

CERTIFICATION

This is to certify that JINADU B.H carried out the project work presented in this report under my supervision

Supervisor

A.C. WAKINBODE

date

Head of department

Dr. J. ODIGURE

date

EXTERNAL EXAMINER

date

DECLARATION

I hereby declare that this particular final thesis is carried out by me and that it is not a copied work from somewhere. It was done under the close supervision of my supervisor Mr. A.O.W Akinbode.

Name

Sign and date

DEDICATION

I dedicate this project to my able father Mr. **JINADU B.A.**, and my brother **NOJEEM**.
Daddy you are too much to quantify, thanks for making this possible. Brother you are
always my source of joy, and each time I look back and see you I always know
mission is certainly going to be accomplished.

ACKNOWLEDGEMENT

My greatest gratitude goes to GOD ALMIGHTY who in his infinite mercies made it possible for me to achieve my goals academically, morally, financially and others.

Words can not express my sincere thanks to my supervisor, Mr. A.O.W Akinbode for his tremendous efforts towards the achievement of this project. I also extend this appreciation to the H.O.D Chemical engineering department Engr. (DR) J.O. Odigure, and to all members of staff of this precious department. May God help you all.

My profound gratitude goes to my parents, god and goddess JINADU. I've always wondered that who can be there for me the way you've been there for me. Daddy and mummy I love you and appreciate all you've done for me, definitely words can not express all you've done for me. Also to my siblings, Nojeem, Kazeem, and Rasheedat, you guys are just too wonderful; you have in one way or the other brought out the stuffs in me. Appreciation also goes to my uncle, Raymond, though far away, but trust me your memory is still in me, and others I cant mention here I thank you all.

A friend in need is a friend indeed, my appreciation goes to my friends; Seun, Seyi, Lukman, Niyi, James kwamme, Dr.Who, Big Dee, Sina, Tayo, Seye sala, Rector, Deola, Obembe Tee, Tkb Baba, Kolapa, Pastor Demmy, Agada, Jimoro, All west siders, Richard and to all 500 level students (1998/99 session) to you all I say thank you and God almighty be with you.

ABSTRACT

The design of a plant for the production of 2000 tons per year of glycerine of 95% purity through the recovery of glycerine from the popular waste of a soap industry, spent lye has been carried out.

The work was carried out by designing and modifying the existing process, which has popularly failed. The major modification of this process designed is the inclusion of ion exchange process. The ion exchange unit was designed based on relevant standards like " Rohm and Haas & Co. summary 11, 1986. BS 4289, 1975" and the safety and the economic considerations carried out. Also a detailed costing of the ion exchange unit was carried out using the open and current market price.

The results of the designs, shows that the plant is safe for operation, the plant also has a lot of economic advantage and also the cost of the ion exchange columns was found to be under a million (N570,000).

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CHAPTER ONE

INTRODUCTION

1.1 GENERAL INTRODUCTION

The advent of new technology has made glycerine, exceptionally important in the process industry. From the primary processing industries to the tertiary processing industry, also from the production of primary, to the production of final or consumable products Glycerine has become inevitable. In recent times the development of processes in both the food industries like food, drugs, cosmetics and tobacco industries has increased the competition for glycerine consumption. Its intoxicity, softening action, and also other physical and chemical properties has aided to an extraordinary quest for its use.

The aim of this work is the design of a recovery plant, to recover glycerin from the waste of a primary plant, "the soap industry". Soap industry is as old as the word technology itself, but the optimum or maximum utilization of this industry has not being met.

Glycerine as earlier mentioned, is a very important product in most process industries. Optimum utilisation of the soap industry automatically makes the industry a producer of glycerine simultaneously with soap. With this the company stands a better chance of diversification, which in today's world any industry that has the power or capacity on potential to diversify has the secret key of industrial boom.

1.2 LITERATURE SURVEY

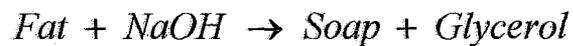
1.2.1 *TECHNOLOGICAL SURVEY.*

The introduction of soaps industry to Nigeria , is as old as the introduction of technology itself to Nigeria ,but it can be clearly stated that untill now, the industry has not really tapped or realised all the advantages or resources that can be accused to it.

The technology of producing of soaps, either laundry or toilet soap is saponification. This process is the reaction of Fats and Oil and Caustic soda or potash to form the product soap and glycerine. To the local soap manufacturer the soap is the ultimate and everything is done in removing the soap in the mixed products to the bearest

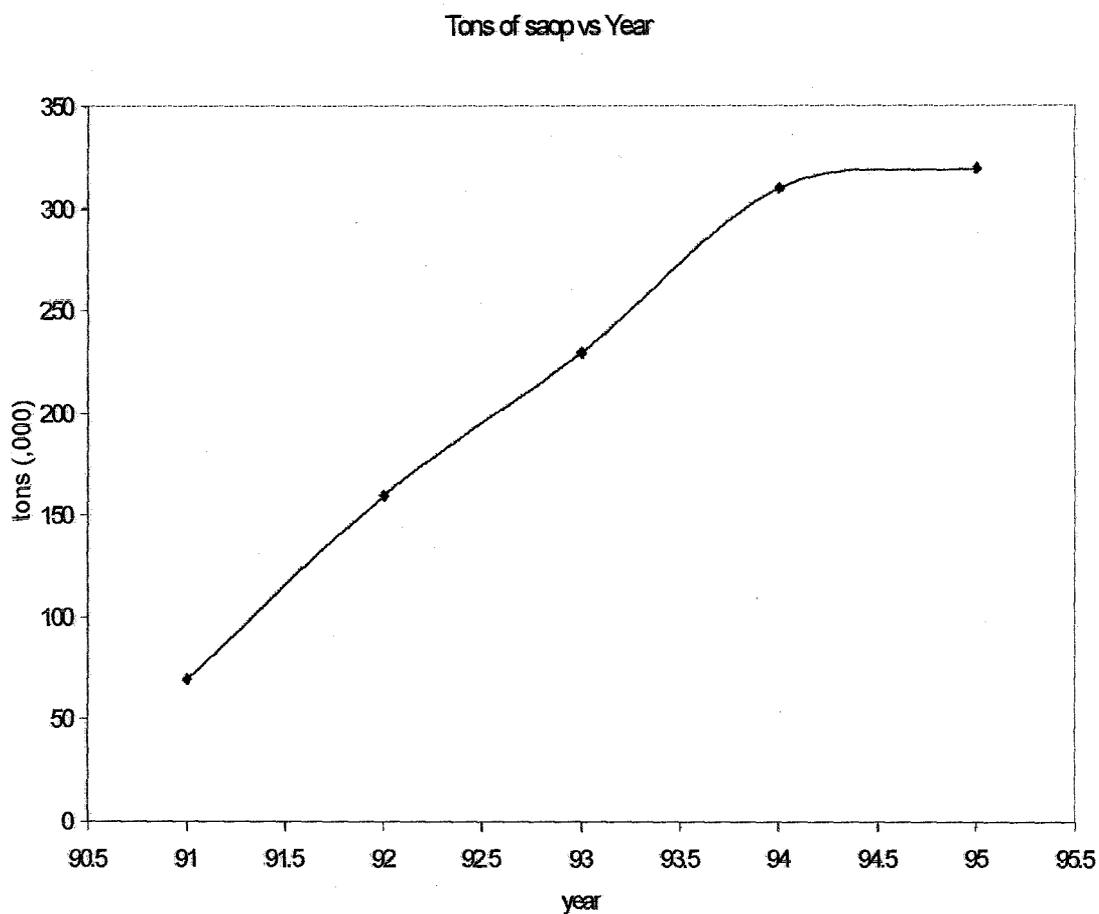
minimum by a process known as salting out (washing out the soap). Salting out as in separation of the soap and glycerine is done by adding salt solution (brine solution). To the soap manufacturer, the only importance of the salted out waste i.e. spent lye, is firstly the unused content of the caustic soda and secondly, use of the spent lye for the catalysis of the other saponification process since the reaction (saponification) is autocatalytic in nature.

From the analysis of soap production through the saponification reaction.



From this reaction, for every 3 moles of soap produced, 1 mole of glycerine is simultaneously. Thus for a gross analysis of fat commonly used showed that for every 1 kg of soap produced, 0.1 kg of glycerine has being produced but wasted.

The graph below shows the amount of soap in tons produced in Nigeria for the period of 5 years. (courtesy, FEDERAL OFFICE OF STATISTICS)



From the graph the production of the soap has decreased geometrically over the years even with the advent of highly powered detergents, the production of soap has continued to increase tremendously.

