# MATHEMATICAL MODELLING OF THE DYNAMIC BEHAVIOUR OF A DOUBLE EFFECT EVAPORATOR SYSTEM

## A MASTER THESIS

PRESENTED TO

## CHEMICAL ENGINEERING DEPARTMENT, FEDERAL UNIVERSITY OF TECHNOLOGY, MINNA, NIGERIA.

**BY**

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# DECLARATION

I hereby declare that this project work is my original work and has never to my knowledge been submitted elsewhere before.

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DATE

## CERTIFICATION

This is to certify that I have supervised, read and approved this project work, which I found adequate both in scope and quality for the partial fulfillment of the requirements for the award of master of engineering degree in chemical engineering.

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# DEDICATION

This project is dedicated to Allah, the creator of the universe and mankind, and to the memory of my late mother Mallama Ramatu Bello.

# ACKNOWLEDGEMENT

My gratitude to Allah Subhana wata'ala is boundless for it is He who gave me the life, the good health, the knowledge and the energy all of which I dedicated in conducting the research and writing this project.

I am hugely indebted in particular, to the tireless effort of my project supervisor, Dr. K.R. Onifade, for the time he spared reading through my manuscripts, his constructive criticism, and offering useful suggestions. My thanks and appreciation to him knows no bound.

Special thanks are due to my father for his prayers, understanding and encouragement during this project work. The roles of Hajia Nana Fatima, Mallam Aliu Umar and Mrs. Comfort Kato are highly appreciated.

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# ABSTRACT

A mathematical model of the dynamic behaviour of a double effect evaporator was developed and numerical technique presented for the application and investigation of evaporation system.

In this research work, models were developed for a single effect and double effect evaporation systems. The models were tested using laboratory experimental data obtained from evaporation of known concentration of sugar solution using a double effect evaporator.

The simulated results especially for the change in concentration of liquor with respect to time were found to be very close to those obtained from experiment. Therefore, the simulated result validates the mathematical model proposed for a double effect evaporator transient behaviour. The model can thus serve as a tool for the analysis of a double effect evaporator dynamic behaviour.

# CHAPTER ONE

## 1.0 INTRODUCTION

Evaporation is a process in which a volatile component is removed from a mixture of non-volatile and volatile component by vaporization. The concentration of dilute liquors by evaporating solvent from the feed stream is an important industrial process used in sugar industries, alumna production, paper manufacturing, pharmaceutical industries and so on.

Traditionally, it was believed that evaporation is one of the oldest techniques known to mankind. It is nevertheless, at the present time, regarded as a key unit operation in the increasingly sophisticated armoury of modem chemical engineering. A clear understanding of the fundamental principles involved in this operation is therefore highly important to all chemical engineers.

Evaporative behaviour is not always easy to predict. Reference must be made to the usefulness of heat and mass balance in the quantitative assessment of this operation. Study must also be made of the differing approaches utilized in different types of individual equipment.

In most industrial processes the concentrated solution of thick liquor is the valuable product, and the vapour is condensed and discarded. However, the reverse situation occurs in the evaporation of mineral bearing water to give a product of solid-free for:

(i) Boiler Feed (ii) Human consumption (iii) Special process requirements.

The conditions under which evaporation is carried out in practice are widely varied and the methods chosen to evaporate a particular solution are very much influenced by the characteristic of the liquid to be concentrated.

(ii) The major liquid characteristics, which are usually considered, are concentration of liquor; scale formation; tendency to foam and temperature sensitivity.

Other liquid characteristics which may influence the choice and design of evaporator are: specific heat; heat of concentration freezing point; gas liberation on boiling; explosive hazards; toxicity, radio activity; and necessary of sterile operation. (Ellison et al.1985)

Most present day evaporators are constructed from alloys or non-ferrous materials depending on the nature of the liquor to be evaporated. In the majority of applications the material to be evaporated flows inside the tubes. The most common types of steam heated tabular evaporators are: -

(1) Short tube evaporators
    (a)Horizontal tube
    (b) Vertical tube
(2) Long tube vertical evaporators.
    (b) Natural and forced circulation.
    (c) Upward flow (climbing film)
    (d) Downward flow (falling film)
(3) Coiled evaporators and
(4) Agitated film evaporators.

Evaporation is a special separation process and its objective, unlike other separation processes is the vaporization of a portion of solvent to produce a concentrated thick liquor solution. Therefore, it differs from drying in that the residue is a liquid, which may be highly viscous rather than a solid. It differs from distillation in that the vapour is usually a single component and even when the vapour is a mixture; there is no attempt to separate the vapours into fractions. The dividing line between evaporation and crystallization is not well defined. An evaporating liquid may sometimes produce crystals but the aim of evaporation is to concentrate a solution rather than to form or build crystals.

Typically, evaporators are heated by condensing steam on the tube surface, thus boiling the solution being concentrated on the opposite surface of the tube. Low-pressure steam is commonly used. In some cases fuel is burned around tubes in a furnace to produce the required energy. In other cases the fuel is burned while it is in direct contact with the liquid being evaporated. Solar energy is often used nowadays in several evaporation processes.

Evaporation is extremely energy intensive, consuming large amounts of energy in many processes. Water has a high heat capacity and large heat of vaporization and large amount of water must often be removed. In some cases, especially the food and pharmaceutical industries, the products are temperature sensitive. Means must then be found to evaporate at modest temperature levels, often requiring vacuum operation and large process vessels.(Luyben and Wenzel.1988)

When evaporators are operated in a series connection it is called a multiple effect system. The major aim of multiple evaporation is to secure steam economy. The effects are arranged in such a way that a vacuum is established in the last effect with the aid of a condenser and an air ejector, which helps withdrawal of non-condensable.

Multiple effect evaporators are fed in many ways. The most commonly used are the forward feed, backward feed, and mixed feed. In the forward feed evaporator system, the feed and the steam flow in the same direction. Pump is required for the feeding of the first effect, while transfer from effect to effect can be achieved without pump. The backward effect system offers a better economy, when the feed is cold liquor, and also gives higher capacity when the thick liquor is viscous. In this system, the uses of pump are vitally necessary because the process fluid is fed to the last effect operating at high pressure. The mixed feed system offers the advantage of eliminating pumps need in the backward feed system. Though the mixed feed system is not commonly employed, but it offers the best performance.

A model is simply the mathematical abstraction of real process. In other words, it is an ordered set of mathematical equations, inequalities, charts, graphs and so on, and numerical solutions to these equations. A model may consist of ordinary differential equations or partial differential equations and algebraic equations. Process modelling has proven to be extremely successful engineering tool for the design and optimisation of physical, chemical and biological processes. It provides an avenue to understanding qualitative and quantitative aspects of the phenomenon of Interest.

Mathematical modelling is versatile and is widely used in practice, and is a recognized and valuable adjunct, and usually a precursor of computer simulation.

In the process of modelling a system critical laboratory experiment at times should be designed and performed in order to determine unknown model parameters. Decisions must be made on which parameters are important and which one can be neglected. Order of magnitude of analysis aids in making these critical simplifying decisions. It is imperative that the engineer be aware of and, not over look nor forget the assumptions made in the development of the mathematical model. (Ramrez. 1997)

The justifications for this research work are for the following reasons:

(1) Evaporation, even though being a unique branch of unit operation is not given enough attention because only few chemical engineers attempt to specialize in the field of evaporation.

(2) Evaporator models are not easy to come by. They are only published in journals and professional magazines. Contact must be made with the owners to have access to them on payment of appropriate charges.

(3) The design of evaporators involve gathering laboratory data, pilot plant experimental data which are in turn scale up to an industrial size design.

(4) Evaporator design calculations are difficult, cumbersome and time consuming because of repeated iterations.

As a result, this necessitates the development of models for application to design and simulation of evaporators. Apart from encouraging local development of evaporator models and simulation packages, the task of design engineers will be made easier and quicker, as well as, troubleshooting and optimisation of the process will be conducted with ease.

The modelling of dynamic behaviour of a double effect evaporator system and the solution of the derived equations using appropriate mathematical method will be presented and the model equations simulated.

# CHAPTER TWO

## 2.0 LITERATURE

## 2.1 EVAPORATOR FUNDAMENTALS

Evaporators are equipments, which are used for evaporation of various liquor. Specific type of evaporator is only suitable for specific liquor evaporation. The characteristic of the aqueous solution is taken into consideration for effective and efficient performance of evaporator. Evaporators can be classified into two major categories, namely, the chemical and power plant evaporators. The plant evaporators are usually the make-up evaporators for boiler feed, heat transfer evaporators and salt-water distiller. And the chemical evaporators are natural and forced circulation types.

Vacuum pumps, surface and contact condensers, steam traps play important role in evaporator performance. The efficiency of evaporator can be affected by salt deposition, foaming, scaling, heat sensitivity, corrosion and space limitation.

## 2.2 EVAPORATION EQUIPMENT

Evaporation equipment ranges from open shallow fans to multiple effect cascades of specially designed heat transfer units. The body construction of evaporators falls into following general categories.

### 2.2.1 THE HORIZONTAL –TUBE EVAPORATOR.

This is one of the classic construction types and has been widely used for many years. The solution to be evaporated boils outside the horizontal tubes within which steam condenses. The horizontal tubes interfere with the natural circulation of the boiling liquid and thus minimize liquid agitation. As a result, the overall heat transfer coefficient is lower than in other forms of evaporators, especially if the solution is viscous. No provision is made for breaking foam that occurs because of the boiling action. Fouling from the evaporating solution builds up on the outside of tubes where it cannot be as easily removed as it could be from the inside of the tube. Traditionally, the horizontal tubes have been inserted into the steam chests with packed glands rather than rolled or welded joints. For these reasons, horizontal tubes are used mainly in small installation where the solution to be treated is diluted and neither foams nor deposits solids on the evaporator tubes, or where materials of construction preclude welding or rolling of tubes. (Foust. 1976)

### 2.2.2 THE VERTICAL TUBE EVAPORATOR.

These consist of the basket and standard vertical varieties, which are distinct improvements over the horizontal tube evaporator. In both of them, the solution boils inside the vertical tube with the heating medium, usually condensing steam, held in a chest through which the tubes pass. In the basket evaporator, the steam chest forms a basket hung in the centre of the evaporator. Boiling, or heating of liquid in the tubes, causes upward flow through the tubes, and un-evaporated liquid flows downward through the annulus around the basket.

In the standard vertical tube evaporator, the steam chest is doughnut shaped. Liquid flows upward through the tubes and downward through the central hole. In large installation there may be several return ducts instead of central ones. In both types, the tubes are inserted into tube sheets by rolling or welding, which considerably reduces the cost compared to the packed glands traditionally used in the horizontal tube evaporator.

These evaporators overcome most of the operational disadvantages of the horizontal tube evaporators. Natural circulation is promoted, as a result, co-efficient are higher than in horizontal tube evaporators. Any solid deposit will build up inside the tubes where it is readily removed by mechanical cleaning. Vertical tube evaporators are completely satisfactory for most evaporation demands and are impractical only where the liquid being evaporated is very viscous, foams markedly, on may be subject to evaporator temperatures for very short period only.

### 2.2.3 FORCED-CIRCULATION EVAPORATOR.

In these evaporators, the evaporating liquid is pumped through a heat exchanger where the heating medium surrounds the tube carrying the solution. Pressure drop and hydrostatic head in combination are frequently great enough to prevent the solution from boiling in the exchanger tubes, so that the vapour generated is flashed off as the liquid enters the disengaging space. Since the velocity of the flashing mixture is high, impingement baffles are important to minimize entrainment. Modern forced circulation evaporators are usually equipped with external heater, rather than with heat transfer surface built into the evaporator body. (Foust. 1976)

This makes cleaning tubes and replacing any corroded and eroded tube much simpler than with the internal heating element.

In forced circulation evaporators, the heat transfer coefficient will depend upon circulation rate as well as overall temperature difference, boiling temperature and the system properties.

Thus the decision to use a forced – circulation evaporator depends on a favourable economic balance that includes the evaporator first cost, the power cost of operating the re-circulating pump, and the cost of maintaining the pump, and the comparative cost of cleaning the evaporator tubes. In the cases of viscous fluids or of solutions that must be kept from boiling in the tubes to minimize solids deposition, the choice of a forced-circulation evaporator may be almost mandatory.

### 2.2.4 THE LONG –TUBE VERTICAL EVAPORATOR

They consist of long tube usually 4m to 6m long. The vapour liquid mixture issues from the top of the tubes and impinges on a baffle. The liquor velocity is high enough that the baffle acts as an effective foam breaker. Among natural circulation evaporators, the long tube vertical evaporator competes most favourably with the forced circulation evaporator for application in large modern installations.

The coefficients are lower than they may be for the forced-circulation evaporator, but no circulating pump is required. The fraction evaporated per pass is usually larger than for forced-circulation operation.

Although viscous fluid cannot be handled, foaming materials can be and, for certain heat sensitive materials, this evaporator has the advantage of being readily operated without re-circulation. Evaporation per pass through the tube is normally far higher than in other types of natural-circulation evaporators and can be increased still more by further lengthening the exchanger tubes or by coiling them.

### 2.2.5 THE FALLING –FILM EVAPORATOR

A falling film evaporator eliminates the hydrostatic head effect and thus operates generally at lower temperature than does the long- tube vertical evaporator. The hold-up of liquid in the evaporation is almost much smaller. For these reasons, falling film evaporators are used for heat-sensitive materials, such as orange juice.

### 2.2.6 SCRAPED – FILM EVAPORATOR

Materials that become viscous or semisolid during evaporation process must be mechanically moved during evaporation. Scraped Film evaporators are used in these applications, but they are very expensive to build and operate. The submerge combustion unit avoids the need for a solid heat transfer surface by using the bubble surface of the combustion gases. This unit is especially useful for evaporating highly corrosive or severely scaling materials because there is no solid heat transfer surface to corrode or foul. (Luyben.1988)

## 2.3 TYPES OF EVAPORATION PROCESSES AND MODE OF OPERATION

Batch evaporation are ones in which filling, evaporating and emptying are consecutive steps. This method of operation is rarely practiced since it requires that the body be large enough to hold the entire charge of feed and the heating element be placed low enough so as not to be uncovered when the volume is reduced to that of the product.

