13	32.01	33.06
2.5	36.00	33.00	36.36	37.11	36.12	37.00
3.0	33.25	41.36	42.34	43.36	43.00	43.29

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	22.55	22.62	23.65	24.64	24.00	24.59
1.0	22.80	23.81	24.66	26.32	26.00	26.31
1.5	22.38	23.36	24.00	26.12	26.08	26.10
2.0	24.20	30.36	31.69	31.63	30.02	31.16
2.5	35.00	32.00	35.66	36.01	35.45	36.00
3.0	30.32	41.26	42.36	43.35	43.02	43.22

Table (4.14) The effect of pressure change of column 1 and 2 on %saving in total energy consumption for feed system a (feed composition 4).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.51	24.00	26.25	26.21	26.89	26.89
1.0	24.01	24.56	27.10	26.85	26.99	27.00
1.5	22.10	26.00	29.99	28.32	28.55	28.56
2.0	26.04	27.66	34.00	32.12	33.00	33.56
2.5	20.00	27.85	35.03	33.52	33.74	33.80
3.0	33.21	33.56	40.21	34.28	35.23	37.00

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	23.00	23.52	27.00	26.45	26.74	26.85
1.0	23.56	24.02	27.45	26.51	26.79	26.89
1.5	20.04	25.32	29.85	27.02	28.42	28.11
2.0	25.36	27.02	33.36	32.22	32.12	32.55
2.5	19.56	29.36	35.21	33.35	33.01	34.45
3.0	32.21	34.32	37.00	34.23	35.02	36.20

R_C= 0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	22.08	22.85	26.36	25.33	25.61	25.88
1.0	23.55	24.00	27.96	26.36	26.45	27.02
1.5	19.02	26.04	32.32	30.21	31.20	31.65
2.0	27.00	27.32	33.00	31.20	32.25	32.64
2.5	18.23	31.45	34.65	32.00	33.70	33.25
3.0	31.00	32.12	36.32	35.32	35.62	35.85

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	20.22	21.30	25.01	24.33	24.63	24.69
1.0	22.83	23.46	27.84	25.16	26.35	26.99
1.5	17.52	25.26	30.85	26.26	28.06	29.32
2.0	26.45	27.36	32.12	27.85	29.00	30.36
2.5	16.00	28.23	33.36	31.32	42.22	33.00
3.0	28.01	29.02	35.32	32.25	33.00	43.25

Table (4.15) The effect of pressure change of column 1 and 2 on % saving in total energy consumption for feed system b (feed composition 1).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	15.36	17.85	20.36	19.25	19.85	19.99
1.0	17.36	18.39	24.36	20.14	21.36	22.56
1.5	18.00	19.21	25.36	21.36	22.00	22.36
2.0	20.36	21.36	28.36	34.36	25.36	26.00
2.5	21.39	22.36	33.36	26.36	37.36	29.00
3.0	23.36	25.01	40.36	28.69	29.00	40.36

R_C= 0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	14.00	16.02	18.00	17.21	17.25	17.36
1.0	15.25	16.85	19.23	18.24	18.36	18.99
1.5	16.36	17.98	19.89	18.45	18.76	19.00
2.0	17.23	18.00	23.21	20.36	21.20	22.00
2.5	18.00	18.52	24.36	21.36	22.52	23.23
3.0	35.00	35.36	39.21	37.63	37.98	38.25

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	13.25	16.00	17.52	16.99	17.00	17.14
1.0	13.36	16.36	17.70	16.99	17.05	17.63
1.5	13.65	16.36	17.99	17.01	17.25	17.39
2.0	16.00	17.98	18.36	18.06	17.32	17.96
2.5	18.14	20.36	22.36	20.31	21.00	22.00
3.0	35.25	36.32	38.12	37.25	37.55	37.36

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	12.14	15.36	16.36	16.00	16.10	16.22
1.0	12.91	15.23	16.36	16.02	16.20	16.30
1.5	13.21	16.32	17.32	16.25	16.88	17.00
2.0	18.36	20.36	21.36	20.00	20.12	20.23
2.5	23.32	25.01	25.41	24.00	24.65	25.10
3.0	31.01	32.52	37.00	32.23	32.21	32.89

Table (4.16) The effect of pressure change of column 1 and 2 on % saving in total energy consumption for feed system b (feed composition 2).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	50.00	52.36	56.58	53.12	54.23	54.25
1.0	53.12	53.69	59.23	55.84	57.36	58.02
1.5	58.36	59.25	60.06	57.36	58.00	58.36
2.0	62.36	67.00	69.36	68.54	59.68	60.18
2.5	63.79	68.36	71.23	66.36	67.45	68.36
3.0	65.36	67.12	78.00	69.45	71.45	76.00

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	42.19	44.78	48.36	44.52	45.12	46.36
1.0	44.33	45.58	50.00	45.36	48.41	48.69
1.5	45.25	46.85	51.25	46.27	48.88	48.89
2.0	46.00	47.58	53.36	5213	51.42	52.36
2.5	50.01	51.23	55.25	51.46	52.40	53.78
3.0	51.25	53.12	57.36	55.45	56.23	57.85

R_C= 0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	41.00	42.52	43.12	43.56	43.62	43.78
1.0	45.36	45.51	46.25	45.84	45.98	45.99
1.5	43.12	44.00	47.23	43.14	43.23	44.56
2.0	55.00	56.78	59.25	54.32	55.64	55.85
2.5	57.36	58.00	60.12	55.63	56.25	57.15
3.0	60.14	61.45	63.36	57.52	58.12	59.00

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	40.12	41.25	43.12	40.56	41.87	42.02
1.0	41.35	41.65	44.00	42.25	43.00	43.65
1.5	42.00	43.25	45.02	42.56	43.51	44.65
2.0	53.63	54.56	56.78	55.12	55.32	55.74
2.5	54.45	55.00	57.36	55.63	56.10	56.85
3.0	55.36	56.51	59.85	57.36	56.12	57.93

Table (4.17) The effect of pressure change of column 1 and 2 on % saving in total energy consumption for feed system b (feed composition 3).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	34.00	34.28	35.02	34.10	34.20	34.52
1.0	34.65	35.04	36.14	34.56	34.85	35.63
1.5	36.00	36.21	37.10	35.12	35.65	36.21
2.0	36.82	36.85	37.74	36.25	36.75	36.86
2.5	37.05	37.89	43.00	39.23	39.81	40.10
3.0	43.00	44.52	48.23	42.10	43.56	44.00

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	33.20	33.65	34.00	33.85	33.85	33.96
1.0	33.85	34.10	34.85	33.63	33.69	33.78
1.5	34.21	35.01	36.00	34.01	35.01	35.64
2.0	35.02	37.36	39.01	37.21	36.14	37.25
2.5	37.01	37.85	42.32	38.36	38.96	39.01
3.0	37.36	37.95	46.00	39.01	39.45	39.85

R_C= 0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	32.00	33.56	34.12	33.46	33.85	33.98
1.0	33.21	34.00	35.23	33.41	33.21	34.00
1.5	34.25	34.58	36.89	35.21	35.56	35.65
2.0	35.04	35.69	37.85	37.00	36.52	36.56
2.5	38.02	38.10	41.00	36.25	37.84	38.99
3.0	36.45	37.00	45.62	37.42	38.36	39.12

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	30.21	31.20	33.52	32.10	32.21	33.00
1.0	31.25	32.00	34.12	33.26	33.52	33.69
1.5	32.52	33.04	35.36	33.56	34.00	34.15
2.0	32.55	34.36	36.00	34.12	34.45	35.21
2.5	38.23	38.99	40.31	36.21	37.24	38.00
3.0	40.00	41.23	43.25	38.21	38.52	39.01

Table (4.18) The effect of pressure change of column 1 and 2 on % saving in total energy consumption for feed system b (feed composition 4).

Possibility No.	Feed composition				
	1	2	3	4	5
1	19.00	41.25	20.03	19.36	58.00*
2	17.05	18.01	42.15	54.26*	42.36
3	15.20	46.36	40.25	42.13	71.00*
4	18.79	25.16	46.36*	33.21	28.36
5	30.12	42.36	52.12	63.22*	13.15
6	19.12	45.10	13.20	64.13*	52.33
7	35.36	19.66	42.12	29.12	43.00*
8	15.00	42.36	53.41	25.52	55.00*
9	46.01	36.12	60.12*	41.10	41.32
10	22.00	34.13	18.36	32.02	44.00*
11	20.23	41.26	62.36*	56.04	56.02
12	27.15	28.21	67.21*	56.22	42.36
13	16.00	32.01	50.13	41.00	74.00*
14	30.36	47.12	55.28	66.32*	53.01
15	41.02	28.20	50.11*	46.20	33.36
16	42.06	23.14	46.25*	16.02	31.02
17	33.26	13.36	74.32*	18.42	42.36
18	42.36	42.01	19.25	45.25	63.23*
19	22.15	18.04	46.23	73.12*	42.00
20	36.01	12.02	71.55*	69.23	56.36
21	25.13	41.32	52.10	68.40*	46.25
22	75.13*	18.36	55.13	62.22	40.20
23	52.14*	26.36	28.16	35.02	30.05
24	46.39	19.82	25.33	41.25	64.23*
25	52.36	71.12	66.12	75.23*	41.25
26	45.00	52.36*	17.33	45.36	48.02
27	41.12	50.00	59.23	70.00	73.25*

Table (4.19) Effect of % saving in total energy consumption with change in feed composition for (27 possibilities) of feed system c.

Feed comp-osit-ion	R_c=0.9		R_c=0.925		R_c=0.95		R_c=0.99	
	Column		Column		column		column	
	1	**2**	**1**	**2**	**1**	**2**	**1**	**2**
1	3.0	2.0	3.0	2.0	3.0	2.0	3.0	2.0
2	3.0	2.0	3.0	2.0	3.0	2.0	3.0	2.0
3	3.0	2.0	3.0	2.0	3.0	2.0	3.0	2.0
4	3.0	2.0	3.0	2.0	3.0	2.0	3.0	2.0

Table (4.20) optimum operating pressures with respect to percentage saving total energy consumption for feed system a (atm.).

Feed comp-osit-ion	R_c=0.9		R_c=0.925		R_c=0.95		R_c=0.99	
	Column		Column		column		column	
	1	**2**	**1**	**2**	**1**	**2**	**1**	**2**
1	3.0	1.5	3.0	1.5	3.0	1.5	3.0	1.5
2	3.0	1.5	3.0	1.5	3.0	1.5	3.0	1.5
3	3.0	1.5	3.0	1.5	3.0	1.5	3.0	1.5
4	3.0	1.5	3.0	1.5	3.0	1.5	3.0	1.5

Table (4.21) optimum operating pressures with respect to percentage saving total energy consumption for feed system b (atm.).

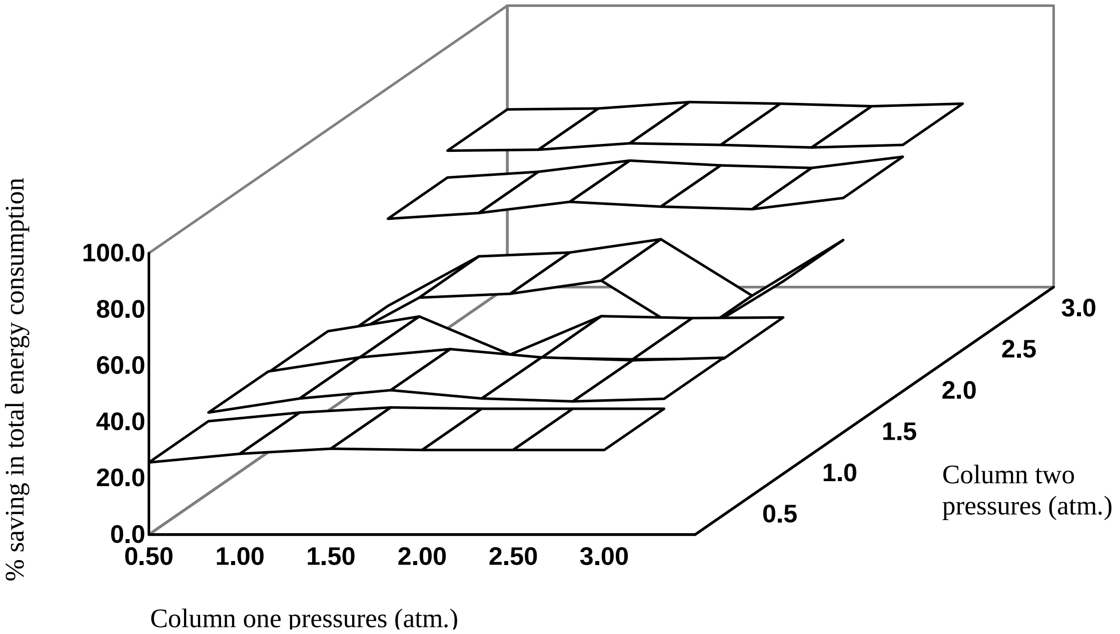

Figure (4.12) the effect of pressure change of column 1 and 2 on total energy consumption for feed system a (feed composition 1 and recovery = 0.9).

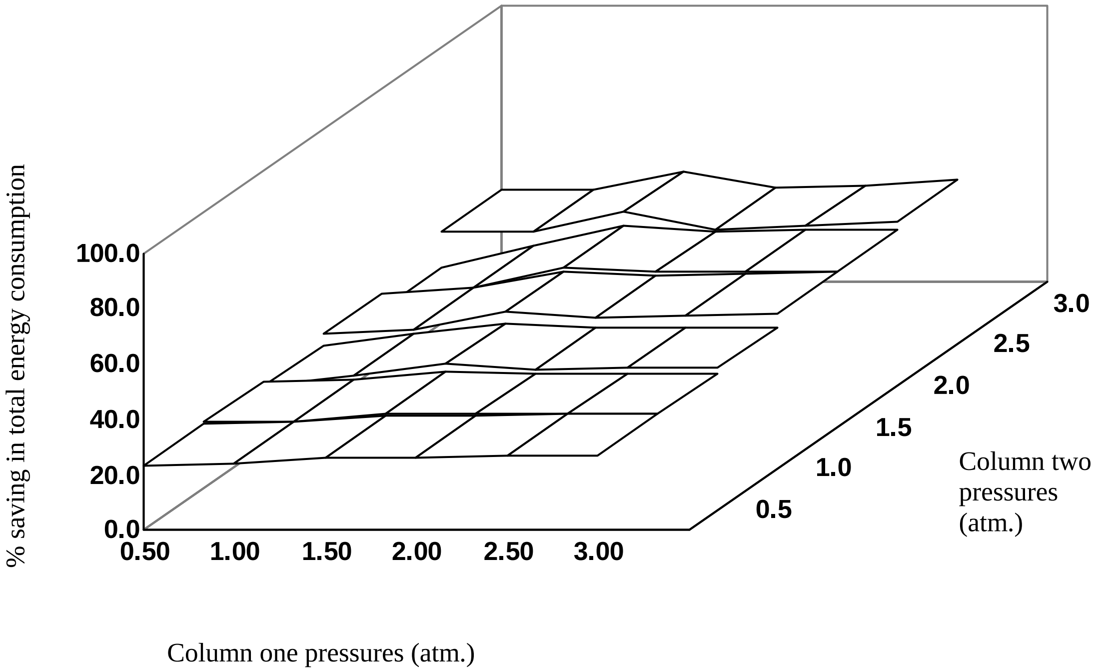

Figure (4.13) the effect of pressure change of column 1 and 2 on total energy consumption for feed system b (feed composition 1 and recovery = 0.9)

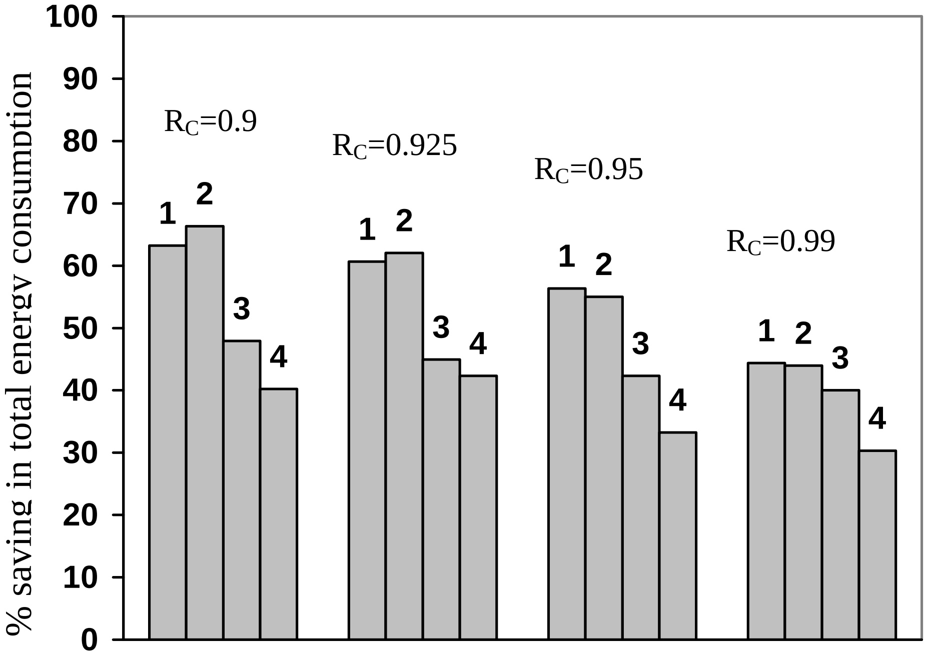

Figure (4.14) the effect of feed composition change on %saving in total energy consumption for feed system a (column one at 3 atm and column two at 0.5 atm).