From fig 1.2, relating the production of soap with glycerine the graph shown below the amount of unrealised glycerine over the last 2 years.

unrealised glycerine curve

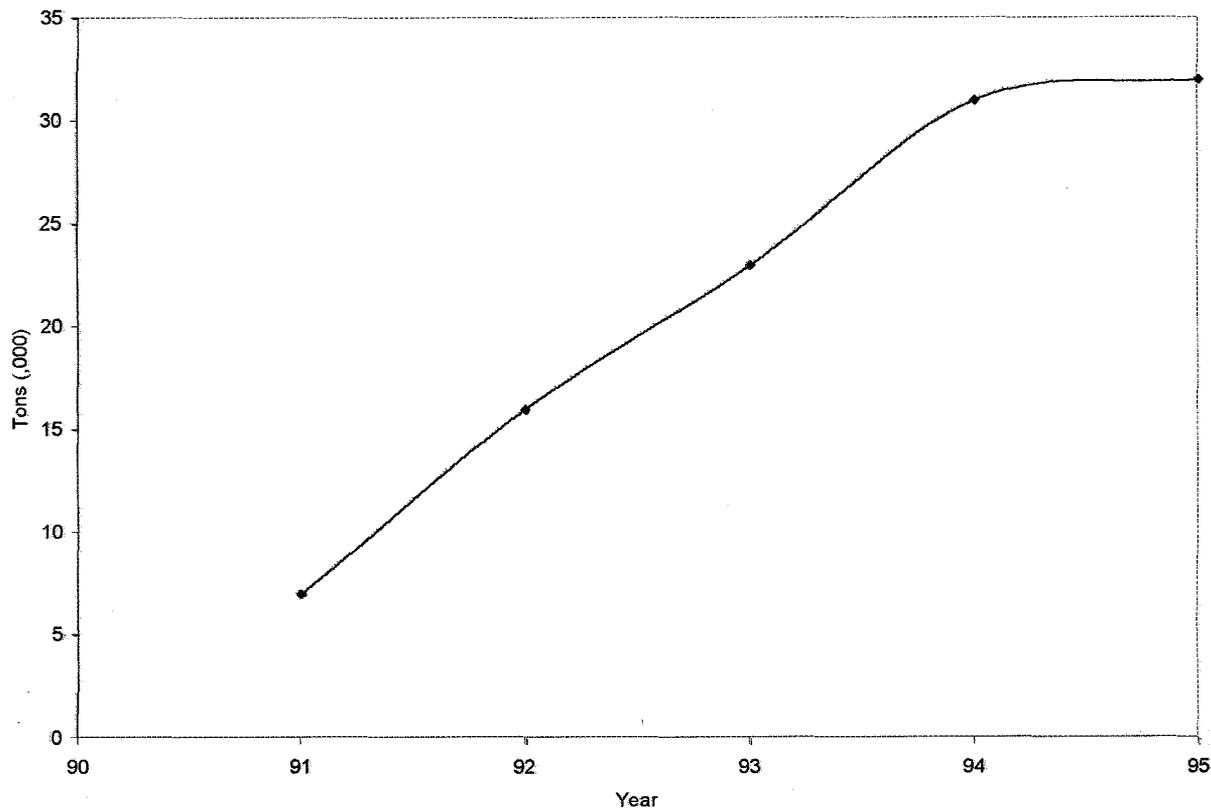


Fig 1.2

Moreover, it should be stated that the recovery of glycerine from spent dye is not a new thing to the technological world even to this country Nigeria. Glycerine recovery has been practiced in Nigeria, but has not been successful, big names like Lever Brother Nig. Ltd., BACL, and Global soap have found recovery of glycerine a wasteful effort. This conclusion was grossly due to the poor quality yield with most plants used, the rate at which their efficiency drops is too high.

The set back in this industry, has also been triggered by the little or no interest which the western world has, in this technology, due to the production of high sophisticated detergents, the production of soap has grossly decreased out there, and this definitely decreases the production of glycerine. Also commercial synthesis of glycerol using chlorine and propylene made this technological in the western world less competitive. But in this part of the world, the production of soap is still very high and still increasing, so this technology still finds its feet.

The technology of glycerine recovery is strictly that of separating the constituent components in the spent lye until the required glycerine is achieved putting in mind that the glycerine should still appear and be natural after all the process it must have undergone. The efficiency of the technology used dictates the kind of glycerine grade recovered. Glycerine recovery goes through the following process.

1. Treatment of the spent lye: This is the removal of the unwanted materials in the spent dye, both organic and inorganic materials. The major reagents used for this treatment is usually the acids solution and the precipitating agents.
2. Evaporation: This is the stage where crude glycerine is produced. Most plants for recovery make use of multiple effect evaporation.
3. Distillation: this is the stage where refined glycerine is produced.

1.2.2 MARKET SURVEY.

Glycerine, being a very important product for most industries, makes it very important to know the effect of market forces on the production of glycerine.

DEMAND.

From the market survey carried out, showed that the demand for glycerine can be represented with the graph below for the last 10 years.

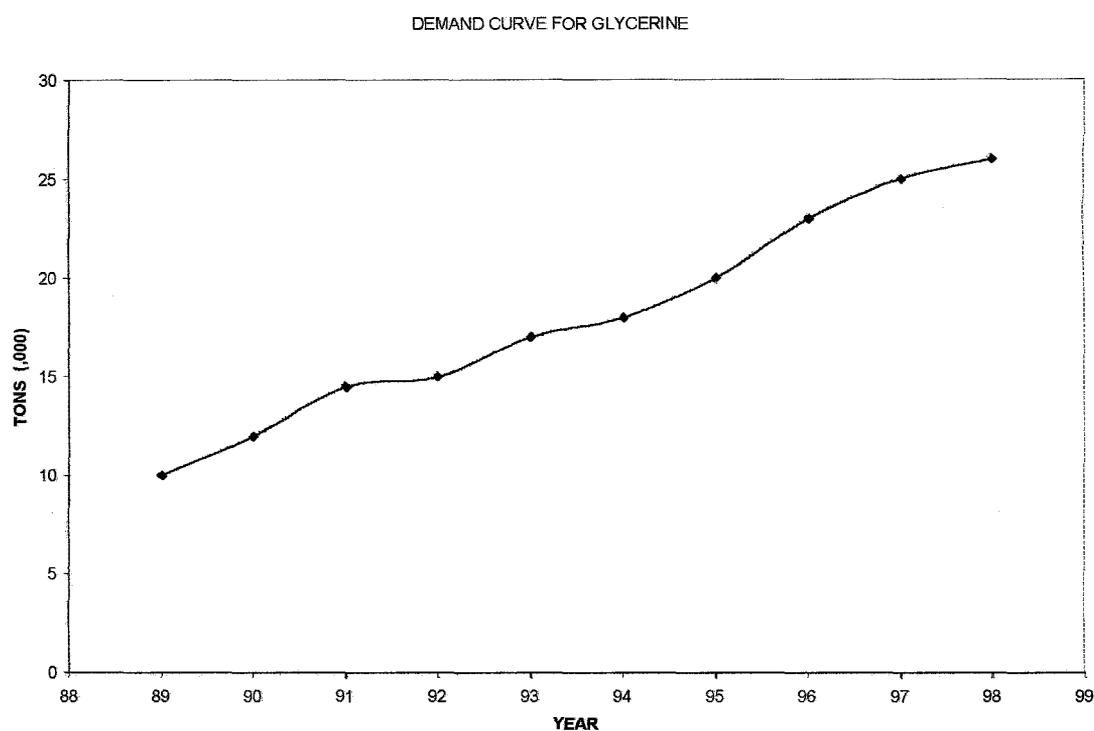


Fig 1.3 Glycerine Demand (Federal Office Of Statistics)

From the graph, the growth in demand started at a very low rate and later picked up at a very high rate, this high increase in demand has seriously been due to the increase in development of food, cosmetics and especially human utilities industry over the last 10 years.

SUPPLY.

From the market survey carried out, this showed that 90% of the demand is met through importation of glycerine. Careful analysis shows that all the demands for glycerine are not met, some of the industries have to go for costly substitutes. Relating the amount of unrealised glycerine produced through the local manufactures (soap industry) and the demand for glycerine, it can be confirmed that 50% of the demand can be met through the local soap industry

1.2.3 COST SURVEY

More than 80% of the raw material associated with soap can be sourced locally and considering the fact that the production or recovery of glycerine is a function of soap production this shows that little or no extra cost is involved apart from the operational cost of recovery. Also most of the materials used in carrying out this operation can be sourced locally.

1.3 SCOPE OF WORK AND DESIGN PROBLEM

The scope of work of this project is strictly subjected to an existing plant. The recovery plant is a second-degree plant that is based on the functioning of the primary plant, the soap plant. With this, before any scope can be analysed or stated the soap plant must be carefully analysed.

From the analysis carried out on the existing plant, the plant is a fully automated and comprehensive plant, having all the required utility facility functioning properly. Also the plant location and layout favours the inclusion of a recovery plant since much considerations has being given to future plant expansion.

Considering the capacity of the plant, the maximum capacity of the existing plant is below 20,000tons per year, but putting into consideration future expansion, the maximum capacity of the plant is put at 20, 000tons per year. With the process and technological relationship between soap and glycerine production, this shows that a recovery plant of 2000tons per year should be designed to meet future expansion.

With this the scope of work can be stated as

- ❖ The design of a process for the recovery of glycerine from spent lye
- ❖ The detailed design of a plant for the recovery of glycerine(the unit to be designed is the ion exchange unit.

- ❖ Safety analysis and considerations of the plant
- ❖ Economic and cost analysis of the plant

Design statement

Design a plant for the production or recovery of 2000tons/year of glycerine of 95% purity.

