PRELIMINARY DESIGN OF A PLANT TO PRODUCE SODIUM HYDROXIDE FROM SODIUM CHLORIDE

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DECLARATION

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

OJUNTA SAM DATE

CERTIFICATION

This is to certify that I have supervised, read and approved this project "PRELIMINARY DESIGN OF A PLANT TO PRODUCE SODIUM HYDROXIDE FROM SODIUM CHLORIDE" which I have found adequately both in scope and quality, for the partial fulfillment of the requirement for award of Bachelor of Engineering degree in chemical Engineering degree in Chemical Engineering Department, Federal University of Technology, Minna.

12/12/2005

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DATE

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DATE

EXTERNAL EXAMINER

DATE

DEDICATION

I wish to dedicate this project work to my darling parents Chief & Mrs. Ebenezer Walter Ojunta, my lovely Elder Brothers and sisters for their love, support and care.

ACKNOWLEDGEMENT

I wish to express my profound gratitude to Almighty God for giving me the wisdom and strength to carry out my work perfectly.

My sincere gratitude goes to my supervisor, Prof. K. R. Onifade who was ready to criticize and offer alternative suggestion at every stage of this project.

I also wish to thank the head of the department, DR. F. A. Aberuagba, members of staff, academic and non-academic as well as the entire students of the department of chemical engineering for their support through out my stay in the department.

I want to specially thank my parents Chief & Mrs. Walter Ojunta and my guidance for their financial and moral support through out my studies. My progress in life in as a result of their encouragement and untiring prayer.

This acknowledgement would be incomplete without saying "ADIEU" to my bosom friend who died after a brief illness 3 years back in school late Tony Achimugu, may your soul rest in perfect peace and to all my graduating students of 2005, to all crew members in the show biz industry especially the Blinx Crew. And to my friends like Prince, Ugo, Vayon, Bobby, Mato, Madu, Sheriff and as many more that I cannot list out right now. I'm saying thank you all for standing by me.

ABSTRACT

`The project is aimed at having a preliminary design of a plant to produce sodium hydroxide from sodium chloride. A basis of per hour of operation was used with the inputs calculated at the start of the material balance based on the production target. The detailed design work comprises of introduction, literature survey, and comparing the chosen method with other processing route. This was followed by detailed calculations material balance, energy balance, equipment design, including detailed chemical and mechanical design of the major equipment and cost estimation were carried out. A CAD program (MathCAD) was been used in the above design calculations. Safety, Control, environmental impact assessment and plant location and layout were also carried out in the project. Standard cost estimation was carried out. From the economic analysis calculated, the equipment cost of the plant is 61.05 million naira. The profit after tax was found to be 4.696 billion naira per annum, with pay back period of approximately two years. Hence the plant designed so far can be said to be economically viable with rate of return on investment (ROI) of 60.039%.

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CHAPTER ONE

1.0 GENERAL INTRODUCTION

Nigeria is endowed with numerous mineral resources. Recent policy reforms have brought the solid mineral sector to the limelight. The emphasis is on encouraging massive foreign investor participation to the sector (consulate, 2000).

The search for industrial development in Nigeria in the 1970's and 1980's led to the setting up of various industries and laboratories, which were imports dependent for basic raw materials. As a result, most of the industries operate with low capacity utilization. Thus, the process industries need essential linkage with local material input for sustainable and appreciable level of plant capacity utilization.

Solutions of caustic soda have been made ever since soda was commercially used, certainly since before the middle of the eighteenth century. Throughout the first half of the Nineteenth century, such solutions were made by soap makers, wool scourers, bleacheries and papermakers for use in their own plants.

Caustic soda is a major raw material used by most process industries (example in Lagos include Paterson Zochonis Industries, Ilupeju, Unilever Nigeria Plc Agbara Industrial Estate, Lagos) etc, the demand for these materials by the industries has necessitated a research project on "Preliminary design of a plant to produce Sodium Hydroxide from Sodium Chloride Nacl."

The high demand for caustic Soda is not restricted to Lagos State alone (example in Akwa Ibom include NNMC Oku Iboku, ALSCON Ikot Abasi etc) hence other states are not left out in the race.

1.1 PHYSICAL AND CHEMICAL PROPERTIES

NaOH is a white solid crystalline solid that is also known as caustic soda, Iye or sodium hydrates. Its chemical formular is NaOH. It is commonly available in the form of pellets, sticks, or chips, and in water solution of various concentrations. It is a strong base and easily soluble in water, alcohol, and glycerine. Sodium Hydroxide is used commonly in the manufacture of other chemicals; for example, in the production of rayon and other textiles, in the production of paper, in etching aluminium, in the production of soaps and detergents.

Caustic soda, as a 50% solution, in an colourless and colourless liquid. In all forms, caustic soda is highly corrosive and reactive. Caustic soda solution reacts readily with metals such as aluminium, magnesium, zinc, tin, chromium, bronze, brass, copper and tantalum. Galvanized (Zinc coated) material should be avoided. Contact with acids. Halogenated organics, organic nitro compound and clycol should be avoided. It reacts with most animal tissue, including leather, human skin, and eyes. It also reacts readily with various reducing sugars (i.e fructose, galactose, maltose, dry whey solids) to produce carbon monoxide.

Some of the physical properties of sodium hydroxide are summarised in Table 1.1 caustic soda in water solution, depending on the concentration can form five hydrates containing 1, 2, 3, 5 and 7 molecules of water, respectively. Hydrates formation is exothermic. Caustic solutions generate heat when further diluted with water. With concentration of 40% or greater, the heat generated can raise the temperature above the boiling point, resulting in sporadic, dangerous eruptions of the solution.

Table1.1: Physical Concentration of Pure NaOH.

Molecular weight	40.005
Specific gravity, $20^0 / 4^0$C	2.130
Melting point, ^{0}C	318
PH	14
Boiling point at 101.3kpa, ^{0}C	1390
Index refraction	1.3576
Latent heat of fusion, J/g	167.4
Heat of transition, alpha to beta J/g	103.3
Heat of formation form the elements: Alpha form, KJ/ Mol.	422.46
Beta form, KJ / mol.	426.60
Transition temperature, ^{0}C	299.60

bility at 20^0C. G/100g H_20	109
our pressure @ 68^0F (0.2KPa@ 20^0C),mmHg	1.5

eous solutions of sodium hydroxide are highly basic and especially useful in reaction weakly acidic materials where weaker bases such as sodium carbonate are effective.

Sodium hydroxide will both burn nor support combustion although in its reaction amphoteric metals such as aluminium, tin and zinc, hydrogen gas is generated which form an explosive mixture.

stic soda reacts with all the mineral acids to form the corresponding salts.

$SO_2 + 2NaOH \longrightarrow Na_2SO_3 + H_2O$……………………..i

$CO_2 + 2NaOH \longrightarrow Na_2CO_3 + H_2O$……………………..ii

stic soda reacts with amphoteric metals and their oxides to from soluble salts.

$Al(OH)_3 + NaOH \longrightarrow NaAlO_2 + 2H_2O$………..…..iii

organic acids also react with sodium hydroxide to soluble salts.

$_5(OOCR)_3 + 3NaOH \longrightarrow C_3H_5(OH)_3 + 3RCOONa$

ause of the complexity of some sodium hydroxide reactions with natural products, ple chemical reactions cannot be written to describe them. In this category are such tion as the solubilization of cotton in rubber reclaiming, the dextrination of starch, ring of cotton, refining of vegetable oils, and the removal of lignin and icelluloses in the craft pulping process.

USES OF SODIUM HYDROXIDE

Caustic soda is used in a wide variety of industrial applications. It is used as a tant in the manufacture of other sodium compounds which themselves may be rmediate or end – use products, such as sodium hypochlorite which is used as sehold bleach and disinfectant and sodium phenolate which is used in antiseptics and the manufacture of Aspirin. It is used in the manufacturing process of soaps and factants for used in soap powders and also in the textile industry to remove contaminants, as bleach in the treatment of scoured cloth and to improve lustier and dye absorption.

Caustic soda use in mainly confined to finishing operations in the textile industry, i.e.

Scouring: Removes contaminants such as pectin's, waxes etc

Beaching: Treatment of scoured cloth with an oxidizing agent.

Mercerizing: Improves lustier and dye absorption.

Oil Drilling: Caustic soda is used in oil drilling industry primarily to control the PH of drilling mud's and secondary as a bactericide and calcium remover.

Petroleum Oil Refining: Products form the refining operation contain impurities which are removed by treatment with various chemicals including caustic soda, which removes sulphur compounds and acidic compounds.

Caustic soda has a variety of uses in the food industry some examples are:

Refining of animal and vegetable oils to remove fatty acids prior to use in food stuff production, dry formulation, for bottle washing, general cleansing operations, cleaning of brewery equipment, Iye peeling of potatoes, fruits and vegetables.

Caustic soda is used extensively in the water industry for PH control and ion exchange resin regeneration. Other uses include effluent neutralization and descaling of pipe work system.

Caustic soda is commonly used as drying agents to remove CO_2 and H_2O simultaneously from products.

1.3 OBJECTIVE OF THE PROJECT

The industrial development strategy pursued in the country lead to establishing various industries, which were import dependent and therefore vulnerable to foreign exchange scarcity. As a result most of the industries operated at low capacity. Particularly the chemical and allied process industries. These industries need local linkages. Caustic soda and chlorine are commonly used chemicals in Nigeria and the ECOWAS sub – region (Accuah 1998) this retarded development of the process industries.

With the discovery of salt spring at awe (Plateau State), Abakaliki (Ebonyi State) and Uburu (Imo State) and rock salt availability in Benue State; hence a total reserve of 1.5 million tones of Rock salt has been identified in Nigeria and availability of energy in form of natural/associated gas, and hydroelectric power plants, the production of these chemicals locally seems an attractive venture, given their extensive application in the industries, this calls for a preliminary of caustic soda plants (RMRDDC 1997).

CHAPTER TWO

2.0 PRODUCTION HISTORY

The rise of the textile industry at the beginning of the industrial revolution meant large quantities of caustic soda were needed to make the soap used to scour the fibres. The caustic was made primary by mixing lumps of lime with a solution of sodium carbonate in an iron vessel and bubbling steam through the mixture. The insoluble chalk formed by this reaction together with the excess lime were allowed to settle and the solution of caustic soda was drawn off and concentration by evaporation, this process is now called solvay or lime soda process. However caustic soda is now produced with chlorine as co-product by the electrolytic method.

The electrolysis of brine to produce caustic soda and chlorine was known since the eighteen century but it was not until 1890 that it was actually produced in this way for industrial consumption. In highly industrialized nation, the electrolytic chlorine is more valuable than the caustic soda as a bye – product. The electrolytic method of production of these products requires a cheap source of electricity since it is energy consuming, and a ready source of salt, and proximity to the markets.

Caustic soda was one of the first chemicals to be manufactured on a commercial scale in the beginning of the industrial revolution (Othmer, 1982)

2.1 PROPERTIES OF SODIUM HYDROXIDE

Caustic soda is available in liquid form, and is available in a variety of concentrations. Caustic solution is a colorless, odorless liquid, and heavier than water. The freezing point of caustic soda solutions varies with concentration. A 20 percent solution will freeze at around -10 degree C; a 50 percent solution will freeze at around 18 degree C; and a 73 percent solution will freeze at around 62 degree C. Due to their freezing points, 50 percent and 73 percent solutions are shipped and stored warm. Caustic solutions react readily with metals such as aluminum, tin and zinc, and in the presence of chemicals like hydrochloric acid. It is a highly corrosive chemical, which is typically stored in carbon steel tanks.

2.2 USES OF SODIUM HYDROXIDE

The main uses are in chemical manufacturing (pH control); pulp and paper manufacturing; in petroleum and natural gas (acidic contaminants in oil and gas processing); manufacture of soaps and detergents; cellulosics such as rayon, cellophane and cellulose ethers.

Other uses include:

- Water treatment
- Food processing
- Flue-gas scrubbing
- Mining
- Glass processing
- Refining vegetable oils
- Rubber reclamation
- Metal processing
- Adhesive preparations
- Paint remover
- Disinfectant
- Sodium hypochlorite

2.3 PROCESSES OF SODIUM HYDROXIDE

The principle method for its manufacture is electrolytic dissociation of sodium chloride; chorine gas is a coproduct. Caustic Soda is produced mainly by one of these three processes: Diaphragm, Membrane or Mercury Cell.

