DESIGN OF A PROCESS PLANT PRODUCING 10 TONNES OF CAUSTIC POTASH FROM COCOA POD HUSK PER DAY

BY

ONIBON MOSES KAYODE
(REG NO. 99/8322EH)

IN PARTIAL FULFILMENT OF THE REQUIREMENT FOR THE AWARD OF DEGREE OF BACHELOR OF ENGINEERING (CHEMICAL ENGINEERING)

DEPARTMENT OF CHEMICAL ENGINEERING
FEDERAL UNIVERSITY OF TECHNOLOGY
MINNA NIGER STATE.
NIGERIA.

NOVEMBER 2005

DECLARATION

I, ONIBON MOSES KAYODE, declare that this design work has been done by me and that it has not been presented either in part or in whole for the award of bachelor of engineering any where. All literature cited have been duly acknowledged in the reference.

ONIBON MOSES KAYODE

SIGNATURE AND DATE

CERTIFICATION

This is to certify that this design project titled "design of a process plant producing 10 tonnes of caustic potash from cocoa pod husk per day" was done by ***Onibon Moses Kayode***, under the supervision of Prof K.R. Onifade and submitted to the department of chemical engineering, Federal University of Technology, Minna in partial fulfillment of the requirement for the award of Bachelor of Engineering (B.Eng) degree in Chemical engineering.

Prof K.R. Onifade (Design supervisor)	Signature and date
Dr. Aberuagba. F (H.O.D. Chem. Engineering Dept)	Signature and date
External Examiner	Signature and date

DEDICATION

To the glory of God and honour of humanity. I dedicate this work to all my loved once. I love you all.

ACKNOWLEDGEMENT

Thank you Jesus for giving the strength to carry on you've being my God in the good times and in the bad times that is why you are God.

I will like to say thank you to the department father and my design supervisor, Prof K.R. Onifade for his fatherly advice, it went a long way helping us.

To my parents Overseer and Deaconess Onibon, I don't need a soothsayer to tell me you are my God sent parents. Thank you a million times.

My siblings Tosin, Bisi, Mide, Philip thank you guys the lord is taking us to greater heights.

To my cousins Ola, Barr Isaac Oginni, Dele, Tomi, Tobi thank you all for been there.

Grandma, mama you prayers are answered oooo.

To my friends, Oluwaseun Omoare, Solo, Siraj, Aewaii, Fulani, nosa, musty, Uwanna, Idi Baba, Sayje, Deen, Elderluaye, Rizo, you guys are worth being called friends.

To all our Chimos, Ebi, Ronnie, Aisha, Enebaby, Ummi, Shaddy but most especially you Bukkie Adelusi thank you for your love and care.

To my classmates you are worth been called classmates and my political allies we shall storm the world by surprise.

And to those I have not mentioned your names your names cant just contain this page cause it is to big in my heart thank you all and most especially you.

ABSTRACT

This project was carried out to design plant for the production KOH plant with a capacity of 10 tonnes per day, in which the Reactor, Evaporator and Filter were the major units designed. Economic Analysis of the project based on 95% plant operating time was also carried out using the cost index method. The project was found to have a pay back period of 10 years, a rate of return of 10% and a Net present value of $50, 896.

TABLE OF CONTENTS

Title page - - - - - - - - - - i
Acknowledgement - - - - - - - - - ii
Certification- - - - - - - - - iii
Dedication - - - - - - - - - - iv
Abstract - - - - - - - - - - v
Table of content - - - - - - - - - vi

CHAPTER ONE

1.0 Introduction - - - - - - - - - 1
1.1 Objectives and aims - - - - - - 3

CHAPTER TWO

2.0 Literature survey- - - - - - - - - 4
2.1 Strength determination in crude potash- - - - 4
2.2 Preparation of caustic potash- - - - - - 6
2.2.1 Effect of caustic potash on the environ- - - - 8
2.3 General principle of leaching operation - - - 11
2.3.1 Process of leaching operation - - - - - 12
2.3.2 Mass transfer in leaching operation - - - - 13
2.3.3 Factors influencing rate of extraction - - - - 15

CHAPTER THREE

3.0 Material balance - - - - - - - 18
3.1 Basis - - - - - - - - - - 18
3.2 Assumptions - - - - - - - - 18
3.3 Material balance of the heater - - - - - 19
3.4 Material balance of the reactor - - - - - 23
3.5 Material balance of the filter - - - - - 27
3.6 Material balance of the evaporator - - - - 30
3.7 Summary of material balance - - - - - 34

CHAPTER FOUR

4.0 Energy balance - - - - - - - 36
4.1 Energy balance of the heater - - - - - - 36
4.2 Energy balance of the reactor - - - - - - 38
4.3 Energy balance of the filter - - - - - - 40
4.4 Energy balance of the evaporator - - - - - 41

CHAPTER FIVE

5.0 Diagram of the flow sheet - - - - - - 43

CHAPTER SIX

6.0 Equipment design - - - - - - - - 44
6.1 Design of a triple forward feed evaporator for the concentration of koh solution - - - - - - - - - 44
6.2 Reactor mixer design - - - - - - - 57
6.3 Filter design- - - - - - - - 63

CHAPTER SEVEN

7.0 Reactor equipment optimization - - - - - 67
7.1 Types of flow reactors - - - - - - - 67
7.1.1 CSTR- Continuous stirred tank reactor- - - - 67
7.1.2 Plug flow reactor (PFR) - - - - - - - 67
7.2 Kinetic experiment for acetone cracking - - - 68
7.3 Reactor volume - - - - - - - - 68
7.4 Determination of activation energy, E - - - - - 68
7.4.1 Arrhenius method - - - - - - - - 69
7.4.2 Shortcut method - - - - - - - - 69
7.5 Temperature profile - - - - - - - 70

CHAPTER EIGHT

8.0 Safety and quality control - - - - - - 74
8.1 Hazards - - - - - - - - - 74
8.2 General safety measures - - - - - - - 75

8.3 Material inventory - - - - - - - - 76
8.4 Equipment inventory - - - - - - - 76
8.6 General quality control measures - - - - - 77

CHAPTER NINE

9.0 Process control and instrumentation - - - - - 80
9.1 Introduction - - - - - - - - - 80
9.2 Instrumentation and control objectives - - - - 81

CHAPTER TEN

10.0 Equipmental acceptability- - - - - - - 84
10.1 Identification of possible pollution - - - - - 84
10.2 Treatment of possible pollutants- - - - - - 84
10.3 Waste water treatment - - - - - - - 85
10.4 Thermal pollution control - - - - - - 85

CHAPTER ELEVEN

11.0 Plant start up and shut down - - - - - - 86
11.1 Start up - - - - - - - - - 86
11.2 Shut down - - - - - - - - - 86

CHAPTER TWELVE

12.0 Plant layout - - - - - - - - - 87
12.1 Building layout - - - - - - - - 88

CHAPTER THIRTEEN

13.0 Economic analysis and project evaluation - - - 89
13.1 Introduction - - - - - - - - - 89
13.2 Purchased equipment cost - - - - - - 90
13.3 Fixed capital investment - - - - - - - 92
13.4 Operating labour - - - - - - - - 94
13.5 Utilities - - - - - - - - - 95
13.6 Raw material and profits - - - - - - - 97

13.7 Financial summary of the project - - - - 98

CHAPTER FOURTEEN

14.0 Conclusion - - - - - - - - - 99

REFERENCES - - - - - - - - - 100

CHAPTER ONE

1.0 INTRODUCTION

Potassium hydroxide also known as caustic potash, potash lye, potassa, potassium hydrate, and lye. It is a white, highly deliquescent caustic solid, which is marketed in several forms, including pellets, flakes, sticks, lumps, and powders. Food grade potassium hydroxide is obtained commercially from the electrolysis of potassium chloride solution in the presence of a porous diaphragm. The reaction can be characterized as follows:

$$KCl + H_2O \text{-------} HCl + KOH$$

Generally, KOH is considered a by-product of hydrochloric acid and chlorine manufacturing (Curlin, Bommarju, and Hansson, 1991).

Its main uses in food processing include use as a direct food additive, formulation aid, pH adjuster, cleaning agent, stabilizer, thickener, and poultry scald agent. It is used in dairy products, baked goods, cocoa, fruits, vegetables, soft drinks, and poultry. Among the main foods that use KOH are: chicken, cocoa, coloring agents, ice cream, and black olives (Ash and Ash, 1995).

Non-food uses include: soap manufacture; electroplating; printing; as a mordant for wood; as a highly reactive source of potassium in a wide variety of industrial chemical syntheses and chemical analyses; in veterinary medicine as a caustic used in disbudding calves horns and in aqueous solution to dissolve scales and hair in skin scrapings; manufacture of cleansers; in wart removal and as a 2.5% solution in glycerol as a cuticle solvent. This type of compound is also used in washing powders, some denture cleaners, some non-phosphate "ecology" detergents, and drain-pipe cleaners (Patnaik, 1992,). The substance is "prohibited for use in lye peeling of fruits and vegetables by a petitioner in 1995, However it is permitted for all other FDA permitted uses, which include as a direct food additive, formulation aid, pH adjuster, cleaning agent, stabilizer, thickener, and poultry scald agent. Original concerns regarding lye peeling included the environmental effects of the waste products, and that mechanical or non-chemical alternatives were available for most fruits and vegetables. The stone fruit (peaches, nectarines, and apricots) do not appear to currently have alternative methods available on a commercial scale to achieve peeling without the use of caustic substances.

The reviewers agree that the substance as used commercially is synthetic, although one points out that it may also be naturally produced and has had historical food use. Two out of three reviewers agree with the petitioner that that this annotation unfairly restricts certain types of operations, and find the environmental affects can be mitigated with the use of good wastewater management practices. The third reviewer finds that the principle of minimizing the use of synthetics should be considered more fundamental than the need for a particular form of a product, and is concerned about lack of international acceptance of this material. This reviewer also believes that prohibitions on products and processes will drive innovation and invention for the development of alternative techniques.

To buttress the foregoing discussion on this introductory aspect potassium hydroxide is a strong base and is alkaline in solution. It is highly corrosive. Caustic peeling is based on the differential solubilization of the cell and tissue constituents. Pectic substances in the middle lamella are particularly soluble. It is in aqueous solution. KOH is used with caramel, annatto, turmeric (Ash and Ash, 1995), and soap. Processors will often combine a number of alkali buffering agents. After taking a close look at potassium hydroxide generally i.e its formation from various sources, its uses and characteristics; this design is specifically to produce potassium hydroxide (KOH) caustic potash from cocoa pod husk.

1.1 OBJECTIVES OF THE PROJECT

The objectives of this project are as follows:

(1) To design a plant that will have a capacity production rate of 10 tonnes per day of caustic potash from cocoa pod husk.

(2) To calculate the material balance values of the components across the individual units.

(3) To carry out the energy balance calculations of the components across the individual units.

(4) Detail design calculations of the major reactor mixer and evaporator equipments used in the plant.

(5) To prepare a block and flow diagrams of the plant.

(6) To carry out detail cost estimation of the plant using plant cost index method.

CHAPTER TWO

2.0 LITERATURE SURVEY

2.1.0 STRENGTH DETERMINATION OF CRUDE POTASSIUM HYDROXIDE IN ASH

Crude caustic from different ash source has different strength/ concentration, this may be due to the plant composition, and besides the ashing temperature and duration of ashing go a long way to determine how strong or concentrated crude caustic will be. This is to test if there is difference between the caustic strength from furnace ashing (at 400^0C) and the local method of ashing, which is simply burning of dry plant to ash. And result are shown in chapter four [13].

The procedures for strength determination are as follows:

1) 50g of ash were weighed and poured into a 2000ml beaker and one (1) liters of distilled water was added.
2) The resulting solution was boiled using hot plate with magnetic stirrer at 100^0C with maximum stirring for three hours.
3) The resulting mixture was centrifuged using Shermond centrifuge machine ad the supernatant solution was filtered using filter paper.
4) The residue obtained form filtration was added to the solid residue left in the beaker after centrifuging.

5) The filtrate from step (4) was collected in weighing stainless plate labeled B1.

6) Another one (1) liter of distilled water was added to the beaker that contained the residue. And step (2) was carried out again.

7) While step (6) was going on, the filtrate from step (5) was boiled off until a constant weight of the stainless plate was obtained (B1).

8) Step 3-5 were carried out on a mixture obtained in step (6) lebeling the filtrate beaker B2.

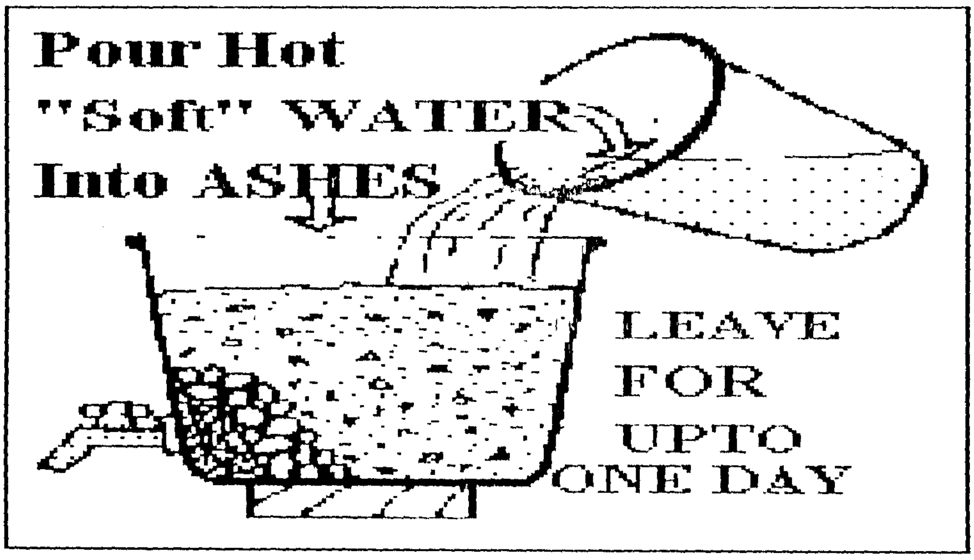
9) Step 2-8 were repeated for one (1)more time with one (1) liters of distilled water. The filtrate beaker was lebelled B1, B2 and B3.

10) The mass of (KOH) residue obtained in each stainless plate (B1-B3) after evaporation to dryness were weighed.

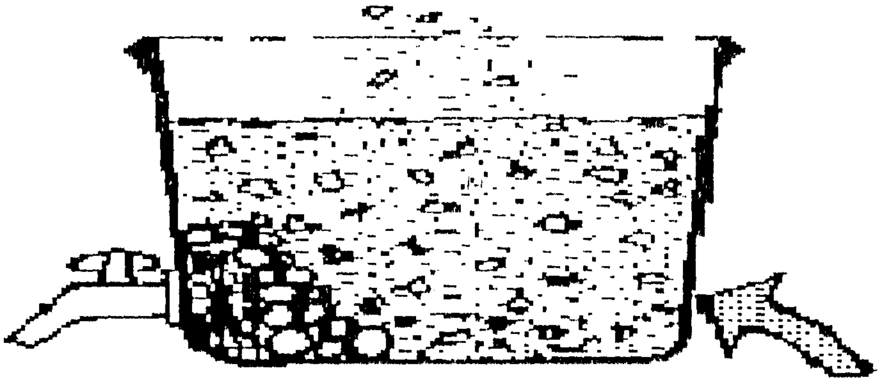
11) Solution of known concentration was prepared from the residues in (B1-B3) the solution were then titrated against 0.1M HCl to determine the caustic potash strength using phenolphthalein indicator.

2.2.0 PREPARATION OF CAUSTIC POTASH

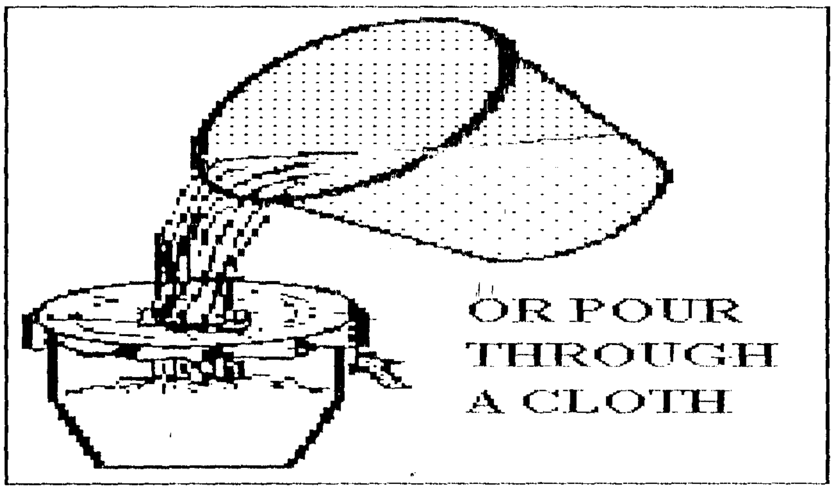
Some reasonable amount of banana peels suspected to have concentration of potassium ion (K^+) were gathered, dried and burned to ashes. The desired quantity of the ashes was then packed in a container. A measured volume of water was poured and boiled, transferred to a porous container and a clean container kept at the lower end for collection of the crude caustic potash. The solution was filtered to remove solid and unwanted dark colour. It was then concentrated by evaporation.

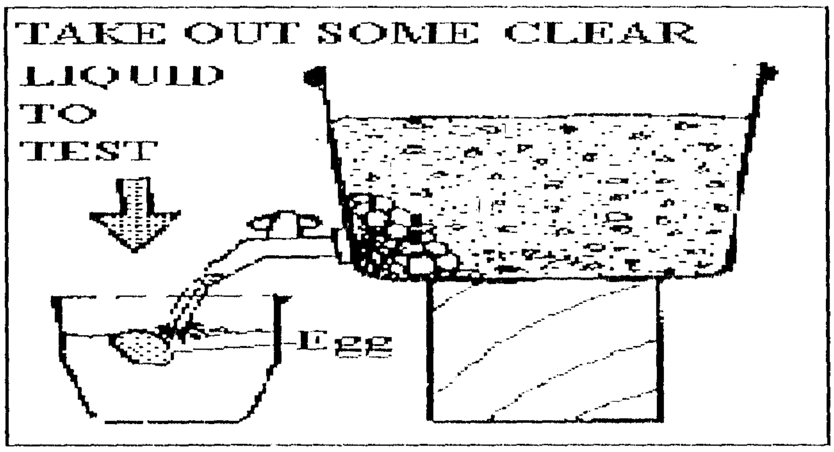


Put Lots of ASHES
In a "Safe" Container.



WOOD or PLASTIC.
NOT Metal Container





2.2.1 EFFECT OF CAUSTIC POTASH ON THE ENVIRONMENT.

Its manufacture, use, and disposal do not have adverse effects on the environment and are done in a manner compatible with organic handling. A lye peeling processing method is of concern to the agro-ecosystem due to handling of waste from the plant. Large volumes of water are used, which enter the waste stream along with the soluble potassium and alkali ions. Lye peeling with sodium hydroxide is more of a disposal problem due to undesirable sodium content that may be soil applied, whereas residual potassium is a plant nutrient, although it would be considered synthetic and not permitted for an organic farming system.

Peach processing plants using lye peeling are generally restricted by state and local waste water treatment requirements, which have resulted in a limited number of plants and sites in operation (O'Bara, 2001). Data supplied by the petitioner indicates that alkalinity of waste is not a factor, due to the natural acidity of the fruit, which must be additionally buffered during on-site treatment (Finn, 2001). Conventional tomato lye peeling processes may use 9800/liters water /ton of tomatoes peeled. Advances in technology to combine lye peeling with mechanical scrubbers reduced the water consumption (Luh, 1988).

Dry caustic peeling was advocated in the 1970s to substantially reduce the amount of plant wastewater discharged (National Canners Association, 1970). This process uses infrared energy at 1650 degrees to condition the surface of fruit that is treated with stronger sodium hydroxide solutions. The peel is removed mechanically by soft rubber scrubbing rolls rather than by water, so that about 90% of the peel is removed as a thick heavy "peanut butter-like" substance, which must be disposed of (Woodroof, 1986). Caustic peeling continues to be considered more effective at peel removal with substantial reduction in wastewater when compared with conventional peeling (Lindsay, 1996).

Disposal of KOH can be potentially dangerous. Mercury cells are used to produce most of the KOH in the United States (Freilich and Petersen, 1996). The stripped mercury is generally recycled and discharge of mercury is forbidden.