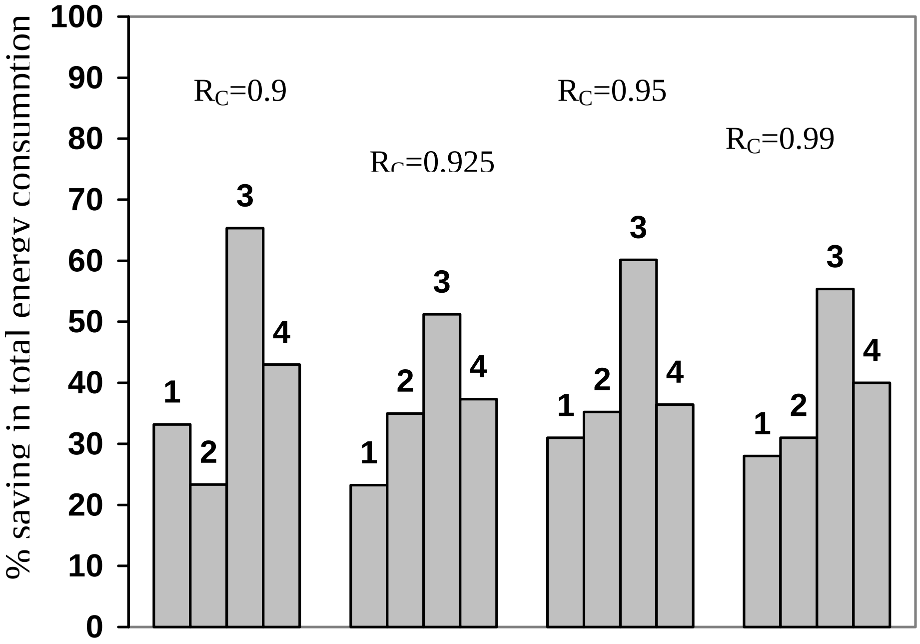

Figure (4.15) the effect of feed composition change on % saving in total energy consumption for feed system b (column one at 3 atm and column two t 0.5 atm).

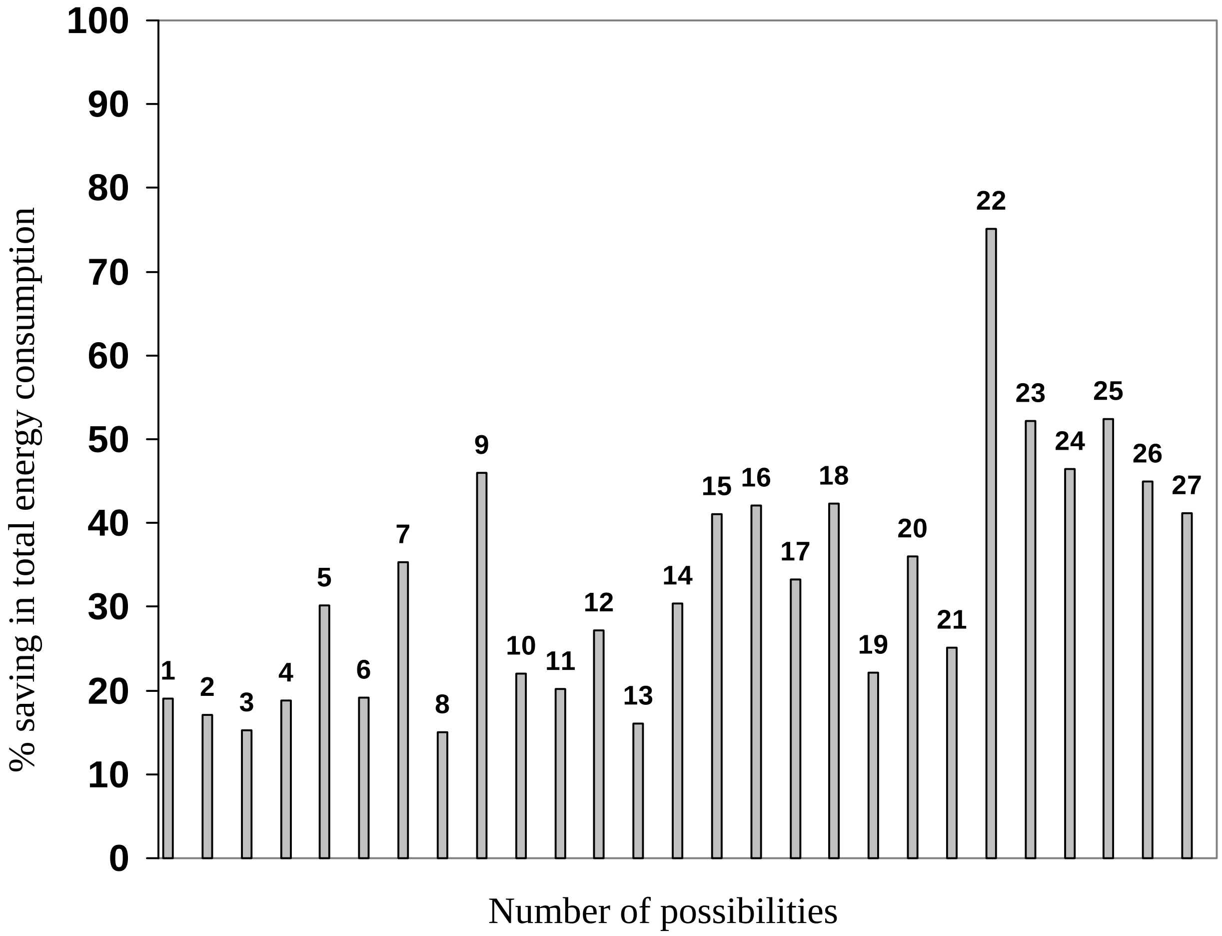

Figure (4.16) the effect of all possibilities of the pressure change for the three columns on % saving in total energy consumption for feed system c (feed composition one).

4.6 Effect of process variables on the percentage saving in total annual cost with heat pump:

Applying heat pump system causes a tangible reduction in total annual cost of the distillation column. In this work the percentage saving in total annual cost was predicted for the feed systems a b and c, and for all feed compositions , fractional recoveries and operating pressure as it was given earlier. The percentage saving in total annual cost was calculated by:

Percentage saving in total annual cost= $\frac{CAN - CAH}{CAN}$ *100

Where:

CAN = Total annual cost for conventional system

CAH = Total annual cost using heat pump system

4.6.1 Effect of pressure:

The effect of changing the pressure on percentage saving in total annual cost was shown in tables (4.22-4.30) for all feed systems and it was also given in figure (4.17) for feed system a, figure (4.18) for feed system b and figure. It was found that increasing the pressure will decrease the percentage saving in total annual cost and this is due to increasing in total annual cost of the system with increasing operating pressure.

4.6.2 Effect of feed composition:

The percentage saving in total annual cost was calculated according to the above equation and it was given in tables (4.22-4.25) for feed system a, tables (4.26-4.29) for feed system b and table (4.30) for feed system c, and shown in figures (4.19) for feed system a and figure (4.20) for feed system b and (4.21) for feed system c, it was found that:

1. For feed system a feed composition 3 show a maximum saving in total cost.
2. For feed system b feed composition 1 shows a maximum saving in total cost.

3. for feed system c, for possibilities number 1, 7, 8 and 27 feed composition 1 shows a maximum percentage saving in total annual cost, also feed composition 2 shows a maximum percentage saving in total annual cost for possibilities number 13, 15, 17, 18, 19, 21, 23 and 25, feed composition 3 has a maximum percentage saving in total annual cost for possibilities number 3, 6, 9, 10, 11, 12, 14, 16, 20 and 26.

4.6.3 Effect of fractional recovery:

The results of changing fractional recovery for four values of 0.9, 0.925, 0.95, 0.9 was given in tables (4.22-4.30) for all feed systems and these results show that by increasing the fractional recovery the percentage saving in total annual cost decrease for all feed systems.

4.6.4 Regions of optimality:

The regions of optimality were given in tables (4.32) and (4.33) for feed system a and b depending on composition and purity changed.

R_C=0.9

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	65.36	63.00	61.46	60.48	58.00	55.46
1.0	64.20	62.48	59.25	59.00	57.46	53.01
1.5	63.00	61.49	58.46	57.42	55.29	51.00
2.0	62.46	58.15	57.15	56.02	54.00	48.25
2.5	60.02	57.96	55.36	54.00	53.12	45.01
3.0	57.56	56.01	54.32	53.01	50.25	43.00

R_C= 0.925

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	63.00	60.00	59.85	58.00	56.00	55.00
1.0	60.15	58.01	57.00	56.13	55.14	53.01
1.5	59.17	56.45	55.63	52.12	52.02	50.15
2.0	58.96	53.01	52.00	49.00	48.00	48.00
2.5	55.78	50.36	49.85	48.25	49.05	47.63
3.0	54.96	49.00	48.00	48.00	47.02	46.12

R_C=0.95

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	62.36	60.25	59.25	56.00	53.25	50.13
1.0	59.12	59.25	56.00	55.58	50.12	49.36
1.5	57.00	57.00	54.12	53.26	49.63	47.01
2.0	56.02	53.01	52.02	50.12	48.25	46.02
2.5	52.36	50.02	48.89	47.58	46.00	45.00
3.0	50.12	48.36	47.21	46.00	43.01	40.36

R_C=0.99

P_2 P_1	0.5	1.0	1.5	2.0	2.5	3.0
0.5	59.25	57.00	55.36	55.00	52.00	50.25
1.0	55.01	52.01	52.02	53.01	50.12	48.12
1.5	50.00	49.36	48.36	47.00	45.00	44.56
2.0	46.00	45.14	43.02	42.69	40.25	39.25
2.5	42.01	40.36	39.01	38.00	35.00	34.00
3.0	40.23	39.03	38.00	37.89	34.01	33.36

Table (4.22) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system a (feed composition 1).

R_C=0.9

P_1 \ P_2	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	60.00	58.00	57.45	56.00	55.23	53.01
1.0	59.02	57.46	54.36	53.23	53.23	51.36
1.5	58.46	56.25	53.12	51.25	51.25	48.36
2.0	58.00	54.00	52.13	49.85	50.36	46.12
2.5	57.02	52.02	50.25	48.12	44.36	43.25
3.0	55.69	50.36	49.36	47.00	43.01	40.46

R_C=0.925

P_1 \ P_2	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	58.00	55.36	52.00	49.00	45.25	44.00
1.0	57.56	53.12	49.01	45.02	43.36	43.36
1.5	55.25	50.36	48.78	43.36	41.25	40.25
2.0	52.13	46.66	45.36	40.25	39.99	38.12
2.5	50.36	44.12	44.25	39.02	38.45	36.02
3.0	49.36	44.00	42.00	38.46	37.00	35.00

R_C= 0.95

P_1 \ P_2	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	55.00	52.12	49.00	45.36	43.12	40.00
1.0	52.32	48.63	47.78	44.36	41.36	37.12
1.5	48.79	47.56	45.69	42.36	39.56	32.36
2.0	45.36	42.32	41.36	39.36	36.46	30.12
2.5	43.12	39.99	38.12	36.45	33.36	29.56
3.0	40.00	38.45	37.69	35.00	30.36	29.00

R_C=0.99

P_1 \ P_2	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	52.12	50.56	47.00	40.21	36.13	34.00
1.0	50.36	46.36	42.36	37.46	33.25	32.13
1.5	46.12	41.23	40.13	35.00	30.13	28.12
2.0	44.36	38.45	37.85	32.12	28.00	26.46
2.5	42.12	35.12	35.00	30.46	27.46	25.13
3.0	40.23	34.00	33.01	29.00	25.00	22.00

Table (4.23) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system a (feed composition 2).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	75.00	74.85	72.01	70.25	69.85	66.52
1.0	73.21	72.15	71.41	68.46	68.14	64.32
1.5	70.15	70.00	69.12	66.52	65.45	63.48
2.0	68.45	68.01	67.14	63.12	63.00	60.21
2.5	66.14	65.23	64.13	61.00	60.00	59.12
3.0	65.00	61.02	63.02	60.23	58.12	58.00

R_C= 0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	73.12	72.00	70.15	69.45	68.25	67.00
1.0	72.46	71.46	69.25	67.42	64.52	66.49
1.5	70.53	70.10	67.17	64.12	63.16	65.23
2.0	68.12	66.23	66.13	62.52	60.48	64.33
2.5	66.23	65.25	63.16	61.36	58.25	57.45
3.0	65.00	62.01	60.12	58.23	57.25	56.00

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	72.00	70.15	68.00	65.36	63.00	60.46
1.0	70.13	69.85	66.49	63.58	62.16	59.40
1.5	69.42	67.18	65.28	62.46	60.28	57.00
2.0	66.14	65.13	63.94	60.59	59.13	55.23
2.5	65.00	63.15	60.23	59.58	57.15	54.16
3.0	63.20	60.36	58.46	57.13	55.36	53.00

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	70.25	69.15	57.00	55.16	54.36	52.05
1.0	69.46	66.16	56.13	52.13	51.00	50.13
1.5	66.15	65.00	55.13	50.36	49.12	48.00
2.0	65.00	64.51	54.26	49.25	48.36	46.25
2.5	63.15	63.25	53.36	48.76	45.00	44.13
3.0	62.13	62.00	51.00	46.13	44.30	43.10