2.3.1 Diaphragm - Named for its use of a Diaphragm to separate the Caustic Soda from the Chlorine; it is the most widely-used process and represents the majority of US capacity. It is the preferred grade in pulp and paper, alumina and refining.

2.3.2 Mercury Cell (Rayon) - This grade was originally created for the rayon fiber industry. It is of higher purity than Diaphragm due to the lower levels of Sodium Chloride and Iron. The negatives of this process include high electrical consumption and the possibility of traces of mercury in the product.

2.3.3 Membrane - Is the newest process and its level of purity is between Diaphragm and Mercury Cell. It is a higher grade material than Diaphragm but due to the levels of Sodium Chloride and Iron, is not as pure as Mercury cell. Membrane Grade is used primarily in applications such as food, photography and ion-exchange regeneration.

Table 2.1: Production Methods Summary

Process	Separation method	Comments
Mercury cell	Mercury cathode forms amalgam with sodium - hydrogen gas is not formed at this stage. Amalgam flows out of the cell where water is used to release the sodium and hydrogen	Oldest method, still very much in use but not chosen for new plant
Diaphragm cell	Steel and asbestos diaphragm separates anode and cathode compartment. Chlorine forms at anode, sodium migrates through diaphragm to cathode compartment - rate of liquid flow prevents hydroxyl ions migrating to anode compartment	More common in the US than EU. Higher energy requirements
Membrane cell	Ion exchange membrane separates anode and cathode. Chlorine forms at anode, only sodium migrates through to cathode compartment - membrane is not permeable to liquid/gas flow and only permits migration of cations (+)	Most new plants opt for this design. Least environmental impact, lowest energy consumption

2.4 BRINE ELECTROLYSIS

Electrolysis is one of the acknowledged means of generating chemical products from their native state. For example, metallic copper is produced by electrolyzing an aqueous solution of copper sulfate, prepared by leaching the copper bearing ores with sulfuric acid. Or, one can prepare chlorine gas and sodium hydroxide solution by electrolyzing an aqueous solution of sodium chloride, which exists in nature in a solid form as rock salt and also available as solar or vacuum evaporated salt. The solution of sodium chloride (common table salt) is often called "brine."

The primary products of electrolysis are chlorine gas, hydrogen gas, and sodium hydroxide solution (commonly called "caustic soda" or simply "caustic"). However, if the electrolyte is maintained at a pH of 6.5 or 10, one can form chlorate or hypochlorite from the electrogenerated chlorine and caustic. This is the basis for the electrolytic production of sodium chlorate or sodium hypochlorite (commonly known as "bleach").

2.4.1 Chlorine and Sodium Hydroxide End Uses

Chlorine and sodium hydroxide are among the top ten chemicals produced in the world and are involved in the manufacturing of a wide variety of products used in day-to-day life. These include: pharmaceuticals, detergents, deodorants, disinfectants, herbicides, pesticides, and plastics.

The first observation of a possible application for chlorine was its bleaching effect on vegetable matter. In 1774, Carl Wilhelm investigated the reactivity of the greenish-yellow gas generated during the reaction involving the oxidation of hydrochloric acid by a manganese dioxide ore (pyrolusite). In 1785, Berthollet tried unsuccessfully to use elemental chlorine for textile bleaching to replace solar bleaching. Elemental chlorine caused discomfort to the workers, *corroded* metal parts, and softened the fabrics. The first use of chlorine in the form of potassium hypochlorite was for bleaching, and dates back to 1789. It was in 1808 that Davy characterized this greenish-yellow gas as an element and named it "chlorine."

The development of chemical bleaching with chlorine and the discovery of calcium hypochlorite bleaching powder as a practical mode of transporting chlorine was of great significance. These technologies made a marked impact on the textile bleaching operations in Great Britain and Europe, who were in the middle of the industrial revolution with expanding production, and hence, the demand for textiles. The invention of the power loom provided the capabilty to produce textiles on a large scale. However, solar bleaching, by spreading the cloth in open fields for months, became increasingly expensive in view of the soaring land values. The chlorine bleaching process not only shortened the operations from months to few days, but also freed vast areas of land for more productive use. Based on the greatly improved efficiency of textile bleaching, the pulp and paper industry also began using bleaching powder. Between 1756 and 1932, the use of chlorine in the pulp making industry increased. Chlorine, in the form of hypochlorites, removed the color or color producing materials from the cellulose fibers, without undue degradation of the fibers.

The first use of chlorine for disinfection dates back to 1823, when it was used in hospitals. Chlorine water was employed in obstetric wards to prevent puerperal fever in 1826, and fumigation with chlorine was practiced during the great European cholera epidemic. Following the discovery that bacteria were responsible for the transmission of certain diseases, several investigators studied chlorination of both sewage and potable water in 1890's in an attempt to destroy these bacteria. By 1912, the use of chlorine for water treatment had become a common practice. There was significant reduction in the incidence of water borne diseases, such as typhoid. For example, from October to December 1909, 549 cases of "winter typhoid" were reported in Montreal, Canada. After chlorination of drinking water was begun in 1910, only 170 cases were reported for the same 4-month period. Thus, virtually all the chlorine manufactured during the 19th century was consumed by these two industries. The major turning event for the growth of the chlorine industry was its use in 1912 for water purification during the Niagara Falls typhoid epidemic. It should be noted that bleaching powder was used in 1897 to clean the polluted mains during a typhoid break in England.

Between 1920 and 1940, several new applications for chlorine were developed, for example, in the manufacture of ethylene glycol, chlorinated solvents, vinyl chloride, and others. World War II triggered the development of new uses for chlorine for military needs. This trend continued to produce new products for civilian use following the war. Progress in synthetic organic chemistry in 19th century had led to the preparation of substitutes for natural products and entirely new and useful compounds including intermediates and final products. Chlorine, because of its reactivity, unique properties, and low price, was used in many of these, including solvents, pharmaceuticals and dyes. In 1795, dichloroethane was produced and in 1831 chloroform was synthesized. By 1848, the anesthetic properties of chloroform were recognized and used in surgical practice.

Presently, the primary uses of chlorine are in the pulp and paper manufacturing operations for bleaching to produce a high quality whitened material and in water treatment operations as a disinfectant (Figure 1). The other uses of chlorine include the production of organic and inorganic chemicals. The largest volume organic chemical manufactured that involves chlorine is polyvinyl chloride (PVC). PVC is a very versatile thermoplastic, used in a wide variety of daily products. The major use of chlorine in the production of inorganic chemicals is for titanium dioxide (a widely used pigment), manufactured from naturally occurring ores (ilmenite or rutile).

The end uses of caustic (sodium hydroxide) are diverse compared to the uses of chlorine. Its primary application are in the neutralization reactions and forming anionic species such as aluminates and zincates. In the manufacture of organic chemicals, caustic is employed for the neutralization of acids, pH control, off-gas scrubbing, dehydrochlorination, and as a source of sodium during various chemical reactions. For example, it is used in the dehydrochlorination stage of the epoxy resin production and hydrolysis reactions involving epichlorohydrin in the formation of glycerin, used in the pharmaceutical, tobacco and food/beverage industries

The major use of caustic for making inorganic chemicals is in the production of hypochlorite for household and industrial bleaching purposes. Also, its use in the pulp and paper industry is in the production of sodium sulfide and sodium hydrosulfide for mechanical pulping. It is also used in the food processing applications, which include skin removal of potatoes, tomatoes etc, for further processing.

2.4.2 Sodium Chlorate End Uses

The electrosynthesis of sodium chlorate dates back to 1802, when von Hisinger and Berzelius prepared sodium chlorate by the electrolysis of sodium chloride solution. The first chlorate cell patent was issued to Watt in 1851. The first chlorate plant was built in 1886 in Villers-St. Sepulchre in Switzerland, where chlorate was electrochemically produced in cells made of wood and equipped with a diaphragm. The energy consumption was about 15,000 kWh/ton potassium chlorate. This may be compared to an energy consumption for a crystal product of about 5,000- 6,000 kWh/ton with modern technologies.

About 93% of sodium chlorate is used for production for bleaching in the pulp and paper industry. The remainder is utilized in the agricultural industry as a cotton defoliant or herbicide (weed killer), as an oxidizer in uranium milling, and in the production of ammonium perchlorate used in rocket propulsion. (It is worth noting that perchlorates are also produced by an electrolytic process, where chlorate is anodically oxidized to perchlorate.) These uses of sodium chlorate have remained unchanged over the past 20 years, although the relative demands have changed. World capacity of sodium chlorate was estimated as about 2.8 million short tons during 1998; the North American share was about 1.95 million tons.

2.4.3 Chlor/alkali Manufacturing Process

The chlor-alkali (also called "chlorine-caustic") industry is one of the largest electrochemical technologies in the world. It is an energy intensive process and is the second largest consumer of electricity (2400 billion kWh) among electrolytic industries. In 1998, about 63% of the total world chlorine capacity of about 43.4 million metric tons was produced electrolytically using diaphragm and membrane cells, while about 35% was made using mercury cells .

Chlorine is produced by the electrolysis of sodium chloride (common table salt) solution, often called "brine." Thus, when sodium chloride is dissolved in water, it dissociates into sodium cations and chloride anions. The chloride ions are oxidized at the anode to form chlorine gas and water molecules are reduced at the cathode to form hydroxyl anions and hydrogen gas. The sodium ions in the solution and the hydroxyl ions produced at the cathode constitute the components of sodium hydroxide formed during the electrolysis of sodium chloride. The electrochemical reactions occurring in Chlor/alkali Manufacturing Process:

Electrochemical and chemical reactions occurring in mercury cells

[1] $2Cl^- \Longrightarrow Cl_2 + 2e^-$ (anodic reaction)

[2] $2Na^+ + 2Hg + 2e^- \Longrightarrow 2Na$ (in Hg) (cathodic reaction)

[3] $2Cl^- + 2Na^+ + 2Hg \Longrightarrow Cl_2 + 2Na$ (in Hg) (overall cell reaction)

[4] $2Na$ (in Hg) $+ 2H_2O \Longrightarrow H_2 + 2NaOH + Hg$ (decomposer reaction)

[5] $2NaCl + 2H_2O \Longrightarrow Cl_2 + 2NaOH + H_2$ (overall process reaction)

Electrochemical and chemical reactions occurring in diaphragm and membrane cells

[1] $2Cl^- \Longrightarrow Cl_2 + 2e^-$ (anodic reaction)

[6] $2H_2O + 2e^- \Longrightarrow 2OH^- + H_2$ (cathodic reaction)

[7] $2Cl^- + 2H_2O \Longrightarrow Cl_2 + H_2 + 2OH^-$ (overall ionic reaction)

[5] $2NaCl + 2H_2O \Longrightarrow Cl_2 + 2NaOH + H_2$ (overall reaction)

[8] $Cl_2 + 2NaOH \Longrightarrow NaOCl + NaCl + H_2O$ (side reaction)

[9] $3NaOCl \Longrightarrow NaClO_3 + 2NaCl$ (side reaction)

Reaction [9] will contaminate the caustic product with chlorate.

Chlorine is produced electrolytically using three types of electrolytic cells. The main difference in these technologies lies in the manner by which the chlorine gas and the sodium hydroxide are prevented from mixing with each other to ensure generation of pure products. Thus, in diaphragm cells, brine from the anode compartment flows through the separator to the cathode compartment, the separator material being either asbestos or polymer-modified asbestos composite deposited on a foraminous cathode. In membrane cells, on the other hand, an ion-exchange membrane is used as a separator. Anolyte-catholyte separation is achieved in the diaphragm and membrane cells using separators and ion-exchange membranes, respectively, whereas mercury cells contain no diaphragm or membrane and the mercury itself acts as a separator. The anode in all technologies is titanium metal coated with an electrocatalytic layer of mixed oxides. All modern cells (since the 1970's) use these so-called "dimensionally stable anodes" (DSA). Earlier cells used *carbon* based anodes. The cathode is typically steel in diaphragm cells, nickel in membrane cells, and mercury in mercury cells. These cell technologies are schematically depicted in Figures 5-7 and are described below.

2.4.3.1 Mercury cells

The mercury cell has steel bottoms with rubber-coated steel sides, as well as end boxes for brine and mercury feed and exit streams with a flexible rubber or rubber-coated steel cover. Adjustable metal anodes hang from the top, and mercury (which forms the cathode of the cell) flows on the inclined bottom. The current flows from the steel bottom to the flowing mercury.