3. If the nutritional quality of the food is maintained and the material itself or its breakdown products do not have adverse effects on human health as defined by applicable Federal regulations.

Potassium is an essential mineral nutrient. Lye peeling with sodium hydroxide has been shown to reduce the amount of the Pru p 1 protein in peaches (Brenna, et al., 2000). This is regarded as the major allergen in peaches and therefore may be considered of nutritional benefit. Allergans in rosaceae fruit are associated with the skin (Fernandez-Rivas, 1999). The petitioner has submitted experimental data showing no increase in potassium content of the fruit due to the use of potassium hydroxide. In data from 1998, samples tested after hand peeling had comparable levels of potassium to those that had been through the treatment line (average 665 ppm and 661 ppm respectively). After blanching, the potassium content drops substantially, to 422 ppm.

Peeling methods can effect product nutrient loss, with the less flesh removed the better the nutrient retention.

Nutrient loss can also occur from leaching out of water soluble constituents or degrading of heat sensitive compounds. Ascorbic acid and thiamin were reduced by 12% by lye peeling, although carotenoids were not reduced.

Fruit that is canned without peeling, for instance, retains more nutrients (Saluhnke, 1990). Mechanical peeling, coring, and slicing has the least effect on nutrients, but is not an option for soft fruits.

Freezing of fruit is not shown to contribute to nutrient loss, whereas canned fruit does lose nutrients (Saluhnke, 1990). Oxygen sensitive nutrients such as vitamin C can decline during storage if the fruit is not properly protected.

Fruit maturity is a key factor in the overall quality and level of nutrients found in fruit. Fruit that is picked earlier for satisfactory texture in freezing may not have as high a content of various nutrients, but other forms of processing such as canning and pureeing, will result in a loss of nutrients as well (Eskin, 1991).

4. Its primary purpose is not as a preservative or used only to recreate/improve flavors, colors, textures, or nutritive value lost during processing except in the latter case as required by law.

KOH does not serve as a preservative nor does it recreate or improve flavor or color. It does aid in preserving texture in the final product, though this is not strictly a recreation of texture.

2.3.0 GENERAL PRINCIPLES OF LEACHING OPERATION

Leaching is concerned with the extraction of soluble constituent from a solid by means of a solvent. The process may be used either for the production of a concentrated solution of a valuable solid material, or in order to remove an insoluble solid, such as a pigment from a soluble material with which it is contaminated. The proportion of soluble constituents present determines the method used for extraction. Its distribution throughout the solid the nature of the solid and the particle size. If the solute is uniformly dispersed in the solid, the material near the surface will be dissolved first. Leaving a porous structure in the solid residue. The solvent will then have to penetrate this outer layer before it can reach further solutes and the process will become progressively more difficult and the extraction rate will fall. If the solute forms a very high proportion of the solid, the porous structure may breakdown almost immediately to give a fine deposit of insoluble residue, and access of solvent to the solute will not impeded. Generally, the process can be considered in the three parts.

1) The change of phase of the solute as it dissolves in the solvent.
2) Its diffusion through the solvent in the pores of the outside of the particle, and
3) The transfer of the solute from the solution in contact with the particles to the main bulk of the solution [15]

2.3.1 PROCESSES OF LEACHING OPRATION

Leaching processes fall into two principal classes; those in which the leaching is accomplished by percolation (seeping of solvent through a bed of solids) and extracting liquid and subsequently separated from it.

Leaching operations are carried out under batch, semi-batch (unsteady state) as well as under completely continuous (steady state) conditions. The unsteady-state operations include those where the solids and liquids are contacted in purely batch wise fashion and also those where a batch of the solids contacted with a continually flowing stream of liquid (semi-batch method). The continuous steady state operation can be broadly classified into two major categories, according to whether it operates in stage wise or in a continuous contact fashion, viz, single-stage and counter-current multiple contact [14].

2.3.2 *MASS TRANSFER IN LEACHING OPERATION*

Mass transfer rates within the porous residue are difficult to assess because it is impossible to define the shapes of the channels through which transfer must take place. It is possible however, to obtain an approximate indication of the rate of transfer from the particles to the bulk of the liquid, using the concept of a thin film as providing the resistance to transfer the equation for mass transfer may be written as

$$\frac{dm}{dt} = \frac{KA(CS - C)}{b} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (i)$$

The diffusion coefficient approximately equal to the liquid phase diffusivity. D_2 and it usually assumed constant for a batch process in which V, the total volume of solution, is assumed to remain constant, then:

$$dm = Vdc \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (ii)$$

The time t taken for the concentration of the solution to rise from initial value to a value is found by integration on the assumption that both b and A remain constant Rearranging equation (iii)

$$\int_{co}^{c} \frac{dc}{Cs - C} = \int_{Cs-C} \frac{(KAdt)}{Vb} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (iv)$$

$$\ln \frac{Cs - Co}{Cs} = \frac{KA}{Vb} t \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (v)$$

$$\frac{I - C}{Cs} = e^{-(KA/Vb)t} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (vi)$$

$$C = Cs(1 - e^{-(KA/Vb)t})$$

where A= area of the solid-liquid interface

b= effective thickness of the liquid film surrounding the particles.

C= concentration of the solute in the bulk of the solution at time t.

Cs= concentration of the saturated solution in the contact of the particle

M= mass of the solid transferred in time t.

H_{4s} -Enthalpy of steam vapor at 12.16°C = 2523 kJ/kg

$$H_{4s} := 2523 \cdot \frac{kJ}{kg} \qquad C_{P3} := 4.18 \cdot \frac{kJ}{kg \cdot {}^{o}C}$$

H_3 -Enthalpy of vapor leaving the third effect = H_{4s} + $(C_p)_{steam}$ × $(BPR_3)_{ssuperheat}$

$$H_3 := H_{4s} + C_{Ps} \cdot BPR_3$$

$$H_3 = 2523.942 \frac{kJ}{kg}$$

h_3 – Enthalpy of outlet from third effect at 22.56°C = C_{p3} ×(t_3 - 0)

$$h_3 := 0.57 \cdot C_{P3} \cdot T_3$$

$$h_3 = 40.945 \frac{kJ}{kg}$$

$$W_2 \cdot \lambda_2 + \left(W_f - W_1 - W_2\right) \cdot h_2 = W_3 \cdot H_3 + \left(W_f - W_1 - W_2 - W_3\right) \cdot h_3$$

$$\lambda_2 = 2392.826 \frac{kJ}{kg} \qquad W_f = 0.698\ kg \cdot s^{-1} \qquad h_2 = 101.359 \frac{kJ}{kg}$$

$$H_3 = 2523.942 \frac{kJ}{kg} \qquad h_3 = 40.945 \frac{kJ}{kg}$$

$$2393 \cdot W_2 + \left(14.165 - W_1 - W_2\right) \cdot 101.359 = 2524 \cdot W_3 + \left(14.165 - W_1 - W_2 - W_3\right) \cdot 40.945$$

$$38.610024166583904393 \cdot W_2 + 14.165000000000000000 - 41.100655477207269838 \cdot W_3$$

$$W_1 = 38.61 \cdot W_2 - 41.10 \cdot W_3 + 14.165$$

$$W_1 - 38.61 \cdot W_2 + 41.10 \cdot W_3 = 14.165$$

The equations are transformed into the following equations:

$$W_s - 0.11645 \cdot W_1 = 0.009126$$

$$W_1 - 1.0615 \cdot W_2 = 0.1313$$

$$W_1 - 38.61 \cdot W_2 + 41.10 \cdot W_3 = 14.165$$

$$W_1 + W_2 + W_3 = 12.9$$

$$\begin{pmatrix} W_s \\ W_1 \\ W_2 \\ W_3 \end{pmatrix} := \begin{pmatrix} 1 & -1.1164 & 0 & 0 \\ 0 & 1 & -1.0615 & 0 \\ 0 & 1 & 38.61 & -41.10 \\ 0 & 1 & 1 & 1 \end{pmatrix}^{-1} \cdot \begin{pmatrix} 0.009126 \\ 0.1313 \\ 14.165 \\ 12.9 \end{pmatrix}$$

$$\begin{pmatrix} W_s \\ W_1 \\ W_2 \\ W_3 \end{pmatrix} := \begin{pmatrix} W_s \\ W_1 \\ W_2 \\ W_3 \end{pmatrix} \cdot \frac{kg}{sec}$$

$$\begin{pmatrix} W_s \\ W_1 \\ W_2 \\ W_3 \end{pmatrix} = \begin{pmatrix} 1.838 \times 10^4 \\ 1.644 \times 10^4 \\ 1.504 \times 10^4 \\ 1.497 \times 10^4 \end{pmatrix} \frac{kg}{hr}$$

Now, $Q_1 =$

$$Q_1 := W_s \cdot \lambda_s$$

$$Q_1 = 41501609.665 \frac{kJ}{hr}$$

But $Q_1 = U_1 \cdot A_1 \cdot \Delta t_1$

$$A_1 := \frac{Q_1}{U_1 \cdot \Delta t_1}$$

$$A_1 = 7.841\ m^2$$

$Q_2 := W_1 \cdot \lambda_1$ W

$$Q_2 = 10572.954 \frac{kJ}{sec}$$

But $Q_2 = U_2 \cdot A_2 \cdot \Delta t_2$

$$A_2 := \frac{Q_2}{U_2 \cdot \Delta t_2}$$

$A_2 = 14.162\ m^2$

But $Q_3 = U_3 \cdot A_3 \cdot \Delta t_3$

$Q_3 := W_2 \cdot \lambda_2$

$Q_3 = 9995.077\,\frac{kJ}{sec}$

$A_3 := \frac{Q_3}{U_3 \cdot \Delta t_3}$

$A_3 = 17.846\ m^2$

Thus the obtained areas are within the acceptable range of 5% difference

Therefore the average area per effect of the evaporator is 232 m².

Tube details:

Most generally used diameters today ranges from 1.25 to 2.00 in. outer diameter and most generally used lengths of tubes ranges from 4 to 15 ft. Let us choose 5/4-in. nominal diameter, 80 schedule, brass tubes of 10-ft length. Therefore Outer diameter d_o = 42.164 mm

Inner diameter d_i = 32.46 mm

Length L = 10 ft

$d_o := 42.164 \cdot mm$

$d_i := 32.46 \cdot mm$

$L := 20 \cdot ft$

$L = 6.096\ m$

Tube pitch P_T = 1.25 × d_o

$A_{Pt} := 1.25 \cdot d_o$

$A_{Pt} = 0.053\ m$

Surface area of each tube $a = \infty d_o L$

$$a := \pi \cdot d_o \cdot L$$

$$a = 0.807 \ m^2$$

Number of tubes required $N_t = A/a$

$$N_t := \frac{A_3}{a}$$

$$N_t = 22.1$$

Area occupied by tubes $= N_t \times (1/2) \times P_T \times P_T \times \sin 60$

$$A_t := N_t \cdot \left(\frac{1}{2}\right) \cdot A_{Pt} \cdot A_{Pt} \cdot \sin(60)$$

$$A_t = 0.009 \ m^2$$

But actual area is more than this. Hence this area is to be divided by factor which varies from 0.8

1.0. Let us choose this factor as 0.9. Therefore actual area required = 0.7445/ 0.9

Actual Area Ac $\quad A_c := \frac{A_t}{0.8}$

$$A_c = 0.012 \ m^2$$

The central downcomer area is taken as 40 to 70% of the total cross sectional area of tubes. Let u

take it as 50%.

Therefore Downcomer area =

$$D_A := 0.5 \cdot \left(N_t \cdot \frac{\pi}{4} \cdot {d_o}^2\right)$$

$$D_A = 0.015 \ m^2$$

Downcomer diameter =

$$D_d := \sqrt{\frac{4 \cdot D_A}{\pi}}$$

$$D_d = 0.14 \ m$$

Total area of tube sheet in evaporator = downcomer area + area occupied by tubes

$A_T := A_c + D_A$

$A_T = 0.027\ m^2$

6.2 Reactor-Mixer Design

Amount of feed:-

$Feed := N_2$

Composition of feed:

% Composition of cocoa pod ash

$y_{BPA} := 29.1 \cdot \%$

Amount of H_2O

$M_{BPA} := y_{BPA} \cdot Feed$

$M_{BPA} = 206.02 \frac{kg}{hr}$

% Composition of H_2O

$y_{H2O} := 70.9 \cdot \%$

Amount of H_2O

$M_{H2O} := y_{H2O} \cdot Feed$

Density of feed:

Density of cocoa pod ash

$\rho_{BPA} := 982 \cdot \frac{kg}{m^3}$

Density of H_2O

$\rho_{H2O} := 1000 \cdot \frac{kg}{m^3}$

Volumetric flowrate of cocoa pod ash

$$V_{BPA} := \frac{M_{BPA}}{\rho_{BPA}}$$

$$V_{BPA} = 0.21 \frac{m^3}{hr}$$

Volumetric flowrate of H_2O

$$V_{H2O} := \frac{M_{H2O}}{\rho_{H2O}}$$

$$V_{H2O} = 0.502 \frac{m^3}{hr}$$

Total volumetric flow rate of the feed is:

$$V_T := V_{BPA} + V_{H2O}$$

$V_T = 0.712 \frac{m^3}{hr}$ Safety Factor (10%) $V_T := \frac{V_T}{0.1}$ $V_T = 170.82 \frac{m^3}{day}$

$number\_tanks := 2$ Each tank capable of handling 1/2 the design flow, tanks to be circular withdepth equal to diameter. The reasoning for using 2 tanks is redundancy. Eventually there will be problems with the tanks themselves or the mixers or both. The plant needs to have a backup. May want to consider stocking an extra mixing unit also. This particular design through may not be optimum since if one basin is out of service all the water will, of necessity go a single basin designed for only 1/2 the flow.

$detention\_time := 40 \cdot sec$ This is somewhat longer than the guidelines given in the text, 0.5 to 10 seconds. It could be argued that if one basin is out of service that the remaining basin will still have a 20 second detention time which is within the guidelines.

$G := 800 \cdot sec^{-1}$ velocity gradient, units are ft/(sec*ft) impeller type : turbine, 6 flat blades, vaned disc: $K_T := 6.3$

The drawing below for what a vaned disk impeller looks like

As part of you design proposal for this facility provide the following:

1. Water power input in kW
2. tank dimensions
3. JWI mixer model number
4. rotational speed of the impeller in rpm.

SOLUTION:

appropriate definitions and constants:

$\mathrm{rps} := 2 \cdot \pi \cdot \frac{\mathrm{rad}}{\mathrm{sec}}$ $\mathrm{rpm} := 2 \cdot \pi \cdot \frac{\mathrm{rad}}{\mathrm{min}}$ $\rho := 998.408 \cdot \frac{\mathrm{kg}}{\mathrm{m}^3}$ $\mu := 1.027 \cdot 10^{-3} \cdot \frac{\mathrm{newton} \cdot \mathrm{sec}}{\mathrm{m}^2}$

$$\mathrm{detention\_time} = \frac{\mathrm{volume}}{\mathrm{flowrate}}$$

$\mathrm{detention\_time} := 3 \cdot \mathrm{day}$

$$G = \sqrt{\frac{P}{\mu \cdot \mathrm{volume}}}$$

$$P = K_T \cdot n^3 \cdot D_{\mathrm{imp}}^{\;5} \cdot \rho$$

AVAILABLE MIXER MODELS

Model	Rotational speed, rpm	power, kW
JTQ50	30, 45	.37
JTQ75	45, 70	.56
JTQ100	45, 110	.75
JTQ150	45, 110	1.12
JTQ200	70, 110	1.50
JTQ300	110, 175	2.24
JTQ500	175	3.74

impeller diameter should be between 0.2 and 0.5 the diameter of the tank

$$\mathrm{volume} := \mathrm{detention\_time} \cdot V_T$$

$\mathrm{volume} = 512.46\mathrm{m}^3$ total volume, the volume of each tank is 1/2 this value. The depth and diameter of each tank is then:

$$\frac{\pi \cdot D^{2}}{4} \cdot D = \frac{volume}{2} \qquad D := 2^{\frac{1}{3}} \cdot \left(\frac{volume}{\pi}\right)^{\frac{1}{3}} \qquad D = 6.884\ m$$

$$G = \sqrt{\frac{P}{\mu \cdot \left(\frac{volume}{2}\right)}}$$

$P := G^{2} \cdot \mu \cdot \frac{volume}{2}$ $P = 168.415kW$ power required to meet required velocity gradient, 800 sec $^{-1}$

Assume that any mixer used is only 80% efficient at transferring power into tank. Therefore req

mixer should have power rating of at least :

$eff := .80 \qquad P_{mixer} := \frac{P}{eff} \qquad P_{mixer} = 210.519\,kW$

At this point some judgements are used. We have rules of thumb regarding the relationship of the size of the mixing impeller to the size of the tank. We have a G value to be met and we have rotational speed of the impeller which is somewhat fixed by the available mixers.

Using a maximum rotational speed of 175 rpm, a six blade, vaned disk impeller $K_T = 6.30$ (as described from the drawing) we can compute the diameter of the impeller. Looking at the tab the smallest unit that will work is the JTQ300 which has a nameplate power level of 2.24 kW

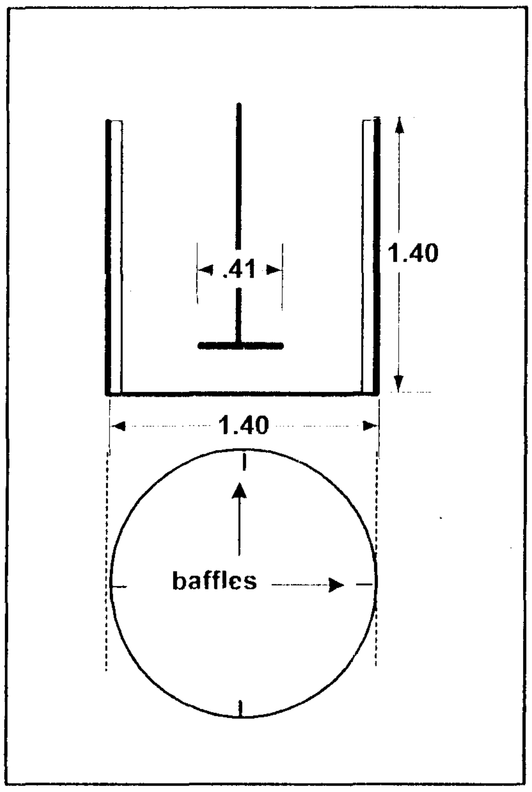
$n := 175 \cdot rpm \qquad P_{actual} := 2.24 \cdot kW \cdot .8$

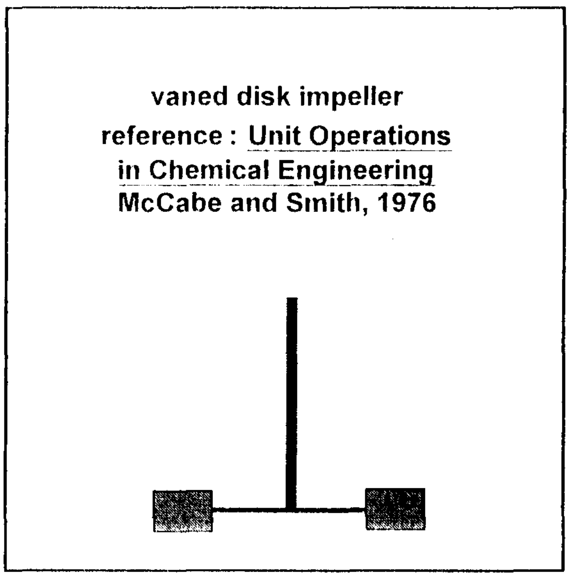
Computing the impeller diameter resulting from using a JTQ300 rated at 2.24 kW is:

$$D_{imp} := \left[\frac{\frac{P_{actual}}{watt}}{K_T \cdot \left(\frac{\frac{n}{60}}{rpm}\right)^{3} \cdot \frac{\rho}{\frac{kg}{m^{3}}}}\right]^{\frac{1}{5}} \cdot m$$

$D_{imp} = 0.409m$ This is the impeller diameter that results if we use a JTQ300 mixing unit

This results in an impeller to tank ratio of $\frac{D_{imp}}{D} = 0.059$ **which is just inside the rule of thumb**




vaned disk impeller
reference : Unit Operations in Chemical Engineering
McCabe and Smith, 1976

The tank has baffles attached to its periphery to break up bulk movement and promote mixing.