CHAPTER TWO

PROCESS DESIGN

2.1 PROCESS DESIGN PHILOSOPHY

The basic chemical engineering knowledge or idea associated with the recovery of glycerine from spent lye which is a by-product of the soap industry is strictly based or influenced by the composition of the lye.

Generally the constituents of the lye are

- (1) Caustic soda
- (2) Soap
- (3) Glycerine
- (4) Matter-non organic and organic
- (5) Salt (sodium chloride)
- (6) Ash & other
- (7) Water

Considering this, the recovery of glycerine means the removal of other components totally or to the nearest minimum. The philosophy of chemical engineering applied now has to do with how to remove this component without affecting the desired product and quality and in what sequence the removal should follow.

Another important factor, is the material or process to be used in removing this process and what effect do they have on the overall objective, process and operations. From this, the recovery of glycerine from spent lye considering the laboratory experiment carried out earlier can be grouped into two

- (1) Treatment
- (2) Concentration (recovery)

Treatment stage has to do with the removal of other components fully or to a tolerable limit except glycerine.

The sequential process, which this follows, is given below

- (1) Removal of caustic soda
- (2) Removal of MONGS and ashes and other.

Concentration is after the treatment, the constituents is reduced to salt solution and glycerine and concentration of the solution, produces the required product and concentrate. (It is at this stage that glycerine can be fully recovered).

2.2 PROCESS DESIGN OPTIONS.

From the process design philosophy associated with recovery of the glycerine and the process path to be taken, the possible options are listed

- (1) Treatment stage
 - (a) Acidification - This is by the use of mineral acid, and the two most common mineral acid that can be used are:
 - (i) Hydrochloric acid (HCl)
 - (ii) Sulfuric acid (H₂SO₄)
 - (b) Precipitation - This is also by the use of inorganic coagulants or precipitants and the most common precipitants are:
 - (i) Alum
 - (ii) Iron per chloride
 - (iii) Aluminium sulphate
- (2) Recovery stage
 - (i) Evaporation with crystallising - This is by evaporating the treated spent lye by providing a means of crystallising the salts presents, the best this process could get is a crude glycerine.
 - (ii) Ion exchange treatment - This is by removing the salts in solution completely, before proceeding to evaporation. This process can produce glycerine of more than 95% conc.
 - (iii) Distillation - This is the process that follows the crude glycerine production, it is used in producing high quality glycerine called refined glycerine of more than 95% after evaporation process.

Process selection and route.

After careful analysis of all the process options for the recovery of glycerine and under proper consideration of the following factors:

- (1) Cost of materials
- (2) Energy requirements
- (3) Reaction involved
- (4) Reusability of materials
- (5) Recoverability of components
- (6) Upstream and down stream effects
- (7) Output of the process and operations
- (8) Feasibility
- (9) Product value and process capacity

**DESIGN OF A PLANT FOR THE RECOVERY OF
2000 TONS/YEAR OF GLYCERINE FROM A SOAP
INDUSTRY.**

By
JINADU B.H

93/3603

**IN PARTIAL FULFILMENT OF THE
REQUIREMENT FOR THE AWARD OF A
BACHELOR OF ENGINEERING (B.Eng.)
DEGREE IN CHEMICAL ENGINEERING**

Submitted to the

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SCHOOL OF ENGINEERING AND
ENGINEERING TECHNOLOGY, FEDERAL
UNIVERSITY OF TECHNOLOGY, MINNA,
NIGERIA**

MARCH 2000

The processes selected are

- (1) Treatment stage
 - (i) Acidification - Hydrochloric acid treatment
 - (ii) Precipitation - Iron per chloride
- (2) Recovery
 - (i) First the treated solution should be diluted with water, and passed through ion exchange unit.
 - (ii) Evaporation - This resulting solution is now concentrated in an evaporator, and since this is free of salt, the product quality can be more than 95%.

2.3 PROCESS DESCRIPTION (SPECIFICATION)

Spent lye solution from the waste reservoir is reacted with hydrochloric acid of 30% weight composition in a batch reactor fitted with an agitator. After the reactor time is completed the resulting solution is pumped out of the reactor into a settling tank where it is allowed to settle and the filtrate is now collected and sent into another batch reactor fitted with an agitator where it is reacted with a precipitating agent, iron per chloride. The resulting solution is passed through filter press. The filtrate is pumped into a dilution tank where it is diluted with distilled water, which is produced from the evaporator through the condenser.

The diluted solution is now passed through the cation exchange column to remove the cations (Na ion and Fe ion). The resulting solution from the cation exchange column is passed through the anion exchange column where the anion Cl, is removed.

The final solution from the ion exchange unit is passed through an evaporator where water is evaporated to give a concentrated glycerine solution.

2.4 MATERIAL DESCRIPTION

UNTREATED SPENT LYE

1. Glycerine	9.00%
2. Sodium chloride	9.00%
3. Caustic soda	2.00%
4. Fatty MONGs	4.00%
5. Ash & others	2.00 %

6. Water 73.50%

HYDROCHLORIC ACID SOLUTION

1. Hydrochloric acid 30.00%
2. Water 70.00%

IRON PER CHLORIDE SOLUTION

1. Iron chloride 40.00%
2. Water 60.00%

CHAPTER THREE

MATERIAL AND ENERGY BALANCE

3.1 SPECIFICATIONS

Overall basis = 2000 tons of glycerine per year
 Total working hours per year = 8000 hours
 Product quality = 95% (percent weight)

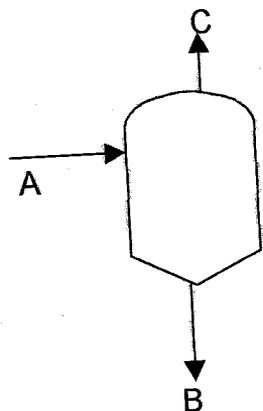
3.2 MATERIAL BALANCE

Basis: $2000 \times 1000\text{kg} = 2,000,000\text{kg} / \text{year}$

$$\text{Per hour} = \frac{2,000,000}{8,000} = 250\text{kg} / \text{hr}$$

$$\text{Per sec} = \frac{250}{3600} = 0.0695\text{kg} / \text{s}$$

BALANCE OVER THE EVAPORATOR



A : glycerine = 1.328%
 water = 98.672%
 B : glycerine = 95%
 water = 5%
 C : water = 100%

Basis: 0.0695 kg/s of B

Overall balance

$$A = B + C \dots\dots\dots(1)$$

component balance

$$\text{glycerine} \Rightarrow 0.01328A = 0.95B \dots\dots\dots(1a)$$

$$\text{water} \Rightarrow 0.98672A = 0.05B + C \dots\dots\dots(1b)$$

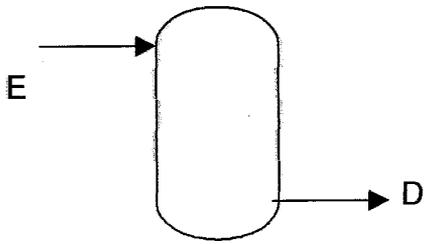
from (1a)

$$A = \frac{0.95 \times 0.0695}{0.01328} = \frac{0.066025}{0.01328} = 4.972\text{kg} / \text{s}$$

from (1)

$$C = A - B \Rightarrow 4.972 - 0.065 = 4.9025\text{kg} / \text{s}$$

BALANCE OVER STORAGE TANK



$$E = D : \text{glycerine} = 1.328\%$$

$$\text{water} = 98.672\%$$

Basis: 4.972kg/s of D=A

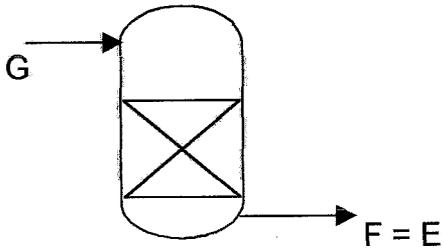
Overall balance

$$E = D + \text{accumulations}$$

allow for 0.0001kg/s accumulations

$$\Rightarrow E = 0.0001 + 4.972 = 4.9721 \text{ kg / s}$$

BALANCE OVER THE ANION EXCHANGER



$$G : \text{glycerine} = 1.3714\%$$

$$\text{HCl} = 0.933\%$$

$$\text{water} = 97.75\%$$

Basis: 4.9721kg/s of F

Overall balance

in = out + accumulation - generation

material accumulated = HCl = s

material generated = H₂O = y

$$G = F + s - y \dots\dots\dots ()$$

component balance

$$\text{glycerine} \Rightarrow 0.013174G = 0.1328F$$

$$\text{water} \Rightarrow 0.9775G + y = 0.98672F$$

$$\text{HCl} \Rightarrow 9.33 \times 10^{-3}G = s$$

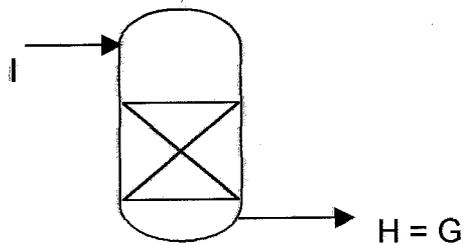
from

$$G - 9.33 \times 10^{-3}G = 4.9721$$

$$G = \frac{4.9721}{0.99067} = 5.019 \text{ kg / s}$$

$$\text{accumulation} = 9.33 \times 10^{-3}(5.019) = 0.0468 \text{ kg / s}$$

BALANCE OVER THE CATION EXCHANGER



$$I : \text{glycerine} = 1.310\%$$

$$\text{NaCl} = 1.477\%$$

$$\text{FeCl}_3 = 0.009\%$$

$$\text{water} = 97.20\%$$

Basis: 5.019kg/s of H

Overall balance

$$\text{in} = \text{out} + \text{generation} - \text{accumulation}$$

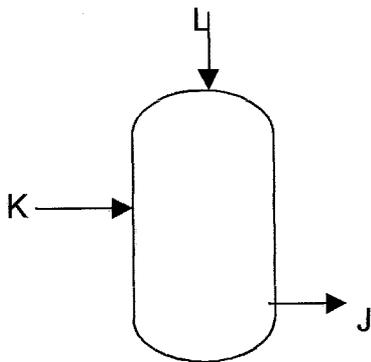
$$\text{Total accumulation} = 0.01486I - 0.0486$$