The more usual method of operation is semi batch process, in which feed is continually added to maintain a constant level until the entire charge reach final density. Continuous batch evaporators usually have a continuous feed and, over at least part of the cycle, a continuous discharge. One method of operation is to circulate from a storage tank to the evaporator and back until the entire tank is up to the desired concentration and then finished in batches.

Continuous evaporators have essentially continuous feed and discharge, and concentration of both feed and product remain substantially constant.

### 2.3.1 MODE OF OPERATION

Any evaporator can be operated in one of two modes namely:

(i) Once through or single pass

(ii) Re-circulation or multipass

### 2.3.2 ONCE-THROUGH MODE

In once-through operation, the feed liquor passes through the steam-heated tubes once only, releases some vapour, and leaves the evaporator as thick liquor.

All evaporation is accomplished in a single pass. The ratio of evaporation to feed is limited in a single pass unit, thus these evaporators are well adapted to multiple effect evaporation where the total amount of concentration can be spread over several effects.

Operation of the once through evaporator under conditions of vacuum results in the reduction of the boiling point of the liquid. Consequently operating temperatures are reduced making this type of evaporation especially suitable for heat sensitive materials.

### 2.3.3 RE-CIRCULATION MODE

Evaporators operated in this mode contain a pool or reservoir of liquid held within the equipment. The incoming feed is allowed to mix with liquid in the reservoir, and then the mixture passes through the tubes. The unevaporated liquid issuing from the tubes returns to the reservoir, and only part of the evaporation is achieved in one pass. Complete evaporation is achieved by the continued re-circulation of the liquid through the tubes. The circulation of the liquid can be achieved in one of two ways.

(1) Natural circulation: Here the flow of liquid through the tube is induced by differences in liquid density between the liquid entering the tubes and the reservoir.

(2) Forced circulation: When density differences are very low and or the liquor is viscous, natural circulation will be inhibited and a pump is used to provide the require circulation.

This is termed forced circulation. Under these conditions the liquid entering the tubes is predominantly composed of the thick liquor in the reservoir and thus its concentration, viscosity, density and boiling points are very close to their maximum values. Consequently the resulting heat transfer coefficient is low, and repeated heating of the liquor in the tube makes the circulation evaporator unsuitable for the concentration of heat sensitive materials.

All short-tube and forced circulation evaporators are operated in this manner, while climbing film evaporators may be either circulation or once through units. (Ellison et al. 1985)

## 2.4 MULTIPLE EFFECT EVAPORATORS

Multiple effect evaporation is the principal means of evaporation in use for economizing energy consumption. Most such evaporators operate on a continuous basis, although for a few difficult materials a continuous-batch cycle may be employed.

In multiple effect operation, several evaporators are connected in series. The vapour or steam produced in the first effect is introduced to the steam chest of the second effect and thus becomes the heating medium from the second effect. Similarly, the vapour from the second effect becomes the steam for the third effect. Also with respect to feed, the thick liquor leaving the first effect becomes the feed for the second effect. For each effect added to the system, approximately one additional kilogram of solvents is evaporated per kilogram of steam fed to the first effect. This increase in the kilogram of solvent evaporated per kilogram of steam fed is achieved at the expense of additional capital outlay required for the additional effects.

To provide the temperature potential required for the heat transfer occurring in each effect, it is necessary that each effect be operated at a successively lower pressure. The operating pressure of the last effect is determined by the condensing capacity of the condenser following this. The term evaporator system is used to mean either one evaporator or any number of evaporators that are connected. In the same prescribed manner, unless otherwise stated, it will be assumed that the evaporators are connected in series with forward feed. (Mc Cabe and Smith. 1986).

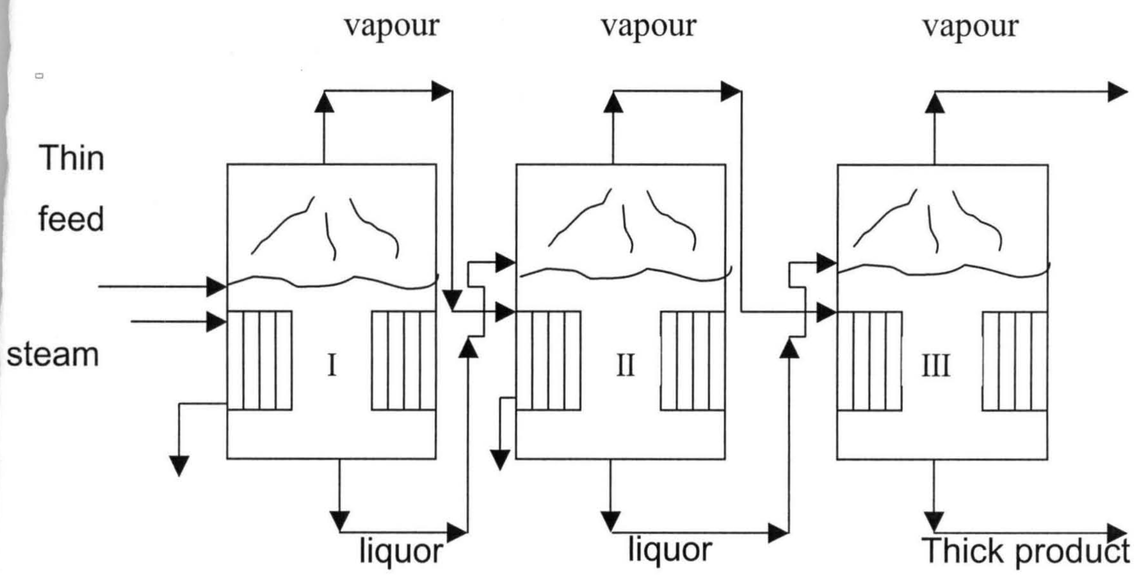

FIGURE. 2.4.1: SCHEMATIC DIAGRAM OF A FORWARD FEED MULTIPLE EFFECT EVAPORATOR

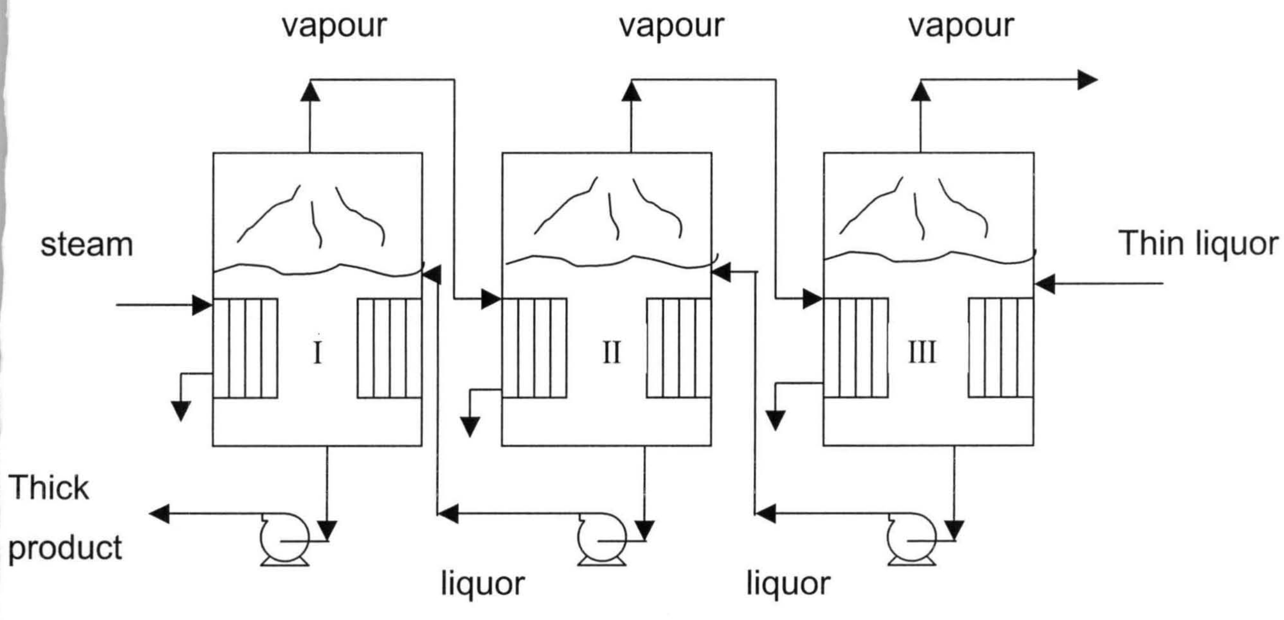

FIGURE. 2.4.2: SCHEMATIC DIAGRAM OF BACKWARD FEED MULTIPLE EFFECT EVAPORATOR

### 2.4.1 METHOD OF FEEDING

The usual method of feeding a multiple effects evaporator is to pump the thin liquid into the first effects and send it in turn through the other effects. This is called forward feed effect. The concentration of the liquid increases from the first effect to the last effect.

It is the simplest pattern of liquid flows. It requires a pump to feed the dilute solution into the first effect, since it is often at about atmospheric pressure, and a pump to remove thick liquor from the last effect. The transfer from effect to effect however can be done without pumps, since the flow is in the direction of decreasing pressure. Control valves in the transfer lines are all that is required. This method of operation is advantageous when the feed is hot or when the concentrated product will be damaged or would deposit scale at high temperature. When the feed is cold, forward feed gives a low steam economy since an appreciable part of the prime steam is needed to heat the feed to effect evaporation. If forward feed is necessary and feed is cold, steam economy can be improved markedly by preheating the feed in stages with vapour bled from intermediate effects of the evaporator.

In a backward feed operation, the raw feed enters the last (coldest) effect, the discharge from this effect becomes the feed to the next-to –the last effect, and so on until products is discharged from the first effect. This method of operation is advantageous when the feed is cold since much less liquid must be heated to higher temperature existing in the early effects.

This is also used when the product is so viscous that high temperatures are needed to keep the viscosity low enough to give reasonable heat-transfer coefficients. When product viscosity is high but a hot product is not needed, the liquid from the first effect is sometimes flashed to a lower temperature in one or more stages and the flash vapour added to the vapour from one or more later effects of the evaporator.

Mixed-feed operation is used only for special applications, as when liquor at an intermediate concentration and a certain temperature is desired for additional processing. In mixed feed, the dilute liquid enters an intermediate effect, flows in forward feed to the end of the series, and then pumped back to the first effect for final concentration. It eliminates some of the pumps needed in backward feed system and yet permits the final evaporation to be done at the highest temperature.

Parallel feed involves the introduction of raw feed and withdrawal of product at each effect of the evaporator. It is used primarily when the feed is substantially saturated and the product is a solid.

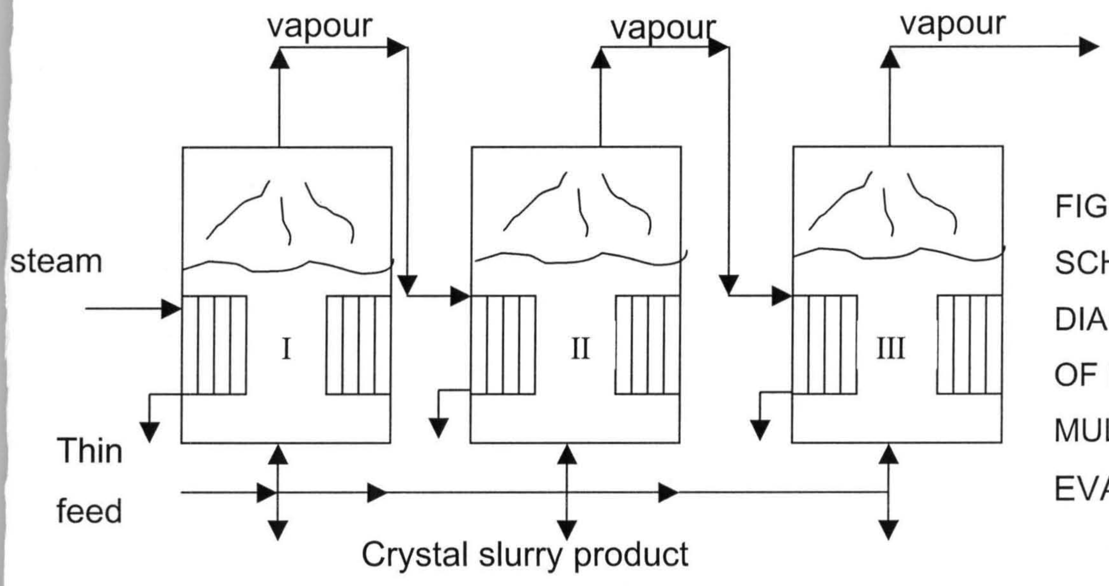

FIGURE. 2.4.3: SCHEMATIC DIAGRAM OF PARALLEL FEED MULTIPLE EFFECT EVAPORATOR

### 2.4.2 OPTIMUM NUMBER OF EFFECTS

The greater the number of effect the larger amount of evaporation per kilogrammes of steam admitted to the first effect. The operating cost will be less, the larger the number of effects. The cost of condenser water too will decrease, the greater the number of effects employed.

In general term, the cost of each effect of an evaporator per square metre of surface is a function of the total area and decreases with area, approaching an asymptote for large installations. Thus the investment required for an N-effect evaporator is about N-times that for a single effect evaporator of the same capacity. The optimum number of effects must be found from an economic balance between the saving in steam obtained by multiple effect operation and added investment required. This is done by computing the process requirements with two, three, four up to six, eight effects and determining the fixed and operating cost resulting from each arrangement. (Badger and Banchero. 1985)

### 2.4.3VAPOUR RECOMPRESSION

The energy in the vapour evolved from a boiling solution can be used to vaporise more water provided there is a temperature drop for heat transfer in the desired directions.

In a multiple effect evaporator this temperature drop is created by progressively lowering the boiling point of the solution in a series of evaporator through the use of lower absolute pressures.

The desired driving force can also be obtained by increasing the pressure of the evolved vapour by mechanical or thermal recompression. The compressed vapour is then condensed in the steam chest of the evaporation from which it comes.