The rule of thumb is that the baffles should extend into the tank a distance of [0.1 x diameter]

$$G := \sqrt{\frac{P_{actual}}{\mu \cdot \frac{volume}{2}}} \qquad G = 82.522 \frac{1}{sec}$$

We can use a slightly smaller impeller to bring G down to 800 sec^{-1} if we care to

$$G \cdot detention\_time = 21389652.952$$

Reynolds and Richards also check the Reynolds number, which should be > 10,000 to assure tur

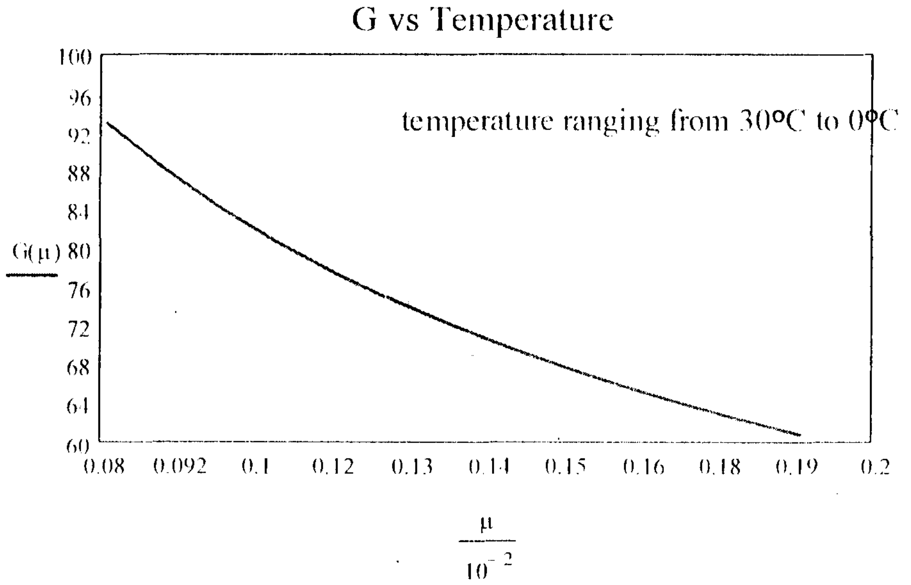
mixing conditions.

$$N_{Re} := \frac{D_{imp}^{2} \cdot n \cdot \rho}{\mu} \qquad N_{Re} = 2984117.22$$ turbulent mixing conditions

By examining mixing intensity, G, as a function of temperature
absolute viscosity at 0° C = 1.89*10^{-2} gm/cm*sec
absolute viscosity at 30° C = .8*10^{-2} gm/cm*sec

$$\mu := \left(1.89 \cdot 10^{-2} \cdot \frac{gm}{cm \cdot sec}\right), \left(1.8 \cdot 10^{-2} \cdot \frac{gm}{cm \cdot sec}\right) .. .8 \cdot 10^{-2} \cdot \frac{gm}{cm \cdot sec}$$

$$G(\mu) := \sqrt{\frac{P_{actual}}{\mu \cdot \frac{volume}{2}}}$$



The actual mixing intensity will range from just over 1000 sec^{-1} at 30°C down to 675 sec^{-1} at ne

freezing water temperatures, 0°C

$$D := \frac{D_{imp}}{\frac{D_{imp}}{D}}$$

$D = 6.884$ m

From literature, Cerebro, 2003 ratio of tank diameter to tank height (D/H) is:

$$\text{Ratio} = \frac{D}{H} = 0.83$$

6.3 Filter Design

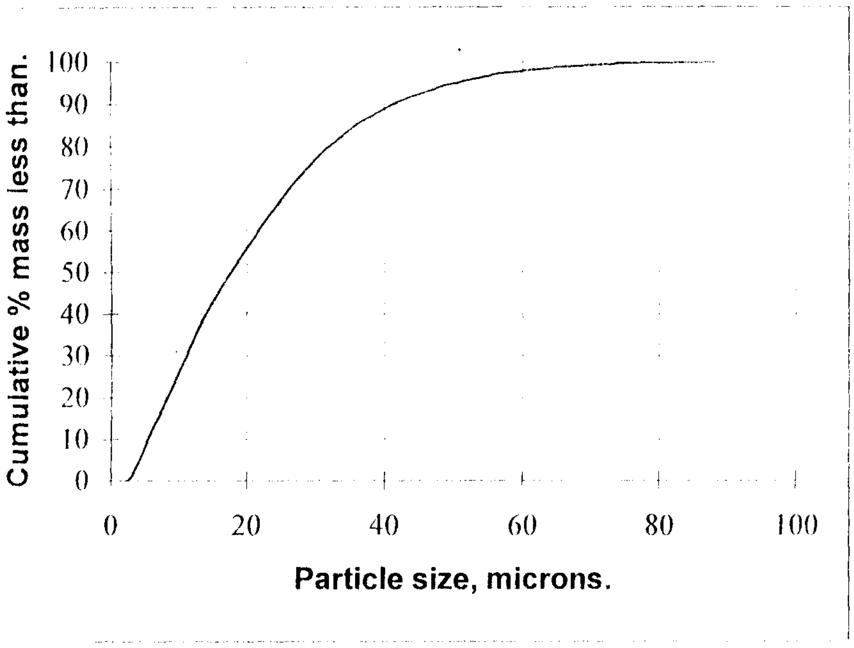
The deign of the filter was carried out by the use of a spreadheet

INPUTS:	**	SECTION FOR SPECIFIC RESISTANCE:					**
	**	Solid density:			2650	kg/m^3	**
	**	Cake concentration:			0.38	v/v	**

Particle size (um)	Cumul've mass undersize (%)	**	Topsize of grade (um)	Bottom of grade (um)	Average size (um)	Mass fraction (-)	Mass fract'n over mid point (1/um)
88.1	100	**	102.1	88.1	95.1	0	0
76	99.7	**	88.1	76	82.05	0.003	3.656E-05
65.6	98.8	**	76	65.6	70.8	0.009	0.0001271
56.6	97.2	**	65.6	56.6	61.1	0.016	0.0002619
48.8	94.5	**	56.6	48.8	52.7	0.027	0.0005123
42.1	90.7	**	48.8	42.1	45.45	0.038	0.0008361
36.3	85.6	**	42.1	36.3	39.2	0.051	0.001301
31.3	79.1	**	36.3	31.3	33.8	0.065	0.0019231
27	71.5	**	31.3	27	29.15	0.076	0.0026072
23.3	63.6	**	27	23.3	25.15	0.079	0.0031412
20.1	55.9	**	23.3	20.1	21.7	0.077	0.0035484
17.4	49.2	**	20.1	17.4	18.75	0.067	0.0035733
15	43	**	17.4	15	16.2	0.062	0.0038272
12.9	36.6	**	15	12.9	13.95	0.064	0.0045878
11.1	29.9	**	12.9	11.1	12	0.067	0.0055833
9.6	24.3	**	11.1	9.6	10.35	0.056	0.0054106
8.3	19.7	**	9.6	8.3	8.95	0.046	0.0051397
7.2	15.9	**	8.3	7.2	7.75	0.038	0.0049032
6.2	12.6	**	7.2	6.2	6.7	0.033	0.0049254
5.3	9.2	**	6.2	5.3	5.75	0.034	0.005913
4.6	6.3	**	5.3	4.6	4.95	0.029	0.0058586
4	3.9	**	4.6	4	4.3	0.024	0.0055814
3.4	2	**	4	3.4	3.7	0.019	0.0051351
3	1	**	3.4	3	3.2	0.01	0.003125
2.6	0.2	**	3	2.6	2.8	0.008	0.0028571
2.2	0	**	2.6	2.2	2.4	0.002	0.0008333
**	**	**					---------
							0.081549

CALCULATED VALUES:

SPECIFIC SURFACE IS:	6 times	0.081549	1/microns
(ie assuming spheres)	=	489293.85	1/metres
Sauter mean diameter:		12.3	microns
Permeability (by K & C):		1.379E-12	square metres
Specific resistance:		720231606	m/kg


Cumulative % mass less than.
100
90
80
70
60
50
40
30
20
10
0
0
20
40
60
80
100
Particle size, microns.

ONSTANT PRESSURE FILTRATION.XLS

Constant pressure filtration readings from vacuum or pressure test

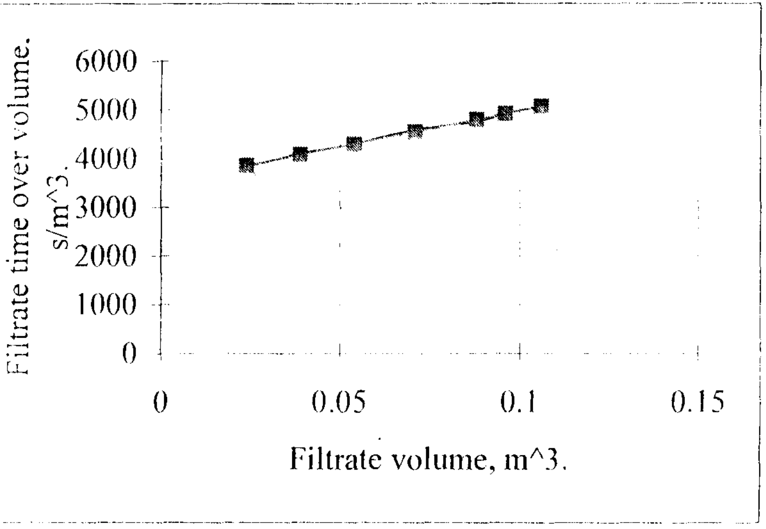
cake and cloth resistance follows from

NPUTS ********* ********* ********* ********* *********

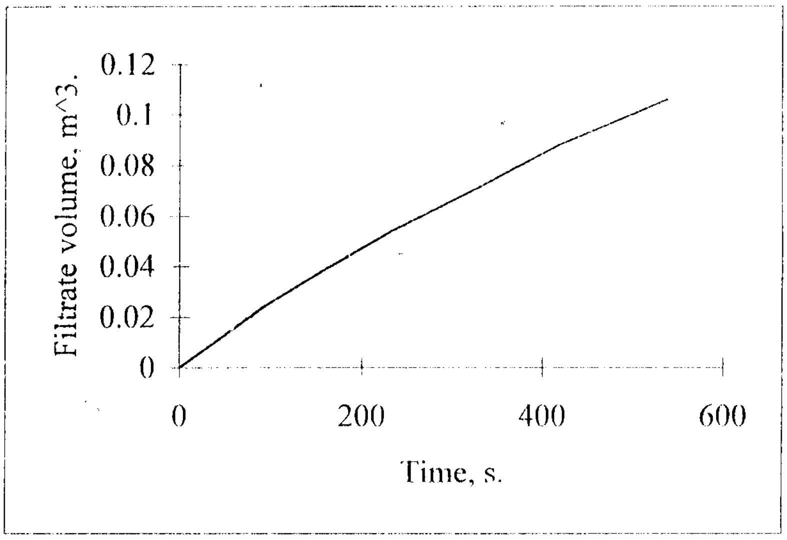
ilter area:	2.72	m^2
iquid viscosity:	0.001	Pa s
iquid density:	1000	kg/m^3
olids in slurry weight fraction:	0.1	-
lass of wet cake sample:	1280	g
lass of dry cake sample:	635	g

			*********	*********	*********	*********
Filtration time	Filtration pressure	Filtrate Volume	********* ********* *********	Time over volume	Pressure forming cake	Fitted t/V
(s)	(bar)	(m^3)	*********	(s/m^3)	(Pa)	(s/m^3)
0	3	0	*******			
92	3	0.024	*********	3833.3333	50295	3859.785231
160	3	0.039	*********	4102.5641	73977	4080.501486
232	3	0.054	*********	4296.2963	93557	4301.217742
327	3	0.071	*********	4605.6338	112013	4551.362832
418	3	0.088	*********	4750	127440	4801.507922
472	3	0.096	*********	4916.6667	133856	4919.223258
538	3	0.106	*********	5075.4717	141235	5066.367429

verage: 3 bar



```
CALCULATED VALUES:                 *********  *********  *********
Moisture ratio of cake:                                  2.015748
Dry cake per filtrate volume (equn 2.18):              125.24655 kg/m^3
Filtration pressure:                                      300000 Pa
Cake specific resistance:                              5.215E+11 m/kg
Cloth resistance:                                      2.861E+12 m^-1
Cake forming pressure:                                 134176.96 Pa
*********  *********  *********  *********  *********  *********
Regression Output:
Constant                        3506.6392
X Coefficient(s)                14714.417
R Squared                       0.9943521
```



CHAPTER SEVEN

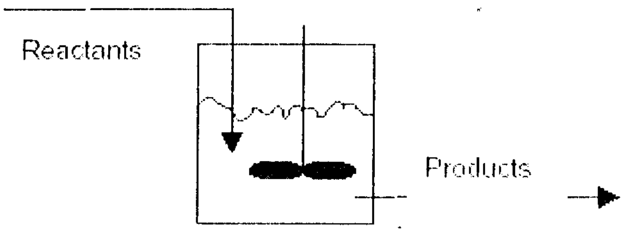
7.0 REACTOR EQUIPMENT OPTIMIZATION

Chemical reaction engineering involves choosing the optimum reacting system in order to produ the greatest amount of desired product in the most cost-efficient manner. It is important to deterr which type of reactor should be used and under what conditions it is to be operated. For exampl one must determine whether to use a batch or continuous process from the production rate. For production rates greater than 1.1E8 lbmol / hr, it is more feasible to use continuous operations. Once the operation process (batch or flow) is chosen the next step is to determine which reactor to use. The selected topics for reaction engineering are: deriva of rate equation, reproducibility of selectivity vs. conversion for Acetone, and a detailed discussion on reactor design

7.1 Types of flow-reactors

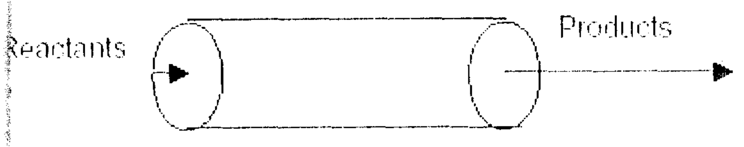
7.1.1 CSTR- Continually stirred tank reactor

In this reactor the contents inside are well stirred and the composition remains uniform througho This is a steady-state operation where the exit stream has the same composition as the fluid insid the reactor.



7.1.2 Plug Flow Reactor (PFR)

The main difference between the plug flow reactor and the mixed flow or CSTR is that the composition changes as the fluid moves through the reactor. The fluid flow through the reactor is orderly with no fluid element overtaking or mixing with any other element ahead or behind. Another identifying characteristic of plug flow reactors is that the residence time in the reactor must be the same for all elements of fluid.



.2 Kinetic Experiments for Acetone Cracking

erivation of rate equations for non-isothermal reactor conditions is an involved process equiring experimental data, trial and error procedures, and calculated guesses. There are hree main parameters are necessary to derive a rate equation:

the order of the reaction, n

the rate constant, k

activation energy, E

Calculations are not perfect, since some reaction takes place in the preheat section of the reactor, t the outlet section, where it is cooler. Our reactor is non-isothermal, but by using the Hougen a Watson method, we can reduce the data to isothermal conditions.

The idea of this section is to find a method for deriving approximate values for the rate constant.

.3 Reactor Volume:

A good place to start is the equation for the equivalent reactor volume, that is the volume under on-isothermal conditions, applicable to our design project, and is approximated by isothermal profiles. The equation, the Hougen-Watson method, from Fogler, 1986, is as follows:

$$V_R = \int_0^V \mathrm{EXP}\left[\frac{E}{R}\cdot\left(\frac{1}{T_1}-\frac{1}{T}\right)\right] dV$$

Where, T is the reactor temperature and E is the Activation Energy. Calculation of VR equires knowledge of temperature profile along the reactor tube, and the Activation Energy, E. ce we solve this equation, the rate is based solely on the continuity equation:

$$F \cdot dX = r \cdot dV_R$$

7.4 Determination of Activation Energy, E

There are two methods used to determine the Activation Energy. There is Arrhenius' formula, and there is a shortcut method.

7.4.1 Arrhenius method

Arrhenius' method is a trial and error procedure calculated at several temperature levels. A value of E is guessed and with the values from a temperature profile, VR is derived by graphical integration. Then from this value a rate constant can be derived. These values are expressed in Arrhenius' formula:

$$k = k_0 \cdot e^{\frac{-E}{R \cdot T}}$$

If this value is consistent with that used in the calculation of VR the whole procedure has to be repeated with a better approximation of E.

7.4.2 Shortcut method

Shortcut method is preferred over the Arrhenius' method, because it can give very accurate results in a shorter amount of time, and does not require knowledge of the rate constant. For this method, experimental data for at least two trials is necessary. There are two preliminary considerations. First, the two temperature profile lines, corresponding to xak and pac values, must be parallel. Second, the feed rates should be adjusted so that equal conversion is obtained for both trials; then the pA vs. VR profiles will be identical. Thus, the Activation Energy, E, can be determined graphically by using the data from two experiments, say one at 710°C and another at 750°C. Figure 1, (Froment, 1960),

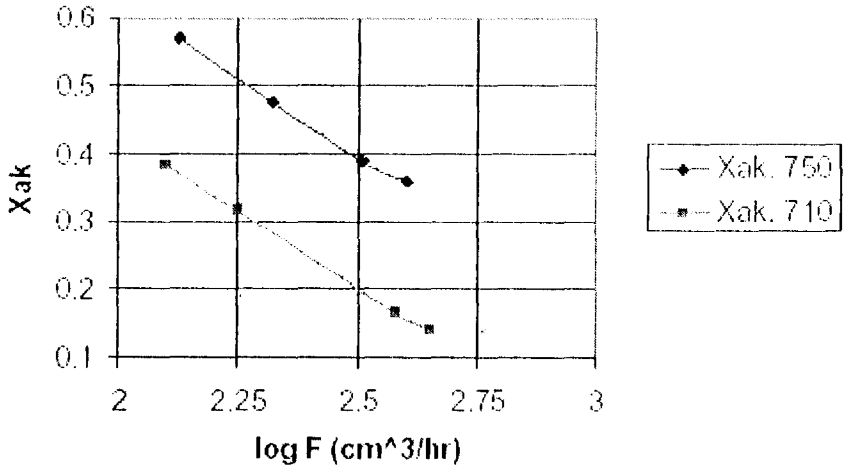


Figure 7.4.2: Estimation of Activation Energy (Froment, 1960)

illustrates the nearly parallel lines at these two temperatures. By measuring the horizontal distance between the two parallel lines on the graph of xak vs. log F, determination of the two feed rates is accomplished. From the following equation, Activation Energy is now found.

$$E = 2.3 \cdot R \cdot \frac{T_1 \cdot T_2}{\Delta T} \cdot \log\left(\frac{F_1}{F_2}\right)$$

7.5 Temperature Profile

Temperature profile along the reactor is the second calculation for determining the volume of the reactor. There are three assumptions:

· wall temperature is known

· physical properties of Acetone/reaction mixture are known

· flow composition is known

A temperature profile can be calculated using the Sieder-Tate equation, (Froment, Pijcke, Goethals, 1960):

$$\Phi = \frac{t_o - t_m}{t_o - t_s} \text{vs} \cdot \frac{m \cdot Cp}{k_R \cdot L}$$

This equation can measure the temperature of the gas, tm, at any point along the reactor; this is done by a graphical method, since the only unknowns are tm and L. The following Figure 2, shows the temperature as a function of the distance along the reactor, by using equation 5. The wall temperature is constant from the preheat section through a good portion of the reactor, and then falls off around 105 cm. However, the calculated temperature profile shows steep drops in the preheat section, and later towards the end of the reactor. This deviation is not taken into account, since little reaction takes place in these sections. It is important to note the heat effect was not taken in account.