$$I - 0.01486I = 5.019$$

$$I = \frac{5.019}{0.98514} = 5.095 \text{ kg / s}$$

$$\text{accumulations} = 0.01486(5.095) - 0.0486 = 0.02958 \text{ kg / s}$$

BALANCE OVER THE DILUTION TANK



Basis: 5.095kg/s of J=I

Overall balance

$$K + L = J$$

component balance

glycerine

$$0.0920K = 0.01310J$$

water

$$L + 0.8041K = 0.9720J$$

from

$$K = \frac{0.01310J}{0.092} = 0.1424J$$

subst.

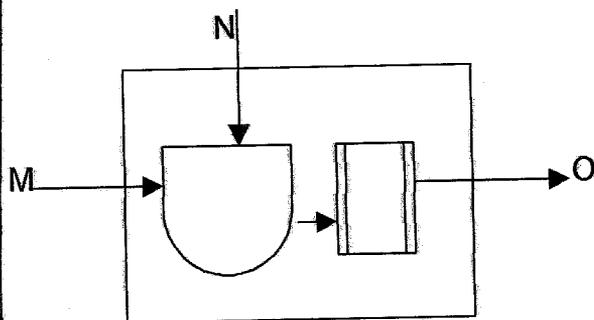
$$L + 0.8041(0.1424J) = 0.9720J$$

$$L + 0.1145J = 0.9720J$$

$$L = 0.8575J \Rightarrow 0.8575(5.095) = 4.369 \text{ kg / s}$$

$$K = 0.1424(5.095) = 0.7256 \text{ kg / s}$$

BALANCE OVER THE COAGULATION UNIT



N : iron per chloride = 40%
water = 60%

M : glycerine = 9.08

NaCl = 10.20

NaOH = 0.00001

MONGS = 3.0

Ash & others = 1.2

Water = 76.5

Basis: 0.7256 kg/s Of O

Overall balance

$$M + N = K + \text{accumulations}$$

component balance

$$\text{glycerine} \Rightarrow 0.0908M = 0.092K$$

$$\Rightarrow M = \frac{0.092(0.7256)}{0.0908} = 0.7352 \text{ kg / s}$$

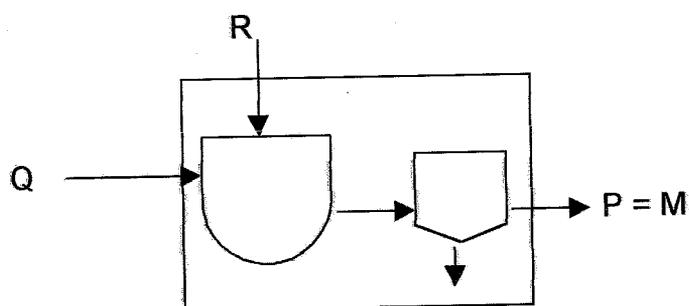
$$\text{water} \Rightarrow 0.765M + 0.60N = 0.8041K$$

$$0.60N = 0.8041(0.7256) - 0.765(0.7352)$$

$$0.60N = 0.5835 - 0.56242 = 0.02104$$

$$N = \frac{0.02104}{0.6} = 0.0351 \text{ kg / s}$$

BALANCE OVER THE ACIDIFICATION UNIT



Q : spent lye

R : HCl = 30%

water = 70%

basis: 0.7352kg/s of P

overall balance

$$Q + R = P$$

component balance

$$\text{glycerine} \Rightarrow 0.09Q = 0.0908P$$

$$0.09Q = 0.0908(0.7352) = 0.06676$$

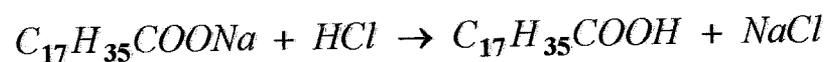
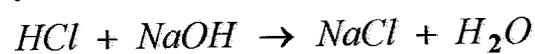
$$Q = \frac{0.06676}{0.09} = 0.7418 \text{ kg / s}$$

to calculate R

$$\text{NaOH} = 0.5\%$$

$$\text{soap} = 2.00\%$$

from the reaction



from the reaction

$$1 \text{ mole of HCl} \equiv 1 \text{ mole of NaOH} \equiv 1 \text{ mole of soap}$$

$$\text{no of moles of NaOH} = \frac{3.709}{40} = 0.09273 \text{ moles}$$

$$\text{no of moles of soap} = \frac{14.836}{306} = 0.0485 \text{ moles}$$

$$\therefore \text{total no of moles of HCl} = 0.09273 + 0.0485 = 0.1412 \text{ moles}$$

$$\text{mass of HCl needed} = 0.1412 \times 36.5 = 5.1543 \text{ g}$$

$$R = \frac{5.1543}{0.3} = 17.181 \text{ g} \cong 0.0172 \text{ kg/s}$$

3.3 ENERGY BALANCE

BALANCE OVER THE ACIDIFICATION UNIT

Overall balance

heat in = heat out

$$\text{heat in} = m_1 c_1 \theta + m_2 c_2 \theta + \Delta H_R$$

$$\text{heat out} = m_3 c_3 \theta$$

$$\Delta H_R = 2.95 \text{kJ}$$

$$c_1 = 2.8 \text{kJ / kgk}, c_2 = 1.7 \text{kJ / kgk}, c_3 = 2.7 \text{kJ / kgk}$$

subst.

$$0.7418(2.8)(25) + 0.0172(1.7)(20) + 2.95 = 0.7352(2.75)\theta_1$$

$$51.926 + 0.5848 + 2.95 = 2.022\theta_1$$

$$\theta_1 = \frac{55.461}{2.022} = 27.43^\circ\text{C} \text{ (temperature of stream leaving the acidification unit)}$$

BALANCE OVER THE COAGULATION UNIT

Overall balance

heat in = heat out

$$\text{heat in} = m_1 c_1 \theta_1 + m_2 c_2 \theta_2 + \Delta H_R$$

$$\text{heat out} = m_3 c_3 \theta_3$$

$$c_1 = 2.75 \text{kJ / kgk}, c_2 = 1.9 \text{kJ / kgk}, c_3 = 2.68 \text{kJ / kgk}$$

$$\Delta H_R = 4.89 \text{kJ}$$

$$0.7352(2.75(27)) + 0.0035(1.9)(20) + 4.89 = 0.7703(2.68)\theta_3$$

$$54.59 + 0.133 + 4.89 = 2.064\theta_3$$

$$\theta_3 = \frac{59.613}{2.064} = 28.88^\circ\text{C} \text{ (temperature of the stream leaving the coagulation unit)}$$

BALANCE OVER THE DILUTION TANK

overall balance

heat in = heat out

$$\text{heat in} = m_1 c_1 \theta_1 + m_2 c_2 \theta_2$$

$$\text{heat out} = m_3 c_3 \theta_3$$

$$c_1 = 2.68 \text{kJ / kgk}, c_2 = 2.42 \text{kJ / kgk}, c_3 = 2.45 \text{kJ / kgk}$$

$$0.7256(2.68(28)) + 4.369(2.42)(30) = 5.095(2.45)\theta_3$$

$$54.45 + 317.19 = 12.48\theta_3$$

$$\theta_3 = \frac{371.64}{12.48} = 29.78^\circ\text{C} \text{ (temperature of diluted solution)}$$

BALANCE OVER THE EVAPORATOR

overall balance

$$m_f c_{pf} (T - T_f) + (m_f - m) \lambda = Q = m_s \lambda_s$$

where

c_{pf} specific heat capacity of the feed = **4.1074 kJ/kgK**

m_f mass flow rate of feed = **4.972 kg / s**

m mass flow rate of concentrate = **0.0695 kg/s**

λ & λ_s latent heat of vapourisation and condensation

m_s mass flow rate of steam

assumption

pressure of vapour space = **10 kN/m²**

steam pressure = **250 kN/m²**

temperature of concentrate = **45°C**

$$4.972(4.1074)(45 - 30) + 4.9025(2392) = 12,033 \text{ kJ}$$

$$12033 = m_s (2181)$$

$$m_s = \frac{12033}{2181} = 5.52 \text{ kg / s (mass of steam needed)}$$

SUMMARY OF MATERIAL BALANCE OVER THE WHOLE PLANT

Unit	In	Generated	Total	Out	Accumulation	Total
Acidification	0.7418 + 0.0172	-	0.759	0.7352	0.0238	0.7589
Coagulation	0.7352 + 0.0351	-	0.7703	0.7256	0.0447	0.7703
Dilution	0.7256 + 4.369	-	5.095	5.095	-	5.095
Cation exchange	5.095		5.095	5.0654*	0.02958	5.095
Anion exchange	5.019		5.019	4.9721	0.0468	5.019
Storage tank	4.9721	-		4.972	0.0001	4.9721
Evaporation	4.972	-	4.972	0.0695 + 4.9025	-	4.972

* Generation included
All values in kg/s

CHAPTER FOUR

PLANT DESIGN (UNIT DESIGN)

The unit to be designed is the ion exchange unit comprising of the cation and anion exchanger, the main function of this unit is the removal of all inorganic salts present in the treated glycerine solution.