### 2.4.4 MECHANICAL RECOMPRESSION

Cold feed is preheated almost to its boiling point by exchange with hot thick liquid and is pumped through a heater as in a conventional forced circulation evaporator.

The vapour evolved, however, is not condensed; instead it is compressed to a somewhat higher pressure by a centrifugal compression and becomes the steam, which is fed to the heater. Since the saturation temperature of the compressed vapour is higher than the boiling point of the feed, heat flows from the vapour to the solution, generating more vapour. The most important applications of the Mechanical recompression evaporation are the concentration of the very dilute radioactive solutions and the production of distilled water.

### 2.4.5 THERMAL RECOMPRESSION

In a thermal recompression system, the vapour is compressed by applying high pressure on it in a jet ejector. This results in more steam than is needed for boiling the solution, so that excess steam must be vented or condensed. Some steam jets can handle large volumes of lower density vapour; thermal recompression is better suited to vacuum evaporation than mechanical recompression.

The major disadvantages of thermal recompression are the low mechanical efficiency of the jets and lack of flexibility in the system towards changed operating conditions.

## 2.5 EVAPORATOR PERFORMANCE

Capacity and economy are the principal measures of performance of steam heated tubular evaporators.

Capacity can be defined as the mass of water vaporized per hour, whilst the economy is defined as the number of kilogrammes of water vaporized per kilogramme of steam feed to the unit.

### 2.5.1 EVAPORATOR CAPACITY

The rate of heat transfer, Q, through the heating surface of an evaporator is the product of three factors:

(i) The heat transfer surface A ($m^2$)
(ii.) The overall heat transfer coefficient, U and
(iii) The overall temperature driving force, $\Delta T$ ($^0C$)

Thus using $Q = UA\Delta T$……………………………………………2.5.1

For any evaporator the heat transfer area (A) is constant and so the rate of heat transfer Q is directly dependent on the overall heat transfer coefficient (U) and the temperature driving force $\Delta T$

### 2.5.2 FACTORS AFFECTING THE TEMPERATURE DRIVING FORCE

The temperature driving force for heat transfer across the heating surface depends on:

i. the liquid feed temperature
ii. the difference in pressure between the steam chest and the separator above the boiling liquid
iii. the boiling point temperature of the liquid

iv. the velocity of liquid flow in the tubes (if this velocity is high, the temperature, driving force is reduced because the frictional loss in the tubes increases the effective pressure in the liquid thus raising its boiling point temperature)

With industrial evaporators two factors, boiling point elevation and liquid head affect the boiling point of the solution. (Ellison et al. 1985).

### 2.5.3 EFFECT OF BOILING POINT ON THE TEMPERATURE DRIVING FORCE ($\Delta T$)

The increase in the boiling point above that of water is defined as the boiling point elevation of the solution. It may be as large as 70°C for concentrated solutions of inorganic salts, but is small for solutions of organic colloids and dilute solutions.

Any boiling point elevation reduces the temperature driving force for heat transfer since the boiling point elevation must be subtracted from the temperature drop that is predicted from steam table.

For concentrated solutions the boiling point elevation may be determined by the use of "DUHRING RULE" which indicates that the boiling point of a given solution is a linear function of the boiling point of pure water at the same pressure. If the boiling point of a solution is plotted against that of water at the same pressure, a straight line is obtained. The rule is not exact especially over wide range of pressure but the use of precalculated *Duhring's* lines especially for aqueous solutions of sodium hydroxide is convenient.

However, if the solution is sufficiently diluted and it can be assumed that the solution has the characteristics of pure water then, if the pressure is known, its boiling point can be read directly from steam tables, as can the temperature of the condensing steam

### 2.5.4 EFFECT OF LIQUID HEAD AND FRICTION ON THE TEMPERATURE DRIVING FORCE ($\Delta T$)

If the depth of liquid in the evaporator tube is large, then the boiling point corresponding to the pressure in the separator is the boiling point of the surface layers of liquid. Liquid at a distance 'h' meters below the liquid/vapour interface is under a pressure of the separator plus the head of 'h' metres of liquid, and thus has a higher boiling point.

Furthermore, if the velocity of the liquid is large then the frictional forces further increase the average pressure in the liquid. Under these conditions the average boiling point of liquid in the tube is higher than the boiling point predicted from the separator pressure. The increase in the boiling point caused by liquid head and friction reduces the temperature driving force for heat transfer, since yet again the resultant boiling point elevation must be subtracted from the temperature drop that is predicted from steam table.

### 2.5.5 FACTORS AFFECTING THE OVERALL HEAT TRANSFER COEFFICIENT

The rate of heat transfer and evaporator capacity is a function of the temperature driving force and the overall heat transfer coefficient (U). The temperature driving force ($\Delta T$) is defined by the properties of the steam and the boiling liquid but, however, the overall heat transfer coefficient is also strongly affected by the design of the evaporator.

In order to appreciate this, the make up of the overall heat transfer coefficient must be examined. From the theory of basic heat transfer for heat exchanger using steam to boil a liquid, it is known that the overall heat transfer is the summation of:

i Steam film-condensing coefficient on the outside of the heat exchanger tube ($h_o$)

ii The thermal resistance of the heat exchanger tube (R)

iii The liquid film heat transfer coefficient on the inside of the heat exchanger tube $h_i$

iv The thermal resistance of scale deposits on both the inside and outside of the tube.

Evaporator design influence the value of the above mentioned components and in consequence affects the overall heat transfer coefficient as follow:

1. **Steam film coefficient –**

This is usually very high even when the condensation is film wise. However, the presence of non-condensable gases drastically reduces this coefficient, thus adversely affecting the overall heat transfer coefficient.

This situation can typically arise through steam containing air, which can build up in the steam chest. Designs normally incorporate air bleed mechanisms to remove trapped air.

2. **Thermal resistance of the exchanger tubes: -**

   Most large industrial evaporators are normally fabricated of metal, so resistance through the tube walls is generally very low and normally ignored. However, when glass is used, the thermal resistance is greater so its effect must be taken into account when calculating the overall heat transfer coefficient.

3. **Liquid film heat transfer coefficient –**

   Both the liquid properties and its velocity through the tubes affect the value of the liquid film coefficient.

   Typically $h \propto V^{0.8}$

Where V is the fluid velocity.

Therefore increasing velocity will increase the liquid film coefficient and thus increase the overall heat transfer coefficient. This may be achieved in practice by incorporating a pump as in forced circulation.

4. **Resistance of scale deposits.**

When scale is deposited either on the inside or the outside of the evaporator tubes, thermal resistance will be increased and the overall heat transfer coefficient reduced. Thermal conductivities of scales are generally low and so the resistances offered by the scale deposits can have a considerable effect on evaporator capacity.

Glass is particularly useful here because it has very smooth surfaces, and so scale deposits are minimised. Otherwise the equipment must be specially designed so that scale deposits can be removed easily. (Ellison et al. 1985).

Overall heat transfer coefficient is dependent on the individual coefficients and resistances. When one of these is predominating, then major changes in the other resistances will have negligible effect on the overall heat transfer coefficient. Typical overall heat transfer coefficient for several designs of evaporators is given in Table (2.5.5) below.

TABLE 2.5.5: TYPICAL HEAT TRANSFER HEAT COEFFICIENT OF COMMON TYPES OF EVAPORATORS

| Evaporator Type | Overall Heat Transfer Coefficient (U) $W/m^2K$ |
|---|---|
| Long-tube vertical evaporators | |
| i. Natural circulation | 1,000 – 3000 |
| ii. Force Circulation | 2,000 – 10,000 |
| Short-tube evaporators | |
| i. Horizontal tube | 1,000 – 2,000 |
| ii. Calendria type | 800 – 25,000 |
| Coil evaporator | 100 – 3000 |

### 2.5.6 EVAPORATOR ECONOMY

The economy of an evaporator is defined as kilograms of solvent evaporated per kilograms of steam used.

As a rule of the thumb it is roughly 0.8 for a single effect, 1.6 for double effect, 2.4 for triple effect evaporators and so on.

The economy of an evaporator is influenced not by the number of effects but also by feed temperatures, crystallization, heat effects, and flow of condensate and head recovery stream.

Evaporator performance is rated on the basis of steam economy. Heat is required,

(i) To raise the feed from its initial temperature to the boiling temperature,

(ii) To provide the minimum thermodynamic energy to separate liquid solvent from the feed.

(iii) To vaporize the solvent.

The first of these functions can be changed appreciable by reducing the boiling temperature or by heat interchange between feed and residual product and / or condensate. The greatest increase in steam economy is achieved by re using the vaporised solvent. This is done in a multiple effect as the heating medium for another effect in which boiling takes place at lower temperature and pressure.

Another method of increasing the utilizations of energy is to employ a thermo-compression evaporator, in which the vapour is compressed so that it will condense at a temperature high enough to permit its use as the heating medium in the same evaporator. (Perry et al. 1988)

## 2.6. EVAPORATOR TROUBLESHOOTING AND DEBUGGING

The operation of a set of multiple effect evaporators is an amazing phenomenon. Once they are started, by introducing the feed and turning on the steam, they reach their equilibrium and attain a steady state. When clean, the steam pressure in the steam chest of the first affect will be below the design pressure, and as they foul, it goes up just enough to increase the $\Delta T$ to the level necessary to produce the heat transfer required. When the maximum allowable pressure is reached, the evaporators must be clean, or the evaporation rate will fall below the design. The time for this cycle might be days or weeks, depending on the service, but during that time they operate pretty much on their own. Even when a change is made that causes an upset, they quickly establish equilibrium and stabilize to the new condition.

However, there are times after cleaning when, instead of stabilizing as before, the liquor carries over to the steam chest of the next effect and goes to the sewer. This might occur in any of the effects, or in two or more at the same time. It is situations like this that convince operators that evaporators are the worst equipment to analyse and debug have any ever made.

Troubleshooting is primarily an art, developed by experience in operating evaporators and the approach will vary from process to process.

Such simple things as checking for leaks, and checking pumps, non-condensable removal, steam and water flows etc. are second nature to an operator.

But when a set of evaporators will not operate after those things have been checked, experience has taught the operator to look for what can be done. The next step is to call the designer and request an immediate visit because science is required to augment the art.

What is required of the engineer is to start up the evaporators and try to obtain temperature and pressure readings, so that a heat balance can be run, usually this will pinpoint the problems area. If necessary, each effect may be investigated to find the exact trouble, but at least the search be narrowed so that all of the effects do not have to be opened. The ability of the engineer to run the heat balance in a comparatively short time, by trial and error, is the key to solving the problem. Solving any problem by trial and error can be very time – consuming unless the skill has been developed by continuous practice, and evaporator heat balances can be among the toughest to solve by this method. Look at the time, and money, which would have been saved, had someone at the plant been able to run a heat balance.

A computer program can be developed to easily carry out the heat balance quickly, once all necessary data are entered. As computer control of evaporators becomes common, this program will enable continuous, or at least frequent, running of heat balance. Even when there is no trouble in operation of the evaporators, this heat balance is a valuable aid in proving the errors in the instrument readings. In many cases, serious arguments have occurred between plant operators and the instrument department concerning the accuracy of the data, and a heat balance has been useful.

Obviously, errors in the feed flow or steam flow will have a greater effect on the heat balance-calculated evaporation rate, but errors in the temperature readings will change the coefficients of heat transfer in each effect, and thus cause false indications of trouble.

Even though one is not an expert in evaporator design. Many tricks-of –the –trade are used by Operators in various industries to stabilize the operation of evaporators. The methods are numerous, but a couple will be mention as illustrations.

(1) As mentioned earlier, evaporators are designed to handle a feed that contains specified solids content. In practical application, it is not uncommon to have variations from the norm and thus create operating problems. Circulating some of the product liquor back to the feed at a controlled rate and thereby maintaining the solids at a uniform level can minimize this. This is called "sweetening" and is widely used by operators.

(2) Another variable that creates operating problems are feed-liquor characteristics. This may take different forms in different industries, but a good example is available in the kraft pulping industry where the alkali content of the black liquor affects foaming characteristics during evaporation. When the alkali content drop below a certain level, carry over increases, so many operators will add caustic or caustic-containing liquor available from the process (white liquor). With only a small addition of caustic, the evaporators settle down to a smooth level. When the feed liquor returns to the normal level, the caustic addition is stopped.

Operators of evaporators in most industries today earn their jobs by learning how to keep equipment running efficiently.

They do not have to know why a certain action helps to correct a problem, as long as the result is obtained. If the level in one effect goes up, and by switching pumps the level returns to normal, the operator does not have to be an engineer to conclude that the first pump needs repairs.

When pressure in the front end starts to climb and the $\Delta T$ in a middle effect is abnormally high, the operator opens the non-condensable line in the middle effect and the evaporator returns to normal, his primary task-to keep the evaporator running has been accomplished. Then the cause can be looked for without the pressure of an imminent shut down.

Operators of today have a greater knowledge of their art (just as designers have a greater knowledge of their science). They want to learn more of the science involved, and the scientist should be just as anxious to learn the art of operating evaporators. Computer will no doubt control the evaporators in the future, but having an “artist” around will be a valuable asset. (Vincent et al. 1979)

## 2.7 THE RATIONAL FOR MATHEMATICAL MODELLING

A model is nothing more than mathematical abstraction of a real process. The equation or sets of equations that comprise the model are best an approximation to the true process. Hence, the model cannot incorporate all of the features, both macroscopic and microscopic of the real process. The engineer normally must seek a compromise involving the cost of obtaining the model, that is, the time and effort required to obtain and verify it.

These considerations are related to the level of physical and chemical details in the model and the expected benefits to be derived from its use. The model accuracy necessary is intertwined in this compromise and the ultimate use of model influences how accurate it needs to be. The importance of mathematical modelling of a process are explain as follows:

1. To improve understanding of the process: - process model can be analysed or used in a computer simulation of the process to investigate process behaviour without the expense and perhaps, without the unexpected hazard of operating the real process.
   This approach is necessary when it is not feasible to perform dynamic experiments in the plant or before the plant is actually constructed.