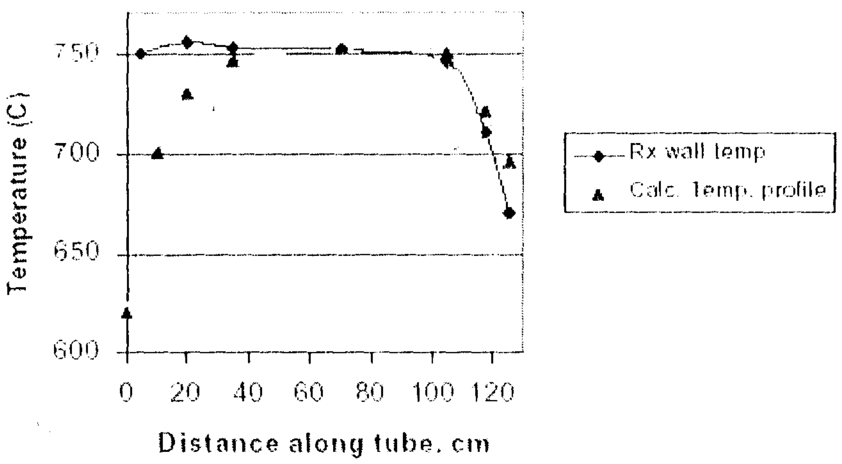


Figure 7.5.1: Temperature vs. Reactor Position

because there is experiment evidence the heat flux from the components offsets the endothermicity of the reaction. With the temperature profile and Activation Energy known, plotting xak vs.

V_R/F_{AO} can solve the reactor volume. Then the continuity equation becomes:

$$\frac{V_R}{F_{AO}} = \frac{1}{k} \cdot \int_0^X \frac{1}{P_A{}^n}\, dX_{ak}$$

When the rate equation is in the form:

$$r_A = k \cdot P_A{}^n$$

Before the continuity equation can be solved, it is necessary to express pac as a function of x_{ak}. Figure 3 shows an empirical fit of p_{ac} as a function of xa shows a curve fitted with the equation:

$$P_{ac} = (1 - 1.05 \cdot X_{ak})^2$$

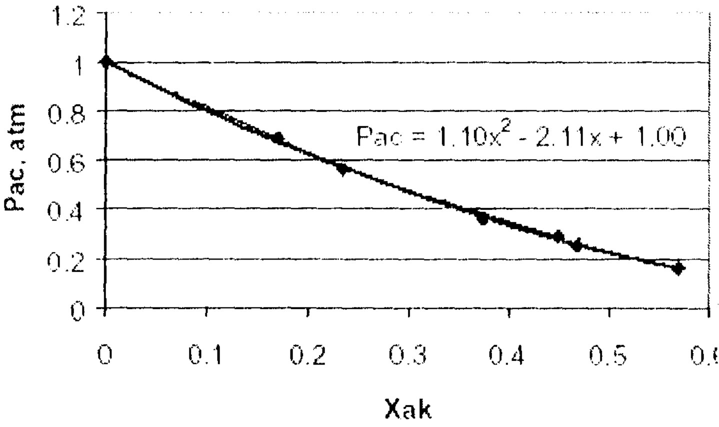


Figure 7.5.2: Empirical Fit of Pac vs Xak

Then the continuity equation becomes:

$$k = \left[\frac{1}{1.05 \cdot \frac{V_R}{F} \cdot \left(2_n - 1\right)} \right] \cdot \left[\frac{1}{\left(1 - 1.05 \cdot X_{ak}\right)^{2n-1}} - 1 \right]$$

The experimental values of V/F and xak are plugged into the above equation; then the order and rate constant can be determined by a plot of graph kp vs. n. The intersection of the curves corresponds to k and n values.

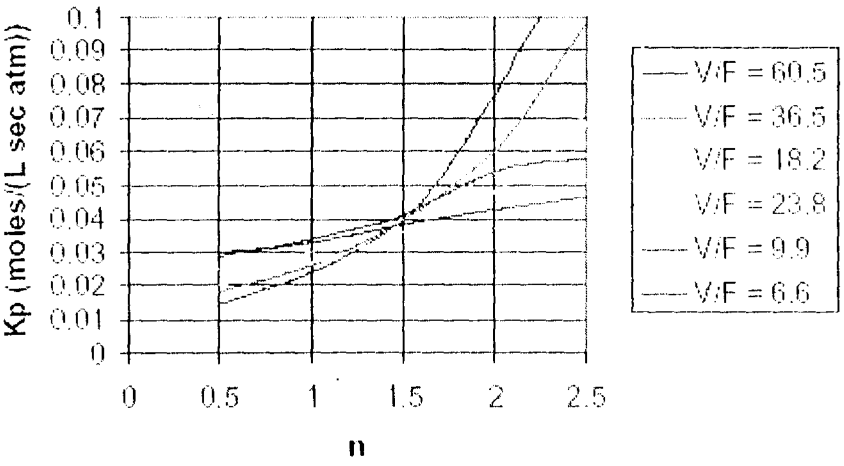


Figure 7.5.3: Determination of rate-constant, k and order, n

The results show at 750°C, k=0.040, and n=1.5. The general rate expression, which is a function of temperature, is determined through a manipulation of Arrhenius' equation, as:

$$r = \exp\left(22.780 - \frac{26600}{T}\right) \cdot P_{ac}^{\ 1.5}$$

Likewise, this whole process is repeated for the decomposition of Ketene. The rate equation from this reaction is:

$$r = \exp\left(19.617 - \frac{25660}{T}\right)$$

Substitution into the continuity equation leads to

$$F \cdot dX_{ke} = \exp\left[\left(22.78 - \frac{26600}{T}\right) \cdot P_{ac}^{\ 1.5} - \exp\left(19.617 - \frac{25660}{T}\right)\right] \cdot dV_R$$

P_{ac} is given earlier as $(1\text{-}0.05X_{ak})^2$ in equation eight.

This last continuity equation can be integrated to:

$$X_{ke} = 0.955\left[1 - \left(2 \cdot \frac{\exp\left(22.78 - \frac{26660}{T}\right) \cdot V_R}{F} + 1\right)^{\frac{-1}{2}}\right] \cdot \exp\left(19.617 - \frac{25660}{T}\right) \cdot \frac{V_R}{F}$$

An application of the derived rate equation, and subsequent continuity equation, is seen in this graph of the selectivity of Ketene (xke vs. VR/F).

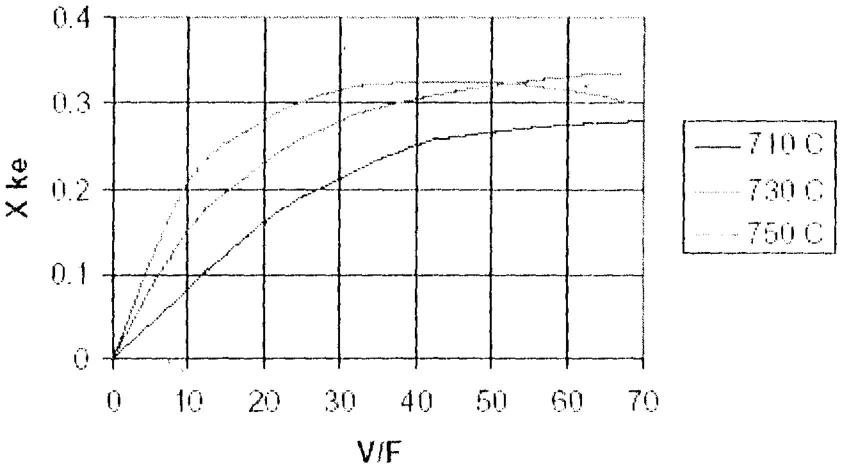


Figure 7.5.4: Experimental values at various temperatures for X_{ke} vs V/F

In the above figure, from Froment, Pijcke, Goethals, 1960; xke vs V/F are calculated at various temperatures. The conversion to KOH occurs at a faster temperatures.

CHAPTER EIGHT

8.0 SAFETY AND QUALITY CONTROL

In designing a plant, safety is one of the major criteria for selection of the best alternative along with economic viability. This is because value placed on the operating personnel and the equipment handled. Operating conditions and equipment in operation are normally dangerous and could lead to serious injury or major damage to the plant as well as disability or death for the victim.

Safety becomes even more pertinent when the materials involved are hazardous. This is the case in the design of the hydrogen synthesis plant. Also sitting of any process plant should be far from the public to avoid environmental pollution. However, the design and quality management plan.

For this design,' the hazards and some of the actions that would lead to these hazards are identified. Safety measures are then recommended to fit into the plant design.

8.1 HAZARDS

To identify hazards present in this design, a material and equipment inventory was taken.

For the material inventory, we have as follows:

(1) KOH, which is volatile at high temperature

(2) Cocoa pod, which is a volatile flammable gas

(3) Steam. KOH which enhances corrosion

Other means of ensuring the safety and quality of process industry, for instance acetic anhydride plant, is by proper maintenance of the equipment involved. Maintenance plays a very important role in optimizing a process also the layout of the plant should be such that there should be reduction in the number of auxiliary equipment and short distance between the feed and the processing section i.e. pipe reduction. The use of natural flow instead of an auxiliary equipment can be part of the optimization i.e. use of gravity instead of pump material for fabrication also plays an important role in process optimization, because a high installation price is cheaper than a higher graded materials for fabrication.

8.2 GENERAL SAFETY MEASURES

1. All materials of construction should be well selected on the basis of corrosion resistance and structural strength. This is to avoid the collapse of any equipment or structures.
2. All pipelines carrying flammable materials must be installed with flame traps.
3. Relief valves should be installed along all lines carrying gases and vapour.
4. Automatic controllers should be installed to control temperature, pressure, and flow rates of the material or equipment involved.
5. A badly maintained plant is a potential hazards. Ensure that a competent staff is responsible for recognizing maintenance and repairs on a planned basis.
6. Signs and placards warning of the hazardous materials should be placed all over the plant.
7. Foam fire extinguishers or those using carbon dioxide should be widely and easily available and ready for immediate use at all times.
8. Plant layout should be such that:
    - i. The storage facilities should be placed away from the plant.
    - ii. All electrical installation should be earthed and insulated and should be kept away from the processing unit.
    - iii. Exit and escape route: - the concept of quality control analysis has been greatly achieved by statistical quality control method. The general intent of SQC is that of sampling units and parts being

produced and essentially determining trends in deviation from production as continuously (affordable and achievable) as possible. There are many important statistical tests, which can be used to determine the quality of products.

8.4 MATERIAL INVENTORY

1 Oxygen (O_2) supports combustion

2 Steam, KOH and Cocoa pod ash enhances corrosion

8.5 EQUIPMENT INVENTORY

1. Reactor:

Allows for catalytic dehydrogenation of acetone at high temperature 700 °C, an endothermic reaction followed by reacting ketene with acetic acid at 80 °C .

With the material and equipment inventory the following hazards were identified as present in the design.

i. Explosion

This is a sudden release of energy as a pressure or a blast wave. It usually occurs on ignition of vapour cloud and also by a pressure build up that leads to the sudden release from a containing vessel. Since, CO, HCl, Cl_2, are in vapour form there is tendency for explosion to occur.

ii. Fire

This is the combustion of material in air the material forms a mixture with are, which would burn once ignited. The combustion of methane could cause if mishandled.

iii. Poisoning

This is related to the toxicity of the material and mainly affects personnel safety. This occurs in two forms first, it exists as health hazard. This continuous handling of these toxic materials exposes personnel to these dangers through emissions and leaks from joints. This leads to accumulation of these materials which would eventually affect the health of such exposed personnel or toxic material identified in the design include KOH, Cocoa pod ash etc.

Other means of ensuring the safety and quality of process industry (e.g. acetic anhydride plant), is by proper maintenance of the equipment involved in optimizing process, also the layout of the plant should be such that there be reduction in the number of auxiliary equipment and short distance between the feed and the processing section i.e. pipe reduction the use of natural flow instead of an auxiliary equipment can be part of the optimization i.e. use of gravity instead of the use of pump material for fabrication also places an important role in process optimization, because a high maintenance price thus the use of a high graded materials for fabrication.

8.6 GENERAL QUALITY CONTROL MEASURES

Maintaining product quality in accordance with acceptable standard has been a major role for industrial instrumentation since its inception decades ago with the ever growing interest in speeding up production, one becomes increasingly aware of the fact that rejects as well as acceptable products can be produced at very high rates.

The concept of quality control analysis has been greatly achieved by statistically quality control (SQC) method. The general intent of SQC is that of sampling units and parts being produced and essentially determining trends in deviation from production as continuously (affordable and achievable) as possible.

There are many important statistical tests, which could be used to determine the quality of products, among techniques is the t-test hypothesis. There are three common hypotheses and their corresponding t equations are given below.

Hypothesis A:

$\mu = \mu_0$ (one sample test)

dF = n-1

$$t = \frac{x - Mo}{s(X)}$$

Hypothesis B:

N1 – N2 = 0 (two sample test – matched pairs)

dF =n-1

$$t = \frac{x - O}{s(X)}$$

N1 = N2 (two sample test)

dF =ni+n2-2

$$t = \frac{x_1 - X_2}{s(X)}$$

Where:

$$s(x) = s(x)/\sqrt{n}$$

$$s(x) = s(x)\sqrt{1/_{n1} + 1/_{n2}}$$

$$s(x) = \sqrt{\frac{\Sigma\left(x - x_1\right)^2 + \Sigma\left(x_2 - x_2\right)}{n1 + n2 - 2}}$$

$$s(x) = \sqrt{\frac{\Sigma(x_1 - x)^2}{n - 1}}$$

df = degree freedom

n = number of observations

However, before a statistical test is applied, the level of significance is generally selected. It is customary to prescribe significance 0f 95% or more before rejecting a specific hypothesis.

2.3.3 FACTORS INFLUENCING RATE OF EXTRACTION

There are four main factors that influence the rate of extraction.

i) Particle size.

Particle size influences the extraction rate in a number of ways. The greater is the inter-facial areas between the solid and liquid and therefore the higher the rate of transfer of material and the smaller is the distance the solute must diffuse within the solid, on the other hand, the surface may not be so effectively used with a very fine material if circulation of the liquid is impede and separation of the particles from the liquid and drainage of the solid residue are more difficult. It is generally desirable that the range of particle size should be small so that each particle requires approximately the same time for extraction and in particular the production of a large amount of the material should be avoided as this may wedge in the interstices of the large particles and impede the flow of solvents.

ii) SOLVENT

The liquid chosen should be a good selective solvent and its viscosity should be sufficiently low for it to circulate freely. Generally, a relatively pure solvent will be used initially but as some extraction process the concentration of solute will increase and the rate of extraction will progressively increase. First, because the concentration gradient will be reduced, and secondly, because the solution will generally become more viscous.

iii) TEMPERATURE

In most cases the solubility of the material which is being extracted will increase with temperature to give a higher rate of extraction. Further, the diffusion coefficient will be expected to increase with rise in temperature and this will also increase the rate of extraction in some cases. The limit temperature is determined by secondary consideration, as for example the necessity to avoid enzyme action during the extraction of sugar.

iv) AGITATION OF THE FLUID

Agitation of the solvent is important because these increase the eddy diffusion and therefore the transfer of material from the surface of the particle to the bulk of the solution further. Agitation of suspension of fine particle prevent sedimentation and more effectively used is made of the interfacial surface.

Table 2.1a Composition percentage % of the 'ash' of cocoa pod husk

Constituents	Percentage (range)
CaO	0.22.059
MgO	0.40-0.52
K_2O	2.85-587
P_2O_5	0.30-0.49

SiO_2	0.06-0.14

The silica (SiO_2) content of c.p.h is low potassium oxide (K_2O), is the only readily soluble oxide in the cocoa pod husk. Extraction of the K_2O in the form of potassium hydroxide (KOH) is achieved by leaching the ash with water.

CHAPTER THREE

3.0 MATERIAL BALANCE

3.1 BASIS: 10,000 kg per day of KOH.

3.2 ASSUMPTIONS:

1. 100 % material recovery

2. Other salts present in the cocoa pod apart from K_2O are small in wt % and considered as inert

3. The dirts present in the peel as dust are considered to be sand particles

The percentage compositions of the cocoa pod in weight percent are:

1. caustic (K_2O) = 61 %

2. water (H2O) = 27 %

3. other salts (Inert) 7 %

4. dirts in form of sand particles 5%

TOTAL 100%

Since K_2O is 61% of the cocoa pod, amount of K_2O to be produced from literature is

10 tonnes/day

Amount of cocoa pod required is:

$$\text{Cocoa\_Pod} := \frac{10 \cdot \frac{\text{tonne}}{\text{day}}}{0.61}$$

cocoa pod (BP) is:

It is assumed that 15% of the KO present in the feed is

$$\text{Cocoa\_Pod} := \frac{\text{Cocoa\_Pod}}{0.95}$$

$\text{CP} := \text{Cocoa\_Pod}$ $\qquad\qquad$ $\text{Cocoa\_Pod} = 17256.255 \frac{\text{kg}}{\text{day}}$

$$\text{CP} = 719.011 \frac{\text{kg}}{\text{hr}}$$

The components present in the cocoa pod and their numbers are:

1 - caustic (K_2O), 2-water , 3 - inert salt , 4 - dirt

efine

Stream indices $j := 1..9$

Species indices (in the order listed for products): $s := 0..3$

BASIS: $CP := 719.011 \frac{kg}{hr}$ of cocoa pod

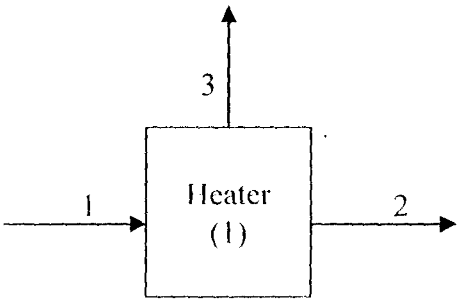
nitialize compositions and flowrates:

$y_{s,j} := 0.00$ $n_{s,j} := 0.00 \cdot \frac{kg}{hr}$ $N_j := 0 \cdot \frac{kg}{hr}$

3.3 Material Balance of the Heater

Stream 1 is the feed charged to the heater. From literature (Andy, 1995), the composition of cocoa pod obtained from laboratory data are:

1. caustic (K_2O) = 61 %
2. water (H2O) = 27 %
3. other salts (Inert) 7 %
4. dirts in form of sand particles 5%



STREAM 1

$y_{s,1} :=$

0.61
0.27
0.07
0.05

1-caustic (K_2O), 2-water , 3 - Inert (salts), 4 - Inert (dirts),

Basis of cocoa pod is:

$N_1 := CP$

Amount of each component in kg:

$n_{s,1} := N_1 \cdot y_{s,1}$

$$n_{s,1} = \begin{pmatrix} 438.596 \\ 194.133 \\ 50.331 \\ 35.951 \end{pmatrix} \frac{kg}{hr} \qquad n_{1,1} = 194.133 \frac{kg}{hr}$$

From literature, (Ossai, 1991), it is assumed that after ashing of the cocoa pod, 99.5% moisture, and 20 % of the inert is removed from the the heater.