Design Data:

Ion exchange resin property (selected resin to be used)

Cation resin:

Trade name: IR132E Amberlite
Exchange capacity = 2.15eq/l
Density (true wet) = 1.31g/cm³
Effective size = 0.58mm
Bed volume /hr = 30/hr
Functional form = Hydrogen form

Anion resin

Trade name: IRA 410 Amberlite
Exchange capacity = 1.35eq/l
Density (true wet) = 1.12g/cm³
Effective size = 0.48mm
Bed volume = 20/hr
Functional form = Hydroxide form

4.1 ANALYSIS

MASS ANALYSIS

Total inorganic salts = 0.07524kd/s

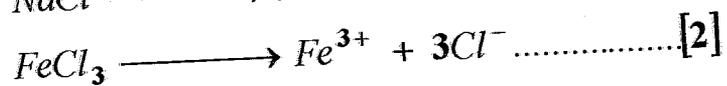
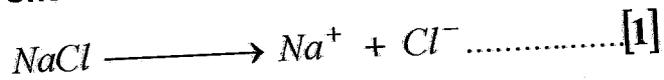
Composition: Na salt = 99.39

Fe salt = 0.61%

Mass of NaCl = 0.9939 × 0.0757 = 0.07524kg/s

FeCl₃ = 0.0061 × 0.0757 = 0.000462kg/s

Chemical analysis



from equation [1] and [2]

1mole of NaCl ≡ 1mole of Na⁺ ≡ 1mole of Cl⁻

1mole of FeCl₃ ≡ 1mole Fe³⁺ ≡ 3moles of Cl⁻

$$\text{no of moles of NaCl} = \frac{0.07524}{58.5} = 1.2862 \times 10^{-3} \text{ kmoles (1.2862 moles)}$$

$$\therefore \text{no of moles of Na}^+ = 1.2862 \text{ moles}$$

$$\text{mass of Na}^+ = 1.2862 \times 23 = 29.582 \text{ g / s}$$

$$\text{no of moles of FeCl}_3 = \frac{0.4618}{162.5} = 0.002842 \text{ moles}$$

$$\therefore \text{no of moles of Fe}^{3+} = 0.002842 \text{ moles}$$

$$\text{mass of Fe}^{3+} = 0.002842 \times 56 = 0.15914 \text{ g / s}$$

$$\text{Total no of moles of Cl}^- = 1.2862 + 3(0.002842) = 1.29473 \text{ moles}$$

$$\text{mass of chloride} = 1.29473 \times 35.5 = 45.963 \text{ g/s}$$

REFERENCES

Perry & Chilton
1973

LORCH

DESIGN CALCULATIONS

CATION EXCHANGER

To remove Na (29.582g/s) and Fe (0.15914g/s)

Flow rate

$$\text{Mass flow rate} = 5.095 \text{ kg / s}$$

$$\text{Specific density} = 1.0003$$

$$\text{Temperature} = 28^\circ\text{C}$$

$$\text{density of water at } 28^\circ\text{C} = 994.2 \text{ kg/m}^3$$

$$\text{density of solution} = 1.0003 \times 994.2 = 994.5 \text{ kg/m}^3$$

$$\text{volumetric flow rate} = \frac{\text{mass flow rate}}{\text{density}} = \frac{5.095}{994.5}$$

$$= 0.005123 \text{ m}^3/\text{s}$$

$$Q = 18.444 \text{ m}^3 / \text{hr}$$

CONVERTING TO EQUIVALENT WEIGHT OF CaCO_3

Na^+

$$\text{mass per litre} = \frac{106.495}{18.444} = 5.774 \text{ kg/m}^3 = 5.774 \text{ g/l}$$

$$\text{equivalent / litre} = \frac{5.774}{23} = 0.251 \text{ eq / l}$$

$$\text{equivalent mass (g CaCO}_3) = 0.251 \times 50 = 12.552 \text{ g / l}$$

Fe^{3+}

$$\text{mass per litre} = \frac{0.573}{18.444} = 0.0311 \text{ kg / m}^3 = 0.0311 \text{ g / l}$$

$$\text{equivalent per litre} = \frac{0.0311}{18.667} = 0.0016643 \text{ eq/l}$$

$$\text{equivalent mass (g CaCO}_3) = 0.0016643 \times 50 = 0.083214 \text{ g/l}$$

$$\text{total cation} = 0.083214 + 12.552 = 12.6352 \text{ g/l}$$

$$\begin{aligned} \text{ion load per hour} &= Q \text{ m}^3/\text{hr} \times \text{ion load / litre} \\ &= 18.444 \times 12.6352 = 233.044 \text{ kg/hr} \end{aligned}$$

Rohm & Haas Co.
summary 11, 1986

$$\text{Resin Volume (bed volume)} = \frac{\text{Total ion load}}{\text{Ion exchange capacity of resin selected}}$$

$$2.15 \times 50 = 107.5 \text{ kg / m}^3$$

$$\text{Bed volume} = \frac{233.044}{107.5} = 2.168 \text{ m}^3$$

allow for 15% safety factor to cover for flow rate
and concentration variation

$$\therefore \text{Bed volume} = 1.15 \times 2.168 = 2.50 \text{ m}^3$$

HYDRAULICS REQUIREMENT

Space velocity (service flow rate)

Rohm & Haas Co.
summary 11, 1986.

$$= \frac{\text{volumetric flow rate}}{\text{bed volume}} = \frac{18.444}{2.5} = 7.38 \text{ m}^3 / \text{h} / \text{m}^3 \text{ of bed}$$

Rohm & Haas Co.
summary 11, 1986

Empty bed contact time

$$= \frac{1}{\text{service flow rate}} = \frac{1}{7.38} = 0.13555 \text{ hr (8.133 minutes)}$$

Rohm & Haas Co.
summary 11, 1986

Linear velocity

$$= \frac{\text{volumetric flow rate}}{\text{area of bed}}$$

$$\text{Area} = \frac{\text{volume of bed}}{\text{height of bed}}$$

height of bed ranges from 1 - 2.5m

$$\text{take } h_B = 1.2 \text{ m}$$

$$\therefore \text{Area} = \frac{2.5}{1.2} = 2.0833 \text{ m}^2$$

$$\therefore \text{linear velocity} = \frac{18.444}{2.0833} = 8.853 \text{ m/h (0.00246 m/s)}$$

Lorch .W, 1986

Sizing and Dimensioning

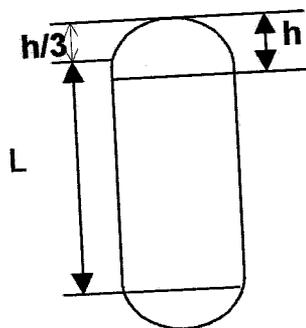
Diameter of bed = D = diameter of column

$$D = \left(\frac{4 \times V_B}{\pi \times h_B} \right)^{1/2} = \left(\frac{4 \times V_B}{\pi \times 1.2} \right)^{1/2} = 1.63 \text{ m}$$

Allow for 100% backwashing of the bed during rinsing

$$h_C = h_B \times 2 = 2.4m$$

Cover head: use of ellipsoidal doomed heads



Take $h = 0.1L$

$$h = 0.1 \times 2.4 = 0.24m$$

$$\therefore \text{minor axis} = \frac{h}{3} = \frac{0.24}{3} = 0.08m$$

$$\text{total height of the column} = 2.4 + 2(0.08) = 2.56m$$

Calculating the thickness of the column

cylindrical section :

$$e = \frac{P_i D_i}{2f - P_i} = \text{minimum thickness}$$

where P_i = Internal pressure

D_i = internal diameter

f = design stress

material of construction : stabilised stainless steel

To choose operating pressure we analyse pressure drop over the whole column

$$\Delta P_T = \Delta P_B + \Delta P_H + \Delta P_E + \Delta P_O$$

$\Delta P_E, \Delta P_O$ are pressure drop at the entrance and outlet

$\Delta P_B, \Delta P_H$ are pressure drop over the bed and free space

take ΔP_E and ΔP_O as negligible.