Table (4.24) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system a (feed composition 3).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	55.46	53.12	50.18	47.45	45.00	43.12
1.0	52.13	50.26	48.00	46.89	44.23	41.25
1.5	47.12	45.00	46.12	42.36	42.12	39.89
2.0	45.00	42.12	42.02	40.13	40.36	37.46
2.5	44.00	40.36	39.99	38.56	37.89	35.88
3.0	42.01	38.00	37.00	35.00	33.22	32.00

R_C= 0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	53.23	50.36	48.46	45.00	44.01	45.00
1.0	50.13	48.46	46.52	43.36	43.25	43.36
1.5	48.00	45.36	43.12	42.15	40.25	38.46
2.0	46.02	42.32	41.21	40.25	38.65	35.56
2.5	44.04	40.36	39.85	38.46	37.65	34.23
3.0	42.99	38.56	37.46	36.45	35.00	33.36

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	50.36	48.13	45.00	44.42	40.13	38.12
1.0	48.36	45.23	42.12	40.13	38.46	35.36
1.5	46.25	40.01	39.13	38.46	37.04	33.12
2.0	39.56	38.36	36.01	35.63	33.12	30.23
2.5	37.12	35.36	35.56	33.00	30.12	29.87
3.0	35.00	33.33	32.00	30.33	28.13	28.00

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	48.00	45.00	43.12	40.23	36.13	35.12
1.0	45.36	44.23	40.25	38.01	34.32	33.23
1.5	43.01	42.16	38.46	35.16	31.02	30.45
2.0	38.13	38.00	36.56	32.36	27.36	28.45
2.5	36.12	35.03	32.10	29.46	25.00	26.12
3.0	35.00	33.36	30.36	27.36	23.00	22.36

Table (4.25) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system a (feed composition 4).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	60.00	58.00	55.13	53.56	50.14	48.00
1.0	58.12	56.36	53.00	50.13	48.41	45.12
1.5	56.13	55.12	51.13	49.41	45.00	44.23
2.0	55.12	53.12	49.46	46.42	42.13	42.12
2.5	53.25	50.12	48.41	44.02	40.17	39.00
3.0	51.00	48.00	45.00	43.00	38.00	37.04

R_C= 0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	58.00	56.12	55.00	53.14	50.11	48.00
1.0	55.45	54.23	52.14	50.22	47.00	45.36
1.5	53.12	50.25	50.17	46.32	45.13	43.12
2.0	49.85	48.41	47.46	45.88	42.10	41.15
2.5	46.02	45.00	45.12	44.00	40.25	39.45
3.0	44.02	43.12	41.36	40.25	39.00	38.12

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	55.12	53.26	52.12	50.18	48.01	46.45
1.0	54.36	50.48	50.44	48.41	45.58	45.00
1.5	53.10	48.41	48.00	46.25	44.10	44.25
2.0	50.17	45.36	44.52	43.01	42.58	41.25
2.5	48.00	43.12	42.00	40.18	38.12	37.00
3.0	47.45	40.33	39.00	38.45	37.00	35.02

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	53.12	50.17	49.45	48.45	45.00	44.00
1.0	50.14	48.52	46.12	45.36	43.12	43.15
1.5	48.45	43.32	43.15	40.12	40.28	40.25
2.0	45.36	40.12	40.00	39.23	38.45	38.00
2.5	44.12	39.45	38.41	37.00	35.00	36.15
3.0	42.12	36.00	35.11	34.18	33.46	35.00

Table (4.26) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system b (feed composition 1).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	55.00	53.22	50.13	48.00	47.25	45.23
1.0	53.12	51.47	49.32	46.52	45.13	43.22
1.5	52.01	48.00	47.00	43.12	42.03	41.25
2.0	50.45	46.45	45.25	40.12	40.36	39.41
2.5	48.52	42.32	43.15	38.15	37.56	35.00
3.0	46.32	41.12	40.00	36.00	35.68	33.78

R_C=0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	52.12	50.23	48.41	37.00	35.02	35.00
1.0	50.36	48.41	46.15	36.48	34.18	34.02
1.5	49.01	46.25	45.00	35.22	33.58	32.45
2.0	47.00	44.41	43.15	34.19	30.12	30.12
2.5	45.17	42.11	40.17	32.00	28.33	27.85
3.0	44.00	40.33	39.23	30.31	27.03	26.45

R_C= 0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	50.00	48.00	47.45	35.25	33.12	32.00
1.0	48.12	46.02	45.63	33.63	31.25	30.45
1.5	46.23	46.00	44.00	31.02	30.15	28.45
2.0	45.12	45.01	42.13	29.12	29.78	27.46
2.5	44.20	43.01	40.45	29.36	28.46	26.25
3.0	43.12	40.00	39.00	28.00	28.00	25.12

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	48.25	45.00	44.00	43.25	40.22	38.10
1.0	46.13	44.20	43.25	41.15	38.52	37.00
1.5	45.22	41.23	40.22	38.52	37.02	36.46
2.0	42.15	39.79	38.15	37.52	36.41	35.00
2.5	40.22	39.22	38.00	36.42	35.36	32.10
3.0	39.48	38.45	35.00	35.00	33.33	30.00

Table (4.27) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system b (feed composition 2).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	52.45	49.00	48.11	45.00	43.18	40.23
1.0	51.00	48.23	46.41	43.10	40.17	38.12
1.5	49.85	46.45	45.00	40.17	38.00	35.32
2.0	47.12	45.00	41.25	38.46	36.12	33.14
2.5	45.12	43.25	40.36	37.04	35.00	30.12
3.0	44.00	40.12	39.00	35.00	32.01	28.33

R_C= 0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	50.14	48.47	45.28	43.01	40.45	38.45
1.0	48.00	46.25	44.00	42.00	38.46	35.25
1.5	45.33	43.21	43.36	40.28	37.00	33.36
2.0	41.11	41.28	42.15	39.17	36.45	30.12
2.5	39.02	38.00	38.00	37.79	35.36	29.00
3.0	39.00	37.46	37.02	35.00	34.22	27.45

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	46.25	43.15	40.15	39.00	38.41	35.12
1.0	43.12	41.28	37.45	37.48	36.00	33.20
1.5	40.18	37.32	36.12	36.46	34.12	31.00
2.0	38.46	37.00	35.00	35.00	33.00	30.36
2.5	37.45	36.56	34.18	33.41	32.45	29.00
3.0	36.00	34.02	33.33	32.00	30.33	28.45

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	43.12	40.02	38.46	35.47	33.14	30.17
1.0	40.17	38.45	35.12	33.23	30.18	28.42
1.5	38.45	37.00	33.17	30.17	28.45	27.63
2.0	37.36	35.28	30.00	28.45	27.12	24.23
2.5	36.14	33.00	28.45	25.47	23.15	22.00
3.0	34.00	31.25	27.28	23.12	21.00	20.01

Table (4.28) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system b (feed composition 3).

R_C=0.9

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	45.00	43.28	40.25	38.00	37.02	35.00
1.0	42.17	40.14	38.74	37.12	35.46	34.12
1.5	40.65	39.01	35.62	35.00	33.47	30.25
2.0	38.49	37.46	33.17	32.48	30.28	28.12
2.5	35.12	34.00	30.47	30.17	28.00	27.45
3.0	33.10	32.17	28.45	28.00	27.25	25.00

R_C= 0.925

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	43.05	40.74	38.11	35.48	33.17	30.16
1.0	40.17	37.85	37.85	33.22	30.23	28.00
1.5	38.45	35.12	33.89	30.36	28.52	26.41
2.0	37.00	33.29	30.17	28.12	27.22	23.36
2.5	33.36	30.17	28.45	27.25	25.36	20.16
3.0	31.17	28.00	24.36	23.00	23.41	19.89

R_C=0.95

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	40.14	38.71	35.25	33.48	30.11	28.52
1.0	37.28	36.25	34.00	30.17	29.23	26.54
1.5	36.42	34.25	33.56	28.03	27.41	23.22
2.0	35.16	33.00	33.12	27.00	25.22	22.03
2.5	33.25	29.02	29.00	26.25	23.10	20.00
3.0	30.00	28.02	28.00	24.22	20.00	18.00

R_C=0.99

P_2 P_1	**0.5**	**1.0**	**1.5**	**2.0**	**2.5**	**3.0**
0.5	39.01	37.00	36.04	35.00	32.28	30.00
1.0	38.41	35.41	35.00	33.18	29.12	28.25
1.5	36.13	4.12	32.13	30.43	27.36	26.41
2.0	33.13	32.28	30.17	27.46	25.23	23.12
2.5	32.00	30.14	28.04	25.12	22.30	20.15
3.0	28.01	28.01	25.02	22.00	20.44	17.89

Table (4.29) the effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system b (feed composition 4).

Possibility No.	Feed composition				
	1	2	3	4	5
1	77.36	62.00	60.33	51.02	60.12
2	77.00	60.36	58.02	50.42	57.25
3	74.00	59.28	56.60	50.13	56.12
4	73.21	58.02	56.56	50.00	50.13
5	71.58	56.13	56.42	49.36	49.41
6	71.58	56.13	56.42	49.36	49.41
7	70.12	53.01	55.22	48.89	46.12
8	70.12	51.13	53.00	48.74	44.15
9	69.39	50.00	52.33	45.00	39.46
10	69.12	49.78	52.36	44.15	36.15
11	69.00	49.35	51.13	43.28	34.46
12	67.82	49.00	50.00	42.58	35.12
13	66.23	8.25	48.56	42.15	31.25
14	64.13	47.46	46.12	41.36	30.15
15	62.22	46.25	44.45	40.12	30.45
16	60.01	43.18	42.13	39.45	30.00
17	59.23	42.12	40.15	38.36	29.46
18	57.58	39.04	37.00	34.58	28.82
19	55.00	38.88	36.10	33.59	26.47
20	54.12	36.18	31.00	32.00	25.11
21	53.011	35.02	29.12	31.12	24.52
22	50.13	34.23	28.00	30.01	20.10
23	45.36	30.13	26.13	29.28	19.13
24	42.00	26.36	24.00	25.13	18.46
25	23.00	25.25	23.36	20.00	17.80
26	20.02	23.10	20.36	18.41	16.25
27	12.30	10.02	11.02	12.00	10.00

Table (4.30) effect of percentage saving in total annual cost with change in feed composition for (27 possibilities) of feed system c.

Feed composition	R_c=0.9		R_c=0.925		R_c=0.95		R_c=0.99	
	Column		Column		column		column	
	1	**2**	**1**	**2**	**1**	**2**	**1**	**2**
1	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
2	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
3	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
4	0.5	0.5	0.5	0.5	0.5	0.5	0.5	05

Table (4.32) optimum operating pressures with respect to percentage saving total annual cost for feed system a (atm.).

Feed composition	R_c=0.9		R_c=0.925		R_c=0.95		R_c=0.99	
	Column		Column		column		column	
	1	**2**	**1**	**2**	**1**	**2**	**1**	**2**
1	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
2	05	0.5	0.5	0.5	0.5	0.5	0.5	0.5
3	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
4	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5

Table (4.33) optimum operating pressures with respect to percentage saving total annual cost for feed system b (atm.).

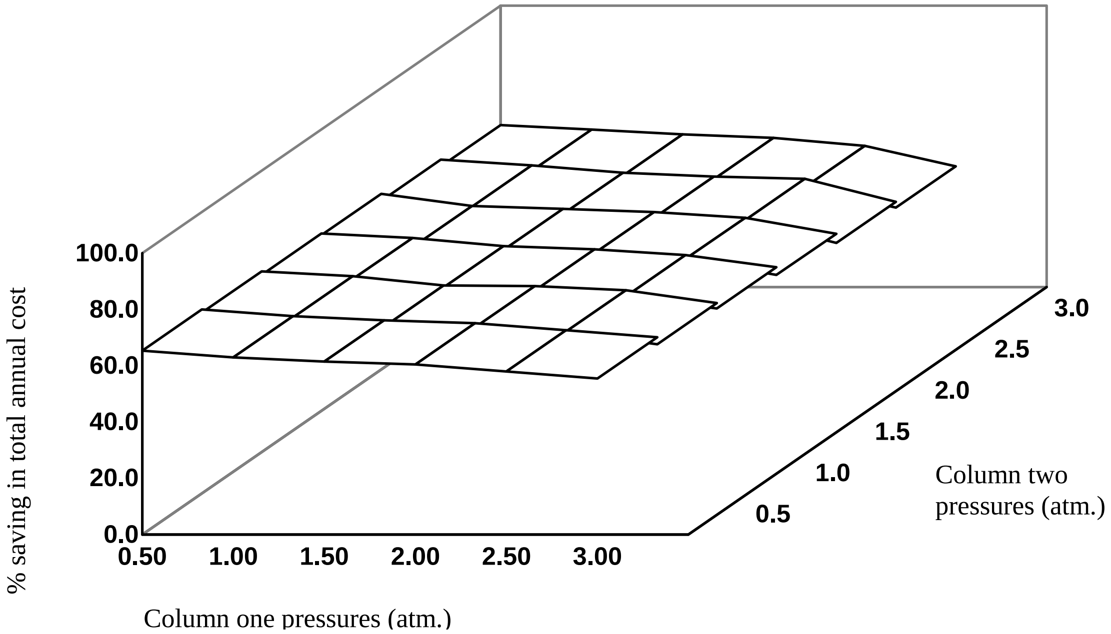

Figure (4.17) the effect of pressure change of column 1 and 2 on total annual cost for feed system a (feed composition 1 and recovery = 0.9).

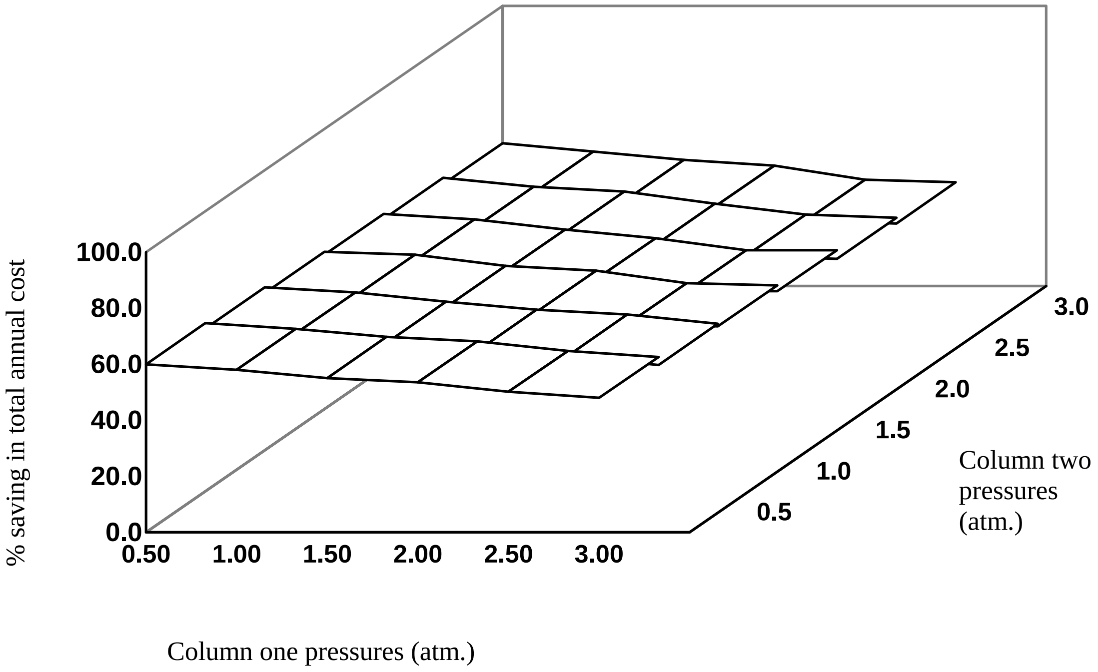

Figure (4.18) the effect of pressure change of column 1 and 2 on total annual cost for feed system b (feed composition 1 and recovery = 0.9).