STREAM 2

Amount of water in stream 2:

$n_{1,2} := 0.995 \cdot n_{1,1}$

$n_{1,2} = 193.162 \frac{kg}{hr}$

Amount of water removed in the heater stream 2:

$n_{1,3} := 0.005 \cdot n_{1,1}$

$n_{1,3} = 0.971 \frac{kg}{hr}$

Amount of inert present in stream 2:

$n_{2,2} := 0.8 \cdot n_{2,1}$

$n_{2,2} = 40.265 \frac{kg}{hr}$

Amount of inert removed in the heater stream 3:

$n_{1,3} := 0.2 \cdot n_{2,1}$

$n_{1,3} = 10.066 \frac{kg}{hr}$

Amount of dirt in strem 2 leaving the heater:

$n_{3,2} := n_{3,1}$

$n_{3,2} = 35.951 \frac{kg}{hr}$

Amount of K_2O in strem 2 leaving the heater:

$n_{0,2} := n_{0,1}$

$n_{0,2} = 438.596 \frac{kg}{hr}$

Total amount of stream 2 is:

$N_2 := n_{0,2} + n_{1,2} + n_{2,2} + n_{3,2}$

$N_2 = 707.974 \frac{kg}{hr}$

The compositions of K_2O in stream 2:

$y_{0,2} := \frac{n_{0,2}}{N_2}$ $\quad y_{0,2} = 0.62$

The compositions of water in stream 2:

$y_{1,2} := \frac{n_{1,2}}{N_2}$ $\quad y_{1,3} = 0$

The compositions of inert in stream 2:

$y_{2,2} := \frac{n_{2,2}}{N_2}$ $\quad y_{2,2} = 0.057$

The compositions of dirt in stream 2:

$y_{3,2} := \frac{n_{3,2}}{N_2}$ $\quad y_{3,2} = 0.051$

$$y_{s,2} := \begin{pmatrix} 0.62 \\ 0.273 \\ 0.057 \\ 0.051 \end{pmatrix}$$

$$\sum_{s=0}^{3} y_{s,2} = 1$$

STREAM 3

Amount of water in stream 3:

$n_{1,3} := 0.005 \cdot n_{1,1}$

$n_{1,3} = 0.971 \frac{kg}{hr}$

Amount of inert present in stream 3:

$$n_{2,3} := 0.2 \cdot n_{2,1}$$

$$n_{2,3} = 0.003\ \mathrm{kg \cdot s^{-1}}$$

Amount of dirt in strem 3 leaving the heater:

$$n_{3,3} := n_{3,3}$$

$$n_{3,3} = 0\ \mathrm{kg \cdot s^{-1}}$$

Amount of K_2O in strem 3 leaving the heater:

$$n_{0,3} := n_{0,3}$$

$$n_{0,3} = 0\ \mathrm{kg \cdot s^{-1}}$$

Total amount of stream 3 is:

$$N_3 := n_{0,3} + n_{1,3} + n_{2,3} + n_{3,3}$$

$$N_3 = 11.037\,\frac{\mathrm{kg}}{\mathrm{hr}}$$

The compositions of K_2O in stream 3:

$$y_{0,3} := \frac{n_{0,3}}{N_3} \qquad y_{0,3} = 0$$

The compositions of water in stream 3:

$$y_{1,3} := \frac{n_{1,3}}{N_3} \qquad y_{1,3} = 0.088$$

The compositions of inert in stream 3:

$$y_{2,3} := \frac{n_{2,3}}{N_3} \qquad y_{2,3} = 0.912$$

The compositions of dirt in stream 3:

$$y_{3,3} := \frac{n_{3,3}}{N_3} \qquad y_{3,3} = 0$$

$y_{s,3} =$

$$\begin{pmatrix} 0 \\ 0.088 \\ 0.912 \\ 0 \end{pmatrix}$$

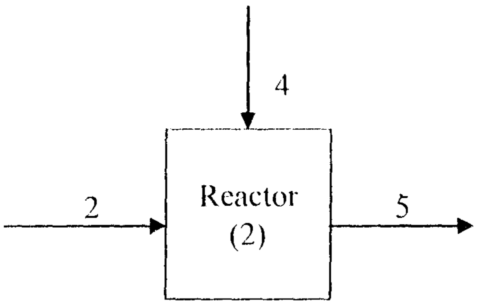
$$\sum_{s=0}^{3} y_{s,3} = 1$$

COMPONENTS	INPUT (kg/hr)	OUTPUT (kg/hr)	
	$n_{s,1} =$ $\frac{kg}{hr}$	$n_{s,2} =$ $\frac{kg}{hr}$	$n_{s,3} =$ $\frac{kg}{hr}$
K_2O,	438.596	438.596	0
H_2O,	194.133	193.162	0.971
Salts,	50.331	40.265	10.066
Dirts,	35.951	35.951	0

$$\sum_{s} n_{s,1} = 719.011\frac{kg}{hr} \qquad \sum_{s} n_{s,2} = 707.974\frac{kg}{hr} \qquad \sum_{s} n_{s,3} = 11.037\frac{kg}{hr}$$

$$\sum_{s} n_{s,1} = 719.011\frac{kg}{hr} \qquad \sum_{s} n_{s,2} + \sum_{s} n_{s,3} = 719.011\frac{kg}{hr}$$

3.4 Material Balance of the Reactor



STREAM 4

From literature the amount of water added for leaching the solute is 6 times the amount of cocoa pod

$N_4 := 3 \cdot N_1$

$N_4 = 2157.032\frac{kg}{hr}$

Amount of water in stream 4:

$n_{1,4} := N_4$

$n_{1,4} = 2157.032 \frac{kg}{hr}$

Amount of inert present in stream 4:

$n_{2,4} := n_{2,4}$

$n_{2,4} = 0 \frac{kg}{hr}$

Amount of dirt in stream 4

$n_{3,4} := n_{3,3}$

$n_{3,4} = 0 \frac{kg}{hr}$

Amount of K_2O in stream 4

$n_{0,4} := n_{0,4}$

$n_{0,4} = 0 \frac{kg}{hr}$

Total amount of stream 4 is:

$N_4 := n_{0,4} + n_{1,4} + n_{2,4} + n_{3,4}$

$N_4 = 2157.032 \frac{kg}{hr}$

The compositions of K_2O in stream 4:

$y_{0,4} := \frac{n_{0,4}}{N_4}$ $\quad y_{0,4} = 0$

The compositions of water in stream 4:

$y_{1,4} := \frac{n_{1,4}}{N_4}$ $\quad y_{1,4} = 1$

The compositions of inert in stream 4:

$y_{2,4} := \frac{n_{2,4}}{N_4}$ $\quad y_{2,4} = 0$

The compositions of dirt in stream 4:

$y_{3,4} := \frac{n_{3,4}}{N_4}$ $y_{3,4} = 0$

$y_{s,4} =$

0
1
0
0

$$\sum_{s=0}^{3} y_{s,4} = 1$$

STREAM 5

Amount of water in streanm 5 is:

$n_{1,5} := N_4 + n_{1,2}$

$n_{1,5} = 2350.194 \frac{kg}{hr}$

Amount of inert present in stream 5:

$n_{2,5} := n_{2,2}$

$n_{2,5} = 40.265 \frac{kg}{hr}$

Amount of dirt in stream 5

$n_{3,5} := n_{3,2}$

$n_{3,5} = 35.951 \frac{kg}{hr}$

Amount of K_2O in stream 5

$n_{0,5} := n_{0,2}$

$n_{0,5} = 438.596 \frac{kg}{hr}$

Total amount of stream 5 is:

$N_5 := n_{0,5} + n_{1,5} + n_{2,5} + n_{3,5}$

$N_5 = 2865.006\,\frac{kg}{hr}$

The compositions of K_2O in stream 4:

$y_{0,5} := \frac{n_{0,5}}{N_5}$ $\qquad y_{0,5} = 0.153$

The compositions of water in stream 4:

$y_{1,5} := \frac{n_{1,5}}{N_5}$ $\qquad y_{1,5} = 0.82$

The compositions of inert in stream 4:

$y_{2,5} := \frac{n_{2,5}}{N_5}$ $\qquad y_{2,5} = 0.014$

The compositions of dirt in stream 4:

$y_{3,5} := \frac{n_{3,5}}{N_5}$ $\qquad y_{3,5} = 0.013$

$$y_{s,5} = \begin{array}{|c|}\hline 0.153 \\ \hline 0.82 \\ \hline 0.014 \\ \hline 0.013 \\ \hline \end{array}$$

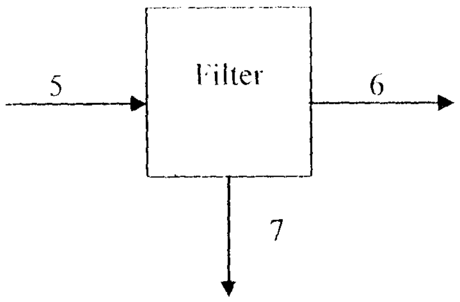
$$\sum_{s=0}^{3} y_{s,5} = 1$$

COMPONENTS	**INPUT (kg/hr)**	**OUTPUT (kg/hr)**	
	$n_{s,2} = \ \frac{kg}{hr}$	$n_{s,4} = \ \frac{kg}{hr}$	$n_{s,5} = \ \frac{kg}{hr}$
K_2O,	438.596	0	438.596
H_2O,	193.162	$2.157 \cdot 10^3$	$2.35 \cdot 10^3$
Salts,	40.265	0	40.265
Dirts,	35.951	0	35.951

$\sum_{s} n_{s,2} = 707.974\,\frac{kg}{hr}$ $\qquad \sum_{s} n_{s,4} = 2157.032\,\frac{kg}{hr}$ $\qquad \sum_{s} n_{s,3} = 11.037\,\frac{kg}{hr}$

$\sum_{s} n_{s,2} + \sum_{s} n_{s,4} = 2865.006\,\frac{kg}{hr}$ $\qquad \sum_{s} n_{s,5} = 2865.006\,\frac{kg}{hr}$

3.5 Material Balance of the Filter



STREAM 6

It is assumed that 10 % of water present in the feed to the filter is removed with the inert (cake)

Amount of water in streanm 6 is:

$$n_{1,6} := 0.90 \cdot n_{1,5}$$

$$n_{1,6} = 2115.175 \frac{kg}{hr}$$

It is assumed that 97 % of inert present in the feed to the filter is removed with the inert (cake)

Amount of inert present in stream 6:

$$n_{2,6} := 0.03 \cdot n_{2,5}$$

$$n_{2,6} = 1.208 \frac{kg}{hr}$$

It is assumed that 99 % of dirt present in the feed to the filter is removed with the inert (cake)

Amount of dirt in stream 6

$$n_{3,6} := 0.01 \cdot n_{3,5}$$

$$n_{3,6} = 0.36 \frac{kg}{hr}$$

It is assumed that 10 % of K_2O present in the feed to the filter is removed with the inert (cake)

Amount of K_2O in stream 6

$$n_{0,6} := 0.9 \cdot n_{0,5}$$

$$n_{0,6} = 394.737 \frac{kg}{hr}$$

Total amount of stream 6 is:

$$N_6 := n_{0,6} + n_{1,6} + n_{2,6} + n_{3,6}$$

$N_6 = 2511.479 \frac{kg}{hr}$

The compositions of K_2O in stream 4:

$y_{0,6} := \frac{n_{0,6}}{N_6} \qquad y_{0,6} = 0.157$

The compositions of water in stream 4:

$y_{1,6} := \frac{n_{1,6}}{N_6} \qquad y_{1,6} = 0.842$

The compositions of inert in stream 4:

$y_{2,6} := \frac{n_{2,6}}{N_6} \qquad y_{2,6} = 0$

The compositions of dirt in stream 4:

$y_{3,6} := \frac{n_{3,6}}{N_6} \qquad y_{3,6} = 0$

$$y_{s,6} = \begin{pmatrix} 0.157 \\ 0.842 \\ 4.81 \cdot 10^{-4} \\ 1.431 \cdot 10^{-4} \end{pmatrix}$$

$$\sum_{s=0}^{3} y_{s,6} = 1$$

STREAM 7

It is assumed that 10 % of water present in the feed to the filter is removed with the inert (cake)

Amount of water in streanm 6 is:

$n_{1,7} := 0.10 \cdot n_{1,5}$

$n_{1,7} = 235.019 \frac{kg}{hr}$

It is assumed that 97 % of inert present in the feed to the filter is removed with the inert (cake)

Amount of inert present in stream 6:

$n_{2,7} := 0.97 \cdot n_{2,5}$

$n_{2,7} = 39.057 \frac{kg}{hr}$

It is assumed that 99 % of dirt present in the feed to the filter is removed with the inert (cake)

Amount of dirt in stream 5

$n_{3,7} := 0.99 \cdot n_{3,5}$

$n_{3,7} = 35.591 \frac{kg}{hr}$

It is assumed that 10 % of dirt present in the feed to the filter is removed with the inert (cake)

Amount of K_2O in stream 7

$n_{0,7} := 0.1 \cdot n_{0,5}$

$n_{0,7} = 43.86 \frac{kg}{hr}$

Total amount of stream 6 is:

$N_7 := n_{0,7} + n_{1,7} + n_{2,7} + n_{3,7}$

$N_7 = 353.527 \frac{kg}{hr}$

The compositions of K_2O in stream 4:

$y_{0,7} := \frac{n_{0,7}}{N_7}$ $y_{0,7} = 0.124$

The compositions of water in stream 4:

$y_{1,7} := \frac{n_{1,7}}{N_7}$ $y_{1,7} = 0.665$

The compositions of inert in stream 4:

$y_{2,7} := \frac{n_{2,7}}{N_7}$ $y_{2,7} = 0.11$

The compositions of dirt in stream 4:

$$y_{3,7} := \frac{n_{3,7}}{N_7} \qquad y_{3,7} = 0.101$$

$$y_{s,7} = \begin{array}{|c|}\hline 0.124 \\ \hline 0.665 \\ \hline 0.11 \\ \hline 0.101 \\ \hline \end{array}$$

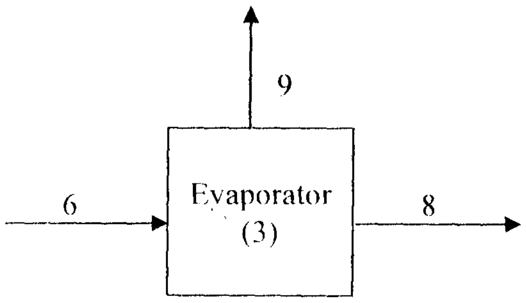
$$\sum_{s=0}^{3} y_{s,7} = 1$$

COMPONENTS	**INPUT (kg/hr)**	**OUTPUT (kg/hr)**	
	$n_{s,5} =$	$n_{s,6} =$	$n_{s,7} =$
K_2O,	438.596	394.737	43.86
H_2O,	$2.35 \cdot 10^3$	$2.115 \cdot 10^3$	235.019
Salts,	40.265	1.208	39.057
Dirts,	35.951	0.36	35.591
	$\frac{kg}{hr}$	$\frac{kg}{hr}$	$\frac{kg}{hr}$

$$\sum_s n_{s,5} = 2865.006 \frac{kg}{hr} \qquad \sum_s n_{s,6} = 2511.479 \frac{kg}{hr} \qquad \sum_s n_{s,7} = 353.527 \frac{kg}{hr}$$

$$\sum_s n_{s,5} = 2865.006 \frac{kg}{hr} \qquad \sum_s n_{s,6} + \sum_s n_{s,7} = 2865.006 \frac{kg}{hr}$$

3.6 Material Balance of the Evaporator



STREAM 8

It is assumed that 95.3 % of water present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of water in streanm 8 is:

$n_{1,8} := 0.047 \cdot n_{1,6}$

$n_{1,8} = 99.413 \frac{kg}{hr}$

It is assumed that 50 % of inert present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of inert present in stream 8:

$n_{2,8} := 0.50 \cdot n_{2,6}$

$n_{2,8} = 0.604 \frac{kg}{hr}$

It is assumed that 90 % of dirt present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of dirt in stream 8

$n_{3,8} := 0.10 \cdot n_{3,6}$

$n_{3,8} = 0.036 \frac{kg}{hr}$

It is assumed that 1 % of K_2O present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of K_2O in stream 8

$n_{0,8} := 0.99 \cdot n_{0,6}$

$n_{0,8} = 390.789 \frac{kg}{hr}$

Total amount of stream 8 is:

$N_8 := n_{0,8} + n_{1,8} + n_{2,8} + n_{3,8}$

$N_8 = 490.843 \frac{kg}{hr}$

The compositions of K_2O in stream 8:

$y_{0,8} := \frac{n_{0,8}}{N_8}$ $y_{0,8} = 0.796$

The compositions of water in stream 8:

$$y_{1,8} := \frac{n_{1,8}}{N_8} \qquad y_{1,8} = 0.203$$

The compositions of inert in stream 8:

$$y_{2,8} := \frac{n_{2,8}}{N_8} \qquad y_{2,8} = 0.001$$

The compositions of dirt in stream 8:

$$y_{3,8} := \frac{n_{3,8}}{N_8} \qquad y_{3,8} = 0$$

$y_{s,8}$

0.796
0.203
$1.23 \cdot 10^{-3}$
$7.324 \cdot 10^{-5}$

$$\sum_{s=1}^{3} y_{s,8} = 0.204$$

STREAM 9

It is assumed that 95.3 % of water present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of water in streanm 9 is:

$$n_{1,9} := 0.953 \cdot n_{1,6}$$

$$n_{1,9} = 2015.762 \frac{kg}{hr}$$

It is assumed that 50 % of inert present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of inert present in stream 9:

$$n_{2,9} := 0.50 \cdot n_{2,6}$$

$n_{2,9} = 0.604 \frac{kg}{hr}$

It is assumed that 90 % of dirt present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of dirt in stream 9

$n_{3,9} := 0.90 \cdot n_{3,6} \quad n_{3,9} = 0 \, kg \cdot s^{-1}$

It is assumed that 1 % of K_2O present in the feed to the evaporator is removed with the condensate leaving the evaoporator

Amount of K_2O in stream 9

$n_{0,9} := 0.01 \cdot n_{0,6}$

$n_{0,9} = 3.947 \frac{kg}{hr}$

Total amount of stream 9 is:

$N_9 := n_{0,9} + n_{1,9} + n_{2,9} + n_{3,9}$

$N_9 = 2020.636 \frac{kg}{hr}$

The compositions of K_2O in stream 9:

$y_{0,9} := \frac{n_{0,9}}{N_9} \qquad y_{1,9} = 0$

The compositions of water in stream 9:

$y_{1,9} := \frac{n_{1,9}}{N_9} \qquad y_{1,9} = 0.998$

The compositions of inert in stream 9:

$y_{2,9} := \frac{n_{2,9}}{N_9} \qquad y_{2,9} = 0$

The compositions of dirt in stream 9:

$y_{3,9} := \frac{n_{3,9}}{N_9} \qquad y_{3,9} = 0$

$$y_{s,9} = \begin{pmatrix} 1.954\cdot10^{-3} \\ 0.998 \\ 2.989\cdot10^{-4} \\ 1.601\cdot10^{-4} \end{pmatrix}$$

$$\sum_{s=0}^{3} y_{s,9} = 1$$

COMPONENTS INPUT (kg/hr) OUTPUT (kg/hr)

	$n_{s,6} =$ (kg/hr)	$n_{s,8} =$ (kg/hr)	$n_{s,9} =$ (kg/hr)
K_2O,	394.737	390.789	3.947
H_2O,	$2.115\cdot10^3$	99.413	$2.016\cdot10^3$
Salts,	1.208	0.604	0.604
Dirts,	0.36	0.036	0.324

$$\sum_s n_{s,6} = 2511.479\,\frac{kg}{hr} \qquad \sum_s n_{s,8} = 490.843\,\frac{kg}{hr} \qquad \sum_s n_{s,9} = 2020.636\,\frac{kg}{hr}$$

$$\sum_s n_{s,6} = 2511.479\,\frac{kg}{hr} \qquad \sum_s n_{s,8} + \sum_s n_{s,9} = 2511.479\,\frac{kg}{hr}$$

SUMMARY OF MATERIAL BALANCE

N (kg/hr)

	0
0	0
1	719.011
2	707.974
3	11.037
4	$2.157\cdot10^3$
5	$2.865\cdot10^3$
6	$2.511\cdot10^3$
7	353.527
8	490.843
9	$2.021\cdot10^3$

N (kg/hr)

	0
0	0
1	719.011
2	707.974
3	11.037
4	$2.157\cdot10^3$
5	$2.865\cdot10^3$
6	$2.511\cdot10^3$
7	353.527
8	490.843
9	$2.021\cdot10^3$

The material balance across each unit:

n = (kg/hr)

	0	1	2	3	4	5	6	7	8	9
0	0	8.596	8.596	0	0	8.596	4.737	43.86	0.789	3.947
1	0	4.133	3.162	0.971	7.032	0.194	5.175	5.019	9.413	5.762
2	0	0.331	0.265	0.066	0	0.265	1.208	9.057	0.604	0.604
3	0	5.951	5.951	0	0	5.951	0.36	5.591	0.036	0.324

The percentage compositions across each unit:

	0	1	2	3	4	5
0	0	0.61	0.62	0	0	0.153
1	0	0.27	0.273	0.088	1	0.82
2	0	0.07	0.057	0.912	0	0.014
3	0	0.05	0.051	0	0	0.013

CHAPTER FOUR

4.0 ENERGY BALANCE

Define: Stream indices $j := 1..9$

Species indices (order: caustic (K_2O), 2-water , 3 - inert salt , 4 - dirt) $s := 0..3$