Saturated brine fed from the end box is electrolyzed at the anode to produce the chlorine gas, which flows from the top portion of the trough and then exits. The sodium ion generated reacts with the mercury to form sodium amalgam (an alloy of mercury and sodium), which flows out of the end box to a vertical cylindrical tank. About 0.25% to 0.5% sodium amalgam is produced in the cell. The sodium amalgam reacts with water in the decomposer, packed with graphite particles and produces caustic soda and hydrogen. Hydrogen, saturated with water vapor, exits from the top along with the mercury vapors. The caustic soda then flows out of the decomposer as 50% caustic. The unreacted brine flows out of the exit end box. Some cells are designed with chlorine and anolyte outlets from the end box, which are separated in the depleted brine tank. The mercury from the decomposer is pumped back to the cell.

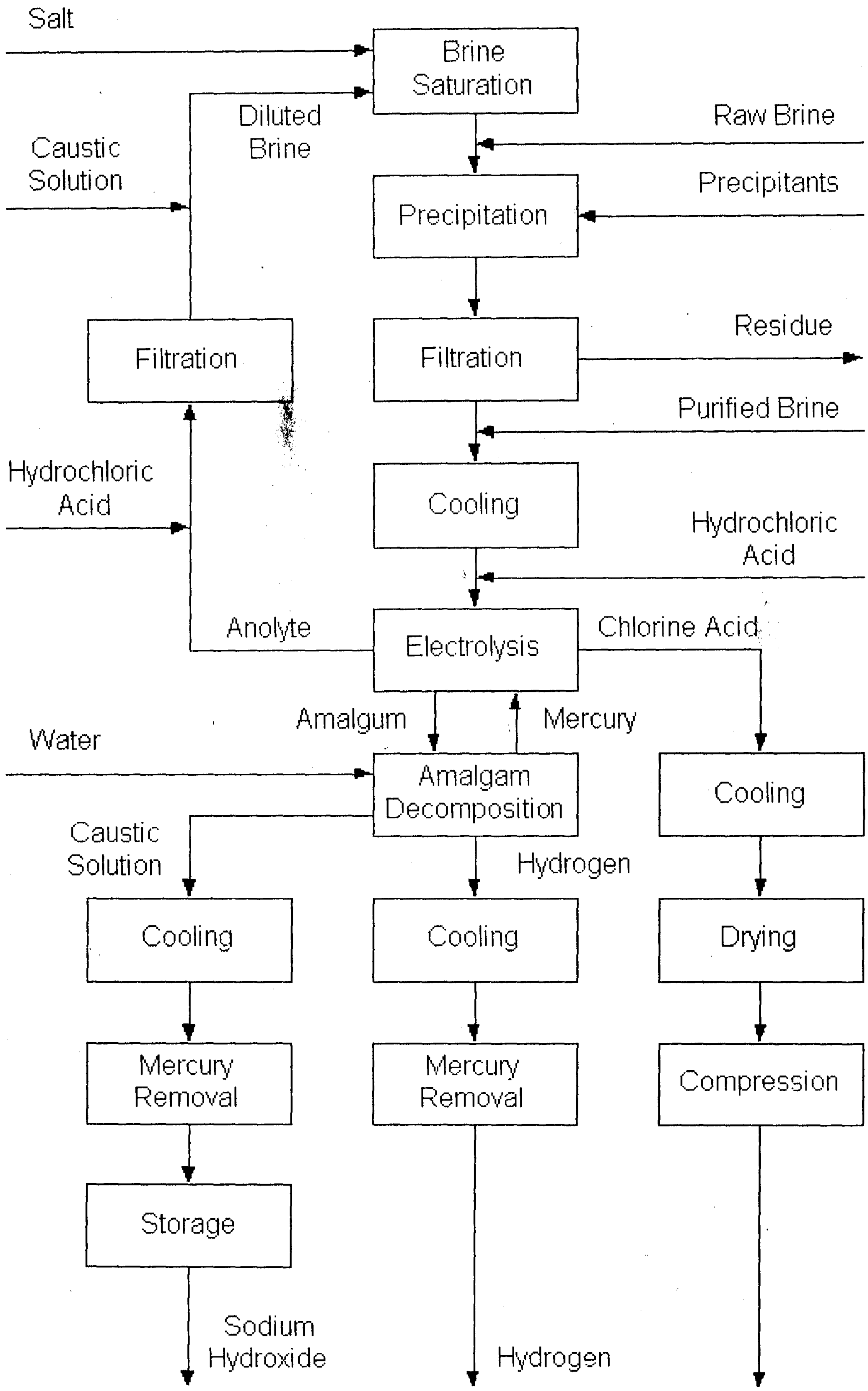


Fig 2.1: Mercury Cell Process Flow Sheet

2.4.3.1 Diaphragm cells

The diaphragm cell is a rectangular box with metal anodes supported from the bottom with copper-base plates, which carries a positive current. The cathodes are metal screens or punch plates connected from one end to the other end of the rectangular tank. Asbestos, dispersed as a slurry in a bath, is vacuum deposited onto the cathodes, forming a diaphragm. Saturated brine enters the anode compartment and the chlorine gas liberated at the anode during electrolysis, exits from the anode compartment. It is saturated with water vapor at a partial pressure of water over the anolyte. The sodium ions are transported from the anode compartment to the cathode compartment, by the flow of the solution and by electromigration, where they combine with the hydroxyl ions generated at the cathode during the formation of the hydrogen from the water molecules. The diaphragm resists the back migration of the hydroxyl ions, which would otherwise react with the chlorine in the anode compartment. In the cathode compartment, the concentration of the sodium hydroxide is ~12%, and the salt concentration is ~14%. There is also some sodium chlorate formed in the anode compartment, dependent upon the pH of the anolyte.

2.4.3.3 Membrane cells

In a membrane cell, an ion-exchange membrane separates the anode and cathode compartments. The separator is generally a bi-layer membrane made of perfluorocarboxylic and perfluorosulfonic acid-based films, sandwiched between the anode and the cathode. The saturated brine is fed to the anode compartment where chlorine is liberated at the anode, and the sodium ion migrates to the cathode compartment. Unlike in the diaphragm cells, only the sodium ions and some water migrate through the membrane. The unreacted sodium chloride and other inert ions remain in the anolyte. About 30-32% caustic soda is fed to the cathode compartment, where sodium ions react with hydroxyl ions produced during the course of the hydrogen gas evolution from the water molecules. This forms caustic, which increases the concentration of caustic solution to ~35%. The hydrogen gas, saturated with water, exits from the catholyte compartment. Only part of the caustic soda product is withdrawn from the cathode compartment. The remaining caustic is diluted to ~32% and returned to the cathode compartment.

Thus, all three basic cell technologies generate chlorine at the anode, and hydrogen along with sodium hydroxide (caustic soda) in the cathode compartment (or in a separate reactor for mercury cells). The distinguishing difference between the technologies lies in the manner by which the anolyte and the catholyte streams are prevented from mixing with each other. Separation is achieved in a diaphragm cell by a separator, and in a membrane cell by an ion-exchange membrane. In mercury cells, the cathode itself acts as a separator by forming an alloy of sodium and mercury (sodium amalgam) which is subsequently reacted with water to form sodium hydroxide and hydrogen in a separate reactor.

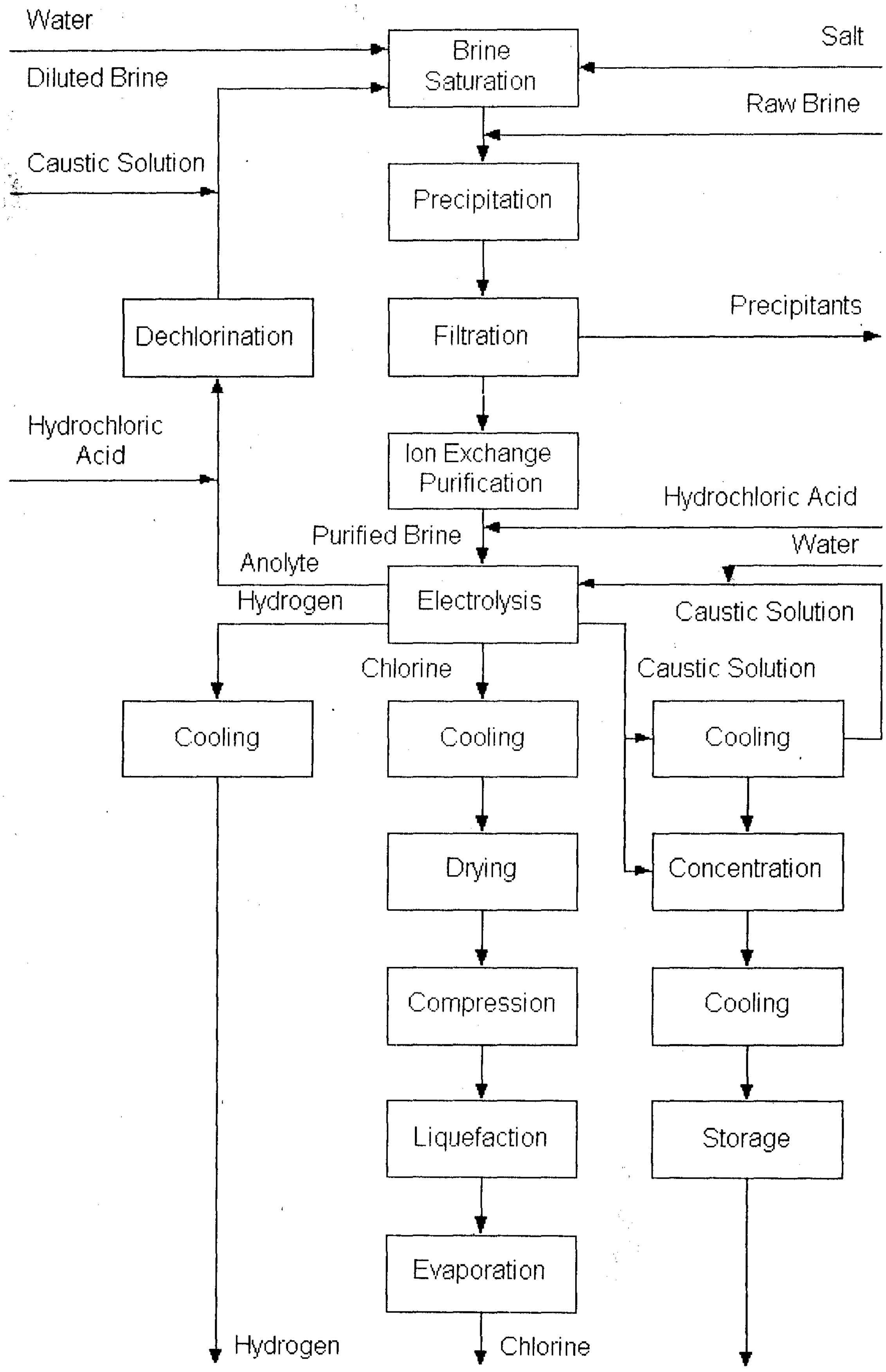


Fig 2.3: Membrane cells Process Flow Sheet

A comparison of the performance characteristics of these three technologies is presented below:

Table 2.2: Comparison of Cell Technologies

	Mercury	**Diaphragm**	**Membrane**
Operating current density (kA/m^2)	8 - 13	0.9 - 2.6	3 - 5
Cell voltage (V)	3.9 - 4.2	2.9 - 3.5	3.0 - 3.6
NaOH strength (wt%)	50	12	33-35
Energy consumption (kWh/MT Cl_2) at a current density of (kA/m^2)	3360 (10)	2720 (1.7)	2650 (5)
Steam consumption (kWh/MT Cl_2) for concentration to 50% NaOH	0	610	180

2.4.4 Chlorine processing

The chlorine gas from the anode compartment contains moisture, by-product oxygen, and some back-migrated hydrogen. In addition, if the brine is alkaline, it will contain carbon dioxide and some oxygen and nitrogen from the air leakage via the process or pipelines.

Chlorine is first cooled to 60°F (16°C) and passed through demisters to remove the water droplets and the particulates of salt and sodium sulfate. The cooled gas goes to sulfuric acid circulating towers, which are operated in series. Commonly, three towers are used for the removal of moisture. The dried chlorine then goes through demisters before it is compressed and liquefied at low temperatures. The non-condensed gas, called snift gas, is used for producing hypochlorite or hydrochloric acid. If there is no market for hydrochloric acid, the snift gas is neutralized with caustic soda or lime (calcium hydroxide) to form hypochlorite. The hypochlorite is either sold as bleach or decomposed to form salt and oxygen.