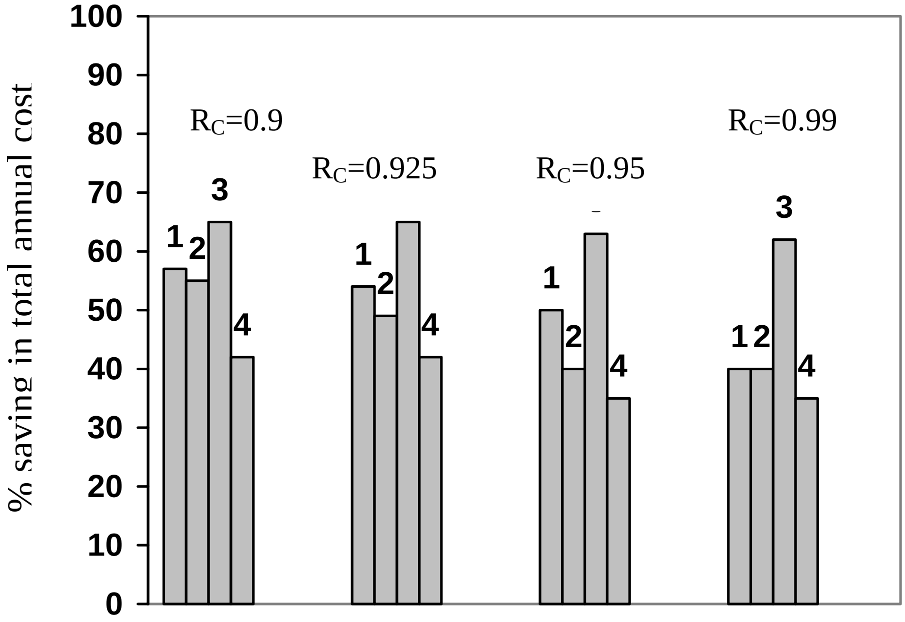

Figure (4.19) effect of changing feed composition % saving in total annual cost for feed system a (column one at 3 atm and column two at 0.5 atm.).

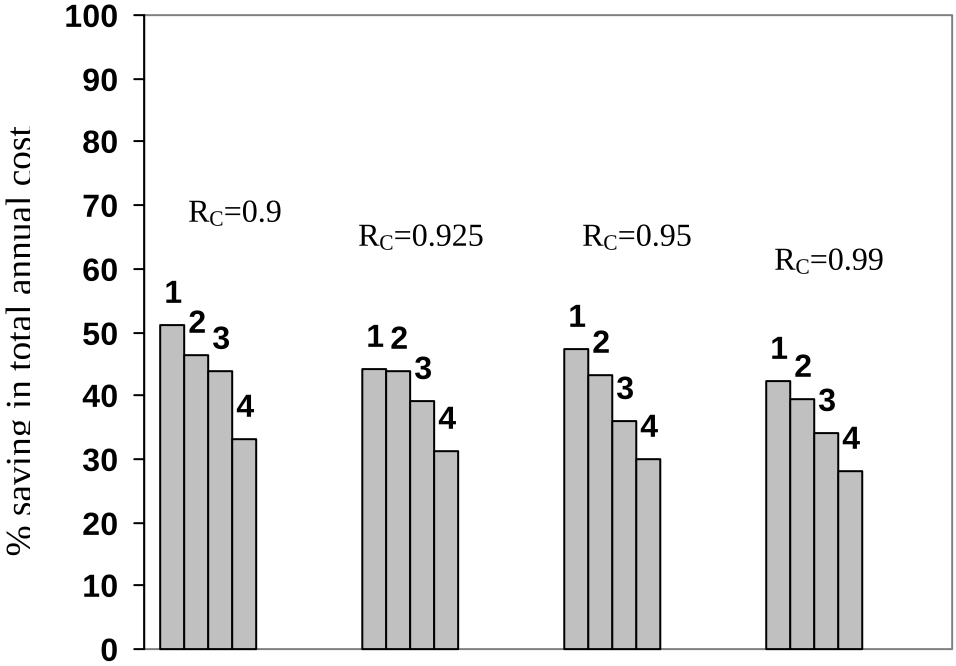

Figure (4.20) effect of changing feed composition % saving in total cost for feed system b (column one at 3 atm and column two at 0.5 atm.).

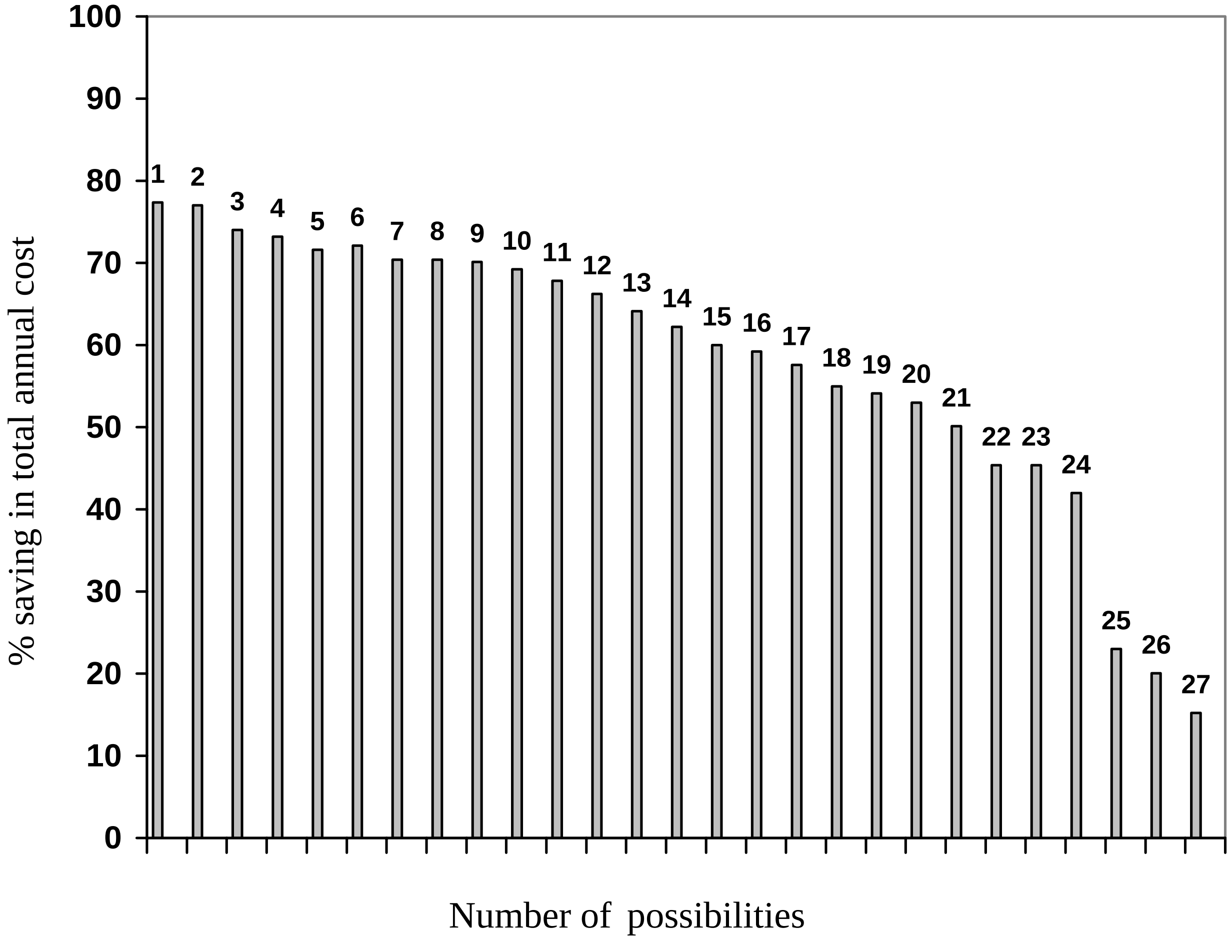

Figure (4.21) Effect of pressure change of column 1 and 2 on % saving in total annual cost for feed system c (feed composition one).

4.7 Discussions of the results:

The effect of changing feed system, operating pressure, feed composition and fractional recovery on total reboiler load, percentage saving in total energy consumption and percentage in total annual cost has been considered for two ternary and one quaternary feed system with and without heat pump with split tower technique.

The results show that as the operating pressure increases the total reboiler load will increase too for all feed systems and feed compositions when the columns operate as the same pressures or using split tower technique, this is due to increasing the temperature at the bottom of the column.

The effect of changing feed composition show that the components which have higher proportion of the less volatile component [feed composition 4 for feed systems a and b feed compositions 4 and 5 for feed system c] require less total reboiler load than the other feed compositions.

The third important variable is fractional recovery that has four different values of 0.9, 0.925, 0.95 and 0.99, it was seen that increasing the fractional recovery would increase the total reboiler load for all cases; the reason is when producing a large amounts of feed as top product requires higher reboiler load.

In the case of using heat pump technique, the results show that increasing the operating pressure is not proportional to the percentage saving in total energy consumption where when operating the two columns at the same pressure the percentage saving in total energy consumption increases while when using split tower technique the percentage saving in total energy consumption was not proportional the pressure change.

Changing the feed composition effect the percentage saving in total energy consumption where feed compositions 2, 3, and 1 for feed systems a, b, c respectively show a maximum percentage saving in total energy consumption.

Increasing the fractional recovery the percentage saving in total energy consumption decrease for all cases.

The total annual cost were studied for all feed systems with and without applying heat pump technique, the percentage saving in total annual cost was obtained and the results show that when increasing the operating pressure the percentage saving in total annual cost decreases for both cases of constant pressure and split tower technique.

Changing feed composition effects the percentage saving in total annual, there were maximum percentage saving in total annual cost for feed compositions 3,1, and 2 for feed systems a, b and c respectively. Also increasing fractional recovery will decrease the percentage saving in total annul cost for all feed systems.

The results show that the percentage saving in total energy consumption for feed systems a, b and c ranging from 22.55%-76.02%, 12.14%-76.00% and 19.00% - 73.25% respectively when the towers were operated at the same pressure with heat pump technique while the percentage saving in total energy consumption ranging from 15.99%-71.26% or feed system a, 12.91%-78.00% for feed system b and 12.02% - 77.36% for feed system c when split tower technique with heat pump were considered.

The previous work on the same systems using split tower technique with heat integration showed that the percentage saving in total energy consumption for feed a ranging from 2.9% to 57.8%, for feed system b 2.6% to 65.8% and feed system c 2.3% to 65.8%. These results are compared with the present work and there were 20% increasing in percentage saving in total energy consumption when heat pump technique was considered.

Chapter Five
Conclusions and Suggestions

5.1 Conclusions:

1. Increasing the operating pressure for any columns in any configuration resulting increasing in total reboiler load

2. Feed composition 4 [0.1, 0.1, 0.8] requires less total reboiler load for feed systems a (methanol, ethanol, water) of a value of ($0.83*10^6$) kcal/hr and b (acetone, methanol, water) of ($0.42*10^6$) kcal/hr, where for feed system c (n-hexane, methylcyclopentane, ethanol, benzene) feed composition 5 [0.1, 0.1, 0.1, 0.7] requires less total reboiler load of value of ($4.55*10^6$) kcal/hr.

3. Increasing the fractional recovery would increase the total reboiler load for all cases; the reason is when producing large amounts of feed as top product requires higher reboiler load.

4. Increasing the operating pressure is not proportional to the percentage saving in total energy consumption when using split tower technique where when operating the two columns at the same pressures the percentage saving in total energy consumption increases.

5. Feed compositions 2 [0.8, 0.1, 0.1] show a maximum percentage saving in total energy consumption of value of 45.63% for feed systems a, feed composition 3 [0.1, 0.8, 0.1] of value of 50.00% for feed system b, and feed composition 1 [0.25, 0.25, 0.25, 0.25] of value of 75.23% for feed systems c.

6. Increasing the fractional recovery the percentage saving in total energy consumption decrease for all cases.

7. Increasing the operating pressure the percentage saving in total annual cost decreases for both cases of constant pressure and split tower technique.

8. Changing feed composition effects the percentage saving in total annual cost, there were maximum percentage saving in total annual cost for feed

compositions 3[0.1, 0.8, 0.1] of value of 75.00% for feed systems a, feed composition 1[0.333, 0.333, 0.333] of value of 60.00% for feed system b, and feed composition 2[0.7, 0.1, 0.1, 0.1]of value of 77.0% for feed system c.

9. Increasing fractional recovery will decrease the percentage saving in total annul cost for all feed systems.

5.2 Suggestions and future work:

1. This work may be extended by using heat pump with energy matching and thermocoupling technique.
2. Heat pump system may be done between intermediate heaters and coolers besides using heat pump between reboilers and condensers for multicomponent distillation.
3. Applying heat pump system on the separation o systems more complicated mixtures (azeotropic).
4. The prediction of vapor-liquid equilibrium can be extended to use other models rather than Wilson model such as UNIFAC and UNIQUAC. In addition, the calculation of fugasity coefficient can be added in order to measure the deviation in vapor phase.
5. Operate the columns at higher pressure in order to study the shifting in composition profile and to study the effect of vapor-liquid equilibrium with high pressure.

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A.1 Mathematical models for short – cut calculations:

1. Minimum reflux ratio:

Underwood has determined minimum reflux ratio [66] relations:

$$\sum_{i=1}^{N} \frac{\alpha_i x_{i,d}}{\alpha_i - \phi} = R_m + 1 \qquad \text{… (A.1.1)}$$

For ternary mixture

$$\frac{\alpha_A x_{A,d}}{\alpha_A - \phi} + \frac{\alpha_B x_{B,d}}{\alpha_B - \phi} + \frac{\alpha_C x_{C,d}}{\alpha_C - \phi} = R_m + 1 \qquad \text{… (A.1.2)}$$

For quaternary mixture

$$\frac{\alpha_A x_{A,d}}{\alpha_A - \phi} + \frac{\alpha_B x_{B,d}}{\alpha_B - \phi} + \frac{\alpha_C x_{C,d}}{\alpha_C - \phi} + \frac{\alpha_D x_{D,d}}{\alpha_D - \phi} = R_m + 1 \qquad \text{… (A.1.3)}$$

ϕ was calculated from the relation :

$$\sum_{i=1}^{N} \left(\frac{\alpha_i x_{if}}{\alpha_i - \phi} \right) = 1 - q \qquad \text{… (A.1.4)}$$

And $\alpha_{LK} \le \phi \le \alpha_{HK}$

For ternary component mixture:

$$\frac{\alpha_A x_{A,f}}{\alpha_A - \phi} + \frac{\alpha_B x_{B,f}}{\alpha_B - \phi} + \frac{\alpha_C x_{C,f}}{\alpha_C - \phi} = 1 - q \qquad \text{… (A.1.5)}$$

For quaternary component mixture:

$$\frac{\alpha_A x_{A,f}}{\alpha_A - \phi} + \frac{\alpha_B x_{B,f}}{\alpha_B - \phi} + \frac{\alpha_C x_{C,f}}{\alpha_C - \phi} + \frac{\alpha_D x_{D,f}}{\alpha_D - \phi} = 1 - q \qquad \text{… (A.1.6)}$$

The value of ϕ is calculated from previous equations by trial and error, and then R_m is calculated from equation (A.1.1).