$$\Delta P_B$$

from the graph

$$\text{pressure drop per m of bed} = 10.02 \text{KN/m}^2$$

$$\text{pressure drop over the bed} = 10.02 \times 1.2 = 12.024 \text{KN/m}^2$$

$$\Delta P_H = \rho gh = 994.5 \times 9.81 \times 1.2 = 11.71 \text{KN/m}^2$$

$$\text{Total pressure drop over the column} = 11.71 + 12.024$$

BS 5000, 1982

BS 5500, 1982

Rohm & Haas Co.
summary 11, 1986

$$= 23.734 \text{ KN} / \text{m}^2$$

take total pressure drop = $30 \text{ KN} / \text{m}^2$

Choose an operating pressure of $100 \text{ KN} / \text{m}^2$

Using a design pressure of $120 \text{ KN} / \text{m}^2$

$$f \text{ at } 0 - 50^\circ\text{C} = 165 \text{ N} / \text{mm}^2$$

$$D_i = 1.63 \text{ m} = 1630 \text{ mm}$$

$$\therefore e = \frac{0.12 \times 1630}{2(165) - 0.12} = 0.593 \text{ mm}$$

$e \ll$ than the minimum thickness required to carry the load
of a cylindrical structure having this diameter.

$$e = 8 \text{ mm}$$

BS 5000, 1982

ellipsoidal head

$$e = \frac{P_i D_i}{4jf - 1.2P_i}$$

$$j = \text{joint factor} = 1$$

$$e = \frac{0.12 \times 1630}{4(165)1 - 1.2(0.12)} = 0.296 \text{ mm}$$

$e \ll$ less than minimum

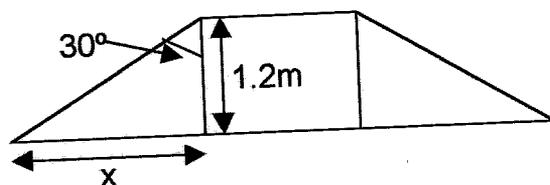
$$e = 8 \text{ mm}$$

BS 4289, 1975

Distributor selections for cation bed

Type: perforated plate

Take maximum angle of spraying as 30°C



From the diagram

$$\tan 30^\circ = \frac{x}{1.2} \Rightarrow x = 1.2 \tan 30 = 0.693 \text{ m}$$

$$\text{Diameter of the distributor} = D_C - 2x$$

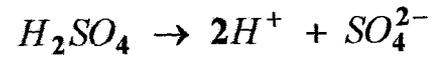
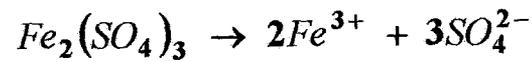
$$1.63 - 2(0.693) = 0.244 \text{ m}$$

Use distributor of **0.25m** diameter

Regeneration and rinsing

Regenerant: Sulphuric acid

Analysis of regeneration



from the equations

$$2 \text{ moles of } Fe^{3+} \equiv 6 \text{ moles of } H^+$$

$$1 \text{ mole of } Na^+ \equiv 1 \text{ mole of } H^+$$

$$\text{total moles of } Fe^{3+} \text{ to be removed} = 0.002842 \text{ moles/s}$$

$$\text{amount of } H^+ \text{ needed} = 3 \times 0.002842 = 0.008526 \text{ moles} \\ (0.0307 \text{ kmol/hr})$$

$$\text{total moles of } Na^+ \text{ removed} = 1.2862 \text{ moles/s}$$

$$\text{amount of } H^+ \text{ needed} = 1.2862 \text{ moles/s} (4.6303 \text{ kmol/hr})$$

$$\text{total moles of } H^+ \text{ needed} = 4.6303 + 0.0307 \\ = 4.661 \text{ kmol/hr}$$

amount of H_2SO_4 to produce the required H^+

$$2 \text{ moles of } H^+ \equiv 1 \text{ mole of } H_2SO_4$$

$$\text{no of moles needed} = \frac{4.661}{2} = 2.3305 \text{ kmol/hr}$$

$$\text{mass of acid required} = 2.3305 \times 98 = 228.39 \text{ kg/hr}$$

Allow for 10% safety factor to cover for excess ion load

$$= 1.1 \times 228.39 = 252 \text{ kg/hr}$$

$$\text{regenerant rate} = \frac{252}{2.5} = 100.8 \text{ kg/m}^3 \text{ of bed}$$

Regeneration flow rate

Rohm & Haas Co.

IX - CU - 79

Use 4% (weight percent) sulfuric acid solution

Density of solution = 1022.3 kg/m^3

$$\text{volumetric flow rate} = \frac{252}{1022.3(0.04)} = 6.16 \text{ m}^3 / \text{cycle}$$

take regeneration time = 20 minutes

$$\therefore \text{flow rate} = 6.16 \times 3 = 18.49 \text{ m}^3 / \text{hr}$$

Rinsing requirement

Rohm & Haas Co.

Bp 932, 126

Range: $3.34 - 6.68 \text{ m}^3 / \text{m}^3$ of bed

take rinsing rate = $3.5m^3/m^3$ of bed

Rinse requirement = $3.5 \times 2.5 = 8.75m^3 / m^3$ of bed

Take rinsing time = 15 minutes

Flow rate requirement = $8.75 \times 4 = 35m^3 / hr$

ANION EXCHANGER

Removal of 45.963g/s of Cl^- (165.47kg/hr)

Flow rate

mass flow rate = 5.019kg / s

specific gravity = 1.0002

temperature = 26°C

density of water = 994.8kg / m^3

density of solution = $1.0002 \times 994.8 = 995kg/m^3$

vol. flow rate = $\frac{5.019}{995} = 0.005044m^3 / s$ (18.16 m^3 / hr)

Converting to equivalent weight

Mass/litre = $\frac{165.47}{18.16} = 9.112g / l$

Equivalent/litre = $\frac{9.112}{35.5} = 0.2567eq / l$

Equivalent mass = $0.2567 \times 50 = 12.833g / l$

Total ion load = $12.833 \times 18.16 = 233.052kg / hr$

Rohm & Haas Co.
summary 11, 1986

Bed volume (resin volume)

$$V_B = \frac{233.052}{67.5} = 3.453m^3$$

allow for 15% safety factor to cover for variation in ion load

$$\therefore V_B = 1.15 \times 3.453 = 3.97m^3$$

HYDRAULICS REQUIREMENT

Rohm & Haas Co.
summary 11, 1986

Space velocity = $\frac{18.16}{3.97} = 4.574m^3 / hr / m^3$ of bed

$$E.B.C.T = \frac{1}{4.574} = 0.2186hr(13.12 \text{ minutes})$$

Rohm & Haas Co.
summary 11, 1986

Linear velocity

$$\text{take } h_B = 1.5m$$

$$\therefore \text{area} = \frac{3.97}{1.5} = 2.65m^2$$

$$\text{linear velocity} = \frac{18.16}{2.65} = 6.86m / hr(0.00191m / s)$$

Lorch W. 1986

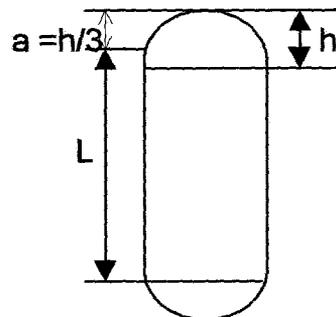
Dimensioning and sizing

$$D = \left(\frac{4 \times 3.97}{\pi \times 1.5} \right)^{1/2} = 1.84m$$

allow for 100% backwashing during rinsing

$$h_C = 1.5 \times 2 = 3m$$

Cover head: Ellipsoidal doomed head



BS 5000, 1982

Take $h = 0.1L$

$$h = 0.1 \times 3 = 0.3m$$

$$\therefore a = \frac{h}{3} = \frac{0.3}{3} = 0.1m$$

$$\text{Total height of column} = 3 + 2(0.1) = 3.2m$$

Calculating the thickness

Cylindrical section:

Choosing operating pressure:

pressure drop over the column

$$\Delta P_B$$

from the graph

$$\text{pressure drop per m of bed} = 8KN/m^2$$

$$\text{pressure drop over the bed} = 8 \times 1.5 = 12KN/m^2$$

$$\Delta P_H = 995 \times 9.81 \times 1.5 = 14.64KN / m^2$$

$$\text{Total pressure drop} = 12 + 14.64 = 26.64 \approx 30KN/m^2$$

Take operating pressure = 100KN/m^2

While design pressure = 120KN/m^2

Material for construction: stabilized stainless steel

$$e = \frac{0.12 \times 1840}{2(165) - 0.12} = 0.66\text{mm}$$

$e \ll$ than the minimum required

$$\therefore e = 8\text{mm}$$

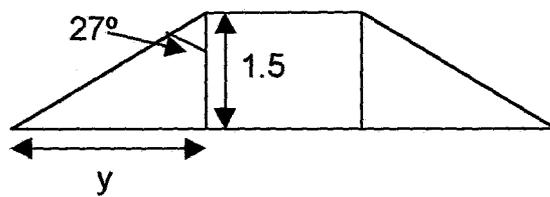
use same thickness for the head cover

BS 5500, 1982.