2.4.5 Hydrogen processing

The hydrogen gas from the chlor-alkali cells is normally used for the production of hydrochloric acid or used as a fuel to produce steam. Hydrogen from mercury cells is first cooled to remove the mercury, which is then returned to the cells. Occasionally, a secondary treatment is used to remove the trace levels of mercury in the hydrogen via molecular sieve columns. The hydrogen gas is then normally compressed. If a customer needs nearly pure hydrogen containing low amounts of oxygen, some plants will heat the hydrogen over a platinum catalyst (to remove the oxygen by reacting it with the hydrogen to form water), cool, and compress the diaphragm or membrane cell hydrogen, before supplying it to the customer. The heat value in the hydrogen cell gas can be recovered in a heat exchanger via heating the brine feed to the cells.

2.4.6 Caustic soda processing

Caustic soda is marketed as 50%, 73%, or anhydrous (dry) beads or flakes. The mercury cell can produce 50% and 73% caustic directly. The caustic from the decomposer is cooled and passed once or twice through an activated carbon filter to reduce the mercury levels in the caustic. After filtration, the mercury concentration is lowered to the parts-per-million (ppm) levels. Even these low levels of mercury may be unacceptable to some customers, who then have to switch to using membrane grade caustic soda. The mercury cell caustic soda has a few ppm salt and <5-ppm sodium chlorate. The mercury cell caustic is the highest purity caustic that can be made electrolytically if trace concentrations of mercury are tolerable in the end use of caustic.

The membrane cell caustic is concentrated in a multiple effect falling film evaporator, which increases the caustic soda concentration to 50% with a high steam economy. Caustic soda from membrane cells generally has 30-ppm sodium chloride and 5-10 ppm sodium chlorate.

The catholyte from the diaphragm cells contains ~12% sodium hydroxide, ~14% sodium chloride, 0.25%-0.3% sodium sulfate, and 100-500 ppm sodium chlorate. The catholyte is evaporated in a multi-effect evaporator. Most of the salt from the catholyte will precipitate during the concentration of the caustic soda to 50% sodium hydroxide. The 50% caustic soda product will contain about 1% sodium chloride. The 50% caustic also has a high chlorate concentration (~0.1%) compared to the caustic from membrane or mercury cells (~10 ppm). The salt, separated from the caustic during evaporation, is used to re-saturate the brine fed to the cell.

An additional single-effect evaporator is needed to produce 73% caustic soda. Anhydrous (dry) caustic soda is produced in a rising film evaporator, operating at 725°F (385°C) and at a few inches (one inch =2.54 cm) of water vacuum.

2.4.7 Brine processing

Sodium chloride is available in the form of solid salt, mined by excavation or by evaporating seawater. It is also available as a liquid by solution mining the salt domes. The salt has varying concentrations of impurities, which should be removed to operate the electrolytic cells at a high current efficiency. The major impurities are calcium, magnesium, and sulfates. The other minor impurities, which are undesirable, depending upon the type of chlor-alkali process selected, are barium, strontium, manganese, aluminum, silica, iron, vanadium, chromium, molybdenum, titanium, etc.

The solution-mined brine or the solid salt dissolved in the salt dissolver is treated in a reactor with sodium carbonate and caustic soda to precipitate calcium carbonate and magnesium hydroxide.

Chemical reactions occurring in brine processing

$CaSO_4 + Na_2CO_3 \Longrightarrow CaCO_3 + NaSO_4$ ($CaCO_3$ precipitates)

$MgCl_2 + 2NaOH \Longrightarrow Mg(OH)_2 + 2NaCl$ ($Mg(OH)_2$ precipitates)

These precipitates are settled in a settler. The underflow carries the solid slurry, which is pumped to a filter to remove it as sludge, or sometimes, it is disposed off along with the rest of the liquid effluent from the plant. The calcium carbonate precipitates are heavy, and drag with it the hydroxides of aluminum, magnesium, strontium, etc. The overflow from the settler, which carries ~10-50 ppm (parts per million) of suspended solids, is filtered. For the mercury and the diaphragm cell process, this brine is adequate, and can be fed to the electrolyzers.

In the all cell processes, the filtered brine is heated and passed through a bed of salt in a saturator in order to increase the salt concentration before feeding it to the electrolyzers. In some plants, the brine feed is acidified to improve the cell current efficiency. The acidification reduces the alkalinity, which would otherwise react with the chlorine in the anolyte compartment forming chlorate.

The membrane cell process brine specifications are more stringent than that of the mercury and diaphragm processes, and calls for impurities to be at the parts-per-billion (ppb) level. This is accomplished by filtering the brine in a pre-coat type secondary filter. An ion-exchange resin is used to remove the calcium, magnesium, barium, and iron impurities. It is also possible to remove aluminum by ion exchange. Often, aluminum and silica are removed by adding magnesium chloride in the brine exiting from the salt dissolver.

The depleted brine from the membrane and mercury cell processes carries dissolved chlorine. This brine is acidified to reduce the chlorine solubility, and then dechlorinated in a vacuum brine dechlorinator. The dechlorinated brine is returned to the brine wells for solution mining or to the salt dissolver. If the membrane and diaphragm processes coexist at a given location, the dechlorinated brine can be sent to a saturator for resaturation before being sent to the diaphragm cells.

The primary technology that is presently being used for future expansions or replacements of existing circuits is the membrane cell technology. There are a number of membrane cell technology suppliers, including: Uhde-deNora, Asahi Chemicals, Asahi Glass, Chlorine Engineers, ICI, and Eltech Systems. Eltech also can supply diaphragm cell technology. It is highly unlikely that anyone will build a new mercury- or diaphragm-cell plant in the future.

2.4.8 Sodium chlorate manufacturing process

One of the energy intensive electrolytic industries is the production of sodium chlorate by the electrolysis of sodium chloride solutions in an electrolytic cell without a separator. The products of the electrode reactions, the chlorine and the caustic, are allowed to intermix and react, producing sodium chlorate as the final product

The major raw material is sodium chloride, either very pure, such as solar rock salt, or partially purified evaporated salt. The salt is stored and dissolved in lixiviators to produce a saturated sodium chloride solution. This solution is purified by removing calcium, magnesium, fluoride, sulfate, and iron as insoluble compounds, through the addition of sodium carbonate, sodium hydroxide, sodium phosphate, and barium chloride.

The impurities or precipitates are removed in a pressure leaf filter with diatomaceous earth as a filter precoat and filter aid. This filter cake, containing approximately 35% water, is the only solid waste stream from the process. A polishing filtration stage and an ion-exchange system follow pressure leaf filtration.

The chemistry and electrochemistry of chlorate formation dictates that an efficient and economical cell should embody several distinct zones. In the electrolysis zone, the electrolytic reactions take place along with the hydrolysis of chlorine. As the chemical chlorate formation proceeds very slowly, a relatively large volume of chemical reaction zone is needed. A cooling zone is also required to remove the excess heat generated from the reaction and control the operating temperature. The cooling zone may be located within the chemical reactor or in an external heat exchanger. Hydrogen gas generated at the cathode must be released from the cell liquor. This hydrogen release takes place in the electrolysis cell, a separate vessel, or the chemical reactor.

A continuous stream of cell liquor flows from the electrolysis system to the "hypo removal" system, where the sodium hypochlorite concentration is reduced to low levels simply by heating the cell liquor to about 185-200°F (85-95°C) under careful pH control. Final traces of hypochlorite can be completely removed by treatment with a reducing agent (such as sodium sulfite or hydrogen peroxide).

Sodium chlorate is usually recovered from cell liquors by concentration, followed by cooling to facilitate crystallization. Hot cell liquor, following hypo removal, is fed continuously into the circulation leg of a draft tube baffle evaporator/crystallizer. Crystal slurry is withdrawn from the bottom of the crystallizer section. The crystals are separated from the mother liquor and washed with water in a pusher centrifuge. They are thoroughly washed to remove sodium dichromate (an additive to the cell solution to increase current efficiency) from the chlorate crystals. Sodium dichromate contains chromium in the hexavalent state, which is a recognized human carcinogen. A white sodium chlorate crystal, containing about 1 to 1.5% moisture, is obtained from the centrifuge. Mother liquor from the centrifuge is mixed with fresh purified brine and recycled to the electrolytic cells.

Approximately 98% of the sodium chlorate capacity in North America is produced directly in sodium chlorate cells. The remaining 2% is produced "chemically" by the reaction of chlorine and caustic.

CHAPTER THREE

3.0 MATERIAL BALANCE, ENERGY BALANCE, EQUIPMENT SPECIFICATION and ECONOMIC EVALUATION

3.1 MATERIAL BALANCE

The major tool in the design or analysis of any process is the application of material balance. A material balance is the flow of material from one unit to another. The output of material from one unit become the input to the next unit. Material balance are useful tools in determining the quantity of raw materials required and products produced. Material are also used to study plant operation, check performance against design, locate the sources of material loss and so on.

The simple balances also aquaints the process designer the maximum yields he can obtain in any process. It also provides a quick check on the profitability of a proposed process, this is often accomplished by assuring complete reaction and a hundred percent recovery and using raw materials costs. If the raw material costs are equal to or greater than the product selling price, the process will not break – even. For the process analysts, the material balance is the first tool applied in diagnosing engineering process.

3.1.1 Principle of Material Balance

The law of conservation of matter is the principle employed in the material balance, and it applies to most of chemical process engineering calculations. The general conservation equation for any process system can be written as:

Input – Output – Depletion = Accumulation – Generation .... (3.1)

For a process with no generation or depletion of material within the system, the term in the equation becomes:

Input – Output = Accumulation .... (3.2)

In a steady state process, the accumulation term is zero, except in nuclear process. Mass is neither generated not consumed, but if a chemical reaction occurs, particular chemical specie may be formed or consumed in the process. If there is no chemical reaction, the steady – state balance reduced to:

Input – Output = 0 .... (3.3)

3.1.2 System

The mass balance is often expressed in terms of the reactions, products and a balance of some sort with respect to an entity (system) which is clearly identified by creating boundary around it.

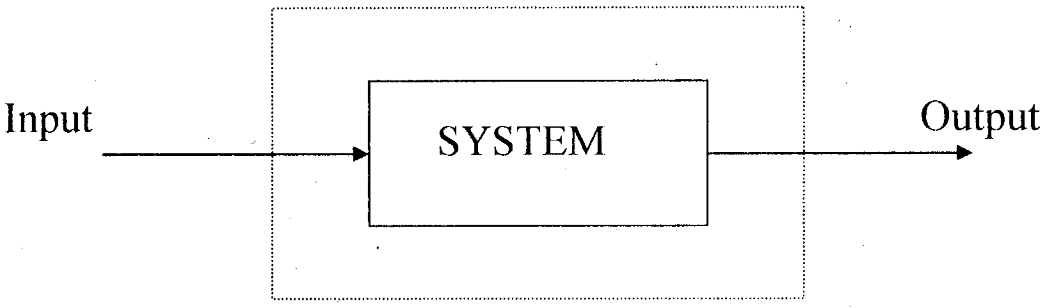


Figure 3.1: A Typical System

3.1.3 Basis

This is the time over which the system is observed. This might be one second, one hour, one day e.t.c depending on the process designer. In some cases where time is not convenient , material input or output is chosen as the basis. In any case, basis must be stated in any mass balance problem.