Liquid flow rate $= L = R * D$ … (A.1.7)

Vapor flow rate $= V = L + D = (D\,(R + 1))$ … (A.1.8)

Theoretical number of plates

Gilliand [20], Van Winkle and Todd [67] relationships were used to calculate the theoretical number of plates:

a. for $(R - R_m)/(R + 1) < 0.125$

$$\frac{N - N_m}{N + 1} = 0.5039 + 0.5068\left(\frac{R - R_m}{R + 1}\right) - 0.0908\log\left(\frac{R - R_m}{R + 1}\right)$$

… (A.1.9)

b. for $(R - R_m)/(R + 1) > 0.125$

$$\frac{N - N_m}{N + 1} = 0.6257 - 0.9868\left(\frac{R - R_m}{R + 1}\right) + 0.516\left(\frac{R - R_m}{R + 1}\right)^2 - 0.1738\left(\frac{R - R_m}{R + 1}\right)^3$$

… (A.1.10)

Number of plates above and below feed

Feed plate location was found by using Underwood [66], Robinson and Gilliand [54] correlations:

$$\frac{x_{D1}}{x_{D2}} = \left(\frac{\Phi_1}{\Phi_2}\right) N_T \frac{x_{F1}}{x_{F2}} \qquad \text{… (A.1.11)}$$

N_T is the number of plates above the feed.

$$x_{D1} = \sum_{i=1}^{N} \frac{\alpha_i x_{i,d}}{\alpha_i - \Phi_1} \qquad \text{… (A.1.12)}$$

$$x_{D2} = \sum_{i=1}^{N} \frac{\alpha_i x_{i,d}}{\alpha_i - \Phi_2} \qquad \text{… (A.1.13)}$$

Φ_1 and Φ_2 were calculated by trial and error from the following equation:

$$\sum_{i=1}^{N} \frac{\alpha_i x_{i,d}}{\alpha_i - \Phi} \cdot \frac{D}{V} = 1.0 \qquad \text{… (A.1.14)}$$

$$x_{F1} = \sum_{i=1}^{N} \frac{\alpha_i x_{i,F}}{\alpha_i - \Phi_1} \qquad \text{… (A.1.15)}$$

$$x_{F2} = \sum_{i=1}^{N} \frac{\alpha_i x_{i,F}}{\alpha_i - \Phi_2} \qquad \text{… (A.1.16)}$$

Condenser and Reboiler load

$$Q_C = V \left(H_V - h_L - h_D\right) \qquad \text{… (A.1.17)}$$

$$Q_R = D\ h_D + S_1 h_{S1} + W\ h_W - Q_C - F_1 h_{F1} - F_2 h_{F2} \qquad \text{… (A.1.18)}$$

A.2 Mathematical models for plate – to –plate calculations:

The MESH equations:

There are generally four sets of equations that must be satisfied in rigorous equilibrium stage calculations they are:

A. Material balance equation (M).
2. The equilibrium equation (E).
3. The summation equation (S).
4. The heat balance equation (H).

the material and heat balances are written around each stage and the mole fraction of liquid $X_{i,j}$, the vapor rate profile and the temperature profile T_j.

For $1 \le i \le m$ and $1 \le j \le N$

1- Material balance equation (M – equation)

$$M_{i,j}\left(X_{i,j}, V_j, T_j\right) = L_{j-1} X_{i,j-1} - \left(V_j + W_j\right) y_{i,j} - \left(L_j + U_j\right) x_{i,j} + V_{j+1} y_{i,j} + F_j Z_{i,j} = 0$$

… (A.2.1)

2- Vapor – liquid equilibrium (E – equation)

$$E_j\left(X_{i,j}, V_j, T_j\right) = y_{i,j} - K_{i,j} X_{i,j} = 0 \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(A.2.2)}$$

Where

$$K_{i,j} = \gamma_{i,j}\left(x_j T_j\right) p_{i,j} / P \ \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(A.2.3)}$$

3- Summation equation (S – equation)

$S_j\left(x_{i,j},V_j,T_j\right)= \sum_{i=1}^{m} y_{i,j}-1.0 = 0$ …………………..… (A.2.4)

$S_j\left(x_{i,j},V_j,T_j\right)=\sum_{i=1}^{m} x_{i,j}-1.0=0$ …………………..… (A.2.5)

4- Heat balance equation (H – equation)

$H_j\left(x_{i,j},V_j,T_j\right)=L_{j-1}h_{j-1}-\left(V_j+W_j\right)H_j-\left(L_j+U_j\right)h_j+V_{j+1}H_{j+1}+F_jH_{Fj}-Q_j=0$

……….. … (A.2.6)

Equation (A.2.1) and (A.2.2) are combined and the liquid low rate (L's) are expressed in terms of vapor flow rate (V's) by an overall material balance of all stages from the condenser through the j-th stage.

$L_j=V_{j+1}+\sum_{k=2}^{j}\left(F_k-W_k-V_k\right)-D$ $2 \le j \le N-1$ … (A.2.7)

Where $D=V_1+U_1$ … (A.2.8)

The M-equation then reduced to tridiagonal matrix form.

$B_1x_{i,,1}+C_1x_{i,2}=D_1$ … (A.2.9)

$A_jx_{i,j-1}+B_jx_{ij}+C_jx_{ij+1}=D$ $2 \le j \le N-1$ … (A.2.10)

$A_Nx_{i,N-1}+B_Nx_{i,n}=D_N$ ... (A.2.11)

Or in matrix notation as:

$$\begin{bmatrix} B_1 & C_1 & & & \\ A_1 & B_2 & C_2 & & \\ & A_j & B_j & C_j & \\ & & A_{N-1} & B_{N-1} & C_{N-1} \\ & & & A_N & B_N \end{bmatrix} \begin{bmatrix} X_{i,1} \\ X_{i,2} \\ X_{i,j} \\ X_{i,N-1} \\ X_{i,N} \end{bmatrix} \begin{bmatrix} D_1 \\ D_2 \\ D_j \\ D_{N-1} \\ D_N \end{bmatrix} \quad \ldots \text{(A.2.12)}$$

or simply

$\left[A_{BC}\right]\left\{x_{i,j}\right\}=\left\{x_{i,j}\right\}=\left\{D_j\right\}$ $1 \le j \le m$ … (A.2.13)

Where

$B_1=-\left(B_1K_{i,1}+U_1\right)$ … (A.2.14)

$$C_1 = V_2 K_{i,2} \qquad \ldots \text{(A.2.15)}$$

$$D_1 = 0 \qquad \ldots \text{(A.2.16)}$$

$$A_j = L_{j-1} = V_j + \sum_{K=2}^{j-1} \left(F_K - W_K - U_K\right) - D \ , \quad 2 \leq \text{j} \leq \text{N-1} \qquad \ldots \text{(A.2.17)}$$

$$B_j = -\left[\left(V_j + W_j\right) K_{i,j}\right] + \left(L_j + U_j\right) \qquad \ldots \text{(A.2.18)}$$

$$= -\left[\left(V_j + W_j\right) K_{i,j} + V_{j+1} \sum_{k=2}^{j} \left(F_k - W_k - U_k\right) - D \ - U_j\right] \quad 2 \leq \text{j} \leq \text{N-1} \ \ldots \text{(A.2.19)}$$

$$C_j = V_{j+1} K_{i,j+1} \qquad 2 \leq \text{j} \leq \text{N-1} \qquad \ldots \text{(A.2.20)}$$

And

$$A_N = V_N + B \qquad \ldots \text{(A.2.21)}$$

$$B_N = -\left(V_N K_{i,N} + B\right); \quad D_N = 0 \qquad \ldots \text{(A.2.22)}$$

With the above manipulation of the M and E equations and further manipulations of the S and H equations the MESH for multicomponent separation at constant pressure in a complex column become:

$$M_{i,j}(x_{i,j}, V_j, T_j) = \left[A_{BC}\right]\left\{x_{ij}\right\} - \left\{D\right\} = 0 \quad 1 \leq \text{i} \leq \text{m and } 1 \leq \text{j} \leq \text{N} \quad \ldots \text{(A.2.23)}$$

$$S_j\left(x_{i,j}, T_j\right) \sum_{i=1}^{m} K_{i,j} x_{i,j} - 1.0 = 0 \qquad \ldots \text{(A.2.24)}$$

$$H_j\left(x_{i,j}, V_j, T_j\right) = \left(H_{j+1} - h_j\right) V_{j+1} - \left(H_j - h_j\right)\left(V_j + W_j\right) - \left(h_j - h_{j-1}\right) L_{j-1} + F_j\left(H_{Fj} - h_j\right) - Q \ = 0$$

$$1 \leq \text{j} \leq \text{N} \qquad \ldots \text{(A.2.25)}$$

The flow rate and composition of feed streams were given, the amount of all product streams are specified, $F_j, Z_{i,j}, W_j, U_j, D \ and \ B$ are all constants. An initial set of V_j and T_j was assumed, $\left[A_{BC}\right]$ and $\left\{d\right\}$ are constant if the equilibrium ratio, $K_{ij}'s$ is expressed as a function of $T_j's$.

Then the M-equation is a linear system by grouping the vector $\left\{D\right\}$ with $\left[A_{BC}\right]$

$$\begin{bmatrix} B_1 & C_1 & & & & & & D_1 \\ A_2 & B_2 & C_2 & & & & & D_2 \\ & A_j & B_j & & & & & D_j \\ & & & A_{N-1} & B_{N-1} & C_{N-1} & & D_{N-1} \\ & & & & A_N & B_N & C_N & D_N \end{bmatrix} \qquad 1 \le \mathrm{j} \le \mathrm{N} \qquad \ldots \text{(A.2.26)}$$

The solution of equation (A.2.23) for $[x_{i,j}]$ can be easily obtained by use of a simple algorithm derived from the Gauss elimination method.

In the algorithm, two auxiliary quantities, $p_j \, and \, q_j$ were calculated by first evaluating $p_1 and\, q_1$ and advancing forward with j increasing that is:

$$P_1 = \frac{C_1}{B_1} \quad ; \qquad q_1 = D_1 / B_1 \qquad \ldots\text{(A.2.27)}$$

$$p_j = C_j / \left(B_j - A_j \, p_{j-1}\right) \quad , \qquad 2 \le \mathrm{j} \le \mathrm{N\text{-}1} \qquad \ldots \text{(A.2.28)}$$

$$q_j = \left(D_j - A_j q_{j-1}\right) / \left(B_j - A_j \, p_{j-1}\right) \quad , \qquad 2 \le \mathrm{j} \le \mathrm{N} \qquad \ldots \text{(A.2.29)}$$

The values of $x_{i.j}'s$ were calculated by first evaluating $x_{i,N}$ and proceeding backward with j decreasing until $x_{i,1}$ is reached.

Thus

$$x_{i,N} = q_N \qquad \ldots \text{(A.2.30)}$$

$$x_{i,j} = q_j - p_i \, x_{i,j+1} \quad , \qquad 1 \le \mathrm{j} \le \mathrm{N\text{-}1} \qquad \ldots\text{(A.2.31)}$$

The obtained $x_j's$ from the above algorithm were substituted into the S-equation and $K_{i,j}'s$ are expressed as a function of temperature and total pressure and composition

$$p_{i,j} = \exp\left(A(I) + \frac{B(I)}{C(I) + T} \right) \qquad \ldots \text{(A.2.32)}$$

$$K_{i,j} = \frac{\gamma_{i,j}(x_j, T_j) p_{i,j}}{P} \quad \dots \text{(A.2.33)}$$

Then

$$S_j(T_j) = \sum_{i=1}^{m} \left(\frac{\gamma_{i,j} p_{i,j}}{P} \right) x_{i,j} - 1.0 = 0, \quad 1 \leq \text{j} \leq \text{N} \quad \dots \text{(A.2.34)}$$

The solution of this equation employed the Newton iteration method. When $x_{i.j}'s$ and $T_j's$ are obtained, then a new value of V_j can be calculated from the heat balance equation (A.2.25).

The enthalpies of the internal vapor streams could be calculated by the use of the following equations:

$$H^E = \sum_{i=1}^{m} Y_{i,j}(b_{1,i} + b_{2,i} T_j + b_{3,i} T_j^2 + b_{4,i} T_j^3) \quad \dots \text{(A.2.35)}$$

The enthalpies of the internal liquid streams could be calculated by the use of the following equation

$$h_j = \sum_{i=1}^{m} x_{i,j} \left(C_{1,i} + C_{2,i} T_j + C_{3,i} T_j^2 + C_{4,i} T_j^3 \right) \quad \dots \text{(A.2.36)}$$

And adding this value h_j to the h^E calculated from Wilson equation in order to get the actual enthalpy content in the liquid phase.

A.3 Mathematical models for heat pump calculation:

After using heat pump technique in multicomponent distillation effect of many variables must be calculated for the condenser and reboiler:

$$Q_C = V_{ac}(H_{VI} - h_{LI}) \quad \dots \text{(A.3.1)}$$

$$Q_R = D \cdot h_D + W \cdot h_w + Q_c - F \cdot H_{iF} \quad \dots \text{(A.3.2)}$$

Where Q_c and Q_R are the heat loads for condenser and reboiler in Kcal/hr.

$$H_{VI} = \sum_{i=1}^{M} \left[\left(b_{1i} + b_{2i} \cdot T_t + b_{3i} \cdot T_t^2 \right) Y_{id} \right] \quad \dots \text{(A.3.3)}$$

Where, b_1, b_2, b_3 are the liquid enthalpy constants and T_t is the top product temperature.

The enthalpy of liquid out of the condenser:

$$h_{LI} = \sum_{i=1}^{M} \left[\left(c_{1i} + c_{2i} \cdot T_t + c_{3i} \cdot T_t^2 \right) x_{id} \right] \qquad \ldots \text{(A.3.4)}$$

Where, c_1, c_2, c_3 the vapor enthalpy constants.

The enthalpy of the distillate is;

$$h_d = h_{LI} \qquad \ldots \text{(A.3.5)}$$

The enthalpy of feed and bottom product, respectively are:

$$h_W = \sum_{i=1}^{M} \left[\left(c_{1i} + c_{2i} \cdot T_b + c_{3i} \cdot T_b^2 \right) Y_{id} \right] \qquad \ldots \text{(A.3.6)}$$

and

$$h_{iF} = \sum_{i=1}^{M} \left[\left(c_{1i} + c_{2i} \cdot T_F + c_{3i} T_F^2 \right) X_{iF} \right] \qquad \ldots \text{(A.3.7)}$$

Where T_b the bottom product temperature and T_F is the feed temperature.

The work of compression in kcal/hr is:

$$W_{comp} = V_{ac} \left(H_{VI} - H_{VO} \right)$$

Where H_{VO} the enthalpy of vapor stream out of the compressor.

A.4 Mathematical models for cost calculation:

Equipment cost, operating cost and total annual cost were calculated depend on the equations of Henry [27] method.

The assumptions made for cost estimation was:

1. Tray efficiency was assumed to be 60%.
2. Tray spacing was (1.5 ft) 0.457 m.
3. Tray area is 85% of the column cross sectional area.
4. All the equipments are assumed to be carbon steel.
5. The equipment operation time is whole of the year.
6. Economical service life is 10 years.