Distributor selection

Take maximum angle of spraying as 27°

BS 4289, 1975



From the diagram

$$\tan 27 = \frac{y}{1.5} \Rightarrow y = 1.5 \tan 27 = 0.765\text{m}$$

$$D_D = 1.84 - 2(0.765) = 0.31\text{m}$$

REGENERATION AND RINSING

Regenerant: caustic soda

Analysis



from the equations

$$1 \text{ mole of } \text{Cl}^- \equiv 1 \text{ mole of } \text{OH}^-$$

Total no of moles of Cl^- to be removed = 4.661 Kmoles/hr

no of moles of OH^- needed = $4.661 \text{ kmoles / hr}$

no of moles of NaOH needed = 4.661 kmoles/hr

mass of NaOH = $4.661 \times 40 = 186.44\text{kg/hr}$

allow for 10% safety factor to cover for variation in ion load

$$= 1.1 \times 186.44 = 205 \text{ kg / hr}$$

$$\text{regenerant rate} = \frac{205}{3.97} = 51.64 \text{ kg / m}^3 \text{ of bed}$$

Flow rate requirement

use of 4% caustic soda solution

$$\text{density of NaOH solution} = 1039.04 \text{ kg/m}^3$$

$$\text{vol. flow rate of regenerant} = \frac{205}{1039.04(0.04)} = 4.932 \text{ m}^3 \text{ / cycle}$$

take regeneration time to be = 20 minutes

$$\therefore \text{flow rate requirement} = 4.932 \times 3 = 14.80 \text{ m}^3 \text{ / hr}$$

Rinsing water

use 4.0 m³ / m³ of bed

$$\therefore \text{rinse water needed} = 4.0 \times 3.97 = 15.88 \text{ m}^3 \text{ / cycle}$$

flow rate requirement

use rinsing time of 15 minutes

$$\therefore \text{flow rate} = 15.88 \times 4 = 63.52 \text{ m}^3 \text{ / hr}$$

PROCESS PIPING ANALYSIS AND DESIGN

Required pipe diameter

Inlet pipe of the cation bed:

Use carbon steel pipe

$$\text{Optimum pipe diameter} = 293G^{0.53}\rho^{-0.37}$$

Sinnott, R.K., 1993

$$G = 5.095 \text{ kg / s}$$

$$\rho = 994.2 \text{ kg / m}^3$$

$$\therefore D = 293 \times 5.095^{0.53} \times 994.2^{-0.37} = 54 \text{ mm}$$

use 60mm inlet pipe

inlet pipe of the column (outlet pipe of the cation)

use of carbon steel pipe

$$D = 293 \times 5.019^{0.53} \times 995^{-0.37} = 53.58 \text{ mm}$$

use 60mm diameter pipe

Pump selection

Analysis

$$\frac{P_1 - P_2}{\rho} + \frac{v_1 + v_2}{2} = z_2 - z_1 - h_l - h_p$$

$$\text{where } \rho = 995 \text{ kg/m}^3, P_1 = 70 \text{ kN / m}^2, P_2 = 120 \text{ kN / m}^2$$

$$z_1 = 0.3 \text{ m}, z_2 = 3.8 \text{ m}, z_3 = 0.3 \text{ m}, Q = 18.16 \text{ m}^3 / \text{hr}$$

$$v_1 = v_2 = \frac{Q}{A_1}, A_1 = \pi \frac{D^2}{4} = \frac{\pi \times 0.06^2}{4} = 0.00283 \text{ m}^2$$

$$v = \frac{18.16}{3600(0.00283)} = 1.784 \text{ m / s}$$

$$h_l = \sum k_1 \frac{v}{2g} + \sum k_2 \frac{v}{2g} + f \frac{L}{D} \frac{v}{2g}$$

use of globe Valves

Chopey N.P, 1994

$$k_1 = 0.8 \text{ (4, } 90^\circ \text{ bends)} \quad k_2 = 1.0 \text{ (valves, globe)}$$

$$\sum k_1 \frac{v^2}{2g} = 4 \times 0.8 \times \frac{1.784^2}{2 \times 9.81} = 0.519m$$

$$\sum k_2 \frac{v^2}{2g} = 2 \times 1 \times \frac{1.784^2}{2 \times 9.81} = 0.325m$$

$$Re = \frac{\rho v D}{\mu} = \frac{995 \times 1.784 \times 0.06}{0.00278} = 38,311.1$$

flow is turbulent

$$f = \frac{64}{Re} = \frac{64}{38,311.1} = 0.0017$$

$$L = 0.3 + 4(0.1) + 2(0.15) + 2.5 + 1.235 + 0.3 = 8.835m$$

$$f \frac{L}{D} \frac{v^2}{2g} = 0.0017 \times \frac{8.835}{0.06} \frac{1.784^2}{2 \times 9.81} = 0.041m$$

$$h_l = 0.041 + 0.519 + 0.325 = 0.885m$$

$$\therefore h_p = 3.8 - 0.6 - 0.885 - \frac{70 - 120}{995} = 2.37m$$

allow for 10% safety factor

$$h_p = 1.1 \times 2.37 = 2.60m$$

pump selected : Centrifugal pump (vertical turbine)

rating: 1 hp

Goulds pumps
chart, 1996

SPECIFICATION SHEET

Ion exchange column			
Identification:			
Item	cation exchange column	Date	1 - 4 - 2000
Item No	CE		
No required	: 2	by	JBH
Function: Removal of cations in treated spent lye solution			
Operation: Continuous			
Materials handled:	Feed	Product	Accumulations
Quantity (kg/s)	5.095	5.019	0.02958
Temperature(°C)	28	25	
Design data:			
Resin used:	IR 132E Amberlite		
Volumetric flow rate of feed:	18.444m ³ /hr		
Diameter of column:	1.63m		
Diameter of resin bed:	1.63m		
Height of column:	2.56m		
Height of resin bed:	1.2m		
Operating temperature:	30°C (maximum)		
Operating pressure:	100kN/m ²		
Design pressure:	120kN/m ²		
E.B.C.T:	8.133minutes		
Superficial velocity:	7.38m ³ /hr m ³ of bed		
Linear velocity:	0.00246m/s		
Feed pipe diameter:	60mm		
Product pipe diameter:	60mm		
Material of construction:	stabilised stainless steel (8mm range)		
Regenerant:	Sulphuric acid		
Regenerant rate:	100.8kg/m ³ of bed		
Utilities: water for rinsing of bed			
Controls: Temperature inside column controlled by ON/OFF controllers to shut down stream flow into the column. Also use of timer controllers for the selection of valves.			
Tolerances: Rohm and Cass co. summary 11, 1986.			

SPECIFICATION SHEET

Ion exchange column			
Identification:			
Item	Anion exchange column	Date	1 - 4 - 2000
Item No	AE		
No required	: 2	by	JBH
Function: Removal of anions (Cl) in treated spent lye solution			
Operation: Continuous			
Materials handled:	Feed	Product	Accumulations
Quantity (kg/s)	5.019	4.9721	0.0468
Temperature(°C)	25	23	
Design data:			
Resin used:	IRA 410 Amberlite		
Volumetric flow rate of feed:	18.16m ³ /hr		
Diameter of column:	1.84m		
Diameter of resin bed:	1.84m		
Height of column:	3.20m		
Height of resin bed:	1.5m		
Operating temperature:	30°C (maximum)		
Operating pressure:	100kN/m ²		
Design pressure:	120kN/m ²		
E.B.C.T:	13.12minutes		
Superficial velocity:	4.574m ³ / hr m ³ of bed		
Linear velocity:	0.00191m/s		
Feed pipe diameter:	60mm		
Product pipe diameter:	60mm		
Material of construction:	Stabilised stainless steel (8mm range)		
Regenerant:	Caustic soda (NaOH)		
Regenerant rate:	51.64kg/m ³ of bed		
Utilities: water for rinsing of bed			
Controls: Temperature inside column controlled by ON/OFF controllers to shut down stream flow into the column. Also use of timer controllers for the selection of valves.			
Tolerances: Rohm and Cass co. summary 11, 1986.			