2. To train plant-operating personnel: - plant operator can be trained to operate a complex process and to deal with emergency situation by use of process simulator. By interfacing a process simulate to standard process control equipment, a realistic environment can be created for operator training without the cost or exposure to dangerous conditions that might exist in a real plant situation.

3. To design the control strategy for a new process: - A process model allow alternative control strategies to be evaluated such as, for example, the selection of the variables that are to be measured (controlled) and those that are to be manipulated.

With more complicated processes or with new processes for which we have little operating experience, the design of an appropriate control strategy seldom is straightforward.

4. To select controller settings. A dynamic model of the process may be used to develop appropriate controller settings, either via computer simulation or by direct analysis of the dynamic model. Prior to start-up of a new process, it is desirable to have reasonable estimates of the controller settings. For some operating processes, it may not be feasible to perform experiments that would lead to better controller settings.

5. To design the control law: a Modern control technique often incorporates a process model into the control law. Such techniques are called model – predictive or model-based control.

6. To optimise process-operating conditions: In most processing plants there is an incentive to adjust operating conditions periodically so that the plant maximizes profits and minimizes costs. A steady state model of the process and appropriate economic information can be used to determine the most profitable process conditions as in supervisory control. (Serborg et al. 1989)

## 2.8 DEVELOPMENT OF A PROCESS MATHEMATICAL MODEL

Because models account more completely for what is happening in a process, fundamental model based on mass and energy balance offers better accuracy and greater insight than do regression models. Thus they are of greater value in improving process yield or through put or in debottlednecking and optimising. Fundamentals models may be in existence for a long time without being used because they involved a lot of calculations and are considered difficult to develop. However, today a calculation can be done quickly and conveniently, at low cost, using an advanced computer having an arithmetic coprocessor and some graphics capability. As for building the model, there is a systematic hand-on approach that will ease the perceived development difficulties.

A model, can be developed by fitting a set of equations to actual plant or pilot plant data, and the range of these data is the range over which the model can be used. For this reason, it is important to have data that covers a fairly wide range of steady state process conditions.

In a pilot plant, such data can be collected by systematic experimentation. The experiment designed should explored high, average and low values for each of the principal process variables. Such a range of data can also be found in plant operating history. Overall, ten (10) to twenty (20) steady state data sets may be required.

The kind of data needed to develop a mathematical model are those that define the material and energy balances around the equipment i.e., the pressure, temperature and composition of feed and product streams etc.

The three major steps for building models are as follows: the first step is to develop a process flow diagram for the process. Next, is to develop a set of material and energy balance equations for the systems or around the equipment being considered and the third step is to fit the model to the data by adjusting the parameters.

For large number of parameters, this step involves a number of iterations and must be done with the help of a computer. If the Mathematical model is accurate and more useful than any single material balance being written, it can be used to predict the behaviour of the process under different conditions. Then it can be used for process optimisation, supervisory control, debottlenecking and dynamic analysis. Obviously, success in any of these areas can pay for the mathematical model development in a very short time. (Wansbrough. 1985)

## 2.9 SOME EXISTING EVAPORATOR MODELS

Owing to the great importance of evaporation technology, many model equation of evaporators performance were being developed but very few and simplify one are the ones that are usually published. This is due to the fact that manufacturers of evaporations in most cases protect their own interest for economic and other reasons by not publishing their detail models of evaporator system.

However, few simplified unsteady state models of evaporator systems are in existence. A general steady state model of evaporator system usually found in standard textbook is illustrated below.

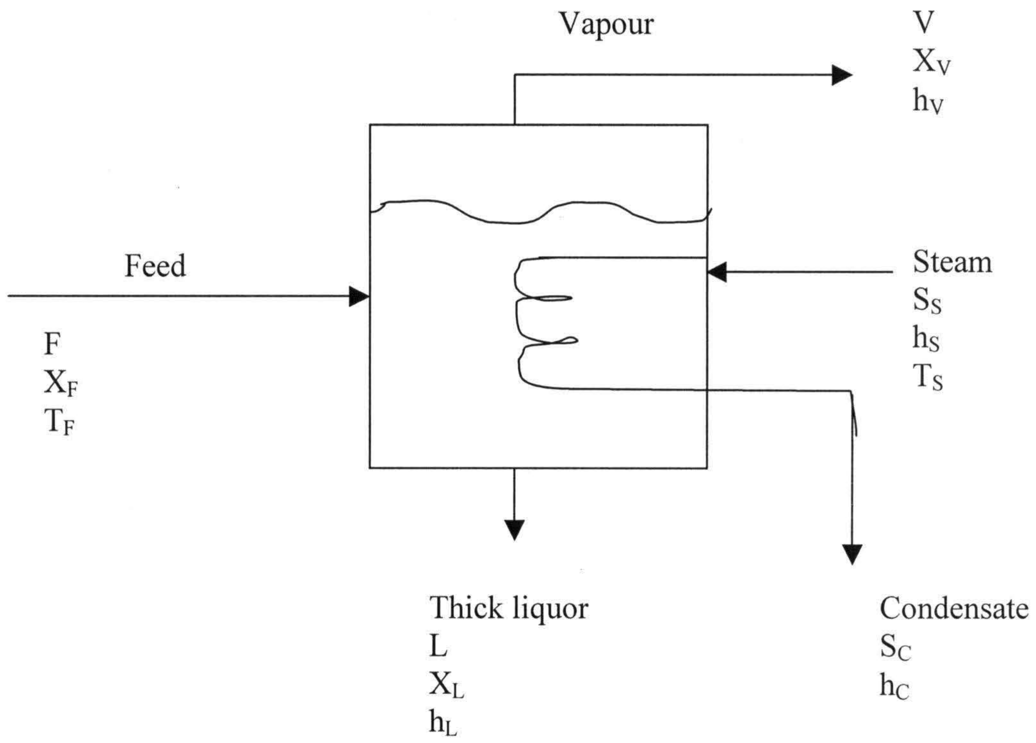

FIGURE: 2.9.0 SCHEMATIC DIAGRAM OF A SINGLE EFFECT EVAPORATOR

MATERIAL BALANCE

Total mass balance

$F= L + V$……………………………………….2.9.1

Solute balance

$FX_F = LX_L + VX_V$……………………………………….2.9.2

ENERGY BALANCE

Enthalpy balance

$Fh_F + Sh_S = Vh_V + Lh_L + Sh_C$……………………….2.9.3

Heat Transfer rate

$Q = UA\Delta t_m$……………………………………….2.9.4

A model for a double effect evaporator is shown in figure 2.9.1 below.

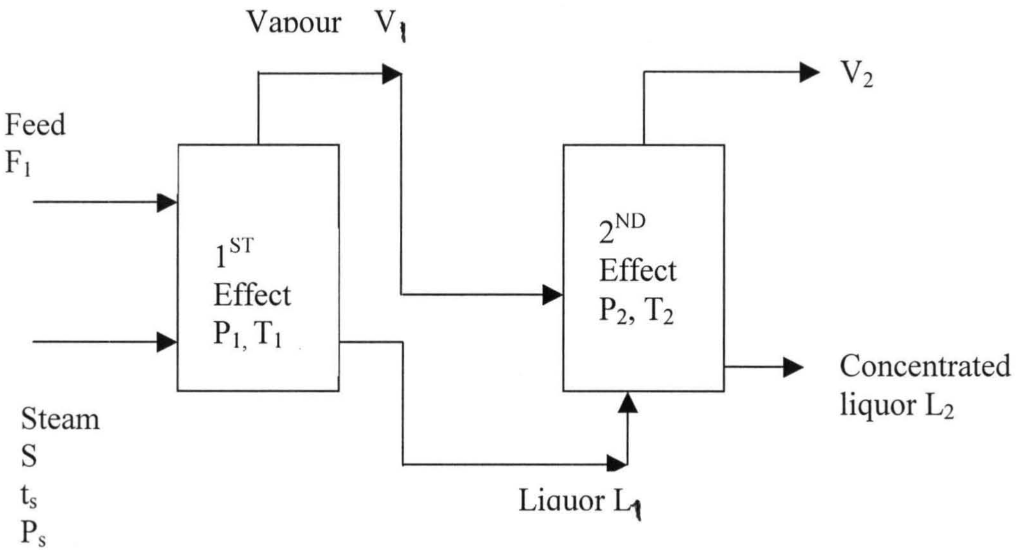

FIGURE. 2.9.1- A SCHEMATIC DIAGRAM A DOUBLE EFFECT EVAPORATOR

From the above figure 2.9.1, the following balance is obtained.

Total balance

$$L_1 = F_1 - V_1 = F_2 \qquad 2.9.5$$

$$L_2 = F_1 - V_1 - V_2 \qquad 2.9.6$$

$$Q_{TOTAL} = Q_1 + Q_2 + Q_3 + \ldots \qquad 2.9.7$$

$$Q_1 = U_1 A_1 \Delta t_1 \qquad 2.9.8$$

Where Q is the heat load, $\Delta t$ is the temperature difference, U is the overall heat transfer coefficient and A is the heat transfer area

$\Delta t_1 = t_s - t_1$ ............................................ 2.9.9

$Q_2 = U_2 A_2 \Delta t_2$ ............................................ 2.9.10

and $\Delta t_2 = t_1 - t_2$ .................................... 2.9.11

$Q_3 = U_3 A_3 \Delta t_3$ ............................................ 2.9.12

and $\Delta t_3 = t_2 - t_3$ .................................... 2.9.13

But $Q_1 = Q_2 = Q_3$........................2.9.14

Usually, $A_1 = A2 = A3$.................2.9.15

Therefore, $U_1 \Delta t_1 = U_2 \Delta t_2 = U_3 \Delta t_3$....................................2.9.16

## 2.10 THE MATHEMATICAL MODEL OF "ARMFIELD" DOUBLE EFFECT EVAPORATOR

The model for a double effect climbing film "Armfield" evaporator of the chemical engineering department, Kaduna Polytechnic, under a steady state condition and assuming negligible heat loss is as follows:

### FIRST EFFECT

Mass Balance

$W_F = W_E + W_C$ ....................................................... 2.10.1

Energy Balance

$W_F h_F + Q_1.H_S = W_E.H_E + W_C.h_C + Q_1.h_S$ ...........2.10.2

SECOND EFFECT

Mass Balance

$W_F+Q_2=W_C+W_{E2}$ .......................................2.10.3

Energy Balance

$W_F.h_F + Q_1.H_S=W_{E2}.H_{E2}+W_C.h_C+Q_1.h_S+Q_2.h_2$...........2.10.4

$W_F=C_1.\Delta L_1$...........................................................2.10.5

$W_E=C_2.\Delta L_2$.........................................................2.10.6

$W_C=C_3.\Delta L_3$.........................................................2.10.7

Where C= calibration constant of tank

$\Delta L$= change in level of liquid in tank

## 2.11 THE NEWELL, SERBORG AND FISHER DOUBLE EFFECT EVAPORATOR MODEL.

This model is based essentially on Newell and Fisher (1972) and Serburg and fisher (1979). The work was developed for a single effect evaporator (figure 2.12.0) and a double effect evaporation (figure 2.12.1) (Balchen and Mumme. 1988)

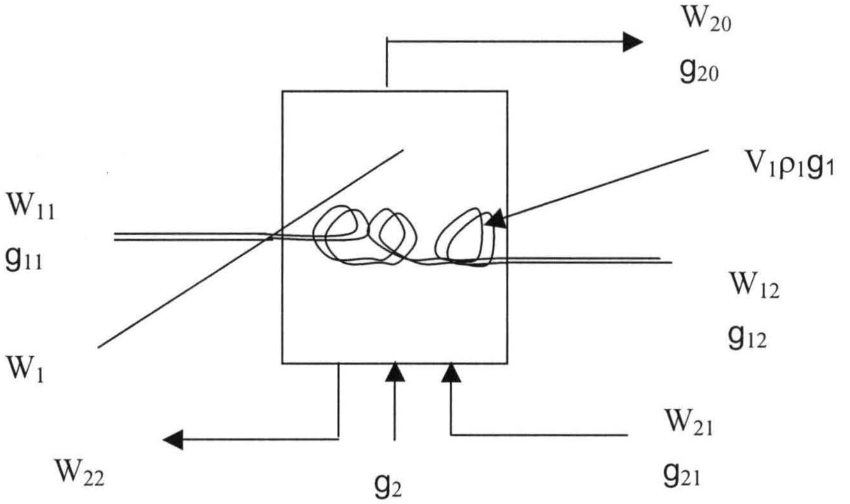

FIGURE 2.11.0: SCHEMATIC DIAGRAM OF A SINGLE EFFECT EVAPORATOR.

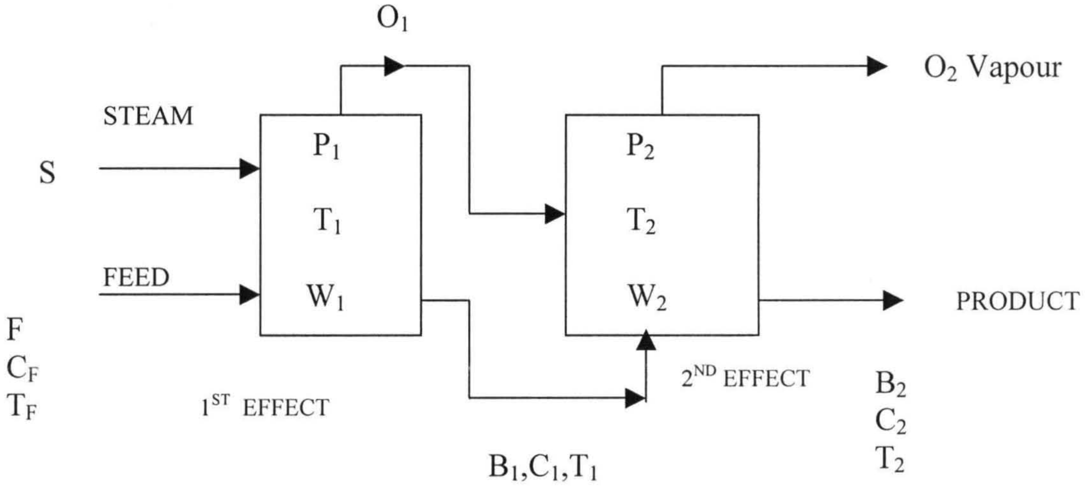

FIGURE 2.11.1 SCHEMATIC DIAGRAM OF A DOUBLE EFFECT EVAPORATOR.