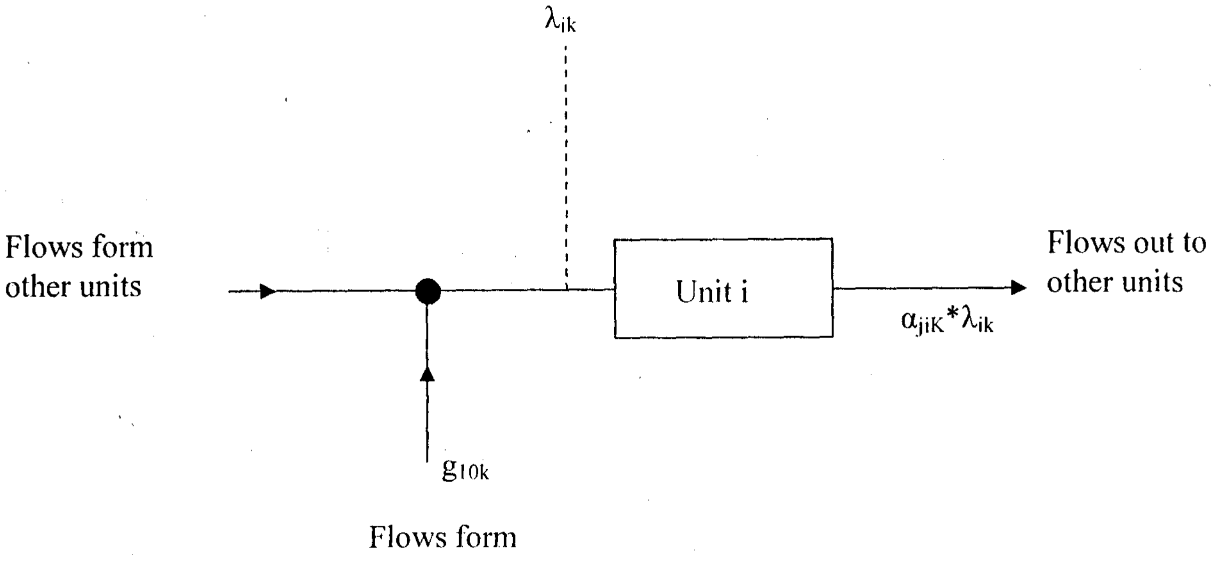
3.1.4 Procedure

The procedure used for the calculation involves:

(1) Conversion of tones per day to kilogram per day

(2) The material balance is based on split fraction method. It follows the following major steps:

If a block shown below represent any units in an information flow diagram and show the nomenclature that will be used in setting up the material balance equation.



i = the unit number

λ_{ik} = The total flow into the unit of the component k

α_{jiK} = The fraction of the total flow of component k entering unit i that leaves in the outlet stream connected to the unit j, the "split – fraction coefficient"

g_{I0k} = Any fresh feed of component k into unit i, flow from outside the system

The values of the split – fraction coefficient will depend on the nature of the unit and the inlet stream composition..

(3). Determining the mass of NaCl converted to NaOH and hence calculating the number of moles of NaCl actually reacted.

(4). Considering the stoichiometric equation, from which moles of $CaCO_3$ (precipitates) and $Mg(OH)_2$ (Precipitated) are formed during brine processing.

(5). Picking a system and a basis of one day, and performing material balance on each system, noting that, calculated flows out of one unit becomes the feed to the next unit.

3.1.5 Assumption of Material Balance

(i) The reaction of H_2O and NaCl taken place in the saturator to form aqueous NaCl and reacted in the proportion of 2.3 moles to 1 mol respectively.

(ii). 5 moles of $CaCO_3$ and $Mg(OH)_2$ enters precipitator in proportions 3 to 2 moles respectively.

(iii) Na_2SO_4 and NaCl produced during brine processing are of negligible quantities.

(iv). The depleted brine (chlorate) in electrolytic cell is being dechlorinated and recycled back into the saturator

(v). The electrolytic mercury cell is assumed a reactor, and the material balance is based on the stoichiometric equation of the reaction occurred.

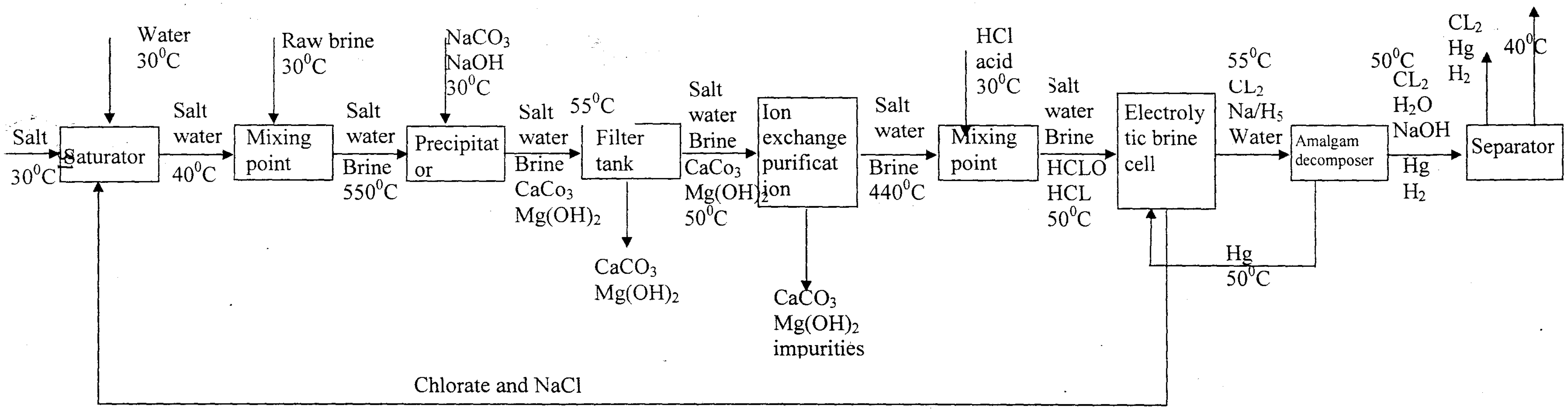
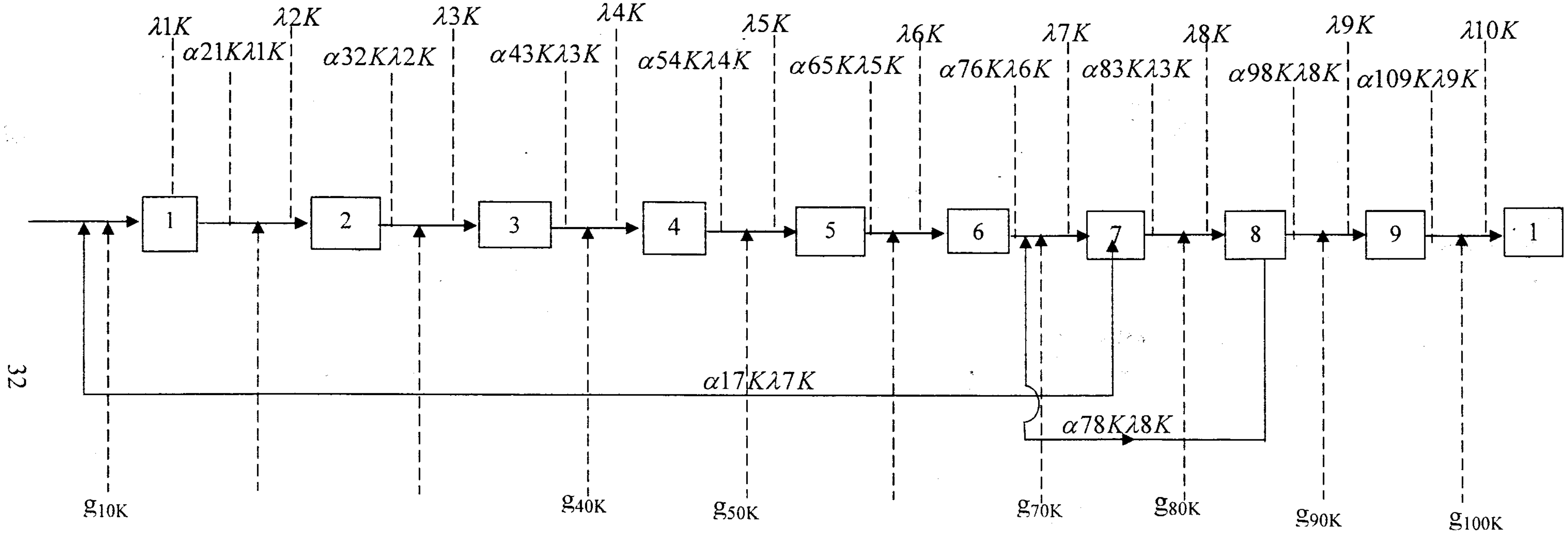


Fig 3.1: Information Flow Diagram of the Mercury Cell Process Flowsheet



SPLIT FRACTIONS AND FRESH FEED

LIST OF COMPONENTS

1. Salt
2. Water
3. Brine
4. Na_2CO_3
5. $CaCO_3$
6. $Mg(OH)_2$
7. Chlorate
8. Hydrogen
9. NaOH
10. Chlorine
11. HCl
12. Mercury

MATERIAL BALANCE USING SPLIT FRACTION WITH THE AID OF NEGIEV'S METHOD

Split Fraction Coefficients and Fresh Feeds (Decision Table)

Alpha/ K	1	2	3	4	5	6	7	8	9	10	11	12
21K	-1	-1	-1	0	0	0	0	0	0	0	0	0
32K	-1	-1	-1	0	0	0	0	0	0	0	0	0
43K	-1	-1	-1	0	0	0	0	0	0	0	0	0
54K	-1	-1	-1	0	-0.2	-0.2	0	0	0	0	0	0
65K	-1	-1	-1	0	0	0	0	0	0	0	0	0
76K	-0.8	-0.7	-0.95	0	0	0	-1	0	0	0	0	0
87K	0	-1	0	0	0	0	0	-1	0	-1	0	0
17K	-0.01	0	-0.02	0	0	0	-1	0	0	0	0	0
98K	0	-1	0	0	0	0	0	-1	-1	-1	0	-0.5
78K	0	0	0	0	0	0	0	0	0	0	0	-0.5
109K	0	-1	0	0	0	0	0	0	-1	0	0	0

Fresh feed	g101	g102	g203	g304	g309	g405	g406	g6011
kmole	603.762813	1388.65447	603.762813	1811.28844	1207.52563	1811.28844	1207.52563	241.5051252

Product	g707	g7010	g7012	g708	g809
kmole	301.881406	301.881406	603.762813	301.881406	603.762813

Matrix Equation

	1	2	3	4	5	6	7	8	9	10	λ	g
1	1	0	0	0	0	0	0	0	0	0	λ_{1k}	0
2	0	1	0	0	0	0	0	0	0	0	λ_{2k}	0
3	0	0	1	0	0	0	0	0	0	0	λ_{3k}	0
4	0	0	0	1	0	0	0	0	0	0	λ_{4k}	0
5	0	0	0	0	1	0	0	0	0	0	λ_{5k}	0
6	0	0	0	0	0	1	0	0	0	0	λ_{6k}	0
7	0	0	0	0	0	0	1	0	0	0	λ_{7k}	0
8	0	0	0	0	0	0	0	1	0	0	λ_{8k}	0
9	0	0	0	0	0	0	0	0	1	0	λ_{9k}	0
10	0	0	0	0	0	0	0	0	0	1	λ_{10k}	0

K=1(Salt)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	-0.01	0	0	0	603.76281	λ_{11}	608.6318679
2	-1	1	0	0	0	0	0	0	0	0	0	λ_{21}	608.6318679
3	0	-1	1	0	0	0	0	0	0	0	0	λ_{31}	608.6318679
4	0	0	-1	1	0	0	0	0	0	0	0	λ_{41}	608.6318679
5	0	0	0	-1	1	0	0	0	0	0	0	λ_{51}	608.6318679
6	0	0	0	0	-1	1	0	0	0	0	0	λ_{61}	608.6318679
7	0	0	0	0	0	-0.8	1	0	0	0	0	λ_{71}	486.9054943
8	0	0	0	0	0	0	0	1	0	0	0	λ_{81}	0
9	0	0	0	0	0	0	0	0	1	0	0	λ_{91}	0
10	0	0	0	0	0	0	0	0	0	1	0	λ_{101}	0

Matrix Inverse

1.00806452	0.00806452	0.008064516	0.00806452	0.00806452	0.00806452	0.01008065	0	0	0
1.00806452	1.00806452	0.008064516	0.00806452	0.00806452	0.00806452	0.01008065	0	0	0
1.00806452	1.00806452	1.008064516	0.00806452	0.00806452	0.00806452	0.01008065	0	0	0
1.00806452	1.00806452	1.008064516	1.00806452	0.00806452	0.00806452	0.01008065	0	0	0
1.00806452	1.00806452	1.008064516	1.00806452	1.00806452	0.00806452	0.01008065	0	0	0
1.00806452	1.00806452	1.008064516	1.00806452	1.00806452	1.00806452	0.01008065	0	0	0
0.80645161	0.80645161	0.806451613	0.80645161	0.80645161	0.80645161	1.00806452	0	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	0	1	0
0	0	0	0	0	0	0	0	0	1

K=2(Water)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	1388.6545	λ_{12}	1388.65447
2	-1	1	0	0	0	0	0	0	0	0	0	λ_{22}	1388.65447
3	0	-1	1	0	0	0	0	0	0	0	0	λ_{32}	1388.65447
4	0	0	-1	1	0	0	0	0	0	0	0	λ_{42}	1388.65447
5	0	0	0	-1	1	0	0	0	0	0	0	λ_{52}	1388.65447
6	0	0	0	0	-1	1	0	0	0	0	0	λ_{62}	1388.65447
7	0	0	0	0	0	-0.7	1	0	0	0	0	λ_{72}	972.0581289
8	0	0	0	0	0	0	-1	1	0	0	0	λ_{82}	972.0581289
9	0	0	0	0	0	0	0	-1	1	0	0	λ_{92}	972.0581289
10	0	0	0	0	0	0	0	0	-1	1	0	λ_{102}	972.0581289