4.3 EQUIPMENT COSTING (CAPITAL COST)

Total capital (fixed) cost of each exchanger column (C_t)

$$C_t = \text{cost of column construction} + \text{cost of ion exchange resins} + \text{workmanship}$$

Cost of column construction

Material for construction: stabilized stainless steel

Thickness of steel plate = 8mm

Price per m^2 of steel plate = N5,500

Cation column :

Total m^2 needed

$$\text{total cross sectional area of the column} = \pi r(2r + h)$$

$$h = 2.56$$

$$r = 0.815$$

$$\Rightarrow \pi \times 0.815(2(0.815) + 2.56) = 10.73m^2$$

$$\text{allow for 10\% safety factor} \Rightarrow 1.1 \times 10.73 = 11.80 \cong 12m^2$$

$$\text{Cost} = 5,500 \times 12 = N66,000$$

Cost of cation resins (IR 132E Amberlite)

Cost of 1 ton of resin (IR 132E) = N35,000

$$\begin{aligned} \text{mass of resin required} &= \text{volume of resin} \times \text{density} \\ &= 2.5 \times 1310 = 3275kg \end{aligned}$$

$$\text{Total cost of resin} = 3.275 \times 35000 = N114,625$$

$$\text{Cost of column} = 114,625 + 66,000 = N180,625$$

$$\text{take workmanship as 30\% of column cost} \Rightarrow 0.3 \times 180,625 = N54,188$$

$$\text{Total cost(capital)} = 180,625 + 54,188 = N234,813$$

$$\begin{aligned} \text{create for contingency, take 5\% of capital cost} &\Rightarrow 1.05 \times 234,813 \\ &= N246,553 \cong N250,000 \end{aligned}$$

$$\text{Capital cost of cation exchange column} = N250,000$$

Anion column :

Total m^2 needed

$$h = 3.2m, \quad r = 0.92m \Rightarrow \pi \times 0.92(2(0.92) + 3.2) = 14.47m^2$$

account for distributors and collector : allow for 10\% safety factor

$$\text{needed } m^2 = 1.1 \times 14.47 = 16.00m^2$$

$$\text{Cost of column(steel plate)} = 5,500 \times 16 = N88,000$$

Cost of anion resin (IRA 410 Amberlite)

$$\text{Cost of 1 ton of resin (IRA 410)} = N32,000$$

$$\text{mass of resin required} = 3.97 \times 1120 = 4446.4\text{kg}$$

$$\text{Total cost of resin} = 4.4464 \times 32,000 = N142,285$$

$$\text{Cost of column} = 142,285 + 88,000 = N230,285$$

$$\text{Take workmanship as 30\% of column cost} \Rightarrow 0.3 \times 230,285 = N69,086$$

$$\text{Total capital cost} = 230,285 + 69,086 = N299,371$$

$$\text{create for contingencies : 5\% of total cost} \Rightarrow 0.05 \times 299,371 = N14,969$$

$$\text{Capital cost} = 299,371 + 14,969 = N314,340 \cong N320,000$$

$$\begin{aligned} \text{Total (capital) cost of ion exchange columns} &= N320,000 + N250,000 \\ &= N570,000 \end{aligned}$$

(All cost based on present open market, viable for 6 month from day of submission)

CHAPTER FIVE

SAFETY AND ECONOMIC CONSIDERATIONS

5.1 SAFETY CONSIDERATIONS

The safety analysis of the process and plant design can be classified as general safety consideration and unit (section) safety consideration.

General safety consideration are those that affect the whole plant or processes a whole while the unit safety consideration are those that affect each unit and later on affect the overall efficiency of the plant.

General safety consideration

This can be analysed under the following

- (1) Health and safety considerations
- (2) Fire and explosion consideration
- (3) Personnel safety

5.1.1 Health and safety considerations

The potential health hazard to an individual by a material used in a chemical process is a function of the inherent toxicity of the material and the frequency and duration of exposure.

The plant yields no material that is toxic but there is the presence of acidic solution tanks. The safety consideration for this

- (1) The tank should be well labelled so that anybody could notice the presence of acid tanks and also only authorised personnels should be allowed to go near the storage tanks.
- (2) When new raw materials are being supplied proper laboratory test should be carried out to make sure the newly supplied raw material does not continue toxic substance or substances that can form toxic substances.

5.1.2 Fire and explosion hazards

Fire and explosion are the two most dangerous events likely to occur in a production plant. For a fire to occur, there must be a fuel, an oxidizer and also an ignition source. In addition the combustion must be self-sustaining.

The plants uses no fuel operation apart from the boiler section, but the presence of electrical driven tools and installation can be a source of fire and latter leads to explosion.

The safety considerations are:

- (1) Proper and periodic installation of all electrical parts and also maintenance.
- (2) The powerhouse (boiler house) should be properly put in order, and only designated and experienced personnel should be allowed to monitor the affair.
- (3) Installation or provision of fire alarms and fire extinguishers in all strategic position.

5.1.3 Personnel safety

The safety considerations are:

- (1) Training of all personnels on the issue of safety
- (2) Plant layouts, enough protected pathways, plat forms, stairs and work areas is provided in the design layout.
- (3) Enough spaces between equipment
- (4) Proper medical facilities has being provided in form of first aid.

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5.1.4 Acidification units and precipitation units

Possible hazards are:

- (1) Spillage
- (2) Electric shock
- (3) Incomplete reaction

Spillage

The following could cause the spillage:

- (1) Over flowing of the batch reactor
- (2) Malfunctioning of the pumps and valves
- (3) Over powering of the agitator

Safety considerations

- (1) Check that the batch reactor is empty before loading new reactants
- (2) Pumps and valves should be checked and maintained periodically
- (3) The reactor is fitted with a level controller, which is shut down the valves at a particular level before spillage.

- (4) Speed of the agitator should be monitored and the motor rolling the agitator should be checked periodically.

Electric shock

- (1) Checking of the mains apply switch periodically to see that they are performing regularly.
- (2) Personnels working at this unit should be provided with safety loop and gloves.

Incomplete reaction

This can be caused by

- (1) Incomplete reaction time
- (2) Variation in stream
- (3) Low efficiency or drop in the efficiency of the agitator
- (4) High temperature (explosion).

Safety considerations

1. For any reaction to be complete there must be a minimum reaction time, proper monitoring of this reaction time should be emphasized. This is done by using an automatic controller that relates the reaction time with the discharge valve.
2. Variations in process streams are corrected by carrying out extensive laboratory and technological test on the streams regularly especially when the material is just been supplied to the plant before usage.
3. Reaction time is a function of mixing, the agitator should be properly fitted to the right dept so as to achieve the required efficiency.

5.1.5 Ion exchange unit

Possible hazards

1. Damage of ion exchange resins
2. Low efficiency of the unit

Damage of ion exchange resins

Causes:

1. Temperature

Safety considerations

1. Proper monitoring of the operation in the dilution tank.

2. Temperature indicator and controller is fitted to the ion exchange unit to shutdown the streams when the temperature of the stream reaches a particular temperature

Low efficiency

Causes

1. Down stream factors
2. Poor regeneration practice

Safety considerations

1. All down stream problems or variations from normal operating condition affects the ion exchange unit. The ion exchange unit is fitted with on and off controllers that immediately shuts down the ion exchange unit when there is any down stream problem that can not be corrected immediately
2. Regeneration initiation is carried out automatically.

5.1.6 Evaporating unit

Likely hazards

1. Explosion

Safety consideration

1. Use condensers instead of steam ejectors.
2. Only trained and designated personnel should be allowed into this unit.
3. Proper and periodic checking of the steam pressure.
4. Fitting of pressure vents.

5.2 GENERAL SAFETY RULES

1. Proper visual signpost to show pathways, and restricted areas
2. Only designated personnel should be allowed into the units
3. Periodic training of staffs on safety rules
4. Good spacing of the units
5. Periodic checking of electrical installations, pumps and fittings
6. Provision of safety gadgets ranging from factory wear, to gloves, to boot and strict adherence should be placed on them in the production hall.
7. Proper communication between the labs (quality control, and control room) and the production floor.

5.3 PLANT ECONOMICS

The production or recovery of glycerine itself is a birth of the economical advantage of the soap industry. The economics of the plant designed ranges from the economical usage of fittings like, pumps to reutilisation of materials.

5.3.1 *General economic consideration*

Piping network

The plant designed involves lot of dual piping network, instead of using multiple pumps to achieve this aims the use of valves where employed which is relatively cheaper than using multiple pumps.

Controllers

Most of the controllers used are automatic on and off controllers because of the nature of the operation, and also these controllers are relatively cheaper

5.3.2 *Acidification unit and precipitating unit*

This unit produces precipitates of organic matter, ash and little or no effort and finance is used in treating this material before disposal with little or no legal restrictions.

5.3.3 *Ion exchange unit*

During regeneration, at the cation exchanger, Na_2SO_4 is produced, and this can be sold to fertiliser companies as a source of income. Also at the anion exchangers NaCl is produced which is taken back to the soap plant to be used for salting out of the soap.

5.3.4 *Evaporating unit*

The distilled water produced from the water vapour removed is used back at the dilution tank and part of it is used as boiler feed.

CHAPTER SIX

CONCLUSION AND RECOMMENDATION

6.1 CONCLUSION

The design of a plant for the production (recovery from spent lye) of 2,000 tons per year of glycerine has been carried out.

The process design shows a better efficiency route than the existing method because the modified processes removes or care for the scaling effect which has been the major problem of the formal method. The new process also looks safer from the safety analysis carried out since there is no production f any material that could be dangerous to the personnels. Also the economic analysis of this new process and operations shows that the plant is more economical.

Ion exchange column has been designed using relevant standards and the results obtained shows accuracy with the standards. The capital costing of the ion exchange column carried out using open market pricing was found to be N570,000 which is relatively cheaper considering the capital cost of a distillation column.

6.2 RECOMMENDATION

Design is all about creativity, and creativity is all about geniusity, but certainly you can no what you don't learn. Proper atmosphere has to be created in other to be creative.

One major problem with design is data, proper data provision should be made so that accurate and more efficient and effective design could be carried out. Also it is not enough to design, but effort should be made in seeing that some of these designed works are being put to reality.

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