From the above figures.

## PRIMARY SYSTEM (STEAM)

### Vapour balance

$V_1 d\rho_1/dt = w_{11} - w_{12}$ ………………………………2.11.1

### Energy Balance

$V_1 d(\rho_1 g_1)/dt = w_{11}g_{11} - w_{12}g_{12} - Q_1 - Q_{01}$ ………………2.11.2

### Heat transfer

$Q_1 = h_1 A\,(\theta_1 - \theta_V) = w_{12}\,\lambda_1$ ………………………………2.11.3

## TUBE SYSTEM (METAL)

$$W_V \cdot C_V \cdot \left(\frac{d}{dt}\theta_v\right) = \left[h_1 \cdot A \cdot \left(\theta_1 - \theta_v\right)\right] - h_2 \cdot A \cdot \left(\theta_v - \theta_2\right) = Q_1 - Q_2 \qquad \text{…………2.11.4}$$

## SECONDARY SIDE (LIQUID SYSTEM)

$$\frac{d}{dt}W_2 = w_{21} - w_{22} - w_{20} \qquad \text{……………………………………2.11.5}$$

$$\frac{d}{dt}\left(W_2 \cdot C_2\right) = w_{21} \cdot C_1 - w_{22} \cdot C_2 \qquad \text{……………………………2.11.6}$$

$$\frac{d}{dt}\left(W_2 \cdot g_2\right) = w_{21} \cdot g_{21} - w_{22} \cdot g_2 - w_{20} \cdot g_{20} + Q_2 - Q_{02} + L_2 \qquad \text{…… 2.11.7}$$

$$Q_2 = h_2 \cdot A \cdot \left(\theta_v - \theta_2\right) \qquad \text{………………………………………2.11.8}$$

## 2.12 THE SERBORG AND FISHER DOUBLE EFFECT EVAPORATOR MODEL

This model was developed using a pilot plant double effect evaporator. It is a modification of their earlier work. The following assumptions were made for the model:

1. The heat capacity of the steam chest, tube walls and other metal parts of the evaporator are sufficiently small that they may be negligible.

2. The pressure controller on the second effect is sufficiently powerful to hold the temperature in the second effect at steady state with negligible dynamic variations.

3. The solute concentration in the vapour leaving each effect of the evaporator is negligibly small compared with the amount of solute leaving the liquid. (Harmon. 1989)

Under these conditions, total material, solute and heat balance on the first effect may be written using the figure 2.13.1 below.

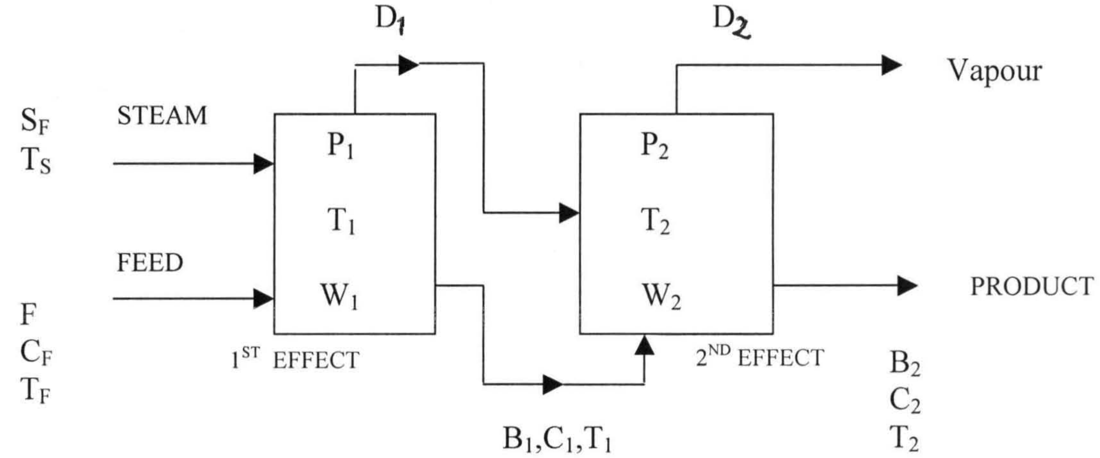

FIGURE 2.13: SCHEMATIC DIAGRAM OF A PILOT PLANT DOUBLE EFFECT EVAPORATOR

## FIRST EFFECT

$$\frac{d}{dt}W_1 = F - B_1 - D_1 \quad \text{.......... 2.}$$

$$W_1 \cdot \frac{d}{dt}C_1 = F \cdot \left(C_F - C_1\right) + D_1 \cdot C_1 \quad \text{.......... 2.1}$$

$$W_1 \cdot \frac{d}{dt}h_1 = F \cdot \left(h_F - h_1\right) - D_1 \cdot \left(H_{1.V} - h_1\right) + Q_1 - L_1 \quad \text{.......... 2.12.3}$$

Similarly the material balance on the second effect gives

$$\frac{d}{dt}W_2 = B_1 - B_2 - D_2 \quad \text{.......... 2.12.4}$$

$$W_2 \cdot \frac{d}{dt}C_2 = B_1 \cdot \left(C_1 - C_2\right) + D_2 \cdot C_2 \quad \text{.......... 2.12.5}$$

While a steady state heat balance on the second effect yields

$$D_2 \cdot \left[\left(H_{2.V}\right) - h_2 + \frac{\partial h_2}{\partial C_2} \cdot C_2\right] = Q_2 - L_2 + B_1 \cdot \left(h_1 - h_2\right) + \frac{\partial h_2}{\partial C_2} B_2 \cdot \left(C_2 - C_1\right) \quad \text{.......... 2.12.6}$$

Here $Q_1$ and $Q_2$ are heat input to each effect given by

$$Q_1 = U_1 \cdot A_1 \cdot \left(T_S - T_1\right) \quad \text{.......... 2.12.7}$$

$$Q_2 = U_2 \cdot A_2 \cdot \left(T_1 - T_2\right) \quad \text{.......... 2.12.8}$$

## 2.13. SOLUTION OF THE MODEL EQUATION

Modelling encompasses much more than solution of differential equations. If the equations are already known, they have to be calibrated and parameters and initial conditions must be optimised. This can be accomplished by solving the equations using standard computational techniques, which is tedious, and time consuming. But with the advent of computers, solutions of complex mathematical models for multiple effect evaporators system can now be accomplished with ease. These solutions are with known data and are used to predict the behaviours of real system, to improve control and facilitate cost calculation and planning of operation.

## 2.14. TESTING OF THE MODEL EQUATION

Testing of any kind of model equation simply means proving that the model is a true representation of the behaviour of the system in question. The problem of testing is indeed a difficult one because; it involves a host of practical, theoretical, statistical and even philosophical complexities. If the model of the process is accurate, when both the model and the process have the same input conditions they should both have the same output conditions. If not, then we must start over again.

# CHAPTER THREE

## 3.0 MATHEMATICAL MODELLING OF DOUBLE EFFECT EVAPORATORS

The mathematical modelling of any process involves the development of a more general equation from fundamental laws governing the system or modification of existing equations describing similar system through the use of necessary assumption to the process or part of the process.

The application of the laws of conservation of mass and energy across the boundaries of the system, are considered in developing the dynamic model of an N-effect evaporator system.

The assumptions made are:

(1). The heat capacities of the steam chest, tube wall and other metal part are negligible.

(2). The process liquid in the hold of the evaporator is perfectly mixed.

(3). The mass of solvent in the vapour space is negligible relative to the mass of hold up of thick liquor in the evaporator.

(4). The solute concentration in the vapour leaving each effect of the evaporator is negligibly small compared with the amount of salute leaving the liquid.

(5). The mass of steam in the steam chest is negligible relative to other terms that appear in the energy balance for this portion of the system.

(6). Heat losses to the environment are negligible.

## 3.1 MATERIAL BALANCE

A material balance for a system is based on the law of conservation of mass consequently, for a system in which the conversion of mass to energy and conversely is not involved it follow that during the time period from $t = t_n$ to $t = t_{n+\Delta t}$

$$\boxed{\begin{array}{l}\text{Input of materials}\\ \text{To the system}\\ \text{During the time}\\ \text{Period } \Delta t\end{array}} - \boxed{\begin{array}{l}\text{Output of materials}\\ \text{from the system}\\ \text{during the time}\\ \text{Period } \Delta t\end{array}} = \boxed{\begin{array}{l}\text{Accumulation}\\ \text{of materials}\\ \text{within the system}\\ \text{during the time period } \Delta t\end{array}} \qquad \ldots\ldots 3.1$$

The accumulation term is defined as follows:

$$\boxed{\begin{array}{l}\text{Accumulation of}\\ \text{materials within}\\ \text{the system during}\\ \text{the time period } \Delta t\end{array}} = \boxed{\begin{array}{l}\text{Amount of}\\ \text{materials in the system}\\ \text{at time } t_{n+\Delta t}\end{array}} - \boxed{\begin{array}{l}\text{Amount of material}\\ \text{in the system}\\ \text{at time } t_n\end{array}} \qquad \ldots\ldots 3.2$$

An unsteady state analysis of the above material balance can be easily restated in the following form

$$\int_{t_n}^{t_{n+\Delta t}} \left| \boxed{\begin{array}{l}\text{Input of}\\ \text{material per}\\ \text{unit time}\end{array}} - \boxed{\begin{array}{l}\text{Output of}\\ \text{material per}\\ \text{unit time}\end{array}} \right| dt = \boxed{\begin{array}{l}\text{Amount of}\\ \text{material in}\\ \text{the}\\ \text{system}\end{array}}\Bigg|_{t_{n+\Delta t}} - \boxed{\begin{array}{l}\text{Amount of}\\ \text{material in}\\ \text{the system}\end{array}}\Bigg|_{t_n} \qquad \ldots 3.3$$

ENERGY BALANCE

Energy balance is based on the first law of thermodynamics, which asserts that energy of the universe is constant. The following formulation of energy balance is applied to systems at unsteady state operations:

$$\int_{t_n}^{t_{n+\Delta t}} \left\| \boxed{\begin{array}{l}\text{Input of}\\ \text{energy to the}\\ \text{system per}\\ \text{unit time}\end{array}} - \boxed{\begin{array}{l}\text{Output of}\\ \text{energy from}\\ \text{the system}\\ \text{per unit time}\end{array}} \right| dt= \boxed{\begin{array}{l}\text{Amount of}\\ \text{energy}\\ \text{within the}\\ \text{system}\end{array}}\Bigg|_{t_{n+\Delta t}} - \boxed{\begin{array}{l}\text{Amount of}\\ \text{energy}\\ \text{within the}\\ \text{system per}\end{array}}\Bigg|_{t_n} \quad \text{........3.4}$$

## 3.3 DYNAMIC MODELLING OF N-EFECT EVAPORATOR SYSTEM

The modelling of N-effect evaporator system is initiated by formulation of the dynamic model for a single effect evaporator at unsteady state operation, based on the assumptions stated above. Supposing at time $t = t_n$ the evaporator is at steady state operation, and at time $t = t_{n+\Delta t}$ an upset in some operating variable, say the concentration $X_F$ of the feed, occurs. The material and Energy balances as well as the rate expressions are obtained as follows:

## MATERIAL BALANCE

A total material balance on the thick liquor has the following form

$$\int_{t_n}^{t_{n+1}} \left(F - V_1 - L_1\right) dt = W_1 \cdot \Big|_{t_{n+1}} - W_1 \cdot \Big|_{t_n} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 3.5$$

Where, F = the feed rate to the evaporator system (kg/s), $V_1$=mass flow rate of the vapour from the first effect (kg/s), $L_1$= mass flow rate of liquid from the first effect (kg/s); and $W_1$= mass hold up liquid in the first effect (kg).

From this integral-difference equation (3.5) as well as those, which follow, the corresponding differential equations are obtained through the use of mean-value theorem of differential and integral calculus and appropriate limiting process.

The left and right hand side of equation 3.5 may be written in the following forms through the use of mean-value theorem of integral calculus.

$$\int_{t_n}^{t_{n+1}} \left(F - V_1 - L_1\right) dt = \left(F - V_1 - L_1\right)\Big|_{tn+\alpha\Delta t} \quad \boxed{\ldots\ldots 3.6}$$

where $0 \le \alpha \le 1$

$$W_1 \cdot \left. \right|_{t_{n+1}} - W_1 \cdot \left. \right|_{t_n} = \Delta t \cdot \frac{d}{dt} W_1 \cdot \left. \right|_{t_{n+\beta\Delta t}} \quad \text{.........} 3.7$$

where $0 \le \beta \le 1$

substituting equation 3.6 and 3.7 in equation 3.5 and dividing the resulting expression by $\Delta t$ gives

$$\left(F - V_1 - L_1\right) \cdot \left. \right|_{t_{n+\alpha\Delta t}} - \left(\frac{d}{dt} W_1\right) \cdot \left. \right|_{t_{n+\beta\Delta t}} = \frac{d}{dt} W_1 \quad \text{.........} 3.8$$

As $\Delta t$ -> 0 equation 3.8 reduces to

$$\left(F - V_1 - L_1\right) \cdot \left. \right|_{t_n} - \left(\frac{d}{dt} W_1\right) \cdot \left. \right|_{t_n} \quad \text{.........} 3.9$$

Since $t_n$ was selected arbitrarily in the time domain $t_n > 0$, equation 3.9 holds for all t>0, and thus it becomes

$$\left(F - V_1 - L_1\right) \cdot \left. \right|_{t_n} = \frac{d}{dt} W_1 \quad (t>0) \quad \text{.........} 3.10$$

Using equation 3.10, The corresponding differential equation representing component material balance on the solute over the time period $t_n$ to $t_{n+1}$

$$FX_F - L_1 X_1 = \frac{d}{dt}\left(W_1 X_1\right) \quad \text{.........} 3.11$$

where $X_F$= concentration of solute in the feed in weight percent

and X = concentration of solute in the liquor.