7. The steam used in the reboiler is saturated steam at 250 psia.

8. The rise in temperature of the cooling water in condensers is $(20F^\circ), 11.1C^\circ$.

9. The space from the top of the column to first plate is 1.83 m (6 ft) and from the final plate to the bottom is (12 ft) 3.66 m.

A.4.1 Equipment cost

1. Cost of distillation column

V_T = dimension parameter of the column

$V_T = DC * HT$ … (A.4.1.1)

C_A= cross sectional area of the column =$\pi . Dc^2/4$ … (A.4.1.2)

And the tray area = $T_A = 0.85 * C_A$ … (A.4.1.3)

Cost of the column $= C_T = C_1 \cdot \left(\frac{V_T}{100}\right)^{E_1} + C_2 \cdot \left(\frac{T_A}{100}\right)^{E_2} .N$ … (A.4.1.4)

Where

C_1 = cost of distillation column with dimension parameter of 100 = ID 900000

C_2 = cost of tray within $(9.29m^2)(100\ ft^2)$=ID 202500.

E_1 = slope of the column diameter parameter versus cost curve.

E_2= slope of the tray area versus cost curve.

N = factor used to calculate the height of the column.

2. Cost of condenser

A_c = area of condenser $= Q_c/(U_c * T_{LM})$ … (A.4.1.5)

Where, U_c is the overall heat transfer coefficient and is taken as $567.8 W/m^2.K^\circ\ (100 Btu/hr\ ft^2.f^\circ)$.

T_{LM} is the logmean temperature difference, and calculated by:

$T_{LM} = (T_t - 80) - (T_t - 100)/[(T_t - 80)/(T_t - 100)]$ (If $T_t > 100 F^\circ$) … (A.4.1.6)

$T_{LM} = (T_t - 40) - (T_t - 45)/[(T_t - 40)/(T_t - 45)]$ (If $T_t < 100 F^\circ$) … (A.4.1.7)

C_c = cost of condenser = $C_3 \cdot \left(\frac{A_C}{1000}\right)^{E_3}$ … (A.4.1.8)

C_3 = cost of heat exchanger with surface area of $92.9 m^2$

E_3 = slope of heat transfer area versus cost curve.

If the surface area of the condenser is greater than $92.9 m^2$ the area was divided by two and the cost was calculated for two units of equal size.

3. Cost of reboiler

A_R = surface area of reboiler $= Q_R / (U_R * T_R)$ … (A.4.1.9)

U_R is the overall heat transfer coefficient for steam and was taken at $710 \ W/m^2 K^{\circ} (125 \ Btu/hr \ ft^2 F^{\circ})$.

T_R = is the logarithm mean temperature difference $T_R = 401 - T_b$ … (A.4.1.10)

C_R =cost of reboiler $= C_3 \cdot \left(\frac{A_R}{1000}\right)^{E_3}$ … (A.4.1.11)

A.4.2 Heat pump system cost:

1. Cost of compressor

$C_{comp} = 150.8 \left(W_{comp}\right)^{0.82}$ ... (A.4.2.1)

2. Cost of reboiler condenser

A_{RC} = area of reboiler condenser $= Q_R / (U \ * T_{1h})$ … (A.4.2.2)

Where, U is the overall heat transfer coefficient and was taken as $567.8 \ W/m^2 K^{\circ} \ 100 \ Btu/hr \ ft^2 F^{\circ}$, T_{1h} is the logmean temperature difference, is calculated by:

$T_{1h} = (T_h - 80) - (T_h - 100) / \ln[(T_h - 80)/(T_h - 100)]$ (If $T_h > 100 \ F^{\circ}$) … (A.4.2.3)

$T_{1h} = (T_h - 40) - (T_h - 45) / \ln[(T_h - 40)/(T_h - 45)]$ (If $T_h < 100 \ F^{\circ}$)

… (A.4.2.4)

C_{RC} = cost of reboiler condenser $= C_3 . \left(\frac{A_{RC}}{1000} \right)^{E_3}$ … (A.4.2.5)

C_3 = cost of heat exchanger with surface area of $92.9 m^2 (1000 ft^2)$ = ID 210000

E_3 = slope of heat transfer area versus cost curve.

3. Cost of subcooler

A_S = area of subcooler $= Q_{SC} / (U_C * T_{1c})$ … (A.4.2.6)

T_{1C} is the logmean temperature difference, is calculated by

$T_{1C} = (T_h - 40) - (T_d - 100) / \ln[(T_h - 80)/(T_d - 100)]$ (If $T_d > 100$ F°)

… (A.4.2.7)

$T_{1C} = (T_h - 40) - (T_d - 45) / \ln[(T_h - 40)/(T_d - 45)]$ (If $T_d < 100$ F°)

… (A.4.2.8)

Where T_d the temperature of the distillate.

Equipment cost = C_{EH} = cost of column + cost of compressor + cost of reboiler condenser +cost of subcooler

A.4.3 annual operating cost

C_S = cost of steam $= 24 * 365 * C_5 * Q_R * 3.9/1000 * H_{VAP}$ … (A.4.3.1)

Where C_S is the cost of 453.6 Kg (1000 lb) of steam =ID 4059.15.

H_{VAP} is the latent heat of steam =825.0 Btu / lb.

C_W =cost of cooling water $= (24 * 365 * C_4 * Q_C * 3.79)/(1000 * 20 * 8.33)$

… (A.4.3.2)

Where C_4 is the cost of 4546 liter (1000 gallon) of cooling water = ID 124.95

Working capital = 18% of equipment cost.

Taxes = 4% of equipment cost.

Insurance = 1% of total equipment cost.

Maintenance = 5% of total equipment cost, the total equal 28% of the total equipment cost.

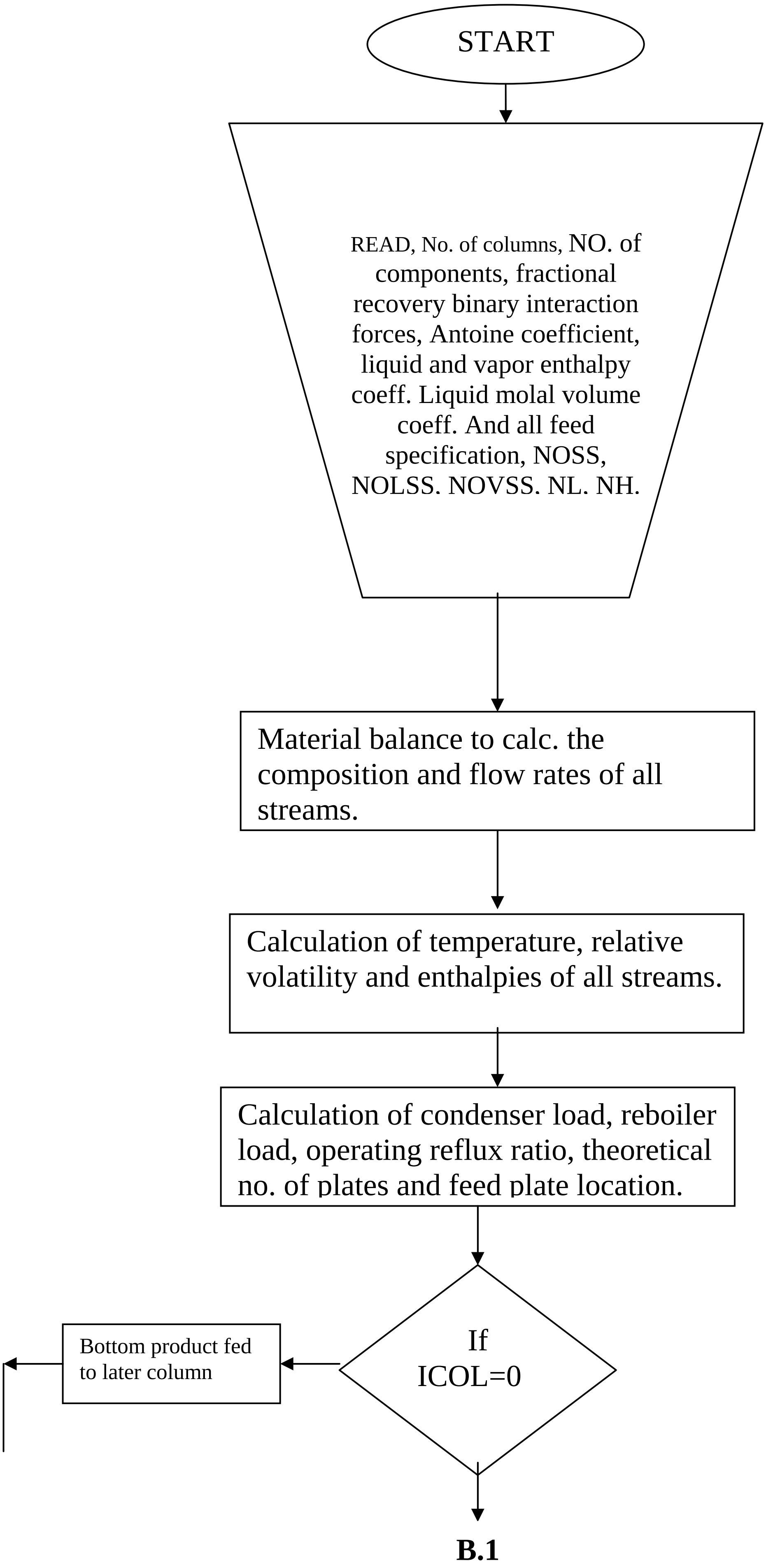
A.4.4 Total annual cost:

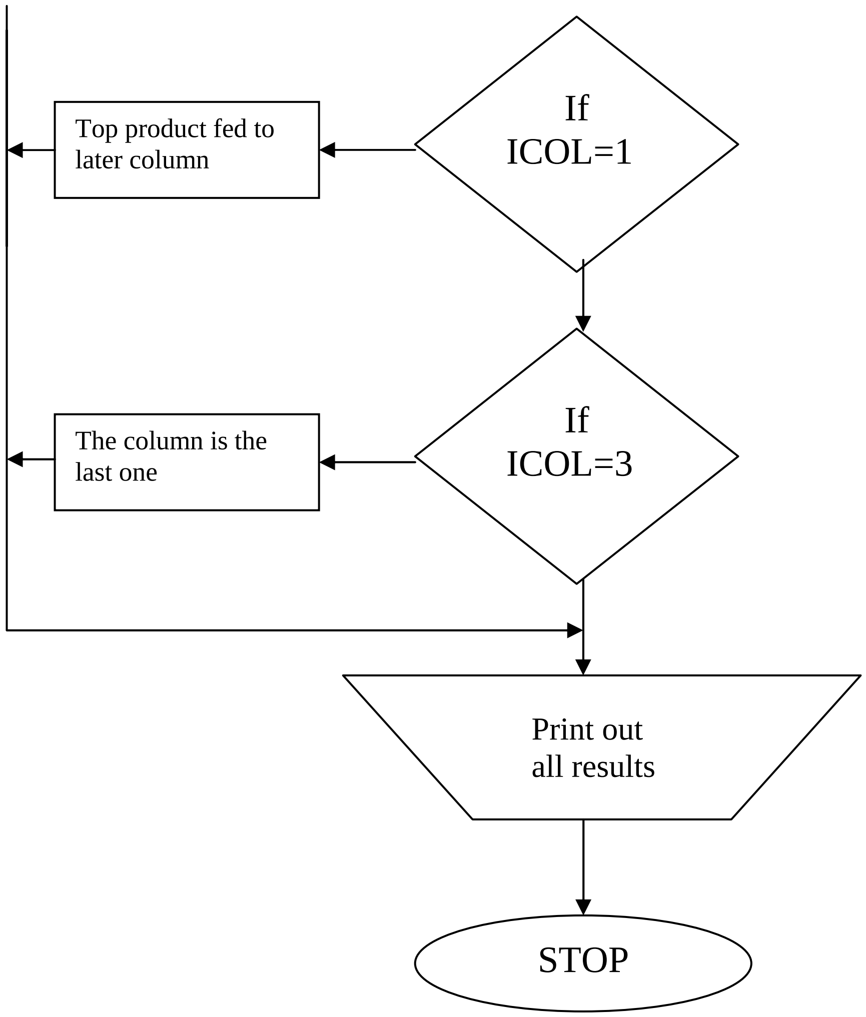
Total annual cost = depreciation + annual operating cost

Total annual cost= can = (equipment cost / service life) +annual operating

Appendix B

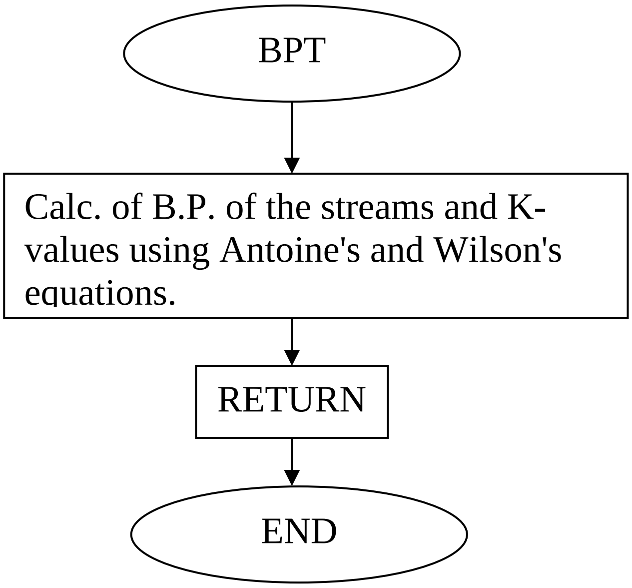
B.1 Flow chart of shortcut programme



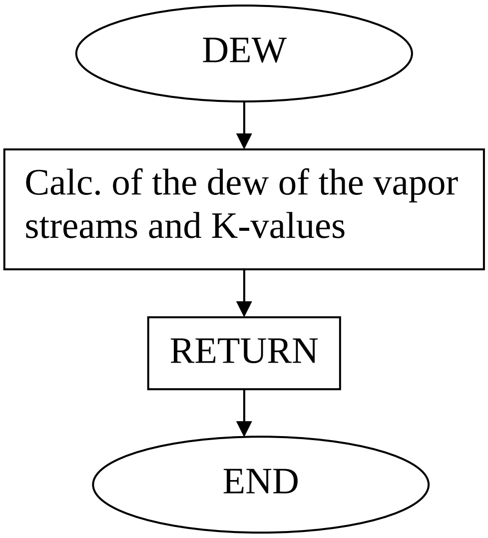

If
ICOL=1
Top product fed to later column
If
ICOL=3
The column is the last one
Print out all results
STOP

B.2 Flow chart of shortcut programmes subroutines

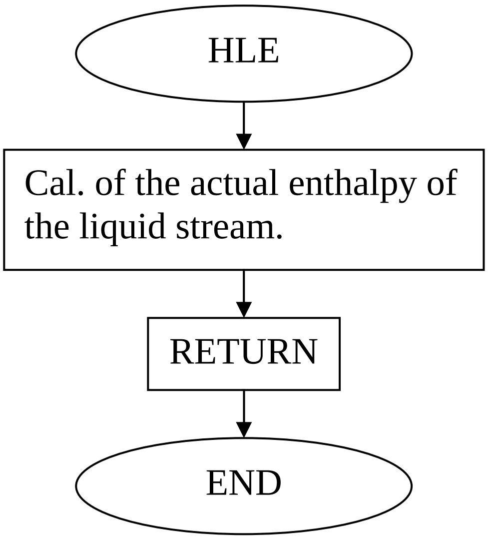
(1)