Index of the heat capacity coefficients $k := 0..4$

Heats of Formation at 25 C

$$h_f := \begin{pmatrix} 1.785 \cdot 10^{3} \\ -1.341164 \cdot 10^{4} \\ -4.94027 \cdot 10^{3} \\ -1.926 \cdot 10^{3} \end{pmatrix} \cdot \frac{kJ}{kg}$$

$$T_i := 0 \cdot K$$

Table 1.Heat capacity constants

$$c := \begin{pmatrix} 29.0063 & 2.49235 \cdot 10^{-3} & -1.86440 \cdot 10^{-5} & 4.79892 \cdot 10^{-8} & -2.87266 \cdot 10^{-11} \\ 34.0471 & -9.65064 \cdot 10^{-3} & 3.29983 \cdot 10^{-5} & -2.04467 \cdot 10^{-8} & 4.30228 \cdot 10^{-12} \\ 19.0223 & 7.96291 \cdot 10^{2} & 0 & 0 & 0 \\ 17.6386 & 6.70055 \cdot 10^{-2} & 0 & 0 & 0 \end{pmatrix}$$

4.1 Energy Balance of the Heater

$T_r := 298 \cdot K$ $T_1 := 303 \cdot K$

Enthalpy of stream 1

$$h(s,T_1) := \left[\sum_{k=0}^{4} \left[\frac{c_{s,k}}{k+1} \cdot \left[\left(\frac{T_1}{K}\right)^{k+1} - \left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right] \cdot \frac{J}{kg} \cdot n_{s,1}$$

$$h(s,T_1) =$$

63.903
32.621
10.809
6.79

$\frac{kJ}{hr}$

Total enthalpy of stream 1

$$H_1 := \sum_{s} h(s,T_1)$$

Enthalpy of stream 2

$T_2 := (273+90)\cdot K$ $\qquad$ $T_r = 298\ K$ $\qquad$ $T_2 = 363\ K$

$$h(s,T_2) := \left[\sum_{k=0}^{4}\left[\frac{c_{s,k}}{k+1}\cdot\left[\left(\frac{T_2}{K}\right)^{k+1}-\left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right]\cdot\frac{J}{kg}\cdot n_{s,2}$$

$$h(s,T_2) = \begin{array}{|r|}\hline 832.079\\ \hline 424.134\\ \hline 118.663\\ \hline 92.966\\ \hline\end{array}\ \frac{kJ}{hr}$$

$$H_2 := \sum_{s} h(s,T_2)$$

$$H_2 = 1467.843\,\frac{kJ}{hr}$$

Enthalpy of stream 3

$T_3 := (273+90)\cdot K$

$$h(s,T_3) := \left[\sum_{k=0}^{4}\left[\frac{c_{s,k}}{k+1}\cdot\left[\left(\frac{T_3}{K}\right)^{k+1}-\left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right]\cdot\frac{J}{kg}\cdot n_{s,3}$$

$$h(s,T_3) = \begin{array}{|r|}\hline 0\\ \hline 2.131\\ \hline 29.666\\ \hline 0\\ \hline\end{array}\ \frac{kJ}{hr}$$

$$H_3 := \sum_{s} h(s,T_3)$$

$$H_3 = 31.797\,\frac{kJ}{hr}$$

Input Enthalpy

$$H_1 = 114.123 \frac{kJ}{hr}$$

Output Enthalpy:

$$H_2 + H_3 = 1499.64 \frac{kJ}{hr}$$

Heat Duty of the heater is:

$$Q_1 := H_2 + H_3 - H_1$$

$$Q_1 = 1385.518 \frac{kJ}{hr}$$

COMPONENTS INPUT (kg/hr) OUTPUT (kg/hr)

	$h(s,T_1) =$	$(h(s,T_2) + h(s,T_2)) - h(s,T_1) =$	$(h(s,T_2)) =$	$h(s,T_3) =$
K_2O,	0 $\frac{kJ}{hr}$	0 $\frac{kJ}{hr}$	0 $\frac{kJ}{hr}$	0 $\frac{kJ}{hr}$
H_2O,	2.131	2.131	2.131	2.131
Salts,	29.666	29.666	29.666	29.666
Dirts,	0	0	0	0

$H_1 = 114.123 \frac{kJ}{hr}$ $Q_1 = 1385.518 \frac{kJ}{hr}$ $H_2 = 1467.843 \frac{kJ}{hr}$ $H_3 = 31.797 \frac{kJ}{hr}$

$H_1 + Q_1 = 1499.64 \frac{kJ}{hr}$ $H_2 + H_3 = 1499.64 \frac{kJ}{hr}$

4.2 Energy Balance of the Reactor

$T_r := 298 \cdot K$ $T_4 := 303 \cdot K$

Enthalpy of stream 4

$$h(s,T_4) := \left[\sum_{k=0}^{4} \left[\frac{c_{s,k}}{k+1} \cdot \left[\left(\frac{T_4}{K}\right)^{k+1} - \left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right] \cdot \frac{J}{kg} \cdot n_{s,4}$$

$h(s,T_4) =$

$$\begin{bmatrix} 0 \\ 362.458 \\ 0 \\ 0 \end{bmatrix} \frac{kJ}{hr}$$

$$H_4 := \sum_{s} h(s,T_4)$$

$$H_4 = 362.458 \frac{kJ}{hr}$$

$$h(s,\theta) := \left[\sum_{k=0}^{4} \left[\frac{c_{s,k}}{k+1} \cdot \left[\left(\frac{\theta}{K}\right)^{k+1} - \left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right] \cdot \frac{J}{kg} \cdot n_{s,5}$$

Estimation of the outlet temperature of the reactor

Estimate: $t := 500 \cdot K$

Objective function: $f(t) := \sum_{s=0}^{3} (h(s,t)) - (H_2 + H_4)$

Solve: $T_5 := \text{root}(f(t),t)$ $T_5 = 317\ K$

Enthalpy of stream 5

$h(s,T_5) =$

$$\begin{bmatrix} 246.352 \\ 1.524 \cdot 10^{3} \\ 33.766 \\ 26.499 \end{bmatrix} \frac{kJ}{hr}$$

$$H_5 := \sum_{s} h(s,T_5)$$

$$H_5 = 1830.301 \frac{kJ}{hr}$$

$$(H_2 + H_4) = 1830.301 \frac{kJ}{hr}$$

COMPONENTS	**INPUT (kg/hr)**		**OUTPUT (kg/hr)**
	$h(s,T_2)$ $\frac{kJ}{hr}$	$(h(s,T_4))$ $\frac{kJ}{hr}$	$h(s,T_5)$ $\frac{kJ}{hr}$
K_2O,	832.079	63.903	246.352
H_2O,	$5.16 \cdot 10^3$	394.916	$1.524 \cdot 10^3$
Salts,	118.663	8.647	33.766
Dirts,	92.966	6.79	26.499

$H_2 = 1467.843 \frac{kJ}{hr}$ $\qquad H_4 = 362.458 \frac{kJ}{hr}$ $\qquad H_5 = 1830.301 \frac{kJ}{hr}$

$H_2 + H_4 = 1830.301 \frac{kJ}{hr}$ $\qquad H_5 = 1830.301 \frac{kJ}{hr}$

4.3 Energy Balance of the Filter

$T_r := 298 \cdot K \qquad T_6 := T_5$

Enthalpy of stream 6

$$h(s,T_6) := \left[\sum_{k=0}^{4}\left[\frac{c_{s,k}}{k+1}\cdot\left[\left(\frac{T_6}{K}\right)^{k+1} - \left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right]\cdot\frac{J}{kg}\cdot n_{s,6}$$

$h(s,T_6) =$

221.717	$\frac{kJ}{hr}$
$1.371 \cdot 10^3$	
1.013	
0.265	

$$H_6 := \sum_s h(s,T_6)$$

$H_6 = 1594.311 \frac{kJ}{hr}$

Enthalpy of stream 7

$T_r := 298 \cdot K \qquad T_7 := T_5$

$$h(s,T_7) := \left[\sum_{k=0}^{4}\left[\frac{c_{s,k}}{k+1}\cdot\left[\left(\frac{T_7}{K}\right)^{k+1} - \left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right]\cdot\frac{J}{kg}\cdot n_{s,7}$$

$h(s,T_7) =$

$$\begin{bmatrix} 24.635 \\ 152.368 \\ 32.753 \\ 26.234 \end{bmatrix} \cdot \frac{kJ}{hr}$$

$$H_7 := \sum_{s} h(s,T_7)$$

COMPONENTS	**INPUT (kg/hr)**	**OUTPUT (kg/hr)**	
	$h(s,T_5) =$	$h(s,T_6) =$	$h(s,T_7) =$
K_2O,	24.635 $\frac{kJ}{hr}$	24.635 $\frac{kJ}{hr}$	24.635 $\frac{kJ}{hr}$
H_2O,	152.368	152.368	152.368
Salts,	32.753	32.753	32.753
Dirts,	26.234	26.234	26.234
	$H_5 = 1830.301 \frac{kJ}{hr}$	$H_6 = 1594.311 \frac{kJ}{hr}$	$H_7 = 235.99 \frac{kJ}{hr}$

$H_5 = 1830.301 \frac{kJ}{hr}$ $\quad$ $H_6 + H_7 = 1830.301 \frac{kJ}{hr}$

4.4 Energy Balance of the Evaporator

$T_r := 298 \cdot K$ $\qquad$ $T_9 := (100 + 273) \cdot K$

Enthalpy of stream 6

$$h(s,T_9) := \left[\sum_{k=0}^{4}\left[\frac{c_{s,k}}{k+1}\cdot\left[\left(\frac{T_9}{K}\right)^{k+1} - \left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right]\cdot\frac{J}{kg}\cdot n_{s,9}$$

$h(s,T_9) =$

$$\begin{bmatrix} 8.644 \\ 5.112\cdot10^{3} \\ 2.072 \\ 0.974 \end{bmatrix} \frac{kJ}{hr}$$

$$H_9 := \sum_{s} h(s,T_9)$$

$H_9 = 5123.633 \frac{kJ}{hr}$

Enthalpy of stream 8

$T_r := 298 \cdot K \qquad T_8 := (100 + 273) \cdot K$

$$h(s,T_8) := \left[\sum_{k=0}^{4}\left[\frac{c_{s,k}}{k+1}\cdot\left[\left(\frac{T_8}{K}\right)^{k+1} - \left(\frac{T_r}{K}\right)^{k+1}\right]\right]\right]\cdot\frac{J}{kg}\cdot n_{s,8}$$

$$h(s,T_8) = \begin{array}{|r|}\hline 855.741\\\hline 252.111\\\hline 2.072\\\hline 0.108\\\hline\end{array}\ \frac{kJ}{hr}$$

$$H_8 := \sum_{s} h(s,T_8)$$

Heat suppply by steam in the evaporator is:

$$Q_2 := (H_8 + H_9) - H_6$$

$Q_2 = 4639.354\,\frac{kJ}{hr} \qquad kgmol \equiv mol \qquad kJ \equiv 1000 \cdot J$

COMPONENTS INPUT (kg/hr) OUTPUT (kg/hr)

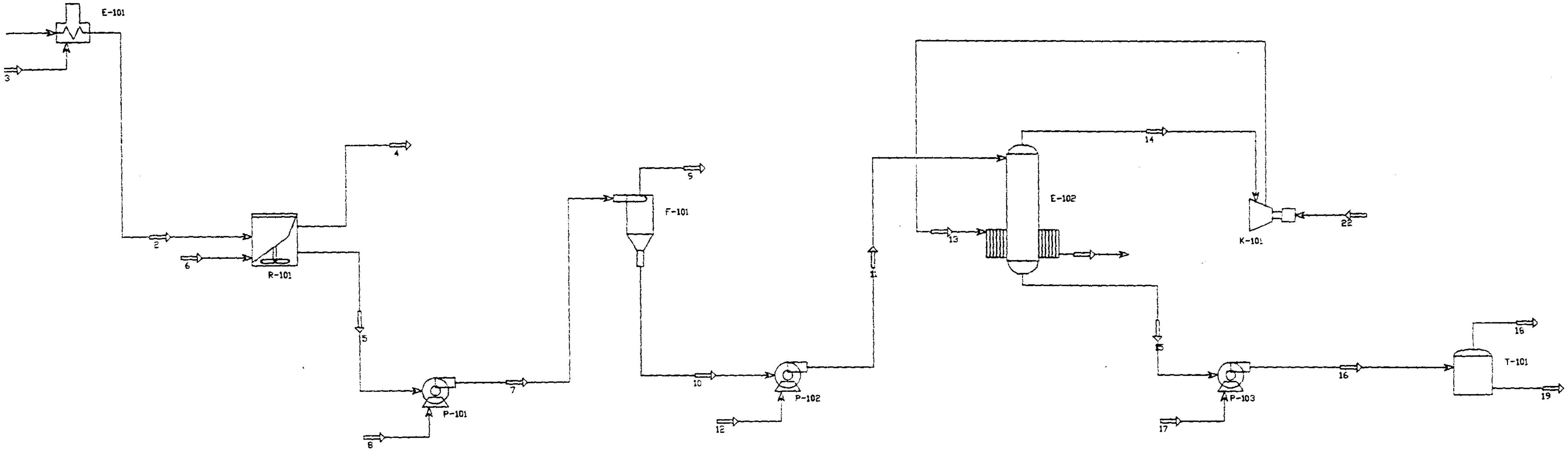
$h(s,T_6) = \qquad (h(s,T_8) + h(s,T_9)) - h(s,T_6) = (h(s,T_8)) = \qquad h(s,T_9) =$

K_2O,
H_2O,
Salts,
Dirts,

$$\begin{array}{|r|}\hline 855.741\\\hline 252.111\\\hline 2.072\\\hline 0.108\\\hline\end{array}\ \frac{kJ}{hr}\quad \begin{array}{|r|}\hline 855.741\\\hline 252.111\\\hline 2.072\\\hline 0.108\\\hline\end{array}\ \frac{kJ}{hr}\qquad \begin{array}{|r|}\hline 855.741\\\hline 252.111\\\hline 2.072\\\hline 0.108\\\hline\end{array}\ \frac{kJ}{hr}\quad \begin{array}{|r|}\hline 855.741\\\hline 252.111\\\hline 2.072\\\hline 0.108\\\hline\end{array}\ \frac{kJ}{hr}$$

$H_6 = 1594.311\,\frac{kJ}{hr} \quad Q_2 = 4639.354\,\frac{kJ}{hr} \qquad H_8 = 1110.032\,\frac{kJ}{hr} \quad H_9 = 5123.633\,\frac{kJ}{hr}$

$H_6 + Q_2 = 6233.665\,\frac{kJ}{hr} \qquad H_8 + H_9 = 6233.665\,\frac{kJ}{hr}$



ter 1	H101		
ctor	R101		
er	F101		
orator	E102		
age Tank	T101		
ressor	K101		
1	P102		
2	P102		
2	P103		

TITLE: DESIGN OF PROCESS PLANT PRODUCING 10 TONNES OF CAUSTIC POTASH FROM COCOA POD HUSK PER DAY
CLIENT: FEDERAL UNIVERSITY OF TECHNOLOGY, MINNA
DESIGNED BY: ONIBON MOSES K: 99/8322EH
CHECKED BY: PROF. K. R. ONIFADE
APPROVED BY: DEPARTMENT OF CHEMICAL ENGINEERING, FEDERAL UNIVERSITY OF TECHNOLOGY, MINNA
DATE: NOVEMBER, 2005

CHAPTER SIX

6.0 EQUIPMENT DESIGN

6.1 DESIGN OF A TRIPPLE EFFECT FORWARD FEED EVAPORATOR FOR THE CONCENTRATION OF KOH SOLUTION:

Design: KOH solution is to be concentrated from 17Wt% to a final concentration of 80Wt% The feed is available at 50993 kg/hr and 27 ^{0}C. Steam is supplied at 103.66 kpa (Abs.) to the first effectand a vacuum of 74.2(Abs.) mm of Hg is maintained in the last effect. Heat losses by radia and by entertainment are neglected. Condensate are assumed to enter each effect at its saturation temperature.

$Feed := N_6$

$X_{F1} := 0.17$

KOH in feed

$GF := Feed \cdot X_{F1}$ $\qquad GF = 426.951 \frac{kg}{hr}$

Feed Rate

$Feed = 0.698\ kg \cdot s^{-1}$ $\qquad W_f := Feed$

Bottom outlet from third effect BO3 =

$$BO3 := \frac{GF}{0.9}$$

$BO3 = 0.132\ kg \cdot s^{-1}$

Total evaporation TE

$TE := Feed - BO3$

$TE = 0.566\ kg \cdot s^{-1}$

Assuming equal evaporation in all the three effects.

W_1 Evaporation rate in first effect

W_2- Evaporation rate in second effect

W_3 Evaporation rate in the third effect

$W_1 = W_2 = W_3 = (12.9/3) = 4.3$ Kg/sec

$$W_1 = W_2 = W_3 = \frac{TE}{3}$$

$$W_1 := \frac{TE}{3}$$

$$W_2 := W_1 \qquad W_3 := W_2$$

$$W_1 = 0.189\ \mathrm{kg \cdot s^{-1}}$$

$$W_2 = 0.189\ \mathrm{kg \cdot s^{-1}}$$

$$W_3 = 0.189\ \mathrm{kg \cdot s^{-1}}$$

Outlet from first effect = $W_F - W_1$

$$OFE := W_f - W_1$$

$$OFE = 0.509\ \mathrm{kg \cdot s^{-1}}$$

Outlet from second effect = $W_F - W_1 - W_2$

$$OFSE := W_f - W_1 - W_2$$

$$OFSE = 0.32\ \mathrm{kg \cdot s^{-1}}$$

Outlet from third effect = $W_F - W_1 - W_2 - W_3$

$$OFTE := W_f - W_1 - W_2 - W_3$$

$$OFTE = 0.132\ \mathrm{kg \cdot s^{-1}}$$

Concentration of KOH in outlet from first effect =

$$CGOF := \left(\frac{W_f \cdot X_{F1}}{9.86 \cdot \frac{kg}{sec}} \right) \cdot 100 \cdot \%$$

$CGOE = 1.203\,\%$

Concentration of KOH solution in outlet from second effect

$$CGOE := \left(\frac{W_f \cdot X_{F1}}{5.56 \cdot \frac{kg}{sec}} \right) \cdot 100 \cdot \%$$

$CGOE = 2.133\,\%$

Concentration of KOH solution in outlet from third effect

$$CGOE := \left(\frac{W_f \cdot X_{F1}}{1.26 \cdot \frac{kg}{sec}} \right) \cdot 100 \cdot \%$$

$CGOE = 9.413\,\%$

Saturation temperature of inlet steam Ts = 100 ^{O}C

Boiling point of water in the third effect = 12 ^{O}C

Temperature difference =100 – 12

$T_s := 100$

$T_{w3} := 12$

$\Delta T := T_s - T_{w3}$

$\Delta T = 88$

BOILING POINT RISE FOR KOH SOULTION

1) BPR for the third effect at 79 Wt% KOH = 10.4^{O}C

2) BPR for the second effect at 17.8 Wt% KOH = 1.5^{O}C

3) BPR for the first effect at 10 Wt% KOH = $0.5^{O}C$

Sum of the BPR of the glycerol solution = 10.4 + 1.5 + 0.5

$BPR_1 := 10.4$

$BPR_2 := 1.5$

$BPR_3 := 0.5$

$S := BPR_3 + BPR_2 + BPR_1$

$S = 12.4$

Effective temperature difference = 88 − 12.4

$\Delta T_E := \Delta T - S$

$\Delta T_E = 75.6$

Now we assume that the amount of heat transfer from steam to the solution is the same in all thethree effects i.e. Q (heat transfer in the first effect) =

Q (heat transfer in the second effect) = Q(heat transfer in the third effect)

OVERALL HEAT TRANSFER COEFFICIENT

Since it is a vertical effect evaporator, the overall heat transfer coefficient can safely be assumed as follows:

In the first effect U_1 = 2325 W/mΥØC

In the second effect U_2 = 1275 $W/m^2\ ^{O}C$

In the third effect U_3 = 1031 $W/m^2\ ^{O}C$

In the first effect U_1 =

$$U_1 := 2325 \cdot \frac{W}{m \cdot {}^{o}C}$$

In the second effect U_2 =

$$U_2 := 1275 \cdot \frac{W}{m \cdot {}^{o}C}$$

In the third effect U_3 =

$$U_3 := 1031 \cdot \frac{W}{m \cdot {}^{o}C}$$

Now as we have assumed heat transfer rate's to be equal , we have

$$Q_1 = Q_2 = Q_3$$

$$U_1 \cdot A_1 \cdot \Delta t_1 = U_2 \cdot A_2 \cdot \Delta t_2 = U_3 \cdot A_3 \cdot \Delta t_3$$