## ENTHALPY BALANCE

similarly, the Integral - difference equation representing an energy balance on th

thick liquor is given by

$$\int_{t_n}^{t_{n+1}} \left(F\,h(T_f,X_f) + Q_1 - V_1 \cdot H(T_1) - L_1 \cdot h(T_1,X_1)\right) dt = \left(W_1 \cdot h(T_1,X_1)\right) \cdot \Big|_{t_{n+1}} - \left(W_1 \cdot h(T_1,X_1)\right) \cdot \Big|_{t_n}$$

...3.1

and the corresponding differential equation representing an energy balance

$$F \cdot h \cdot (T_1,X_1) + Q_1 - V_1 \cdot H(T_1) - L_1 \cdot h(T_1,X_1) = \frac{d}{dt}\left(W_1 \cdot h(T_1 \cdot X_1)\right)$$

$t > 0$...................................3.13

Where T = temperature(oC), h = enthalpy of the liquid (kJ/kg)

H = enthalpy of vapour (KJ/Kg), and Q = heat load or heat input (Watt or KiloWatt)

## HEAT TRANSFER RATE

In the steam chest, the steam hold up is negligible relative to the other hold ups of the

system, hence, the enthalpy balance on the steam is given by

$$\int_{t_n}^{t_{n+1}} \left(V_s \cdot H_s - V_s \cdot h_s - Q\right) dt = 0$$ ...........................................................................................3.14

where subscript S = steam

since the integral is equal to zero for any choice of the upper and lower limit, it follows that the integrand is identically equal to zero for all t in the time domain o interest, that is

$$V_s \cdot (H_s - h_s) - Q = 0 \quad t > 0 \qquad \text{......} 3.15$$

or

$$V_s \cdot \lambda_s = Q_1 \qquad \text{......} 3.16$$

where $\lambda_s = H_S - h_S$ ...... 3.17

where $\lambda$ = latent heat of vaporization

Also, since the hold up of energy by the metal through which the energy is regarded as negligible, therefore the heat transfer rate can be expressed as

$$Q_1 = U_1 \cdot A_1 \cdot (T_s - T_1) \qquad \text{......} 3.18$$

or

$$U_1 \cdot A_1 (T_s - T_1) - Q_1 = 0 \qquad \text{......} 3.19$$

Equation 3.18 is applicable for each t in the time interval $t_n \leq t \leq t_{n+1}$ under consideration. Equation 3.16 may be used to eliminate Qwherver it appears above. Hence the set of equation to describe completely the set of unsteady state operation of a single effect evaporator are as follows:

## Total Mass Balance:

$$F - V_1 - L_1 = \frac{d}{dt} W1 \qquad \text{......} 3.20$$

Mass of the liquid W can be expressed as the product of liquid level K, cross-sectional area of vessel A, and the density of liquid $\rho$.Therefore equation 3.20 can be expressed as:

$$F - V_1 - L_1 = \rho \cdot A_1 \cdot \frac{d}{dt} K_1 \tag{3.21}$$

If $\rho A = C$, therefore the equation of the total mass balance will become

$$F - V_1 - L_1 = C \cdot \left(\frac{d}{dt} K_1\right) \tag{3.22}$$

Where the unit of K = m, A = $m^2$, $\rho$ = $kg/m^3$ and C = kg/m

## Component mass Balance

$$F \cdot X_F - L_1 \cdot X_1 = \frac{d}{dt}(W_1 \cdot X_1) \tag{3.23}$$

## Enthalpy Balance

$$F \cdot h \cdot (T_f, X_f) + V_s \cdot \lambda_s - [V_1 \cdot H_1 \cdot (T_1)] - L_1 \cdot h \cdot (T_1, X_1) = \frac{d}{dt}[W_1 \cdot h \cdot (X_1, T_1)] \tag{3.24}$$

## Heat Transfer rate

$$U_1 \cdot A_1 \cdot (T_s - T_1) - V_s \cdot \lambda_s = 0 \tag{3.25}$$

Equation describing multiple effect evaporator system can be formulated in a manner analogous to those shown for the single effect system.

For instance for the second effect.

## Total Mass Balance

$$\left(L_1 - V_2 - L_2\right) = \frac{d}{dt} W_2 \quad \ldots\ldots 3.26$$

## Component Mass Balance

$$L_1 \cdot X_1 - L_2 \cdot X_2 = \frac{d}{dt}\left(W_2 \cdot X_2\right) \quad \ldots\ldots 3.27$$

## Enthalpy Balance

$$L_1 \cdot h \cdot \left(T_1, X_1\right) + V_1 \cdot \left(H\left(T_1\right) - h\left(T_1\right)\right) - \left[V_2 H\left(T_2\right) - L_2 \cdot h_2\left(T_2, X_2\right)\right] = \frac{d}{dt}\left(W_2 \cdot h\left(T_2, X_2\right)\right) \quad \ldots\ldots 3.28$$

## Heat Transfer Rate:

$$U_3 \cdot A_3\left(T_2 - T_3\right) = V_1\left(H\left(T_1\right) - h\left(T_1\right)\right) = 0 \quad \ldots\ldots 3.29$$

and for the third effect.

## Total Mass Balance

$$\left(L_2 - V_3 - L_3\right) = \frac{d}{dt} W_3 \quad \ldots\ldots 3.30$$

## Component Mass Balance

$$L_2 \cdot X_2 - L_3 \cdot X_3 = \frac{d}{dt}\left(W_3 \cdot X_3\right) \quad \ldots\ldots 3.31$$

## Enthalpy Balance

$$L_2 \cdot h\left(T_2, X_2\right) + V_2\left(H\left(T_2\right) - h\left(T_2\right)\right) - V_3 \cdot H\left(T_3\right) - L_3 \cdot h\left(T_3, X_3\right) = \frac{d}{dt}\left[W_3 h \cdot \left(T_3, X_3\right)\right] \quad \ldots\ldots 3.32$$

## Heat Transfer rate

$$U_3 \cdot A_3\left(T_2 - T_3\right) - V_2 \cdot \left(H\left(T_2\right) - h\left(T_2\right)\right) = 0 \quad \ldots\ldots 3.33$$

and for the nth effect,

### Total Mass Balance

$$L_{n-1} - V_n - L_n = \frac{d}{dt} W_n \quad \ldots\ldots 3.34$$

### Component Mass Balance

$$L_{n-1} \cdot X_{n-1} - L_n \cdot X_n = \frac{d}{dt}\left(W_n \cdot X_n\right) \quad \ldots\ldots 3.35$$

### Enthalpy Balance

$$L_{n-1} \cdot h\left(T_{n-1}, X_{n-1}\right) + V_{n-1} \cdot \left(H\left(T_{n-1}\right) - h\left(T_{n-1}\right)\right) - \left(V_n \cdot H\left(T_n\right) - L_n \cdot h\left(T_n, X_n\right)\right) = \frac{d}{dt}\left(W_n h\left(T_n, X_n\right)\right)$$

$$\ldots\ldots 3.36$$

### Heat Transfer Rate

$$U_n \cdot A_n \cdot \left(T_{n-1} - T_n\right) - V_{n-1} \cdot \left(H\left(T_{n-1}\right) - h\left(T_{n-1}\right)\right) = 0 \quad \ldots\ldots 3.37$$

Note that to obtain the equaytion for single effect for those representing n effect, for $n = 1$,

that is single effect evaporator system, $L_0 = F$, $X_0 = X_F$, $T_0 = T_F$ and

$H(T_0) - h(T_0) = \lambda_0$. The single effect evaporator system is described by four independent equations. All the variables except seven at $t_{n+1}$ must be fixed.

It is of course suppose that the values of all variables are known from the beginning of the time period under consideration. Problems may be formulated in terms of the values of the variables which are fixed and those which are to be found at $t_{n+1}$ in th following manner.

The above specification correspond to the case where the variables F, $X_f$, $T_f$, $T_s$, $T_1$ W1 and $L_1$ are either controlled or fixed at the prescribed value at time $t_{n+1}$, they may differ from those at at time $t_n$. The overall heat transfer coefficient must be known. Simiarly for the second effect the fixed variables and those to be found at tin $t_{n+1}$ are in the following manner .

Specification: $L_1$, $V_1$, $L_2$, $X_1$, $T_1$, $T_2$, and $W_2$

To find : $V_2$, $h_2$, $X_2$

For N-effect evaporator system.

The specification: $L_{n-1}$ , $L_n$, $V_{n-1}$, $X_{n-1}$, $T_{n-1}$, $T_n$, and $W_n$

To find: $V_n$, $h_n$, and $X_n$

# CHAPTER FOUR

## 4.0 LABORATORY EXPERIMENTS TO VALIDATE THE MATHEMATICAL MODEL

An important aspect of mathematical modelling exercise is the validation of the model. This assists in the determination of the level of accuracy of the model and indeed ensures that the model is a true representation of the real system. The system in question is a double effect evaporator and the equipments used for the experiment is the Armfield Double effect evaporator and a recfractometer. The samples material is a sugar solution.

Six sets of experiments were conducted with varied feed flow rate and steam flow rate while other operation parameters were held constant.

## 4.1 EXPERIMENT

### PRELIMINARY OPERATION

Prior to the commencement of the experimental work, 10 percent w/w concentration of sugar solution was prepared in a vessel and the concentration was determined with the aid of a refractometer.

The condensate and the concentrate tanks (2,4) of the double effect evaporator were drained and the equipment connected to a source of power supply, steam supply, and available cooling water services. The valves V1 and V4 were open while valves V2, V3, V5, C6, C8, C12 and C14 were shut. Then the sample solution was poured into the feed tank (3).

## PROCEDURE

### DOUBLE EFFECT FORWARD FEED OPERATION

After the feed tank was filled with sample solution. The double effect evaporator was set in the forward feed mode by opening the following valves V6, V8, V9, V11 V16, C4, and C10 while valves V7, V10, V12, V13, V14, and V15 were shut.

Next, the feed pump was switched on and valve C13 was fully open to prime the first effect. The feed pre-heater was switched on. When the liquid appeared in flow meter F4, valve C13 was adjusted so as to obtain the desired feed rate. Also valve C2 was opened and adjusted so as to obtained cooling water flows rate on F1 of 40 times that on F4.

Valve C15 was opened and adjusted so as to obtain the desired steam pressure P3. The valve V14 was opened and C11 and C10 were adjusted to obtain a circulation feed rate for the first effect on flow meter F5.

The thermostat control on the feed pre-heater was adjusted so as to maintain temperature T10 and T11 as closed as possible. Valve V13 was open and V9 shut immediately, it was observed that the feed to the second effect was established on flow meter F2.

When liquid appeared in the level vessel of the second effect, the concentrate / recirculation pump was commissioned. Then valve V10 was open and valves C4 and C5 were adjusted so as to obtain the desired circulation rate for the second effect on flow meter F3. At steady state operation, feed rate, concentrated liquor flow rate and steam flow rate were taken at three minutes interval.

Also the samples of concentrated liquor from the first effect and the second effect were taken, as well at three minutes interval and their concentration determined with the aid of a refractometer. The results were tabulated as shown in the next chapter.

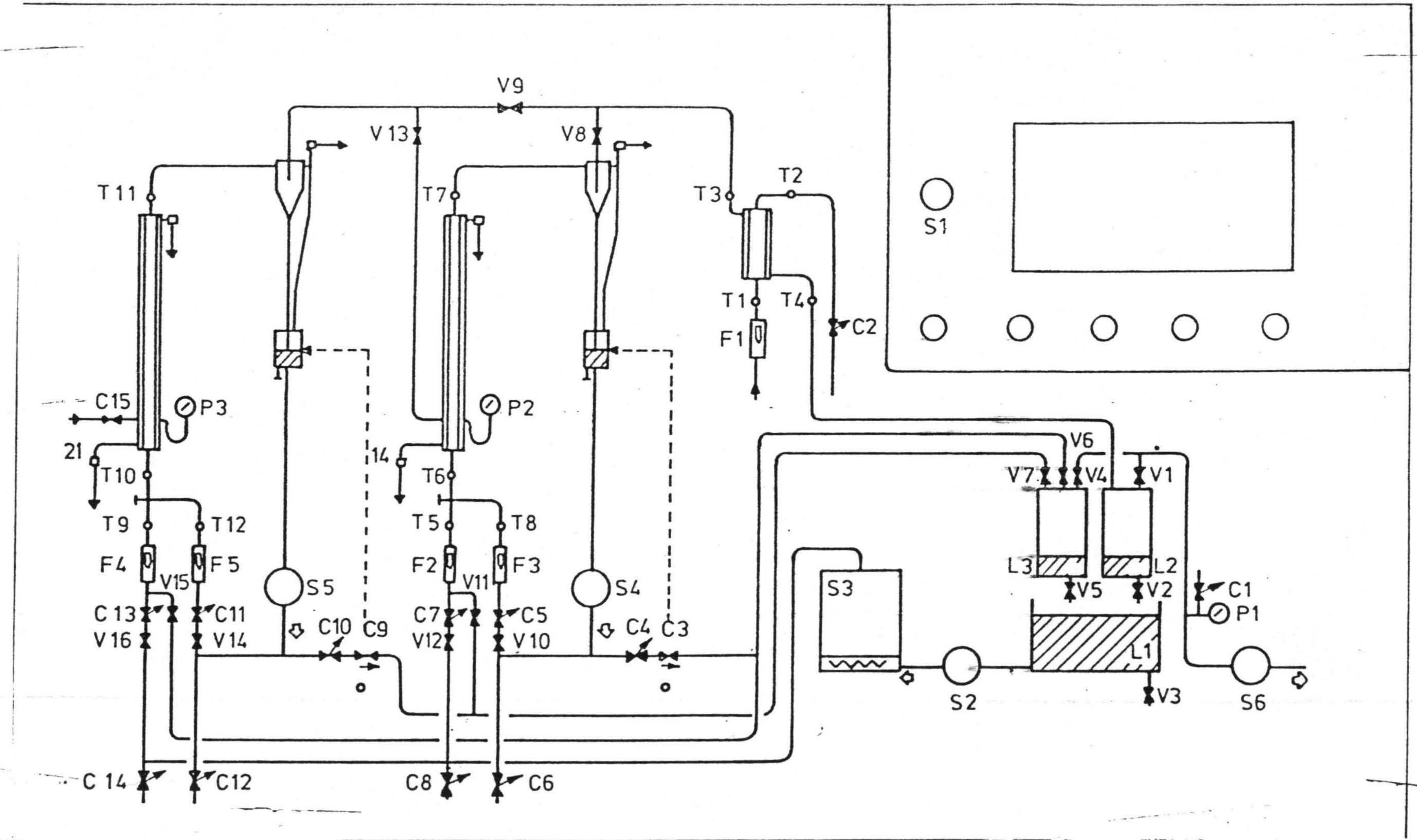

FIGURE 4.0: SCHEMATIC DIAGRAM OF ARMFIELD DOUBLE EFFECT EVAPORATOR

# CHAPTER FIVE

## 5.0 RESULTS

## 5.1 EXPERIMENTAL RESULTS

Table 5.1 to 5.8 present the experimental and simulation result of double effect evaporation of 10.0 wt% sugar solution showing an important parameter namely, the change in concentration of liquor with respect to time. In these tables the feed rate and the steam rate were varied while other operating parameters were held at constant rate. Tables 5.9 and 5.10 present the simulation result of double effect evaporation of 10.0 wt% sugar solution under varied vapour rate. The results show change in concentration, change in enthalpy and the change in mass liquid hold-up with respect to time. The experiment is described in Chapter four