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
1	1	0	0	0	0	0	0	0	0
1	1	1	0	0	0	0	0	0	0
1	1	1	1	0	0	0	0	0	0
1	1	1	1	1	0	0	0	0	0
1	1	1	1	1	1	0	0	0	0
0.7	0.7	0.7	0.7	0.7	0.7	1	0	0	0
0.7	0.7	0.7	0.7	0.7	0.7	1	1	0	0
0.7	0.7	0.7	0.7	0.7	0.7	1	1	1	0
0.7	0.7	0.7	0.7	0.7	0.7	1	1	1	1

K=3(Brine)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	-0.02	0	0	0	0	λ_{13}	11.69367324
2	-1	1	0	0	0	0	0	0	0	0	603.76281	λ_{23}	615.4564862
3	0	-1	1	0	0	0	0	0	0	0	0	λ_{33}	615.4564862
4	0	0	-1	1	0	0	0	0	0	0	0	λ_{43}	615.4564862
5	0	0	0	-1	1	0	0	0	0	0	0	λ_{53}	615.4564862
6	0	0	0	0	-1	1	0	0	0	0	0	λ_{63}	615.4564862
7	0	0	0	0	0	-0.95	1	0	0	0	0	λ_{73}	584.6836619
8	0	0	0	0	0	0	0	1	0	0	0	λ_{83}	0
9	0	0	0	0	0	0	0	0	1	0	0	λ_{93}	0
10	0	0	0	0	0	0	0	0	0	1	0	λ_{103}	0

Matrix Inverse

1.01936799	0.01936799	0.019367992	0.01936799	0.01936799	0.01936799	0.02038736	0	0	0
1.01936799	1.01936799	0.019367992	0.01936799	0.01936799	0.01936799	0.02038736	0	0	0
1.01936799	1.01936799	1.019367992	0.01936799	0.01936799	0.01936799	0.02038736	0	0	0
1.01936799	1.01936799	1.019367992	1.01936799	0.01936799	0.01936799	0.02038736	0	0	0
1.01936799	1.01936799	1.019367992	1.01936799	1.01936799	0.01936799	0.02038736	0	0	0
1.01936799	1.01936799	1.019367992	1.01936799	1.01936799	1.01936799	0.02038736	0	0	0
0.96839959	0.96839959	0.968399592	0.96839959	0.96839959	0.96839959	1.01936799	0	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	0	1	0
0	0	0	0	0	0	0	0	0	1

K=4(Na_2CO_3)													
	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{14}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{24}	0
3	0	0	1	0	0	0	0	0	0	0	1811.2884	λ_{34}	1811.288439
4	0	0	0	1	0	0	0	0	0	0	0	λ_{44}	0
5	0	0	0	0	1	0	0	0	0	0	0	λ_{54}	0
6	0	0	0	0	0	1	0	0	0	0	0	λ_{64}	0
7	0	0	0	0	0	0	1	0	0	0	0	λ_{74}	0
8	0	0	0	0	0	0	0	1	0	0	0	λ_{84}	0
9	0	0	0	0	0	0	0	0	1	0	0	λ_{94}	0
10	0	0	0	0	0	0	0	0	0	1	0	λ_{104}	0
								Matrix Inverse					
	1	0	0	0	0	0	0	0	0	0			
	0	1	0	0	0	0	0	0	0	0			
	0	0	1	0	0	0	0	0	0	0			
	0	0	0	1	0	0	0	0	0	0			
	0	0	0	0	1	0	0	0	0	0			
	0	0	0	0	0	1	0	0	0	0			
	0	0	0	0	0	0	1	0	0	0			
	0	0	0	0	0	0	0	1	0	0			
	0	0	0	0	0	0	0	0	1	0			
	0	0	0	0	0	0	0	0	0	1			

K=5($CaCO_3$)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{15}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{25}	0
3	0	0	1	0	0	0	0	0	0	0	0	λ_{35}	0
4	0	0	0	1	0	0	0	0	0	0	1811.2884	λ_{45}	1811.288439
5	0	0	0	-0.2	1	0	0	0	0	0	0	λ_{55}	362.2576878
6	0	0	0	0	0	1	0	0	0	0	0	λ_{65}	0
7	0	0	0	0	0	0	1	0	0	0	0	λ_{75}	0
8	0	0	0	0	0	0	0	1	0	0	0	λ_{85}	0
9	0	0	0	0	0	0	0	0	1	0	0	λ_{95}	0
10	0	0	0	0	0	0	0	0	0	1	0	λ_{105}	0

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0.2	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	0	1	0	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	0	1	0
0	0	0	0	0	0	0	0	0	1

$K=6(Mg(OH)_2)$

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{16}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{26}	0
3	0	0	1	0	0	0	0	0	0	0	0	λ_{36}	0
4	0	0	0	1	0	0	0	0	0	0	1207.5256	λ_{46}	1207.525626
5	0	0	0	-0.2	1	0	0	0	0	0	0	λ_{56}	241.5051252
6	0	0	0	0	0	1	0	0	0	0	0	λ_{66}	0
7	0	0	0	0	0	0	1	0	0	0	0	λ_{76}	0
8	0	0	0	0	0	0	0	1	0	0	0	λ_{86}	0
9	0	0	0	0	0	0	0	0	1	0	0	λ_{96}	0
10	0	0	0	0	0	0	0	0	0	1	0	λ_{106}	0

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0.2	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	0	1	0	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	0	1	0
0	0	0	0	0	0	0	0	0	1

K=7(Chlorate)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	-1	0	0	0	0	λ_{17}	301.8814065
2	0	1	0	0	0	0	0	0	0	0	0	λ_{27}	0
3	0	0	1	0	0	0	0	0	0	0	0	λ_{37}	0
4	0	0	0	1	0	0	0	0	0	0	0	λ_{47}	0
5	0	0	0	0	1	0	0	0	0	0	0	λ_{57}	0
6	0	0	0	0	0	1	0	0	0	0	0	λ_{67}	0
7	0	0	0	0	0	-1	1	0	0	0	301.88141	λ_{77}	301.8814065
8	0	0	0	0	0	0	0	1	0	0	0	λ_{87}	0
9	0	0	0	0	0	0	0	0	1	0	0	λ_{97}	0
10	0	0	0	0	0	0	0	0	0	1	0	λ_{107}	0

Matrix Inverse

1	0	0	0	0	1	1	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	1	1	0	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	0	1	0
0	0	0	0	0	0	0	0	0	1

K=8(Hydrogen)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{18}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{28}	0
3	0	0	1	0	0	0	0	0	0	0	0	λ_{38}	0
4	0	0	0	1	0	0	0	0	0	0	0	λ_{48}	0
5	0	0	0	0	1	0	0	0	0	0	0	λ_{58}	0
6	0	0	0	0	0	1	0	0	0	0	0	λ_{68}	0
7	0	0	0	0	0	0	1	0	0	0	0	λ_{78}	0
8	0	0	0	0	0	0	-1	1	0	0	301.88141	λ_{88}	301.8814065
9	0	0	0	0	0	0	0	-1	1	0	0	λ_{98}	301.8814065
10	0	0	0	0	0	0	0	0	0	1	0	λ_{108}	0

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	0	1	0	0	0
0	0	0	0	0	0	1	1	0	0
0	0	0	0	0	0	1	1	1	0
0	0	0	0	0	0	0	0	0	1

K=9(NaOH)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{19}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{29}	0
3	0	0	1	0	0	0	0	0	0	0	1207.5256	λ_{39}	1207.525626
4	0	0	0	1	0	0	0	0	0	0	0	λ_{49}	0
5	0	0	0	0	1	0	0	0	0	0	0	λ_{59}	0
6	0	0	0	0	0	1	0	0	0	0	0	λ_{69}	0
7	0	0	0	0	0	0	1	0	0	0	0	λ_{79}	0
8	0	0	0	0	0	0	0	1	0	0	0	λ_{89}	0
9	0	0	0	0	0	0	0	-1	1	0	603.76281	λ_{99}	603.762813
10	0	0	0	0	0	0	0	0	-1	1	0	λ_{109}	603.762813

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	0	1	0	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	1	1	0
0	0	0	0	0	0	0	1	1	1

K=10(Chlorine)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{110}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{210}	0
3	0	0	1	0	0	0	0	0	0	0	0	λ_{310}	0
4	0	0	0	1	0	0	0	0	0	0	0	λ_{410}	0
5	0	0	0	0	1	0	0	0	0	0	0	λ_{510}	0
6	0	0	0	0	0	1	0	0	0	0	0	λ_{610}	0
7	0	0	0	0	0	0	1	0	0	0	0	λ_{710}	0
8	0	0	0	0	0	0	-1	1	0	0	301.88141	λ_{810}	301.8814065
9	0	0	0	0	0	0	0	-1	1	0	0	λ_{910}	301.8814065
10	0	0	0	0	0	0	0	0	0	1	0	λ_{1010}	0

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	0	1	0	0	0
0	0	0	0	0	0	1	1	0	0
0	0	0	0	0	0	1	1	1	0
0	0	0	0	0	0	0	0	0	1

K=11(HCl)

	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{111}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{211}	0
3	0	0	1	0	0	0	0	0	0	0	0	λ_{311}	0
4	0	0	0	1	0	0	0	0	0	0	0	λ_{411}	0
5	0	0	0	0	1	0	0	0	0	0	0	λ_{511}	0
6	0	0	0	0	0	1	0	0	0	0	241.50513	λ_{611}	241.5051252
7	0	0	0	0	0	0	1	0	0	0	0	λ_{711}	0
8	0	0	0	0	0	0	0	1	0	0	0	λ_{811}	0
9	0	0	0	0	0	0	0	0	1	0	0	λ_{911}	0
10	0	0	0	0	0	0	0	0	0	1	0	λ_{1011}	0

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	0	1	0	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	0	1	0
0	0	0	0	0	0	0	0	0	1

K=12(Mercury)	1	2	3	4	5	6	7	8	9	10	g	λ	Stream Flows
1	1	0	0	0	0	0	0	0	0	0	0	λ_{112}	0
2	0	1	0	0	0	0	0	0	0	0	0	λ_{212}	0
3	0	0	1	0	0	0	0	0	0	0	0	λ_{312}	0
4	0	0	0	1	0	0	0	0	0	0	0	λ_{412}	0
5	0	0	0	0	1	0	0	0	0	0	0	λ_{512}	0
6	0	0	0	0	0	1	0	0	0	0	0	λ_{612}	0
7	0	0	0	0	0	0	1	-0.5	0	0	0	λ_{712}	301.8814065
8	0	0	0	0	0	0	0	1	0	0	603.76281	λ_{812}	603.762813
9	0	0	0	0	0	0	0	-0.5	1	0	0	λ_{912}	301.8814065
10	0	0	0	0	0	0	0	0	0	1	0	λ_{1012}	0

Matrix Inverse

1	0	0	0	0	0	0	0	0	0
0	1	0	0	0	0	0	0	0	0
0	0	1	0	0	0	0	0	0	0
0	0	0	1	0	0	0	0	0	0
0	0	0	0	1	0	0	0	0	0
0	0	0	0	0	1	0	0	0	0
0	0	0	0	0	0	1	0.5	0	0
0	0	0	0	0	0	0	1	0	0
0	0	0	0	0	0	0	0.5	1	0
0	0	0	0	0	0	0	0	0	1

MOLAR FLOWS AND COMPOSITION(kmole/hr)

UNITS	COMPONENTS 1	2	3	4	5	6	7	8	9	10	11	12	Total
1	608.6318679	1388.65447	11.6936732	0	0	0	301.881406	0	0	0	0	0	2310.861417
2	608.6318679	1388.65447	615.456486	0	0	0	0	0	0	0	0	0	2612.742824
3	608.6318679	1388.65447	615.456486	1811.28844	0	0	0	0	1207.5256	0	0	0	5631.556889
4	608.6318679	1388.65447	615.456486	0	1811.28844	1207.525626	0	0	0	0	0	0	5631.556889
5	608.6318679	1388.65447	615.456486	0	362.257688	241.5051252	0	0	0	0	0	0	3216.505637
6	608.6318679	1388.65447	615.456486	0	0	0	0	0	0	0	241.50513	0	2854.247949
7	486.9054943	972.058129	584.683662	0	0	0	301.881406	0	0	0	0	301.88141	2647.410098
8	0	972.058129	0	0	0	0	0	301.881406	0	301.88141	0	603.76281	2179.583755
9	0	972.058129	0	0	0	0	0	301.881406	603.76281	301.88141	0	301.88141	2481.465161
10	0	972.058129	0	0	0	0	0	0	603.76281	0	0	0	1575.820942