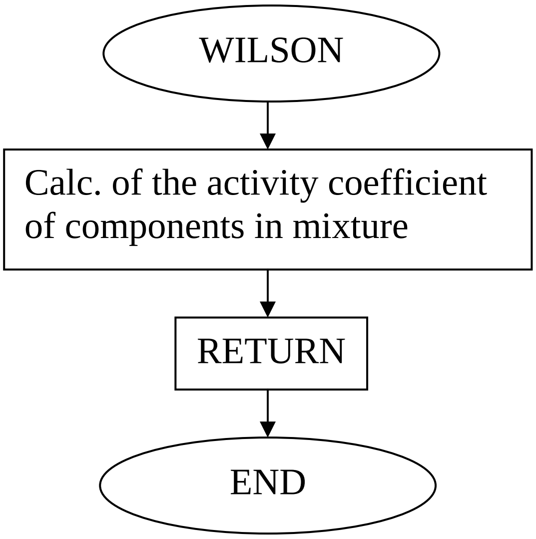
(2)



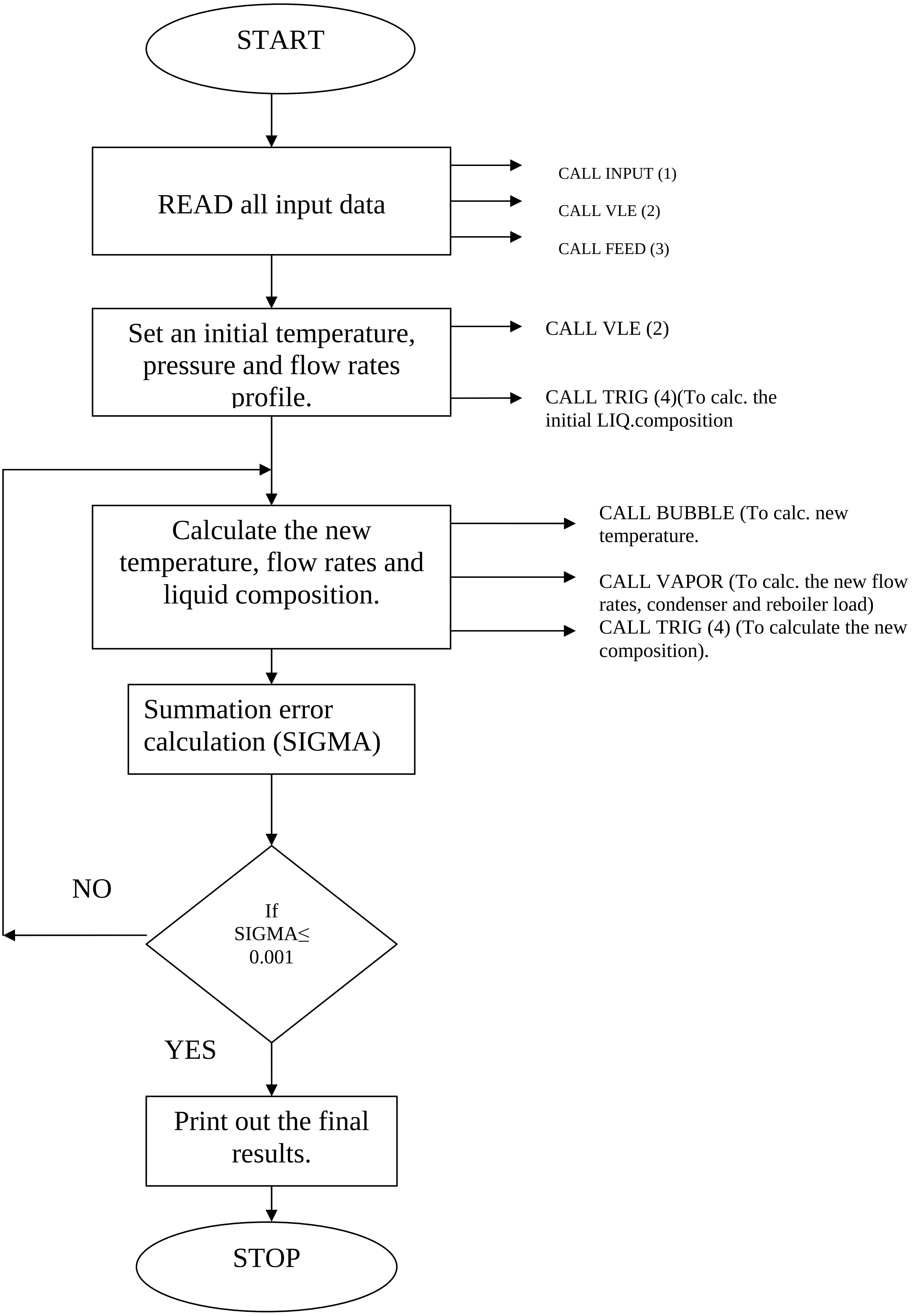
(3)



(4)

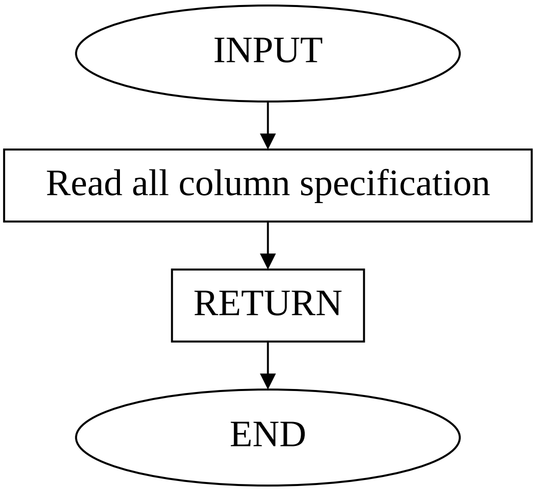

WILSON
Calc. of the activity coefficient of components in mixture
RETURN
END

B.3 Flow chart of plate-to-plate programme

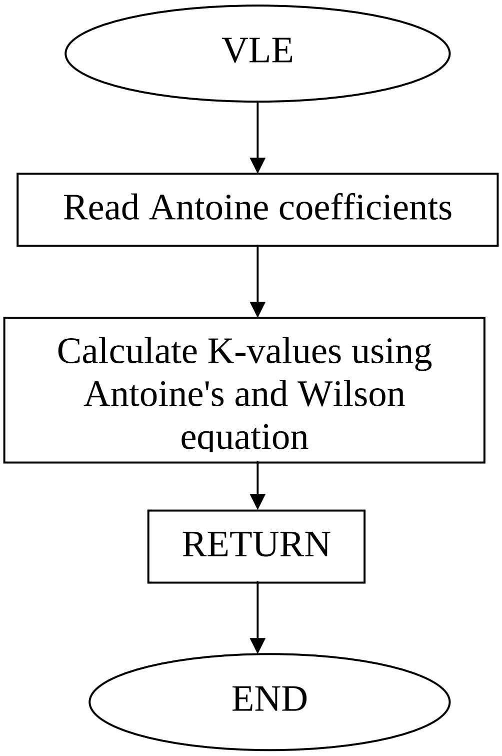


B.4 Flow charts of plate-to-plate programmes subroutines

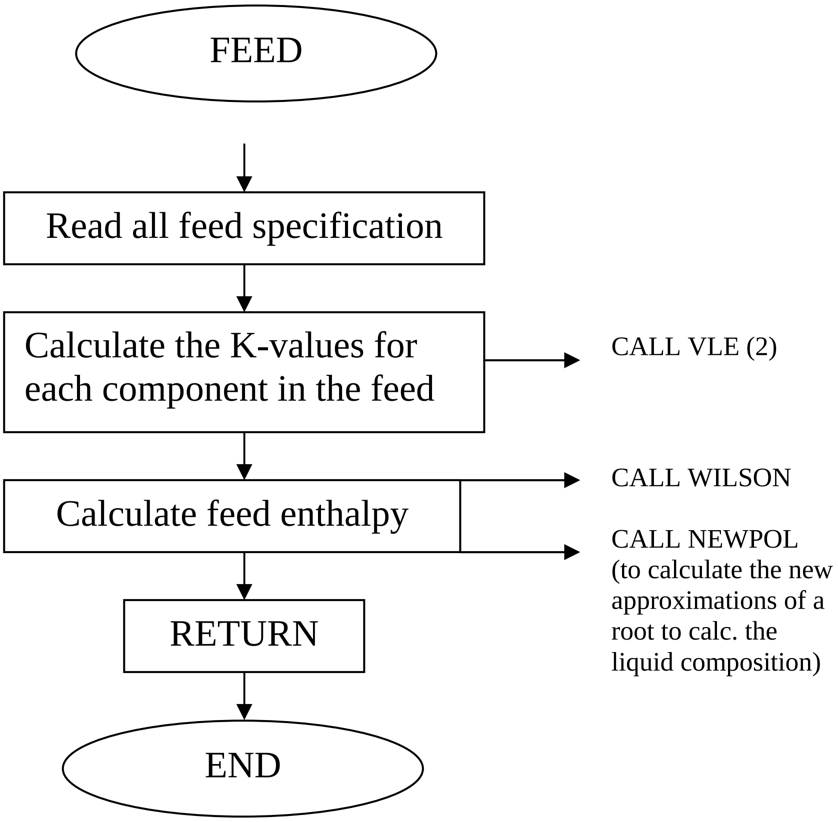
(1)



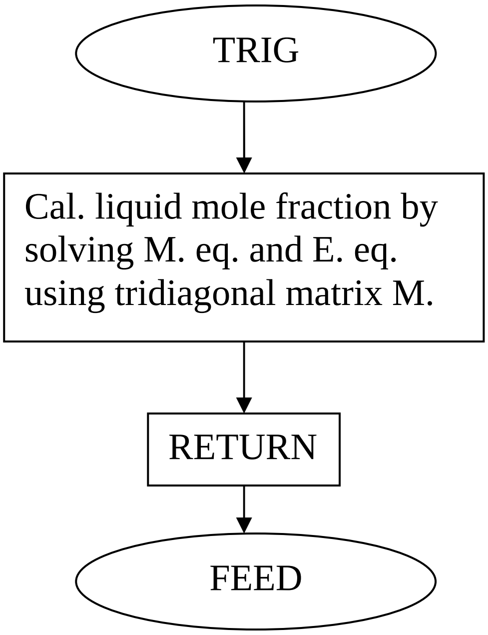
(2)



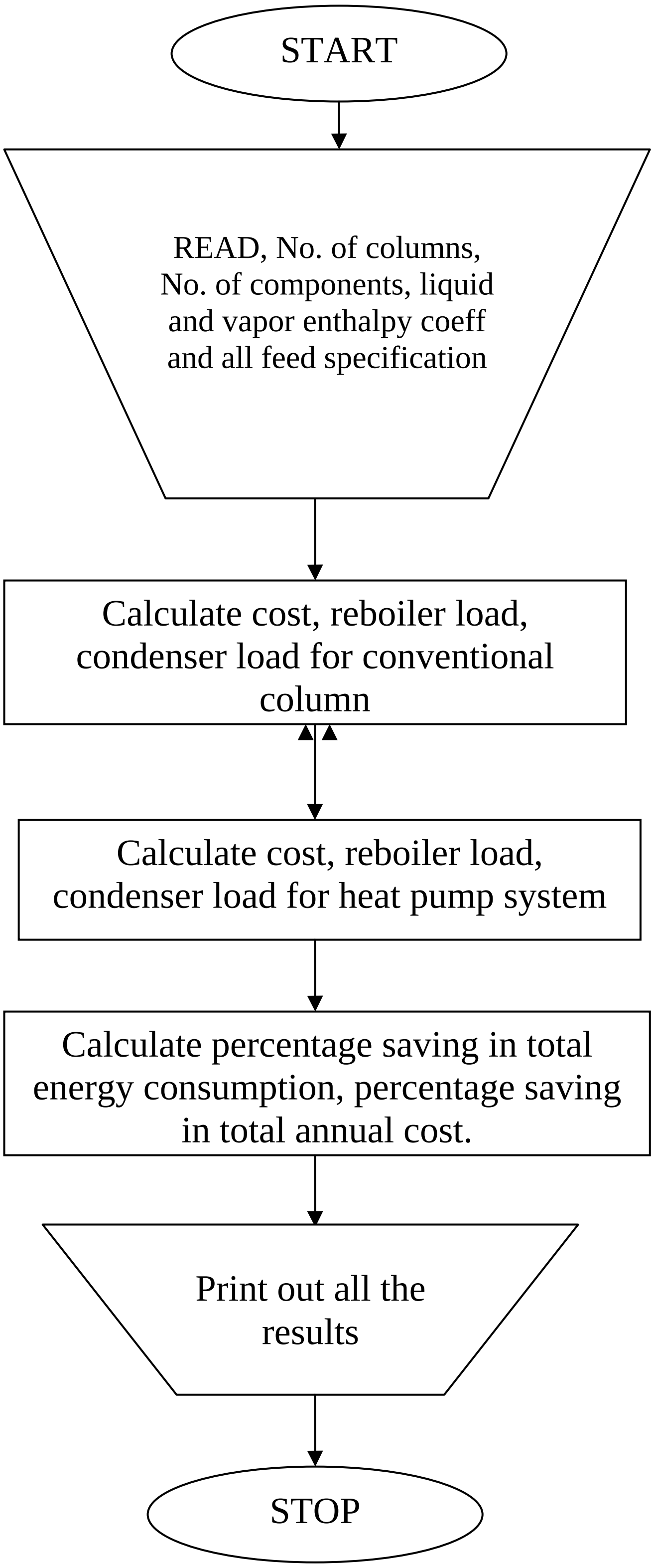
(3)


FEED
Read all feed specification
Calculate the K-values for each component in the feed
CALL VLE (2)
Calculate feed enthalpy
CALL WILSON
CALL NEWPOL (to calculate the new approximations of a root to calc. the liquid composition)
RETURN
END

(4)


TRIG
Cal. liquid mole fraction by solving M. eq. and E. eq. using tridiagonal matrix M.
RETURN
FEED

B.5 Flow chart of cost calculation programme



Appendix C

Composition profile for feed system b feed composition (3)

```
--------------------------------------------------------------------------------------------------------------------------------------
STAGE       TEMPERATURE  PRESSURE     VAPOR         VAPOR          LIQUID       LIQUID          HEAT       SUMX (J)
 NO.                                FLOW RATE    SIDE-STREAM     FLOW RATE   SIDE-STREAM        DUTY
--------------------------------------------------------------------------------------------------------------------------------------
CONDENSER        40.113     .500          .000          .000          31.671       10.000    152 4521.00    1.001509
   2             41.604     .500        41.671          .000          27.352         .000           .000    1.000999
   3             43.346     .500        37.352          .000          23.914         .000           .000    1.000756
   4             44.852     .500        33.914          .000          21.278         .000           .000    1.000367
   5             46.040     .500        31.278          .000         121.308         .000           .000     .999781
   6             46.038     .500        31.308          .000         121.249         .000           .000     .999635
   7             46.071     .500        31.249          .000         121.115         .000           .000     .999521
   8             46.146     .500        31.115          .000         120.782         .000           .000     .999540
   9             46.325     .500        30.782          .000         120.205         .000           .000     .999357
  10             46.675     .500        30.205          .000         118.496         .000           .000     .999681
 REBOILER        47.628     .500        28.496          .000          90.000         .000    155 4521.00     .999832
--------------------------------------------------------------------------------------------------------------------------------------
COMPOSITION PROFILE:
---------------------------------------------------------------------------------------------
  .46188   .53438   .00374
  .31446   .67557   .00997
  .20079   .77645   .02276
  .13646   .81658   .04696
  .10607   .80323   .09070
  .10532   .80392   .09076
  .10383   .80530   .09087
  .10084   .80805   .09111
  .09486   .81345   .09169
  .08286   .82301   .09412
  .05972   .82956   .11071
---------------------------------------------------------------------------------------------
```