We design the triple effect evaporator such that the heating area in all the three effect is the same

$$A_1 = A_2 = A_3$$

$$U_1 \cdot \Delta t_1 = U_2 \cdot \Delta t_2 = U_3 \cdot \Delta t_3$$

$$\frac{\Delta t_2}{\Delta t_1} = \frac{U_1}{U_2}$$

$$\Delta t_{2\_1} = \frac{\Delta t_2}{\Delta t_1}$$

$$\Delta t_{2\_1} := \frac{U_1}{U_2}$$

$\Delta t_{2\_1} = 1.824$ $\qquad$ $\Delta t_1 = 1.8 \cdot \Delta t_2$

$$\frac{\Delta t_3}{\Delta t_2} = \frac{U_2}{U_3}$$

$$\Delta t_{3\_2} = \frac{\Delta t_3}{\Delta t_2}$$

$$\Delta t_{3\_2} := \frac{U_2}{U_3}$$

$$\Delta t_{3\_2} = 1.237$$

$$\Delta t_3 = 1.237 \cdot \Delta t_2$$

$\Delta t_1 + \Delta t_2 + \Delta t_3 = \Delta T_E$

$1.8 \cdot \Delta t_2 + \Delta t_2 + 1.237 \cdot \Delta t_2 = \Delta T_E$

$4.3037 \cdot \Delta t_2 = \Delta T_E \qquad \Delta T_E = 75.6$

$$\Delta t_2 := \frac{\Delta T_E}{4.3037}$$

$\Delta t_2 = 17.566$

$\Delta t_1 := 1.8 \cdot \Delta t_2$

$\Delta t_1 = 31.619$

$\Delta t_3 := 1.237 \cdot \Delta t_2$

$\Delta t_3 = 21.729$

ACTUAL BOILING POINTS IN EACH EFFECT

First effect:

$T_1 := T_s - \Delta t_1$

$T_1 = 68.381$

Second effect:

$T_2 := T_1 - BPR_1 - \Delta t_2$

$T_2 = 40.414$

Third effect:

$T_3 := T_2 - BPR_2 - \Delta t_3$

$F_3 = 17.185$

HEAT BALANCE

FIRST EFFECT:

$$W_s \cdot \lambda_s + W_f \cdot H_f = W_1 \cdot H_1 + \left(W_f - W_1\right) \cdot h_1$$

latent heat of steam

$$\lambda_s := 2257.86 \cdot \frac{kJ}{kg}$$

H_F - Enthalpy of feed at inlet temperature (27°C)

$$H_f = Cp_f \cdot \left(T_f - 0\right)$$

$$Cp_f := 4.18$$

$$T_f := 27$$

$$H_f := \left[0.578 \cdot Cp_f \cdot \left(T_f - 0\right)\right] \cdot \frac{kJ}{kg}$$

$$H_f = 65.233 \frac{kJ}{kg}$$

H_1- Enthalpy of vapor leaving the first effect = H_{2S} + $(C_p)_{steam}$ × $(BPR_1)_{supethea}$

$$H_{2s} := 2640$$

$$C_p := 1.884$$

$$BPR_1 := 0.5$$

$$H_1 := \left(H_{2s} + C_p \cdot BPR_1\right) \cdot \frac{kJ}{kg}$$

$$H_1 = 2640.942 \frac{kJ}{kg}$$

H_{2s}- Enthalpy of steam at 78°C = 2640 kJ / kg

$$H_{2s} := 2640 \cdot \frac{kJ}{kg}$$

$$H_{2s} = 2640\,\frac{kJ}{kg}$$

$(C_p)_{steam}$ at 78°C = 1.884 kJ/ kg

$$C_{Ps} := 1.884\,\frac{kJ}{kg}$$

h_1 - enthalpy of outlet from first effect at 78.5°C = $C_{p1} \times (t_1 - 0)$

$$h_1 := \left(0.62 \cdot C_{Ps} \cdot T_1\right) \cdot \frac{kJ}{kg}$$

$$h_1 = 79.874\,\frac{kJ}{kg} \qquad C_{Ps} := 1.884 \cdot \frac{kJ}{kg}$$

$$W_s \cdot \lambda_s + W_f \cdot H_f = W_1 \cdot H_1 + \left(W_f - W_1\right) \cdot h_1$$

$$\lambda_s = 2257.86\,\frac{kJ}{kg} \qquad W_f = 0.698\,\frac{kg}{sec} \qquad H_f = 65.233\,\frac{kJ}{kg}$$

$$h_1 = 79.874\,\frac{kJ}{kg} \qquad H_1 = 2640.942\,\frac{kJ}{kg}$$

$$W_s \cdot 2258 + (14.165 \cdot 65.233) = W_1 \cdot 2641 + \left(14.165 - W_1\right) \cdot 79.874$$

$$22598 \cdot W_s + 924.025 = 2641 \cdot W_1 + 1131 - 79.874 \cdot W_s$$

$$9.1267373652397927601 \cdot 10^{-3} + .11645712468461549791 \cdot W_1$$

$$W_s = 0.11645 \cdot W_1 + 0.009126$$

$$(14.165 \cdot 79.874) = 1131.415$$

SECOND EFFECT:

$$W_1 \cdot \lambda_1 + \left(W_f - W_1\right)h_1 = W_2 \cdot H_2 + \left(W_f - W_1 - W_2\right)h_2$$

$$4.308 \cdot 2258 + (14.165 - 4.308) \cdot 79.874 = 4.308 \cdot 2640 + (14.165 - 4.308 - 4.308)362$$

$$h_{2s} := 325 \cdot \frac{kJ}{kg}$$

Latent heat of steam at 78°C =

$$\lambda_1 := H_1 - h_{2s}$$

$$\lambda_1 = 2315.942\,\frac{kJ}{kg}$$

H_{3s} - Enthalpy of steam vapor at 44.2 °C = 2580 kJ /kg

$H_{3s} := 2580 \cdot \frac{kJ}{kg}$

H_2 - Enthalpy of vapor leaving the second effect = H_{3s} + $(C_p)_{steam} \times (BPR_2)_{supreheat}$

$H_2 := H_{3s} + C_{ps} \cdot BPR_2$

$H_2 = 2582.826 \frac{kJ}{kg}$ $\qquad Cp_2 := 4.18$

h_2-Enthalpy of outlet from the second effect at 45.7°C = $C_{p2} \times (t_2 - 0)$

$h_2 := (0.6 \cdot Cp_2 \cdot T_2) \cdot \frac{kJ}{kg}$

$h_2 = 101.359 \frac{kJ}{kg}$

$W_1 \cdot \lambda_1 + (W_f - W_1)h_1 = W_2 \cdot H_2 + (W_f - W_1 - W_2)h_2$

$\lambda_1 = 2315.942 \frac{kJ}{kg}$ $\qquad W_f = 0.698\ kg \cdot s^{-1}$ $\qquad h_1 = 79.874 \frac{kJ}{kg}$

$W_f = 0.698\ kg \cdot s^{-1}$ $\qquad H_2 = 2582.826 \frac{kJ}{kg}$ $\qquad h_2 = 101.359 \frac{kJ}{kg}$

$2316 \cdot W_1 + (14.165 - W_1) \cdot 79.874 = W_2 \cdot 2583 + (14.165 - W_1 - W_2) \cdot 101.539$

$.13127831618302879155 + 1.0615126632772446009 \cdot W_2$

$W_1 = 0.1313 + 1.0615 \cdot W_2$

THIRD EFFECT:

$W_2 \cdot \lambda_2 + (W_f - W_1 - W_2) \cdot h_2 = W_3 \cdot H_3 + (W_f - W_1 - W_2 - W_3) \cdot h_3$

Latent heat of steam at 44.2°C =

$\lambda_2 = H_2 - h_{3s}$

$h_{3s} := 190 \cdot \frac{kJ}{kg}$

$\lambda_2 := H_2 - h_{3s}$

$\lambda_2 = 2392.826 \frac{kJ}{kg}$

CHAPTER NINE

9.0 PROCESS CONTROL AND INSTRUMENTATION

9.1 INTRODUCTION

In the design of an industrial plant, the methods, which will be used, for plant operation and control help determine many of design variables, for example, the extent of instrumentation can be a factor in choosing the type of process and setting the labor requirements. It should be remembered that maintenance work would be necessary to keep the installed equipment and facilities in good operating condition. The importance of such factors which are directly related to plant operation and control must be recognized and take into proper account during the development of a design project.

Processes may be controlled more precisely to give more uniform and higher - quality products by the application of automatic control, often leading to higher profits. Therefore, any process with an input and output which may be flow, pressure, liquid level, temperature, composition or any other inventory, environmental or quality variable that is to be held at a desired value must have some measure of control applied to it.

Change in output may occur:

1. Randomly as caused by changes in weather or raw material quality.
2. Diurnally with ambient temperature
3. Manually when operators change production rate.
4. Stepwise when equipment is switched in or out of service, or
5. Cyclically as a result of oscillations in other control loops.

Variation in any of the ways stated above would drive the output (controlled variable) further away from the set point (desired value) thus requiring a corresponding variable to bring it back (manipulative variable).

9.2 INSTRUMENTATION AND CONTROL OBJECTIVES

The key objectives adhered to in the specification of the instrumentation and control schemes are:

1. Safe plant operation
    - i. To keep the process variables within known safe operations limits.
    - ii. To defect dangerous situations as they develop and to provide alarms and automatic shut – down systems
    - iii. To provide interlocks and alarms to prevent dangerous operation procedures
2. Production rate. To achieve the design producer output
3. Product quality. To maintain the product composition within the specific quality standards
4. Cost. They operate at the lowest production cost, but not to the detriment of the product quality.

In the plant design some of the variables needed to be monitored and controlled are, the flow rate, temperature, pressure and composition.

- i. Flow rates/feed ratio control: a feed ratio controller is applied between the methane stream and the oxygen stream with the aim of maintaining a complete conversion of methane to the desired products. This is necessary to avoid excess oxygen being built up in the reactor.
- ii. Temperature control reaction: temperature is controlled by regulating the flow rate of the reaction into it. This can also be controlled by heat transfer. The reaction is exothermic. It produces heat, which tends to raise reaction temperature thereby increasing reaction rate and producing more heat. This positive feedback is countered by negative feedback in the cooling system, which removes more heat as reactor temperature rises.

The temperature controller, which in turn operates the coolant value, to counter the rise or drop in temperature in the sector.

iii. Pressure control: pressure sensing is quite straightforward with the aid of pneumatic instrumentation such as Bourdon gauge, diaphragms and bellows. These sensors measure absolute pressure and pressure differences between two levels. Therefore, pressure control is achieved by manipulating the airflow rate in the compressor to avoid deviation from set point.

iv. Composition control: first requirement here is to establish proper stoichiometry of the reactants in propositions needed to satisfy the reaction chemistry and also the desired output product. This is achieved by setting input flow rates in ratio to one another, or a composition measurement (analyzer) can be used to trim the ratios to the right proportion

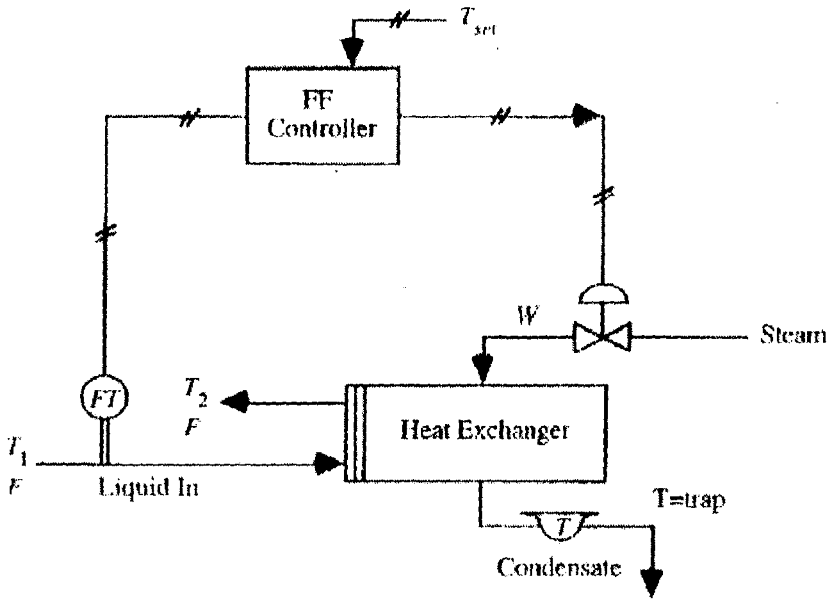


Fig. 9.1 Feedforward control of a heat exchanger.

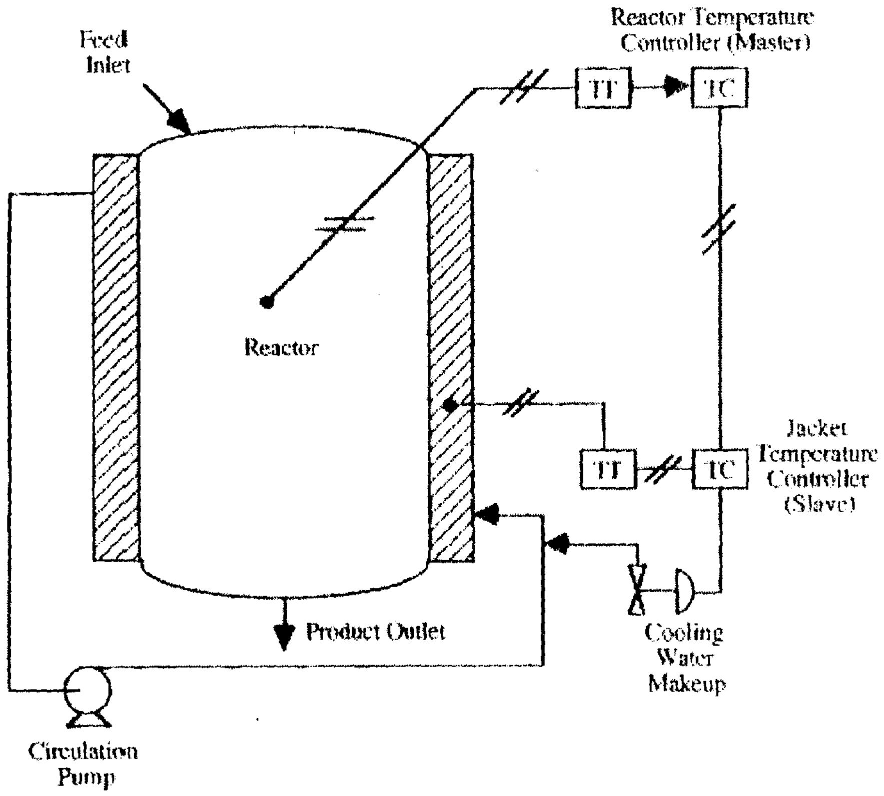


Fig. 9.2 Cascade control of an exothermic chemical reactor.

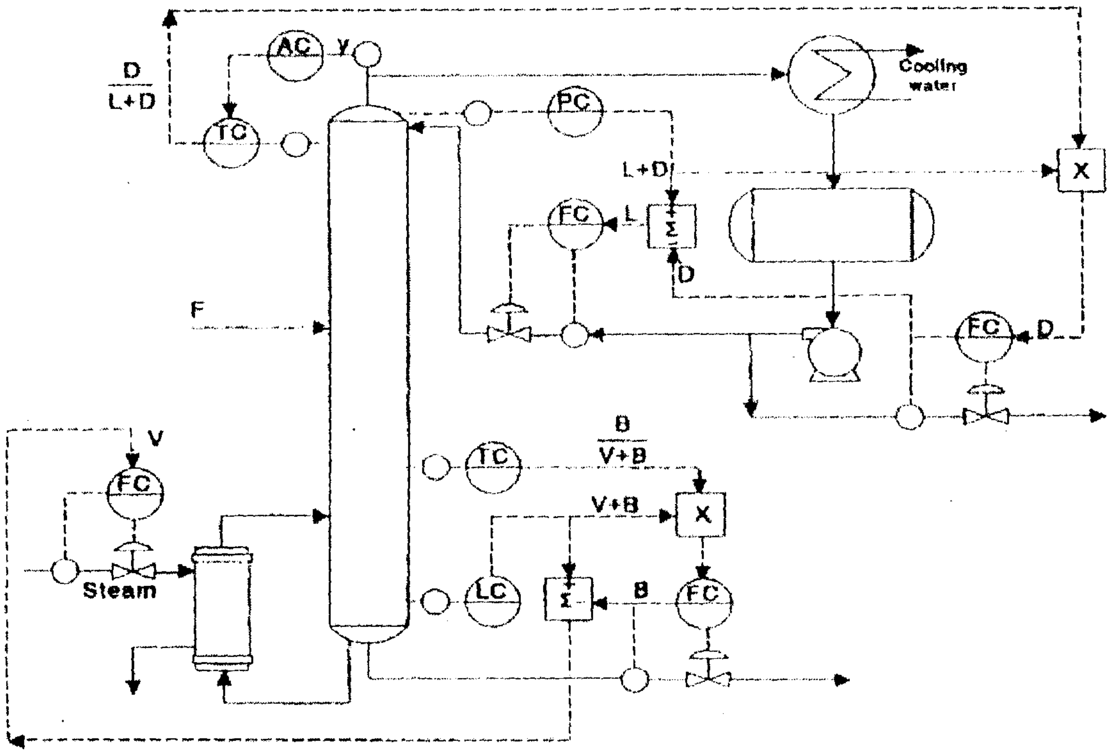


Fig. 9.3 Cascade control of the distillation column.

CHAPTER TEN

10.0 ENVIRONMENTAL ACCEPTABILITY

10.1 IDENTIFICATION OF POSSIBLE POLLUTANTS

Any project involving new technology or hazardous material such as hydrogen synthesis plant requires a rational approach of assessing the methodologically sound in order to identify systematically all possible pollutants. Suitability or acceptability of any process plant depends on its impact to the life, properties and environment. The possible pollutants emanating from hydrogen synthesis gas plant are: carbon monoxide, ethylene, methane and acetone.

10.2 TREATMENTS OF POSSIBLE POLLUTANTS

Therefore to make this process suitable and acceptable to the environment, there is need for these pollutants to be controlled. Presently, the following measures can be used to control possible pollutants from acetic anhydride plant.

i. Recycling of unconverted acetone minimizes loss of acetone as waste

ii. Absorption process this is used in based on dissolution of poisonous impurities liquid (physical absorption), or selective absorption using chemical active absorbent such as water, solutions of NaOH, Na_2CO_3 and $KMnO_4$ the major equipment used in absorption purification process is packed bed absorption reactor

iii. Catalyzed conversion of pollutants in this method, the pollutants are converted to fewer sometimes harmless and products in presence of a catalyst. These products are than concentrated removed from the gas stream or passed to the atmosphere.

The major advantage of this process is its high purification coefficient of about 99%. Depending on the operating condition.

iv. Absorption using solid absorbents: the purification technique based on selective removal of poisonous impurities from the gas, using solid absorption can be physical or chemical (chemisorptions) process. Commonly used absorbents include highly porous materials such as activated coal, silica gel, synthetic zeolite, etc.

Absorption of gaseous impurities could be performed using the batch reaction with the absorbent placed on the grid. The flue gases are passed to top-down at a velocity determined by the hydraulic resistance of the absorbent layer with the absorbent lose their absorptive due to saturation of their active sites by the solute gas (impurities) they can therefore be regenerated by passing through them hot water vapour air or inert gas-Nitrogen.

10.3 WASTE WATER TREATMENT

This is carried out by using both physically and chemical treatments. The physical treatment involves the removal of large floating or suspended particles from the waste water. This is achieved by gravity settling. This is further processed by filtration to clarify the effluents. The next stage is the chemical treatment. It involves the neutralization of acids and alkali, conversion of ions to poorly soluble compounds, co-precipitation of inorganic substances, oxidation, Electrolysis, catalytic oxidation, etc. these methods are mainly used to deactivate and remove the impurities of inorganic compounds.

10.4 THERMAL POLLUTION CONTROL

The waste heat should be reclaimed by raising the thermal potentials of the gas, by setting up cogeneration technology; the water heat could be converted to mechanical power.