TABLE (5.1): THE RESULT OF DOUBLE EFFECT EVAPORATION OF 10 wt% SUGAR SOLUTION AT:

FEED RATE (F) = 16kg/min

STEAM SUPPLY (S) = 10kg/min

LIQUOR RATE FROM THE FIRST EFFECT (L1) = 9.0kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 3.0kg/min

| TIME | FIRST EFFECT | | SECOND EFFECT | |
|---|---|---|---|---|
| t | EXPERIMENTAL RESULT (x1) | SIMULATED RESULT (x1) | EXPERIMENTAL RESULT (x2) | SIMULATED RESULT (x2) |
| min | wt% | wt% | wt% | wt% |
| 0 | 10 | 10 | 15.43 | 13.82 |
| 3 | 17.37 | 15.76 | 25.54 | 23.84 |
| 6 | 18.88 | 17.26 | 31.83 | 30.22 |
| 9 | 19.25 | 17.64 | 35.91 | 34.3 |
| 12 | 19.36 | 17.74 | 38.51 | 36.89 |
| 15 | 19.38 | 17.77 | 40.18 | 38.55 |
| 18 | 19.39 | 17.77 | 41.21 | 39.61 |
| 21 | 19.4 | 17.77 | 41.9 | 40.28 |
| 24 | 19.41 | 17.78 | 42.23 | 40.71 |
| 27 | 19.41 | 17.78 | 42.6 | 40.98 |
| 30 | 19.41 | 17.78 | 42.76 | 41.15 |

$r_1 = 0.999$ $r_2 = 0.996$

CORRELATION COEFFICIENT = 0.997

TABLE (5.2): THE RESULT OF DOUBLE EFFECT EVAPORATION OF 10 wt% SUGAR SOLUTION AT:

FEED RATE (F) = 17kg/min

STEAM SUPPLY (S) = 10kg/min

LIQUOR RATE FROM THE FIRST EFFECT (L1) = 9.0kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 3.0kg/min

| TIME | FIRST EFFECT | | SECOND EFFECT | |
|---|---|---|---|---|
| t | EXPERIMENTAL RESULT (x1) | SIMULATED RESULT (x1) | EXPERIMENTAL RESULT (x2) | SIMULATED RESULT (x2) |
| min | wt% | wt% | wt% | wt% |
| 0 | 10 | 10 | 16.07 | 14.36 |
| 3 | 18.25 | 16.58 | 26.38 | 24.77 |
| 6 | 20.09 | 18.29 | 33.01 | 31.4 |
| 9 | 20.34 | 18.73 | 37.26 | 35.64 |
| 12 | 20.46 | 18.85 | 39.95 | 38.33 |
| 15 | 20.49 | 18.88 | 41.67 | 40.05 |
| 18 | 20.55 | 18.89 | 42.76 | 41.15 |
| 21 | 20.57 | 18.89 | 43.46 | 41.85 |
| 24 | 20.58 | 18.89 | 43.91 | 42.3 |
| 27 | 20.58 | 18.89 | 44.98 | 42.58 |
| 30 | 20.58 | 18.89 | 44.99 | 42.76 |

$r_1 = 0.999$ $r_2 = 0.999$

CORRELATION COEFFICIENT = 0.999

TABLE (5.3): THE RESULT OF DOUBLE EFFECT EVAPORATION OF 10 wt% SUGAR SOLUTION AT:

FEED RATE (F) = 16kg/min

STEAM SUPPLY (S) = 8kg/min

LIQUOR RATE FROM THE FIRST EFFECT (L1) = 9.0kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 5.0kg/min

| TIME | FIRST EFFECT | | SECOND EFFECT | |
|---|---|---|---|---|
| t | EXPERIMENTAL RESULT (x1) | SIMULATED RESULT (x1) | EXPERIMENTAL RESULT (x2) | SIMULATED RESULT (x2) |
| min | wt% | wt% | wt% | wt% |
| 0 | 10 | 10 | 15.42 | 13.82 |
| 3 | 17.78 | 15.76 | 22.72 | 21.11 |
| 6 | 19.05 | 17.26 | 26.16 | 24.56 |
| 9 | 19.34 | 17.71 | 27.68 | 26.18 |
| 12 | 19.37 | 17.74 | 28.45 | 26.95 |
| 15 | 19.38 | 17.77 | 28.83 | 27.32 |
| 18 | 19.39 | 17.78 | 28.99 | 27.46 |
| 21 | 19.4 | 17.78 | 29.07 | 27.57 |
| 24 | 19.41 | 17.78 | 29.11 | 27.61 |
| 27 | 19.41 | 17.78 | 29.12 | 27.62 |
| 30 | 19.41 | 17.78 | 29.15 | 27.63 |

$r_1 = 0.996$ $r_2 = 0.999$

CORRELATION COEFFICIENT = 0.997

TABLE (5.4): THE RESULT OF DOUBLE EFFECT EVAPORATION OF 10 wt% SUGAR SOLUTION AT:

FEED RATE (F) = 16kg/min

STEAM SUPPLY (S) = 9kg/min

LIQUOR RATE FROM THE FIRST EFFECT (L1) = 8.0kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 3.0kg/min

| TIME | FIRST EFFECT | | SECOND EFFECT | |
|---|---|---|---|---|
| t | EXPERIMENTAL RESULT (x1) | SIMULATED RESULT (x1) | EXPERIMENTAL RESULT (x2) | SIMULATED RESULT (x2) |
| min | wt% | wt% | wt% | wt% |
| 0 | 10 | 10 | 16.01 | 14.51 |
| 3 | 18.59 | 16.99 | 28.38 | 26.78 |
| 6 | 20.6 | 19.09 | 36.11 | 34.6 |
| 9 | 21.23 | 19.73 | 41.08 | 39.59 |
| 12 | 21.43 | 19.92 | 44.17 | 42.77 |
| 15 | 21.48 | 19.98 | 46.2 | 45.52 |
| 18 | 21.5 | 19.99 | 47.4 | 46.09 |
| 21 | 21.55 | 20 | 48.13 | 46.92 |
| 24 | 21.56 | 20 | 48.6 | 47.44 |
| 27 | 21.58 | 20 | 48.82 | 47.78 |
| 30 | 21.58 | 20 | 48.94 | 47.99 |

$r_1 = 0.999$ $r_2 = 0.999$

CORRELATION COEFFICIENT = 0.999

TABLE (5.5): THE RESULT OF DOUBLE EFFECT EVAPORATION OF 10 wt% SUGAR SOLUTION AT:

FEED RATE (F) = 16kg/min

STEAM SUPPLY (S) = 5kg/min

LIQUOR RATE FROM THE FIRST EFFECT (L1) = 10.0kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 6.0kg/min

| TIME | FIRST EFFECT | | SECOND EFFECT | |
|---|---|---|---|---|
| t | EXPERIMENTAL RESULT (x1) | SIMULATED RESULT (x1) | EXPERIMENTAL RESULT (x2) | SIMULATED RESULT (x2) |
| min | wt% | wt% | wt% | wt% |
| 0 | 10 | 10 | 15.26 | 13.64 |
| 3 | 16.58 | 14.88 | 20.35 | 18.64 |
| 6 | 17.35 | 15.75 | 22.37 | 20.67 |
| 9 | 17.58 | 15.94 | 22.7 | 21.5 |
| 12 | 17.64 | 15.99 | 23.55 | 21.84 |
| 15 | 17.73 | 16 | 23.83 | 21.97 |
| 18 | 17.75 | 16 | 24.1 | 22.03 |
| 21 | 18 | 16 | 24.36 | 22.05 |
| 24 | 18 | 16 | 24.6 | 22.06 |
| 27 | 18 | 16 | 24.61 | 22.06 |
| 30 | 18 | 16 | 24.62 | 22.07 |

$r_1 = 0.998$ $r_2 = 0.990$

CORRELATION COEFFICIENT = 0.994

TABLE (5.6): THE RESULT OF DOUBLE EFFECT EVAPORATION OF 10 wt% SUGAR SOLUTION AT:

FEED RATE (F) = 16kg/min

STEAM SUPPLY (S) = 10kg/min

LIQUOR RATE FROM THE FIRST EFFECT (L1) = 10.0kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 6.0kg/min

| TIME | FIRST EFFECT | | SECOND EFFECT | |
|---|---|---|---|---|
| t | EXPERIMENTAL RESULT (x1) | SIMULATED RESULT (x1) | EXPERIMENTAL RESULT (x2) | SIMULATED RESULT (x2) |
| min | wt% | wt% | wt% | wt% |
| 0 | 10 | 10 | 15.42 | 13.82 |
| 3 | 17.56 | 15.76 | 22.92 | 21.11 |
| 6 | 19.05 | 17.26 | 26.27 | 24.56 |
| 9 | 19.36 | 17.64 | 27.89 | 26.18 |
| 12 | 19.44 | 17.74 | 28.74 | 26.95 |
| 15 | 19.57 | 17.77 | 29.01 | 27.32 |
| 18 | 19.59 | 17.78 | 29.09 | 27.49 |
| 21 | 19.6 | 17.78 | 29.16 | 27.57 |
| 24 | 19.62 | 17.78 | 29.21 | 27.61 |
| 27 | 19.62 | 17.78 | 29.24 | 27.62 |
| 30 | 19.62 | 17.78 | 29.26 | 27.63 |

$r_1 = 0.999$ $r_2 = 0.999$

CORRELATION COEFFICIENT = 0.999

TABLE5.7: SIMULATED RESULT OF DOUBLE EFFECT EVAPORATION OF 10wt% SUGAR SOLUTION AT:

FEED RATE (F) = 16kg/min

STEAM SUPPLY = 10kg/min

LIQUOR RATE FOR THE FIRST EFFECT (L1) = 9kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 4kg/min

VAPOUR RATE FROM THE FIRST EFFECT (V1) = 4kg/min

VAPOUR RATE FROM THE SECOND EFFECT (V2) = 3kg/min

| TIME | FIRST EFFECT | | | SECOND EFFECT | | |
|---|---|---|---|---|---|---|
| t | conc. | enthalpy | liq.Hold up | conc. | enthalpy | liq.Hold up |
| min | x1(wt%) | h1(kJ/kg) | w1(kg) | x2(wt%) | h2(kJ/kg) | w2(kg) |
| 0 | 10 | 449 | 20 | 13.82 | 377 | 20 |
| 3 | 15.75 | 1514.5 | 20.23 | 21.59 | 1526.5 | 20.15 |
| 6 | 17.24 | 1791.4 | 20.45 | 25.85 | 2155.4 | 20.3 |
| 9 | 17.63 | 1864.5 | 20.68 | 28.18 | 2500.9 | 20.45 |
| 12 | 17.74 | 1884 | 20.9 | 29.47 | 2691.7 | 20.6 |
| 15 | 17.77 | 1889.3 | 21.13 | 30.19 | 2797.4 | 20.75 |
| 18 | 17.78 | 1890.8 | 21.35 | 30.58 | 2856.2 | 20.9 |
| 21 | 17.78 | 1891.2 | 21.58 | 30.81 | 2889.1 | 21.05 |
| 24 | 17.78 | 1891.3 | 21.8 | 30.93 | 2907.6 | 21.2 |
| 27 | 17.78 | 1891.3 | 22.03 | 31 | 2918 | 21.35 |
| 30 | 17.78 | 1891.3 | 22.25 | 31.04 | 2923.9 | 21.5 |

TABLE5.8: SIMULATED RESULT OF DOUBLE EFFECT EVAPORATION OF 10wt% SUGAR SOLUTION AT:

FEED RATE (F) = 16kg/min

STEAM SUPPLY = 10kg/min

LIQUOR RATE FOR THE FIRST EFFECT (L1) = 9kg/min

LIQUOR RATE FROM THE SECOND EFFECT (L2) = 4kg/min

VAPOUR RATE FROM THE FIRST EFFECT (V1) = 8kg/min

VAPOUR RATE FROM THE SECOND EFFECT (V2) = 7kg/min

| TIME | FIRST EFFECT | | | SECOND EFFECT | | |
|---|---|---|---|---|---|---|
| t | conc. | enthalpy | liq.Holdup | conc. | enthalpy | liq.Hold up |
| min | x1(wt%) | h1(kJ/kg) | w1(kg) | x2(wt%) | h2(kJ/kg) | w2(kg) |
| 0 | 10 | 449 | 20 | 13.83 | 377 | 20 |
| 3 | 15.77 | 632.94 | 19.93 | 21.64 | 582.51 | 19.85 |
| 6 | 17.26 | 680.6 | 19.85 | 25.94 | 647.5 | 19.7 |
| 9 | 17.65 | 692.88 | 19.78 | 28.3 | 757.69 | 19.55 |
| 12 | 17.74 | 696.03 | 19.7 | 29.58 | 791.53 | 19.4 |
| 15 | 17.77 | 696.84 | 19.63 | 30.28 | 809.9 | 19.25 |
| 18 | 17.78 | 697.04 | 19.55 | 30.66 | 819.83 | 19.1 |
| 21 | 17.78 | 697.1 | 19.48 | 30.86 | 825.17 | 18.95 |
| 24 | 17.78 | 697.11 | 19.4 | 30.97 | 828.03 | 18.8 |
| 27 | 17.78 | 697.11 | 19.33 | 31.03 | 829.55 | 18.65 |
| 30 | 17.78 | 697.11 | 19.25 | 31.06 | 830.36 | 18.5 |

# CHAPTER SIX

## 6.1.0 DISCUSSION OF RESULT

The aims of this research work is the formulation of a set of mathematical model equations which predicts the dynamic behaviour of a double effect evaporator system and a numerical technique for investigating the transient behaviour of it's operating parameters, as well as it's application to evaporators.