Table (C.1) at recovery = 0.9

```
------------------------------------------------------------------------------------------------------------------------------------------
 STAGE     TEMPERATURE  PRESSURE      VAPOR          VAPOR          LIQUID         LIQUID        HEAT       SUMX (J)
  NO.                               FLOW RATE     SIDE-STREAM     FLOW RATE     SIDE-STREAM     DUTY
------------------------------------------------------------------------------------------------------------------------------------------
CONDENSER      39.578      .500          .000          .000          41.103        10.000    1591256.00    1.004738
   2           40.690      .500        51.103          .000          36.171          .000          .000    1.004902
   3           42.272      .500        46.171          .000          31.952          .000          .000    1.000293
   4           43.800      .500        41.952          .000          28.953          .000          .000     .996885
   5           44.969      .500        38.953          .000          26.178          .000          .000     .997988
   6           46.017      .500        36.178          .000         126.236          .000          .000     .999134
   7           46.005      .500        36.236          .000         126.135          .000          .000     .999043
   8           46.042      .500        36.135          .000         126.115          .000          .000     .998699
   9           46.073      .500        36.115          .000         125.814          .000          .000     .998982
  10           46.210      .500        35.814          .000         125.315          .000          .000     .999304
  11           46.454      .500        35.315          .000         124.381          .000          .000     .999650
  12           46.919      .500        34.381          .000         122.379          .000          .000     .999772
 REBOILER      47.918      .500        32.379          .000          90.000          .000    1601256.00     .999473
------------------------------------------------------------------------------------------------------------------------------------------
COMPOSITION PROFILE:
------------------------------------------------------------------------------------------------------------------------------------------
  .52424   .47416   .00160
  .39283   .60307   .00410
  .26803   .72231   .00966
  .17942   .79934   .02124
  .13032   .82551   .04416
  .10590   .80565   .08845
  .10516   .80632   .08851
  .10395   .80748   .08857
  .10159   .80966   .08875
  .09732   .81360   .08907
  .08956   .82058   .08985
  .07588   .83138   .09274
  .05261   .83639   .11099
```

Table (C.2) at recovery = 0.925

```
----------------------------------------------------------------------------------------------------------------------------------
  STAGE        TEMPERATURE  PRESSURE    VAPOR        VAPOR       LIQUID       LIQUID       HEAT       SUMX (J)
   NO.                                FLOW RATE   SIDE-STREAM  FLOW RATE   SIDE-STREAM    DUTY
----------------------------------------------------------------------------------------------------------------------------------
CONDENSER        39.091      .500        .000        .000        54.034       10.000   1724125.00   1.001980
     2           39.790      .500      64.034        .000        49.926         .000         .000   1.001807
     3           40.834      .500      59.926        .000        45.028         .000         .000   1.001049
     4           42.188      .500      55.028        .000        40.348         .000         .000   1.000215
     5           43.604      .500      50.348        .000        36.554         .000         .000    .999740
     6           44.822      .500      46.554        .000        33.190         .000         .000    .999571
     7           45.910      .500      43.190        .000       133.127         .000         .000    .999909
     8           45.936      .500      43.127        .000       133.022         .000         .000    .999929
     9           45.978      .500      43.022        .000       132.851         .000         .000    .999942
    10           46.047      .500      42.851        .000       132.572         .000         .000    .999944
    11           46.161      .500      42.572        .000       132.124         .000         .000    .999935
    12           46.347      .500      42.124        .000       131.380         .000         .000    .999963
    13           46.659      .500      41.380        .000       130.156         .000         .000    .999955
    14           47.181      .500      40.156        .000       127.752         .000         .000    .999905
 REBOILER        48.208      .500      37.752        .000        90.000         .000   1742543.00    .999780
----------------------------------------------------------------------------------------------------------------------------------
COMPOSITION PROFILE:
----------------------------------------------------------------------------------------------------------------------------------
   .60019   .39916   .00065
   .49694   .50150   .00155
   .38008   .61629   .00362
   .26830   .72336   .00834
   .18480   .79657   .01863
   .13498   .82469   .04032
   .10817   .80666   .08517
   .10726   .80751   .08523
   .10575   .80892   .08533
   .10326   .81123   .08550
   .09922   .81500   .08579
   .09272   .82100   .08628
   .08253   .83017   .08730
   .06703   .84218   .09079
   .04430   .84464   .11106
----------------------------------------------------------------------------------------------
```

Table (C.3) at recovery = 0.95

```
----------------------------------------------------------------------------------------------------------------------------------
  STAGE        TEMPERATURE  PRESSURE    VAPOR        VAPOR       LIQUID       LIQUID       HEAT       SUMX (J)
   NO.                                FLOW RATE   SIDE-STREAM  FLOW RATE   SIDE-STREAM    DUTY
----------------------------------------------------------------------------------------------------------------------------------
CONDENSER        38.247      .500        .000        .000        86.082       10.000   1934521.00   1.005081
     2           38.351      .500      96.082        .000        85.362         .000         .000   1.005109
     3           38.494      .500      95.362        .000        84.296         .000         .000   1.005106
     4           38.694      .500      94.296        .000        82.700         .000         .000   1.005045
     5           38.983      .500      92.700        .000        80.296         .000         .000   1.004870
     6           39.412      .500      90.296        .000        76.742         .000         .000   1.004468
     7           40.052      .500      86.742        .000        71.824         .000         .000   1.003652
     8           40.973      .500      81.824        .000        65.944         .000         .000   1.002255
     9           42.155      .500      75.944        .000        60.149         .000         .000   1.000604
    10           43.417      .500      70.149        .000        55.156         .000         .000    .999428
    11           44.571      .500      65.156        .000        50.337         .000         .000    .999219
    12           45.690      .500      60.337        .000       150.251         .000         .000    .999750
    13           45.715      .500      60.251        .000       150.133         .000         .000    .999763
    14           45.749      .500      60.133        .000       149.971         .000         .000    .999780
    15           45.796      .500      59.971        .000       149.748         .000         .000    .999800
    16           45.860      .500      59.748        .000       149.446         .000         .000    .999821
    17           45.949      .500      59.446        .000       149.040         .000         .000    .999842
    18           46.068      .500      59.040        .000       148.500         .000         .000    .999858
    19           46.229      .500      58.500        .000       147.794         .000         .000    .999866
    20           46.443      .500      57.794        .000       146.890         .000         .000    .999856
    21           46.722      .500      56.890        .000       145.762         .000         .000    .999824
    22           47.079      .500      55.762        .000       144.380         .000         .000    .999763
    23           47.528      .500      54.380        .000       142.631         .000         .000    .999674
    24           48.107      .500      52.631        .000       139.729         .000         .000    .999562
 REBOILER        49.039      .500      49.729        .000        90.000         .000   1924512.00    .999436
```

```
------------------------------------------------------------------------------------------------------------------------------------------------------------
COMPOSITION PROFILE:
------------------------------------------------------------------------------------------------------------------------------------------------------------
   .75545    .24451    .00004
   .71890    .28103    .00007
   .67441    .32545    .00013
   .61915    .38059    .00026
   .54987    .44959    .00054
   .46458    .53425    .00117
   .36699    .63033    .00268
   .27098    .72273    .00630
   .19433    .79091    .01477
   .14400    .82185    .03415
   .11405    .80794    .07801
   .11312    .80881    .07806
   .11185    .81001    .07814
   .11010    .81165    .07825
   .10771    .81389    .07840
   .10449    .81691    .07860
   .10020    .82094    .07887
   .09455    .82623    .07922
   .08727    .83304    .07969
   .07812    .84159    .08029
   .06698    .85191    .08111
   .05394    .86352    .08255
   .03927    .87360    .08713
   .02332    .86551    .11117
----------------------------------------------------------------------------------------------------------------------------------------------------------
```

Table (C.4) at recovery = 0.99

Component	Antoine coefficients		
1	9.80800	-2804.77188	229.14050
2	11.67470	-3460.90750	231.32812
3	11.58040	-3754.17459	224.27734

Component	Molal volume		
1	56.86600	0.00843	0.00017
2	64.51000	-0.19720	0.00039
3	22.88000	-0.3642	0.00007

Component	Enthalpy coefficient of liquid		
1	0.00002	31.12990	0.00001
2	0.00002	18.09900	0.00001
3	0.00000	18.00000	0.0000

Component	Enthalpy coefficient of vapor		
1	7392.85999	-1.69420	0.02375
2	8464.59998	3.36319	0.01076
3	10749.90002	7.85602	-0.00036

Binary interaction forces
$\lambda_{12}- \lambda_{11} = 25$
$\lambda_{13}- \lambda_{11} = 383$
$\lambda_{21}- \lambda_{22} = 390$
$\lambda_{23}- \lambda_{22} = 216$
$\lambda_{31}- \lambda_{33} = 1474$
$\lambda_{32}- \lambda_{33} = 453$

Table C.5 physical properties of feed system one

Component	**Antoine coefficients**		
1	11.67470	-3460.90750	231.32812
2	11.28410	-3261.37628	210.3125
3	11.58040	-3754.17459	224.27734

Component	**Molal volume**		
1	64.51000	-0.19720	0.00039
2	53.86600	-0.03111	0.00016
3	22.89000	-0.03642	0.00007

Component	**Enthalpy coefficient of liquid**		
1	0.00002	18.09900	0.00001
2	-0.19696	26.35200	-0.00011
3	0.00000	18.00000	0.00000

Component	**Enthalpy coefficient of vapor**		
1	8464.59998	3.36319	0.10760
2	11312.90002	-1.59611	0.02433
3	10749.90002	7.85602	-0.00064

Binary interaction forces
$\lambda_{12}- \lambda_{11} = -175$
$\lambda_{13}- \lambda_{11} = 280$
$\lambda_{21-} \lambda_{22} = 216$
$\lambda_{23}- \lambda_{22} = 435$
$\lambda_{31}- \lambda_{33} = 442$
$\lambda_{32}- \lambda_{33} = 901$

Table C.6 physical properties of feed system two

Component	Antoine coefficients		
1	9.04530	-2613.02252	220.15625
2	9.28170	-2789.43430	228.75000
3	12.42110	-3847.62738	231.25000
4	9.24230	-2773.57233	220.00000

Component	Molal volume		
1	125.96000	-0.14456	0.00055
2	104.27000	-0.08676	0.00039
3	53.70000	-0.03110	0.00016
4	70.86000	0.01490	0.00016

Component	Enthalpy coefficient of liquid		
1	-0.00105	51.70000	0.00001
2	-0.10636	38.80230	0.03725
3	-0.19696	26.35200	-0.00011
4	-0.00002	27.94500	-0.00001

Component	Enthalpy coefficient of vapor		
1	9463.23242	26.09780	0.04898
2	8353.41797	18.44516	0.04340
3	11312.90002	-1.59611	0.02433
4	10539.90002	-16.25300	0.04270

Binary interaction forces
$\lambda_{12}- \lambda_{11} = 425$
$\lambda_{13}- \lambda_{11} = 320$
$\lambda_{21}- \lambda_{22} = 390$
$\lambda_{23}- \lambda_{22} = 216$
$\lambda_{31}- \lambda_{33} = 1474$
$\lambda_{32}- \lambda_{33} = 453$

Table C.7 physical properties of feed system three

الخلاصة

تهتم هذه الدراسة بتقليل احتياجات الطاقة في عمليات التقطير متعدد المكونات باستعمال طريقة المضخة الحرارية.

تم استعمال مبدأ المضخة الحرارية مع طريقة الفصل الحراري وبدونه لخليطين ثلاثيين (ميثانول-ايثانول-ماء) و (أسيتون – ميثانول –ماء) وخليط واحد رباعي(هكسان –مثيل سايكلوبنتان-ايثانول –بنزين)وتمت دراسة تأثير عدة متغيرات مثل الضغط، النسبة المولية لمكونات المغذي و نسبة النقاوة.

تم الفصل باستخدام شكل واحد لكل نظام تبعاً لخواص النظام من حيث كونه غير مثالي.

كل الخلائط اعتبرت سوائل في درجة حرارة الغليان وتمت دراسة أربع قيم للنقاوة هي 0.9,0.925,0.95,0.99 للخلائط الثلاثية و 0.9 للخليط الرباعي.

الضغط المستخدم في الأبراج يتراوح بين 0.5-3.0 جو لغرض تقليل احتياجات الطاقة الكلية المستخدمة.

استخدمت طريقة فصل الابراج حيث الضغط في البرج الاول 0.5جو والبرج الثاني يعمل بضغط يتراوح بين 0.5-3.0 جو ثم يعمل البرج الأول تحت ضغط1.0 جو ويتراوح ضغط الثاني بين 0.5-3.0 جو وهكذا حتى يصل ضغط البرج الأول إلى 3.0 جو وهذا يعني 36 احتمالاً مدروساً للخليطين الثلاثيين و 216 احتمالاً للخليط الرباعي.

تمت دراسة نسب الطاقة الكلية للمرجل، نسبة الادخار في الطاقة الكلية، نسبة الادخار في الكلفة السنوية الكلية.

تم أعداد ثلاث برامج بلغة فورتران 90 باستخدام الحاسبة الالكترونية من اجل تصميم أبراج التقطير متعدد المكونات مع أو بدون استخدام مبدأ المضخة الحرارية باستخدام مبدأ فصل أبراج التقطير.

أظهرت النتائج ان استخدام مبدأ المضخة الحرارية يعطي نسبة ادخار في الطاقة تتراوح بين12.14% -76.02% عند عدم استخدام مبدأ فصل أبراج التقطير و من12.02%- 78.00% عند استخدام هذا المبدأ وكذلك وجد ان استخدام هذا المبدأ ينتج ادخارا في الكلفة السنوية الكلية يتراوح بين 10.00% -77.36% لكل الحالات المدروسة.

تمت مقارنة نتائج هذا البحث ببحث أخر يستخدم طريقة التكامل الحراري باستخدام نفس الخلائط ووجد ان استخدام مبدأ المضخة الحرارية يعطي زيادة 20% في قيمة الادخار في الطاقة الكلية المستهلكة.

شكر و تقدير

أود ان اعبر عن خالص شكري و تقديري وامتناني العميق للمشرفة الدكتورة ندى بهجت نقاش لما قدمته لي من توجيهات ونصائح قيمة ومساعدة كبيرة في إتمام هدا البحث.

أود أيضا ان اشكر رئاسة قسم الهندسة الكيمياوية بما فيه من اساتدة ومنتسبين للمساعدة في إنجاز هدا العمل.

ولا أنسى ان اشكر والداي العزيزان وأخواتي لمساندتهم لي ولصبرهم وتفهمهم خلال فترة دراستي.

المضخة الحرارية في التقطير متعدد المكونات غير المثالي

رسالة
مقدمة إلى كلية الهندسة في جامعة النهرين وهي جزء
من متطلبات نيل درجة ماجستير علوم في
الهندسة الكيمياوية

من قبل
ندى ضياء علي المعتصم

(بكالوريوس في الهندسة الكيمياوية 2002)

رمضان 1426

تشرين الأول 2005