MASS FLOWS AND COMPOSITION(kg/nr)

UNITS	COMPONENTS 1	2	3	4	5	6	7	8	9	10	11	12	Total
1	35568.44636	25023.5535	683.378264	0	0	0	32132.5588	0	0	0	0	0	93407.93696
2	35568.44636	25023.5535	35967.2771	0	0	0	0	0	0	0	0	0	96559.27696
3	35568.44636	25023.5535	35967.2771	191976.65	0	0	0	0	48301.025	0	0	0	336836.9523
4	35568.44636	25023.5535	35967.2771	0	181286.426	70422.41149	0	0	0	0	0	0	348268.1144
5	35568.44636	25023.5535	35967.2771	0	36257.2852	14084.4823	0	0	0	0	0	0	146901.0445
6	35568.44636	25023.5535	35967.2771	0	0	0	0	0	0	0	8814.9371	0	105374.214
7	28454.75709	17516.4875	34168.9132	0	0	0	32132.5588	0	0	0	0	60554.391	172827.1079
8	0	17516.4875	0	0	0	0	0	608.592915	0	21405.203	0	121108.78	160639.0661
9	0	17516.4875	0	0	0	0	0	608.592915	24150.513	21405.203	0	60554.391	124235.1872
10	0	17516.4875	0	0	0	0	0	0	24150.513	0	0	0	41667

MOLE FRACTION

COMPONENTS	1	2	3	4	5	6	7	8	9	10	11	12	Total
UNITS													
1	0.263378783	0.60092503	0.00506031	0	0	0	0.13063588	0	0	0	0	0	1
2	0.232947484	0.53149298	0.23555954	0	0	0	0	0	0	0	0	0	1
3	0.108075241	0.24658447	0.10928709	0.32163192	0	0	0	0	0.2144213	0	0	0	1
4	0.108075241	0.24658447	0.10928709	0	0.32163192	0.214421278	0	0	0	0	0	0	1
5	0.189221452	0.43172767	0.1913432	0	0.11262461	0.075083072	0	0	0	0	0	0	1
6	0.21323721	0.48652202	0.21562825	0	0	0	0	0	0	0	0.0846125	0	1
7	0.183917669	0.36717323	0.22085119	0	0	0	0.11402895	0	0	0	0	0.114029	1
8	0	0.44598338	0	0	0	0	0	0.13850416	0	0.1385042	0	0.2770083	1
9	0	0.39172749	0	0	0	0	0	0.1216545	0.243309	0.1216545	0	0.1216545	1
10	0	0.61685824	0	0	0	0	0	0	0.3831418	0	0	0	1

MASS FRACTION

COMPONENTS	1	2	3	4	5	6	7	8	9	10	11	12	Total
UNITS													
1	0.380786125	0.26789537	0.00731606	0	0	0	0.34400245	0	0	0	0	0	1
2	0.368358665	0.25915225	0.37248909	0	0	0	0	0	0	0	0	0	1
3	0.105595441	0.07428981	0.10677949	0.5699394	0	0	0	0	0.1433959	0	0	0	1
4	0.102129494	0.07185141	0.10327468	0	0.52053696	0.202207462	0	0	0	0	0	0	1
5	0.242125211	0.17034292	0.24484017	0	0.24681435	0.095877346	0	0	0	0	0	0	1
6	0.33754412	0.23747322	0.34132902	0	0	0	0	0	0	0	0.0836536	0	1
7	0.164642905	0.10135266	0.19770575	0	0	0	0.18592314	0	0	0	0	0.3503755	1
8	0	0.10904251	0	0	0	0	0	0.00378857	0	0.1332503	0	0.7539186	1
9	0	0.14099458	0	0	0	0	0	0.00489872	0.1943935	0.1722958	0	0.4874174	1
10	0	0.42039234	0	0	0	0	0	0	0.5796077	0	0	0	1

PERCENTAGE MOLE(%)

COMPONENTS UNITS	1	2	3	4	5	6	7	8	9	10	11	12	Total
1	26.3378783	60.0925031	0.50603092	0	0	0	13.0635876	0	0	0	0	0	100
2	23.29474843	53.149298	23.5559536	0	0	0	0	0	0	0	0	0	100
3	10.8075241	24.658447	10.9287094	32.1631917	0	0	0	0	21.442128	0	0	0	100
4	10.8075241	24.658447	10.9287094	0	32.1631917	21.44212781	0	0	0	0	0	0	100
5	18.92214523	43.1727666	19.1343201	0	11.2624608	7.508307227	0	0	0	0	0	0	100
6	21.32372095	48.6522017	21.5628249	0	0	0	0	0	0	0	8.4612525	0	100
7	18.39176691	36.7173235	22.0851187	0	0	0	11.4028955	0	0	0	0	11.402895	100
8	0	44.598338	0	0	0	0	0	13.8504155	0	13.850416	0	27.700831	100
9	0	39.1727494	0	0	0	0	0	12.1654501	24.3309	12.16545	0	12.16545	100
10	0	61.6858238	0	0	0	0	0	0	38.314176	0	0	0	100

PERCENTAGE MASS(%)

COMPONENTS UNITS	1	2	3	4	5	6	7	8	9	10	11	12	Total
1	38.0786125	26.7895367	0.73160621	0	0	0	34.4002446	0	0	0	0	0	100
2	36.83586651	25.9152247	37.2489088	0	0	0	0	0	0	0	0	0	100
3	10.55954405	7.42898111	10.6779487	56.9939399	0	0	0	0	14.339586	0	0	0	100
4	10.21294942	7.18514056	10.3274677	0	52.0536961	20.22074619	0	0	0	0	0	0	100
5	24.21252108	17.0342925	24.4840172	0	24.6814346	9.587734622	0	0	0	0	0	0	100
6	33.75441201	23.7473217	34.1329019	0	0	0	0	0	0	0	8.3653645	0	100
7	16.46429049	10.1352662	19.7705751	0	0	0	18.5923141	0	0	0	0	35.037554	100
8	0	10.9042513	0	0	0	0	0	0.37885735	0	13.32503	0	75.391862	100
9	0	14.0994575	0	0	0	0	0	0.48987161	19.43935	17.229582	0	48.741739	100
10	0	42.0392336	0	0	0	0	0	0	57.960766	0	0	0	100

ESTIMATION OF SCALE UP FACTOR

Total Amount of NaOH obtained from 100 kmol/hr of salt 6901.22 kg/hr

Targeted Production Capacity of NaOH 1000 tons / day

But ton = 10^3 kg/hr and day = 24 hr

Therefore,

Targeted Production Capacity of NaOH 41667 kg/hr

Scale - up Factor is thus. 6.03762813

3.2 ENERGY BALANCE

Define: Stream indices

1- Salt, 2-Water, 3- Brine, 4- Na_2CO_3, 5- $CaCO_3$, 6- $Mg(OH)_2$, 7- Chlorate

8- Hydrogen, 9-NaOH, 10- Chlorine, 11- HCl, 12- Hg

$k := 1..4$

Heat Capacity Coefficients for Various Components

Heat capacity constants for Salt

$c_{1,1} := 36710$ $c_{1,2} := 62.77$ $c_{1,3} := -0.06667$ $c_{1,4} := 2.8 \cdot 10^{-5}$

Heat capacity constants for Water

$c_{2,1} := 276370$ $c_{2,2} := -2090.1$ $c_{2,3} := 8.125$ $c_{2,4} := -0.014116$

Heat capacity constants for Brine

$c_{3,1} := 51110$ $c_{3,2} := 72.24$ $c_{3,3} := -0.07583$ $c_{3,4} := 2.31 \cdot 10^{-5}$

Heat capacity constants for Na_2CO_3

$c_{4,1} := -25700$ $c_{4,2} := 1224.8$ $c_{4,3} := -4.442$ $c_{4,4} := 0.007642$

Heat capacity constants for $CaCO_3$

$c_{5,1} := -15500$ $c_{5,2} := 582.3$ $c_{5,3} := -1.067$ $c_{5,4} := 0.0001983$

Heat capacity constants for $Mg(OH)_2$

$c_{6,1} := -8000$ $c_{6,2} := 246.9$ $c_{6,3} := -0.3766$ $c_{6,4} := 0.0001983$

Heat capacity constants for Chlorate

$c_{7,1} := 15.127$ $c_{7,2} := 0.069703$ $c_{7,3} := 0.0018169$ $c_{7,4} := -1.4383 \cdot 10^{-6}$

Heat capacity constants for Hydrogen

$c_{8,1} := 33.5$ $c_{8,2} := 29.6$ $c_{8,3} := 1.52$ $c_{8,4} := 27.2$

Heat capacity constants for NaOH

$c_{9,1} := 71.427$ $c_{9,2} := 8.7015 \cdot 10^{-4}$ $c_{9,3} := 0$ $c_{9,4} := 0$

Heat capacity constants for Chlorine

$c_{10,1} := 0 \quad c_{10,2} := 0.8415 \quad c_{10,3} := -9.88\cdot 10^{-5} \quad c_{10,4} := 1.1398\cdot 10^{-7}$

Heat capacity constants for HCl

$c_{11,1} := -2.283 \quad c_{11,2} := 0.952 \quad c_{11,3} := -2.811\cdot 10^{-4} \quad c_{11,4} := 6.552\cdot 10^{-7}$

Heat capacity constants for Hg

$c_{12,1} := 3.1402 \quad c_{12,2} := 0.6318 \quad c_{12,3} := 0.0015534 \quad c_{12,4} := -1.755\cdot 10^{-6}$

Reference Temperature $T_r := 298\cdot K$

Stream Number n

Stream Temperature T_n

Enthalpy Equation for reacting components :

$$h(s,T_r) = \left[h_{f_s} + \left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}^{\ 2}\right) + c_{s,3}\cdot\left(T^3 - T_{ref}^{\ 3}\right) \ldots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}^{\ 4}\right) \end{array} \right]\cdot\frac{kJ}{kmol} \right]\cdot z_{s,j}$$

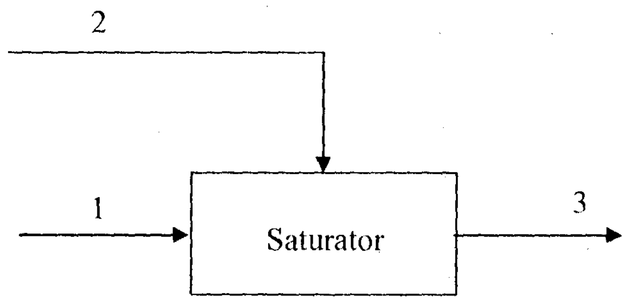
$$h(s,T_r) = \left[h_{f_s} + \left[\begin{array}{l} c_{s,1}\cdot\left(\frac{T}{K} - \frac{T_{ref}}{K}\right) + c_{s,2}\cdot\left(\frac{T^2}{K^2} - \frac{T_{ref}^{\ 2}}{K^2}\right) + c_{s,3}\cdot\left(\frac{T^3}{K^3} - \frac{T_{ref}^{\ 3}}{K^3}\right) \ldots \\ + c_{s,4}\cdot\left(\frac{T^4}{K^4} - \frac{T_{ref}^{\ 4}}{K^4}\right) \end{array} \right]\cdot\frac{kJ}{kmol} \right]\cdot z_{s,j}$$

Enthalpy Equation for non-reacting components :

$$h(s,T_r) = \left[\left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}^{\ 2}\right) + c_{s,3}\cdot\left(T^3 - T_{ref}^{\ 3}\right) \ldots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}^{\ 4}\right) \end{array} \right]\cdot\frac{kJ}{kmol} \right]\cdot z_{s,j}$$

$$h(s,T_r) = \left[\left[\begin{array}{l} c_{s,1}\cdot\left(\frac{T}{K} - \frac{T_{ref}}{K}\right) + c_{s,2}\cdot\left(\frac{T^2}{K^2} - \frac{T_{ref}^{\ 2}}{K^2}\right) + c_{s,3}\cdot\left(\frac{T^3}{K^3} - \frac{T_{ref}^{\ 3}}{K^3}\right) \ldots \\ + c_{s,4}\cdot\left(\frac{T^4}{K^4} - \frac{T_{ref}^{\ 4}}{K^4}\right) \end{array} \right]\cdot\frac{kJ}{kmol} \right]\cdot z_{s,j}$$