In table 5.1 where thin liquor feed rate of 16kg/min was used, the experimental and the simulated concentration change from 10.0wt% to 42.76wt% and 41.15wt% respectively. In table 5.2, when the feed rate of thin liquor is 17kg/min, concentration change was from 10.0 wt% to 44.99 wt% and 42.76 wt% respectively. These sets of results shows that an increase in feed rate of thin liquor obviously leads to a corresponding increase in concentration of thick liquor obtained. This trend agrees with the theory which states that change in feed rate is directly proportional to change in concentration of thick liquor (Perry 1988).

The results in table 5.3 show that a feed rate of 16kg/min into the first effect resulted in a discharge rate of 5kg/min in the second effect. The change in concentration for the experimental and the simulated result are from 10 wt% to 29.15 wt% and 27.6 wt% respectively.

Similarly, from table 5.4, a feed rate of 16kg/min into the first effect resulted in a discharge rate of 3kg/min in the second effect. The corresponding change in concentration for the experimental and simulated results are 10.0 wt% to 48.94 wt% and 47.99 wt% respectively. For the first set of results the difference between the feed rate into the first effect and discharge rate from the second effect is 11.0 kg/min while it is 13kg/min for the second sets of results.

In theory, the second set of results is expected to produce greater change in concentration.

This is corroborated by the results, because the change in concentration for the first sets of experimental and simulated results are 19.5 wt % and 17.63 wt% respectively. While the experimental and simulated result for concentration change for the second set of results are 38.94 wt% and 37.91 wt% respectively.

When the steam rate into the evaporator is 5kg/min as shown in table 5.5. The concentration for experimental and simulated results changed from 10.0 wt% to 24.62 wt% and 22.07 wt% respectively. On increasing the steam supply to 10 kg/min as shown in Table 5.6, the concentration for the experimental and simulated result changed from 10.0 wt% to 29.26 wt% and the 27.63 wt% respectively. These set of results shows that a significant increase in steam rate will lead to an increase in the concentration of thick liquor. The result in Table 5.1 to 5.6 implies that the feed rate of thin liquor and steam supply can be manipulated to obtain a desired thick liquor concentration.

In Table 5.7, when the vapour rate was reduced from the normal flow rate of 7 kg/min to 4 kg/min in the first effect and 5 kg/min to 3 kg/min in the second effect, the simulated result shows an increase in the mass liquid hold-up from 20 kg to 22kg in the first effect and 20 kg to 21.5 kg in the second effect. Also, in Table 5.8, an increase in the vapour rate from the normal 7 kg/min to 8 kg/min in the first effect and 5 kg/min to 7 kg/min in the second effect, give rise to a fall in mass liquid hold-up from 20 kg to 19.25 kg in the first effect and 20kg to 18.5 kg in the second effect respectively.

The result in this table imply that, a reduction in vapour rate below the normal rate will lead to an increase in mass liquid hold-up in the evaporator while an increase in vapour rate above the normal rate will result in fall in mass liquid hold-up in the evaporator. This observation is possible under abnormal operating condition when there is an obstruction to vapour flow or there is liquid carry over along the vapour line from effect to effect in the evaporator system (Vincent et. al. 1979).

Also in Table 5.7, the simulated result shows that enthalpy of liquor in the first effect and the second effect increased from 449 kJ/kg to 1891.3 kJ/kg and 377 kJ/kg to 2923.9 kJ/kg respectively. Similarly, in Table 5.8, the enthalpy of liquor in the first effect increased from 449 kJ/kg to 697.11 kJ/kg and from 377 kJ/kg to 830.36 kJ/kg in the second effect. These sets of results above show that change in enthalpy is directly proportional to change in concentration (Perry and Chilton, 1988).

The computed correlation coefficients for the experimental and simulated results as shown in Table 5.1 to 5.6 are 0.997, 0.999, 0.997, 0.999, 0.995 and 0.999. This gives an average value of 0.998. This range of correlation coefficient suggests a strong positive relationship between the two sets of results. Thus the developed model adequately represents the real process.

## 6.20 CONCLUSION

The dynamic behaviour of a double effect evaporator system was modelled. The comparison of the experimental result with the simulated result shows a negligible difference as confirmed by the correlation coefficient.

From the simulated results, it was found that the concentration of thick liquor obtained is a function of two operating variable namely; the feed rate of thin liquor into the evaporator system, and steam supplied to the evaporator system. Also change in the enthalpy is a function of concentration of liquor and steam flow rate. This phenomenon is also true for a real process.

Therefore, it can be concluded that the proposed mathematical model truly represents and can be used for the prediction of the transient behaviour of a real process.

## 6.3 RECOMMENDATION

(1) The accuracy of this model can still be tested and improved upon by using data obtained from sugar industries salt industries, paper industries etc. to simulate the model equation.

(2) Instead of investigating change in feed rate and change in steam rate and its effect. The investigation can be extended to change in concentration of feed and temperature and its effect.

(3) Incorporating a condenser heat load, temperature and pressure distribution across an evaporator system, in addition to this model to predict the transient behaviour of double effect evaporator can develop this work further.

(4) Further work should be conducted to develop a model, which will accommodate the transient behaviour of more than double effect evaporator system.

(5) Research should be conducted to develop a reliable simulation and optimisation computer package for multiple effect evaporation processes.

## NOMENCLATURE

$F$ = Feed, mass/time (kg/s) or kg/min

$L$ = Liquid rate, mass/time (kg/min) or (kg/s)

$T$ = Temperature in $^{O}C$

$X$ or $X_L$ = solute concentration in weight percent

$S_f$ or $S$ = steam flow rate (kg/min) or (kg/s)

$P$ = Pressure (Bar)

$W$ = mass holdup of liquid (kg)

$Q$ = Heat transfer rate (kW)

$h$ or $h_L$ = enthalpy of liquid (kJ/kg)

$H$ = enthalpy of vapour (kJ/kg)

$V$ = Vapour flow rate in (kg/min) or (kg/s)

$U$ = overall heat transfer coefficient (kJ/g $^{0}C$)

$A$ = Area of heat transfer in $m^2$

$\lambda$ = Latent heat of vaporization

$t$ = time (minutes or seconds or hours)

$N$ = Number of effects (1,2,3,4…)

$T_f$ = Temperature of the feed ($^{O}C$)

$X_f$ = Feed composition (wt %)

$H_f$ = enthalpy of feed in (kJ/Kg)

$H_N$ = enthalpy of vapour in N effect (kJ/kg)

$h_N$ = enthalpy of liquid in N effect (kJ/kg)

$h_c$ = enthalpy of condensate in (kJ/kg)

$h_s$ = enthalpy of steam in (kJ/kg)

$W_f$ = mass of liquid fed into the evaporator, (kg)

$W_E$ = mass of water evaporated, (kg)

$W_C$ = mass of water in concentrate, (kg)

$Q_1$ = mass of steam condensed in first effect (kg)

$H_E$ = Enthalpy of vapour leaving first effect (kJ/kg)

$H_s$ = Enthalpy of steam leaving the evaporator (kJ/kg)

hc = enthalpy of concentrate (kJ/kg)

hc = enthalpy of condensate leaving the evaporator, (kJ/kg)

V = volume of primary side stream, $m^3$

$\rho$ = density of primary side stream, ($kg/m^3$)

$w_{11}$ = input flow of primary stream, (kg/s)

$w_{12}$ = output flow of primary stream, (kg/s)

$w_{21}$ = input of primary thin liquor, (kg/s)

$w_{22}$ = outflow of secondary thick liquor (kg/s)

$w_{20}$ = outflow of secondary thick steam (kg/s)

$g_{11}$ = enthalpy of inlet primary steam (kJ/kg)

$g_{12}$ = enthalpy of primary condensate (kJ/kg)

$g_{21}$ = enthalpy of incoming thin liquor (secondary side) (kJ/kg)

$g_2$ = enthalpy of secondary thick liquor

$Q_1$ = heat flow from primary side to metal tube, kJ/s)

$Q_{o1}$ = heat loss from primary side, (kJ/ s)

$Q_2$ = heat flow from metal to secondary side, (kJ/s)

$Q_{o2}$ = heat loss from secondary side, (kJ/S)

$H_1$ = heat transfer coefficient between primary side and metal, (kJ/S$^{O}$C)

$H_2$ = heat transfer coefficient between metal and secondary side, (kJ/S$^{O}$C)

$\lambda_1$ = heat of evaporation of primary steam (kJ/kg)

$W_v$ = mass of metal in the tube, (kg)

$W_2$ = mass of secondary liquid, (kg)

$C_v$ = specific heat of metal, (kJ/kg$^{O}$C)

$C_{21}$ = concentration of non – volatile components in the secondary feed flow, (kg/kg)

$C_2$ = concentration of non – volatile components in the secondary side, (kg/kg)

A = area of heat transfer of tubes, ($m^{2)}$

$L_2$ = heat of solution of non – volatile components on the secondary side, (kJ/kg.s)

C = calibration constant of tank

$\Delta$ L = change in level of liquid in tank (m)

$B_1$ , $B_2$ = Bottom product flow rate, (kg/s)

$C_f$ , $C_1$ , $C_2$ = Solute concentration, (wt%)

$D_1$, $D_2$ = Over head vapour flow rate (kg/s)

$L_1$ , $L_2$ = Heat loss from first and second effect (kW)

$H_{1v}$ , $H_{2v}$ = Vapour enthalpy in first and second effect (kJ/kg)

$\Delta$ = Change

$\lambda_S$ = Latent heat of vaporization of steam. (kJ/kg)

$\Delta$ tm = Log. Mean temperature difference ($^{O}$C)

## APPENDIX 1.0

## POLYMATH SIMULATION OF DYNAMIC MODELLING OF DOUBLE EFFECT EVAPORATOR

### KEY TO SYMBOLS

F = Feed rate, kg/min

$M_N$ = mass liquid holdup in N effects, kg

$X_N$ = solute concentration in N effect, wt%

$V_N$ = Vapour flow rate out of N effects, kg/min

$L_N$ = liquid flow rate out of N effects, kg/min

$X_F$ = solute concentration in feed

$h_N$ = Liquid enthalpy in N effect, kJ/kg

$H_N$ = Vapour enthalpy in N effect, kJ/kg

L = latent heat of vaporization of steam, kJ/kg

S = steam flow rate, kg/min

t = time, (min, sec, hr)

N = number of effect (1,2,3,4,............)

$t_0$ = initial value of time

$h_{N0}$ = initial liquid enthalpy in N-effect.

$X_{N0}$ = initial solute concentration in N-effect

$M_{N0}$ = initial liquid mass holdup in N-effect

$T_F$ = final time

## FIRST EFFECT EQUATIONS

d(h1)/d(t)=((F*hf)+(L*S)-(V1*H1)-(L1*h1))/M1

d(X1)/d(t)=((F*Xf)-(L1*X1))/M1

d(M1)/d(t)=F-V1-L1

V1=F-L1

INPUT VARIABLES: F, L1, S, HF, Hf, H1, h1, L, $X_F$, $M_{10}$, $X_{10}$, $t_0$, and $X_F$

## SECOND EFFECT EQUATIONS

d(h2)/d(t)=((L1*h1)+V1*(H1-h1)-(V2*H2)-(L2*h2))/M2

d(X2)/d(t)=((L1*X1)-(L2*X2))/M2

d(M2)/d(t)=L1-V2-L2

V2=L1-L2

INPUT VARIABLES: F, L1, L2, h1, H1, H2, X1, $t_0$, $h_{20}$, $X_{20}$, $M_{20}$, and $t_F$

```
DOUBLE EFFECT EVAPORATOR1
d(h1)/d(t)=((F*hf)+(L*S)-(V1*H1)-(L1*h1))/M1
d(X1)/d(t)=((F*Xf)-(L1*X1))/M1
d(M1)/d(t)=F-V1-L1
F=16
L1=9
S=8
hf=355
L=2209
H1=2687
V1=F-L1
Xf=10
t(0)=0, h1(0)=49, X1(0)= 0, M1(0)=20
t(f)=30
```

```
DOUBLE EFFECT EVAPORATOR 2
d(h2)/d(t)=((L1*h1)+V1*(H1-h1)-(V2*H2)-(L2*h2))/M2
d(X2)/d(t)=((L1*X1)-(L2*X2))/M2
d(M2)/d(t)=L1-V2-L2
L1=9
L2=5
V2=L1-L2
h1=449
V1=7
H1=2687
H2=2660
X1=13.82
t(0)=0, h2(0)=377, X2(0)=13.82, M2(0)=20
t(f)=30
```

CALCULATION OF CORRELATION COEFFICIENT

Correlation coefficient r is expressed mathematically as

$$r = \frac{n\Sigma xy-\Sigma x\Sigma y}{\sqrt{[n\Sigma x^2-(\Sigma x)^2][n\Sigma y^2-(\Sigma y)^2]}}$$

where n = number of sample data or runs of experiments

x and y are the variables whose relationship is being determined

Σ = summation symbol

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