CHAPTER ELEVEN

11.0 PLANT START-UP AND SHUT DOWN

11.1 START-UP

1. Put all the valves in manual position
2. Air-purge all the equipment and the process line with compressed air.
3. steam purge all steam bearing lines to debris which must have been contacted during construction
4. Adjust controls to set point.
5. Start water into the tanks with enough water to run for about 2 hours, and commission all the pumps.
6. Fill entire mixer (reactor) with cocoa pod ah and water and start operation by commissioning P01
7. Start cooling water to heat exchangers, and start steam to boilers.
8. Read all indicators and ensure process is in control
9. Turn all valves to automatic position

11.2 SHUT-DOWN

1. Turn all valves to manual position
2. Put burner off by isolating P01
3. Isolate all pumps, and stop cooling water and steam supply to heat exchangers and boilers respectively.
4. Discharge reactors content.
5. Flush all line with air to inert the system

CHAPTER TWELVE

12.0 PLANT LAYOUT

The choice of site for plant has a long lasting effect on the cost of production and distribution room for expansion and safe living condition for plant operations. The following factors should be considered when sitting a plant.

1 Availability of raw materials: hydrogen synthesis gas plant should be sited near its raw material such as oil delta zone: so as to cut down the cost of transporting and storing raw materials

2 Market Area: The plant should be located close to its market in order to reduce cost of transporting products and time of delivery

3 Availability of energy: Power and steam requirements are high in hydrogen synthesis gas plant. Hence, it should be sited where there is constant supply of energy.

4 Climate: Excessive humidity or extremes of hot or cold weather can have serious effect on the economic operation of a plant. Therefore, this factor should not be neglected during site consideration.

5 Water Supply: water, in hydrogen synthesis gas plant, is required for cooling, washing steam generation and also as a raw material. The plant therefore must be located where a dependable supply of water is available.

6 Transportation: The plant should be sited where it's accessible to roads, railways and waterways for effective transportation of raw materials and product.

7 Waste Disposal: The site chosen for plant should have adequate capacity and facilities for correct waste disposal.

8 Labour Supply: The type of supply of labour available in vicinity of a proposed site must be examined. Skilled labour could be brought from outside site area. But there should be adequate pool of unskilled labour in the area.

9 Land considerations: The topography of land must be evaluated. The land should be relatively flat, well drained and have suitable load bearing characteristics. It should also have allowance for real estate expansion.

10 Political and strategic considerations: Capital grants and other inducements are often given by government to direct new investment to area or locations such as a high unemployment prone zone. The availability of such grants can be overriding considerations in site selection.

11 Community factors: The character and facilities of a community such as churches, libraries, schools, civic theatres concert association, etc for satisfactory living of plant personnel should be considered when choosing a site.

12.1 BUILDING LAYOUT

The hydrogen synthesis gas plant and auxiliary buildings should be laid out to give the most economical flow of materials and personnel around the site. Consideration must also be given to the future expansion of the plant. The auxiliary building s and services buildings and services required on the site in addition to the processing units(buildings) includes:

1. Storage for raw materials and products
2. Materials workshop
3. Stores for maintenance and operating supplies
4. Laboratory for process control
5. Fire station and other emergency services
6. Effluent disposal plant
7. Offices for general administration
8. Canteen, car park, security post, etc.

Building layout is arranged in such a way that its processing steps from raw material to the final product and location of auxiliary buildings should be in a way that minimum time is spent by personnel in traveling between buildings.

CHAPTER THIRTEEN

3.0 ECONOMIC ANALYSIS AND PROJECT EVALUATION

3.1 Introduction

Chemical plants are designed to make profit. The profitability or otherwise of the plant is determined from the economic analysis and there are a number of techniques for estimating this. Some of these are:

- Pay back Period (PBP): This is the time required after the start of the project to pay off the initial investment from income generated. Payback period is usually between 2.5 to 5 years (12).

$$PBP = \frac{\text{Total cash Invested}}{\text{Annual cash flow}} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (10.1)$$

- Return on Investment (ROI): Return on Investment is the return of annual profit to investment is a way of measuring the performance of the capital invested. The higher the ROI, the better.

$$ROI = \frac{\text{Profit after Tax}}{\text{Total investment}} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (10.2)$$

- Discounted Cash Flow Rate of Return (DCFRR): This is a measure of the maximum rate that the project could pay and still break-even by the end of the project life. It takes into consideration the time value of money.
- Purchased Equipment Cost: This is the purchased price of the equipments.

The purchased cost of stripper and absorber is given below as:

$$PC = \left(\frac{M \& S}{280}\right)\left(101.9D^{1.066}H^{0.802}F_C\right)$$

The purchased cot of heat exchanger and cooler is given below as:

$$PC = \left(\frac{M \& S}{280}\right)\left(101.3A^{0.65}F_C\right)(\$)$$

The cost estimation calculation was carried out using Microsoft excel spreadsheet.

13.2 Purchased Equipment Cost

Reactor			
M_S	1100		
Area	139		
Fm	2.81	Fp	1.25
CS/CS	1	<500	1
CS/SS	2.81	1000	1.25
SS/SS	3.75	2000	1.52
CS/Ti	8.95	3000	1.55
Ti/Ti	13.05		
FC	3.5125		
Purchased Cost of Reactor	$72,956.23		

Evaporator			
M_S	1100		
Area	87.9325		
Fm	2.81	Fp	1.25
CS/CS	1	500	1
CS/SS	2.81	1000	1.25
SS/SS	3.75	2000	1.52
CS/Ti	8.95	3000	1.55
Ti/Ti	13.05		
FC	3.5125		
Purchased Cost of Evaporator	$25,654.15		

Filter			
M_S	1100		
Area	59		
Fm	13.05	Fp	1.55
CS/CS	1	500	1
CS/SS	2.81	1000	1.25
SS/SS	3.75	2000	1.52
CS/Ti	8.95	3000	1.55
Ti/Ti	13.05		
FC	20.2275		
Purchased Cost of Filter	$113,983.01		

Heater			
M_S	1100		
Area	161.46		
Fm	3.75	Fp	1.25
CS/CS	1	<500	1
CS/SS	2.81	1000	1.25
SS/SS	3.75	2000	1.52
CS/Ti	8.95	3000	1.55
Ti/Ti	13.05		
FC	4.6875		
Purchased Cost of Heater	$50,819.17		

Compressor

M_S	1100		
Area	87.9325		
Fm	2.81	Fp	1.25
CS/CS	1	500	1
CS/SS	2.81	1000	1.25
SS/SS	3.75	2000	1.52
CS/Ti	8.95	3000	1.55
Ti/Ti	13.05		
FC	3.5125		
Purchased Cost of Compressor	$25,654.15		

Chlorine Storage Tank

Volume of Tank			
Cost of Tank	$0.00		
Fm	2.81	Fp	1.25
CS/CS	1	500	1
CS/SS	2.81	1000	1.25
SS/SS	3.75	2000	1.52
CS/Ti	8.95	3000	1.55
Ti/Ti	13.05		
Fc	3.5125		
Volume of Tank	353.2		
Cost of Tank	$558.85		

Pump 1

Capacity 1	10.00
Capacity 2	0.917
Cost 1	$1,000.00
M_S_1979	545
M_S_2005	1098

Cost 2 =Cost 1*((Capacity 2/Capacity 1)$^{0.34}$)*(M_S_2005/M_S_1979)

Cost of Pump 1	$894.15

Pump 2

Capacity 1	10.00
Capacity 2	0.1
Cost 1	$1,000.00
M_S_1979	545
M_S_2005	1098

Cost 2 =Cost 1*((Capacity 2/Capacity 1)$^{0.34}$)*(M_S_2005/M_S_1979)

Cost of Pump 2	$420.93

Pump 3

Capacity 1	10.00
Capacity 2	0.338
Cost 1	$1,000.00
M_S_1979	545
M_S_2005	1098

Cost 2 =Cost 1*((Capacity 2/Capacity 1)$^{0.34}$)*(M_S_2005/M_S_1979)

Cost of Pump 3	$636.85

13.3 Fixed Capital Investment

Equipment	
Name	**Installed Cost**
Reactor	$72,956
Evaporator	$25,654
Filter	$113,983
Heater	$50,819
Compressor	$25,654
Storage Tank	$559
Pump 1	$894
Pump 2	$421
Pump 3	$637

Name	Installed Cost

Total	$291,577	
Fixed Capital Investment Multiplier=	4.74	(Default is 1.5)
FCI	$1,382,077	

List your equipment by name and **installed** cost. Then enter the Fixed Capital Investment Multiplier. This multiplier will take into account the additional cost associated with building the plant.

There is a large difference between ***purchased*** cost and the ***installed*** cost of a piece of equipment. If you only know the purchased cost, you can multiply this cost by a "Lang" factor as shown below:

4.74 for Fluid Processing Plants
3.63 for Solid-Fluid Processing Plants
3.10 for Solid Processing Plants

13.4 Operating Labor

Scroll Down

Equipment	
Name	Installed Cost
Reactor	$72,956
Evaporator	$25,654
Filter	$113,983
Heater	$50,819
Compressor	$25,654
Storage Tank	$559
Pump 1	$894
Pump 2	$421
Pump 3	$637
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0

Name	Installed Cost
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0
0	$0

Equipment	Operators per Shift
Auxiliary Facilities	
Air Plants	1.0
Boilers	1.0
Chimneys and Stacks	0.0
Cooling Towers	1.0
Water Demineralizers	0.5
Electric Generation Plants	0.5
Portable Generation Plants	3.0
Electric Substations	0.0
Incinerators	2.0
Mechanical Refrigeration Units	0.5
Waste Water Treatment Plants	2.0
Water Treatment Plants	2.0
Process Equipment	
Evaporators	0.30
Vaporizers	0.05
Furnaces	0.50
Fans	0.05
Blowers and Compressors	0.15
Heat Exchangers	0.10
Towers	0.35
Vessels	0.00
Pumps	0.00
Reactors	0.50

Enter the expected annual salaries of operators in the plant $51,000 US

Now based on the equipment list that you have above, enter the total number of each type of unit in the space below.
You may not have auxiliary equipment in your list, but you should enter the number of pieces below.
For example, if you have 6 heat exchangers in your equipment list, you should enter "6" under "heat exchangers".
Also, if you have heat exchangers with cooling water in them, you'll probably have at least one cooling tower as well.

Equipment	# of units	Operators per shift
Auxiliary Facilities		
Air Plants		0.00
Boilers		0.00
Chimneys and Stacks		0.00
Cooling Towers		0.00
Water Demineralizers		0.00
Electric Generation Plants		0.00
Portable Generation Plants		0.00
Electric Substations		0.00
Incinerators		0.00
Mechanical Refrigeration Units		0.00
Waste Water Treatment Plants		0.00
Water Treatment Plants		0.00
Process Equipment		
Evaporators		0.00
Vaporizers		0.00
Furnaces		0.00
Fans		0.00
Blowers and Compressors		0.00
Heat Exchangers		0.00
Towers		0.00
Vessels		0.00
Pumps		0.00
Reactors		0.00

Total Number of Operators Needed = 0

Annual Cost of Operating Labor (Col) = $0

13.5 Utilities

Utility costs vary widely from region to region, you cah change these values as you see fit

Utility	Description	Cost $/GJ	Cost $/common unit	
Air Supply	Pressurized and Dried			
	a. Process		$2.30 per 100 m3	
	b. Instrument		$4.70 per 100 m3	
Steam from Boilers	Process Steam Latent Heat			
	a. Low Pressure (5 barg, 160 0C)	$3.17	$6.62 per 1000 kg	
	b. Medium Pressure (10 barg, 184 0C)	$3.66	$7.31 per 1000 kg	
	c. High Pressure (41 barg, 254 0C)	$5.09	$8.65 per 1000 kg	
Cooling Tower Water	Process Cooling Water			
	30 0C to 40-45 0C		$6.70 per 1000 m3	
Other Water	High purity water for			
	a. Process Use		$0.04 per 1000 kg	
	b. Boiler Feed Water		$2.54 per 1000 kg	**This can also be used as condensate
	c. Potable (drinking)		$0.26 per 1000 kg	return credit amount
	d. Deionized Water		$1.00 per 1000 kg	
Electrical Substation	Electric Distribution			
	110V, 220V, 440V	$16.80	$0.06 per kWh	
Fuels	a. Fuel Oil (no. 2)	$4.00	$170.00 per m3	
	b. Natural Gas	$2.50	$0.09 per std m3	**Based on lower heating value
	c. Coal (FOB mine mouth)	$1.20	$31.00 per metric tonne	
Refrigeration	a. Moderately Low Temp. 5 0C	$20.00	Process cooling duty	
	b. Low Temp. 20 0C	$32.00	Process cooling duty	
	c. Very Low Temp. -50 0C	$60.00	Process cooling duty	
Thermal Oil Systems	a. Moderately High Temp. to 330 0C	$4.90	Process heating duty	
	b. High Temp. to 400 0C	$5.20	Process heating duty	
	c. Very High Temp. to 600 0C	$5.90	Process heating duty	
Waste Disposal	a. Non-Hazardous		$36.00 per metric tonne	
	b. Hazardous		$145.00 per metric tonne	
Waste Water Treatment	a. Primary (filtration)		$39.00 per 1000 m3	
	b. Secondary (filtration + activated sludge)		$41.00 per 1000 m3	
	c. Tertiary (filtration, activated sludge, chem proc)		$53.00 per 1000 m3	

Scroll Down

Equipment	(enter a description)	(Calc from Table at Left)
Name	**Utility Type**	**Annual Utility Cost**
Reactor	Fuel	$1,480,000.00
Evaporator	Process Cooling Water	$58,738.90
Filter	b. Medium Pressure (10 barg, 184 0C)	$70 29
Heater		
Compressor		
Storage Tank	Process Cooling Water	$278 79
Pump 1	Process Cooling Water	$1,079 84
Pump 2	Process Cooling Water	$7,778 63
Pump 3		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		
0		

Utility Cost Not Associated with Equipment

	Description	**Utility Cost**

As an example, let's say that your reactor needs to be mainted at 350 0C by a Therminol or Dowtherm loop within the plant The heating duty turns out to be about 75,000 GJ per year Your utility calculation for the reactor would be

(75,000 GJ/year) x ($5 2/GJ) $390,000 per year

The $5.2/GJ is from the utility table under thermal systems

Total Utility Cost per Year (Cut) = **$1,547,946.45**

13.6 Raw Materials and Profits

Enter any raw material costs. Remember that sometimes, utilities can be used as raw materials.
Some processes may utilize low pressure steam as a reactor feed for example.

Raw Material	Annual Cost
Cocoa Pod	$63,000.00

Total Annual Raw Material Cost (Crm) = $63,000.00

Enter any profits. These should consists of your main, salable reaction products as well as any generated utilities that can be sold. Low pressure steam, for example, can be generated in heat exchangers and sold to other facilities. Or condensed steam can be credited as boiler feed water at times

Profit Material	Annual Profit
KOH	$2,811,000.00

Total Annual Profits = $2,811,000.00

13.7 Financial Summary of the Project

Total Installed Cost:	
	$291,577
FCI	$1,382,077

Operating Labor	
Col =	$0

Raw Materials	
Crm=	$63.000

Utilities	
Cut=	$1.547.946

Profit from Products	
Annual Revenue=	$2.811.000

The following assumptions are made in the cash sheet shown below:

1. **Two years for construction of plant**
2. **Five year depreciation schedule**
3. **10 year plant operating life**
4. **Cost of constructing the plant is spread over two years**
5. **Taxation rate is defaulted to 40%.**
6. **The discounted cash flow rate is set at 8%.**
7. **Optimization function used is "Net Present Value"**

Financial Outlook

COM= $2,230,238.04 (without depreciation)

assume 15% of FCI as salvage value

Year	Depreciation	FCI-deprec.	Revenue	COM	Cash Flow	Discount Cash	Cumulative Discounted
0	-	$1,382.077	-	-	($691.039)	($691.039)	($691.039)
1	-	$1,382.077	-	-	($691.039)	($639,851)	($1,330,889)
2	$276,415	$1,105.662	$2,811,000	$2.230.238	$459.023	$393.539	($937,351)
3	$353,812	$751.850	$2,811.000	$2.230.238	$489.982	$388,963	($548,387)
4	$144,355	$607.495	$2,811,000	$2.230.238	$406.199	$298,569	($249,819)
5	$116,639	$490.856	$2,811,000	$2.230,238	$395.113	$268,907	$19,088
6	$47,122	$443.734	$2,811.000	$2.230,238	$367.306	$231,465	$250,554
7	-	$443.734	$2,811.000	$2.230.238	$348.457	$203,321	$453,875
8	-	$443.734	$2,811,000	$2.230.238	$348.457	$188,261	$642,136
9	-	$443.734	$2,811.000	$2.230.238	$348.457	$174,315	$816,451
10	-	$443.734	$2,811.000	$2.230.238	$348.457	$161.403	$977,854
11	-	$443.734	$2,811.000	$2.230.238	$555.769	$238.360	$1,216.214

Taxation rate= 0.4

Discount rate= 0.08

NPV= $1,216,214
(Net Present Value)

PBP= 4
(Pay Back Period)

ROR 25%
(Rate of Return)

CHAPTER FOURTEEN

14.0 CONCLUSION

From the aspects of the design of the KOH plant from cocoa pod has been carried out, the following conclusions can be drawn:

1. Economic analysis of the project gave the following results:

Return of Return	25%
Pay back period	4 years
Total capital investment	$ 1,387,077
Net Present Value (NPV)	$1,216,214

2. The plant is therefore profitable to invest in and shall be beneficial to the country because it shall cater for the perennial shortage of the KOH related chemicals and its precursor, provide employment, and enhance industrial development.
3. The plant being designed can be expanded to increase production rate.
4. The plant has a satisfactory safety level as provisions for ensuring this have been put in place.

REFERENCES

(1) W.L. McCabe, J.C. Smith UNIT OPERATION OF CHEMICAL ENGINERING, fourth edition, 1986, McGraw Hill Book Company New York, U.S.A. pages 413-443.

(2) Robert. H. Perry, Don.W. Green PERRY'S CHEMICAL ENGINEERS' HAND BOOK Sixth edition, 1987, McGraw Hill Book, USA, section 11.31-11.37.

(3) W.L. Badger and J.T. Banchero, INTRODUCTION TO CHEMICAL ENGINEERING, International student edition, 1985 McGraw Hill U.S.A., page 170-229

(4) Coulson and Richardson, CHEMICAL ENGINEERING VOL. 2 Second edition, 1999, Butter worth Heinemann, Linacre House, Jordan Hill, oxford, page 622-664.

(5) Jens G. Balchen and K.I. Mumme, PROCES CONTROL STRUCTURE AND APPLICATIONS First edition, 1988 Blackie Academic and professional, wester cleddens Road, Bishop brigs, Glasgow G64 2NZ, UK, page 171-174, 508-513

(6) Gerald Coren, PROFESSIONAL ENGINEER'S EXASMINATION CHEMICAL ENGINEERING, ARCO, New York, 1979, page 185-187.

(7) Philip .A. Schweitzer, HAND BOOK OF SEPARATION TECHNIQUES FOR CHEMICAL ENGINEERS, McGraw Hill Book Company ST. Lows, 1979, section 2.130-2.135.

(8) Robert .W. Wansbrough, "MODELLING CHEMICAL REACTORS" CHEMICAL ENGINEERING / AUGUST 5, 1985,Page 95 102.

(9) William. L. Luyben and L.A. wenzel, CHEMICAL PROCESS ANALYSIS, MASS AND ENERGY BALANCES, Prentice Hall, Englewood cliffs, New Jersey 1988, page 422-445.

(10). Vincent, C., PROCESS HEAT EXCHANGE, Chemical Engineering, McGraw Hill Publication Company, New York 1979, page 471-476.

(11) W.F. Ramirez, COMPUTATIONAL METHOD FOR PROCESS SIMULATION, Second edition, 1997, Butter worth Heinemann, Linacre House Jordan Hill, oxford, page 5-8.

(12) R. Bronson, THEORY AND PROBLEMS OF DIFFERENTIAL EQUATIONS, Second edition 1994, Schaums outline series, McGraw Hill Inc USA.

(13) F Madron and D. Sharp, PROCESS PLANT PERFORMANCE, measurement and data processing for optimization and retrofits First edition 1992, ELLIS HORWOOD LTD, Chichester, West Sussex, England, page 27-49.

(14) Ellison M.A., Taylor and co authors, CTS-5 DOUBLE EFFECT EVAPORATORS, Corning Process System, Stone, England, 1985, section 1.1-3.2 and 8.1-8.5.

(15) W.H., Ray, ADVANCED PROCESS CONTROL, Butter worth Publisher, U.S.A., 1989, page 331-333.