1. ENERGY BALANCE OF THE SATURATOR



Stream 1

Enthalpy of Species in Stream 1

Reference Temperature $T_{ref} := 298$

Stream Number $n := 1$

Stream Temperature $T_1 := (30 + 273) \quad T := T_1$

$$h_1(s,T_1) := \left[\left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}{}^2\right) + c_{s,3}\cdot\left(T^3 - T_{ref}{}^3\right) \cdots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}{}^4\right) \end{array}\right]\cdot\frac{J}{kmol}\right]\cdot\left(z_{s,1}\cdot\frac{kmol}{hr}\right)$$

Total Enthalpy of Species in Stream 1

$$H_1 := \sum_{s=1}^{12} h_1(s,T_1)$$

$$H_1 = -1.98 \times 10^6 \frac{kJ}{hr}$$

Stream 2

Enthalpy of Species in Stream 2

Reference Temperature $T_{ref} := 298$

Stream Number $n := 2$

Stream Temperature $T_2 := (55 + 273) \quad T := T_2$

$$h_2(s,T_2) := \left[\left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}{}^2\right) + c_{s,3}\cdot\left(T^3 - T_{ref}{}^3\right) \cdots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}{}^4\right) \end{array}\right]\cdot\frac{J}{kmol}\right]\cdot\left(z_{s,2}\cdot\frac{kmol}{hr}\right)$$

Total Enthalpy of Species in Stream 2

$$H_2 := \sum_{s=1}^{12} h_2(s,T_2)$$

$$H_2 = 3.948 \times 10^4 \frac{kJ}{hr}$$

Total Enthalpy of Species in the Inlet Streams

$$H_{inlet1} := H_1 + H_2$$

$$H_{inlet1} = -1.94 \times 10^6 \frac{kJ}{hr}$$

Stream 3

Enthalpy of Species in Stream 3

Reference Temperature $T_{ref} := 298$

Stream Number $n := 3$

Stream Temperature $T_3 := (40 + 273)$ $T := T_3$

$$h_3(s,T_3) := \left[\left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}^{\ 2}\right) + c_{s,3}\cdot\left(T^3 - T_{ref}^{\ 3}\right) \cdots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}^{\ 4}\right) \end{array}\right]\cdot\frac{J}{kmol}\right]\cdot\left(z_{s,3}\cdot\frac{kmol}{hr}\right)$$

Total Enthalpy of Species in Stream 3

$$H_3 := \sum_{s=1}^{12} h_3(s,T_3)$$

$$H_3 = -6.444 \times 10^6 \frac{kJ}{hr}$$

Total Enthalpy of Species in Outlet Stream

$$H_{outlet1} := H_3$$

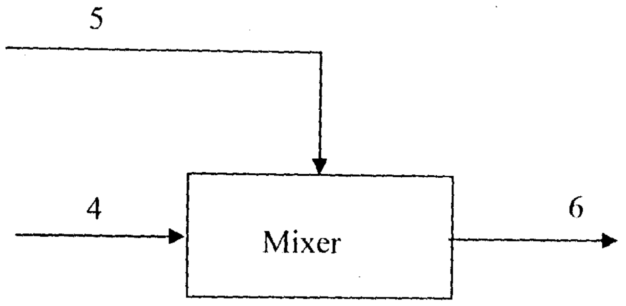
$$H_{outlet1} = -6.444 \times 10^6 \frac{kJ}{hr}$$

Total Heat Load

$$\Delta H_1 := H_{outlet1} - H_{inlet1}$$

$$\Delta H_1 = -4.504 \times 10^6 \frac{kJ}{hr}$$

2. ENERGY BALANCE OF THE MIXER



Stream 4

Enthalpy of Species in Stream 4

Reference Temperature $T_{ref} := 298$

Stream Number $n := 4$

Stream Temperature $T_4 := (30 + 273) \quad T := T_4$

$$h_4(s,T_4) := \left[\left[\begin{array}{l} c_{s,1}\cdot\left(T - T_{ref}\right) + c_{s,2}\cdot\left(T^2 - T_{ref}{}^2\right) + c_{s,3}\cdot\left(T^3 - T_{ref}{}^3\right) \ldots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}{}^4\right) \end{array}\right]\cdot\frac{J}{kmol}\right]\cdot\left(z_{s,4}\cdot\frac{kmol}{hr}\right)$$

Total Enthalpy of Species in Stream 4

$$H_4 := \sum_{s=1}^{12} h_4(s,T_4)$$

$$H_4 = 2.309 \times 10^5 \, \frac{kJ}{hr}$$

Stream 5

Enthalpy of Species in Stream 5

Reference Temperature $T_{ref} := 298$

Stream Number $n := 5$

Stream Temperature $T_5 := (40 + 273) \quad T := T_5$

$$h_5(s,T_5) := \left[\left[\begin{array}{l} c_{s,1}\cdot\left(T - T_{ref}\right) + c_{s,2}\cdot\left(T^2 - T_{ref}{}^2\right) + c_{s,3}\cdot\left(T^3 - T_{ref}{}^3\right) \ldots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}{}^4\right) \end{array}\right]\cdot\frac{J}{kmol}\right]\cdot\left(z_{s,5}\cdot\frac{kmol}{hr}\right)$$

Total Enthalpy of Species in Stream 5

$$H_5 := \sum_{s=1}^{12} h_5(s,T_5)$$

$$H_5 = -6.445 \times 10^6 \frac{kJ}{hr}$$

Total Enthalpy of Species in the Inlet Streams

$$H_{inlet2} := H_4 + H_5$$

$$H_{inlet2} = -6.214 \times 10^6 \frac{kJ}{hr}$$

Stream 6

Enthalpy of Species in Stream 6

Reference Temperature $T_{ref} := 298$

Stream Number $n := 6$

Stream Temperature $T_6 := (50 + 273)$ $T := T_6$

$$h_6(s,T_6) := \left[\left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}^{\ 2}\right) + c_{s,3}\cdot\left(T^3 - T_{ref}^{\ 3}\right) \ldots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}^{\ 4}\right) \end{array}\right]\cdot\frac{J}{kmol}\right]\cdot\left(z_{s,6}\cdot\frac{kmol}{hr}\right)$$

Total Enthalpy of Species in Stream 6

$$H_6 := \sum_{s=1}^{12} h_6(s,T_6)$$

$$H_6 = -1.052 \times 10^7 \frac{kJ}{hr}$$

Total Enthalpy of Species in Outlet Stream

$$H_{outlet2} := H_6$$

$$H_{outlet2} = -1.052 \times 10^7 \frac{kJ}{hr}$$

Total Heat Load

$$\Delta H_2 := H_{outlet2} - H_{inlet2}$$

$$\Delta H_2 = -4.308 \times 10^6 \frac{kJ}{hr}$$

3.3 EQUIPMENT DESIGN CALCULATIONS

This gives the size of each equipment which is being determined using the various reactants into each equipment. And also the choices of metal for construction as well as type of specific equipment are considered

3.3.1 DESIGN OF SATURATOR

Total flow rate of component to the Saturator

$$F_p := 93407.937\frac{kg}{hr}$$

$$F_p = 8.188 \times 10^8 \frac{kg}{yr}$$

Average density of component to the saturator

$$\rho_{salt} := 3465\cdot\frac{kg}{m^3} \quad \rho_{water} := 1000\cdot\frac{kg}{m^3} \quad \rho_{brine} := 3465\cdot\frac{kg}{m^3} \quad \rho_{chlorate} := 4250\cdot\frac{kg}{m^3}$$

$$M_{w\_salt} := 58.44\cdot\frac{g}{mol} \quad M_{w\_water} := 18.02\cdot\frac{g}{mol} \quad M_{w\_brine} := 58.44\cdot\frac{g}{mol} \quad M_{w\_chlorate} := 106.44\cdot\frac{g}{mol}$$

$$\rho_{avg} := \frac{\rho_{salt}\cdot M_{w\_salt} + \rho_{water}\cdot M_{w\_water} + \rho_{brine}\cdot M_{w\_brine} + \rho_{chlorate}\cdot M_{w\_chlorate}}{M_{w\_salt} + M_{w\_water} + M_{w\_brine} + M_{w\_chlorate}}$$

$$\rho_{avg} = 3.627 \times 10^3 \frac{kg}{m^3}$$

Volumetric flow rate of the saturator

$$V_p := \frac{F_p}{\rho_{avg}}$$

$$V_p = 25.752\frac{m^3}{hr}$$

Storage time, τ:

$$\tau := 1\cdot hr$$

Volume of the Saturator required for the storage time:

$$V := V_p\cdot\tau$$

$$V = 25.752\,m^3$$

The total volume of the tank can be calculated as thus:

$$V_{total} := V$$

$$V_{total} = 25.752\,m^3$$

Allowing for 30% safety allowance:

$$V_{total} := (V_{total} \cdot 0.3) + V_{total} \qquad V_{total} = 33.478\,m^3$$

The optimum proportions for a cylindrical container. A classical example of the optimization of a simple function.

The surface are, A, of a closed cylinder is:

$$A = \pi \times D \times L + 2\frac{\pi}{4}D^2$$

Where D = vessel diameter

L = vessel length (or height)

This will be the objective function which is to be minimized; simplified:

$$f(D \times L) = D \times L + \frac{D^2}{2}$$

For a given volume, V, the diameter and length are related by:

$$V = \frac{\pi}{4}D^2 \times L$$

$$L = 4\frac{V}{\pi D^2}$$

and the objective function becomes

$$f(D) = 4\frac{V}{\pi D} + \frac{D^2}{2}$$

Setting the differential of this function zero will give the optimum value for D

$$-4\frac{V}{\pi D^2} + D = 0$$

$$D := \sqrt[3]{4\frac{V_{total}}{\pi}}$$

$$D = 3.493\,m \qquad D = 11.461\,ft$$

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

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

$$Ratio := 0.83$$

$$H := \frac{D}{Ratio}$$

$$H = 4.209\,m \qquad H = 13.808\,ft$$

3.4 COST ESTIMATION EQUIPMENT AND ECONOMIC ANALYSIS

Acceptable plant design must present a process that is capable of operating under conditions which will yield profit. Since net profit equals total value minus all expenses, it is essential that the chemical engineer be aware of the many different types of cost involved in the manufacturing processes. Capital must be allocated for the direct, plant expenses, such as those for raw material, labour and equipment. Besides direct expenses many others indirect expenses are incurred, and these must be included if a complete analysis of the total cost is to be obtained. Some examples of these indirect expenses are administrative salary, product distribution cost and cost for interplant communication. A capital investment is required for every industrial process and determination of necessary investment is an important part of a plant design process. The total investment for any process consist of fixed capital investment for practical equipment and facilities in the plant plus working capital, which must be available to pay salaries, keep raw material and products at hand, and handle other special items requiring the direct cost outline.

When the cost for any type of commercial process is to be determined, sufficient accuracy has to be provided for reliable decision. There are many factors affecting investment and production cost. These are;

1. source of equipment
2. price fluctuation
3. company policies
4. operating and rate of production
5. governmental policies

Before an industrial plant can be put into operation, a large sum of money must be supplied to purchase and install the necessary machinery and equipment.

3.4.1 Costing of Equipment

The cost are based on cost data of 2004 which were available in Dollars

Exchange rate ER $150Naira = 1Dollar$

Therefore $ER := 150Naira$ $Naira \equiv 1$

3.4.2 Purchased Cost Data:

The purchased cost data for process equipments is given below as:

3.4.2.1 Purchased Cost data for Pressure Vessels, Reactors and Columns

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

where

D = Diameter of the column (ft)

H = Height of the column (ft)

M_S = Marshall and Smith Index

$$F_C = 1.00 + F_m + F_p$$

F_m		F_p	
CS	1.00	<50 psia	1.00
SS	3.67	200	1.15
Monel	6.34	400	1.35
Titanium	7.89	600	1.60

Purchased Cost of the Saturator:

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34$ $F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the tank:

$D_{saturator} := 11.461\ ft$

Height of the tank:

$H_{saturator} := 13.808\ ft$

$$PC_{saturator} := \frac{M\Omega S}{280}\left[101.9 \cdot \left(D_{saturator}\right)^{1.066} \cdot \left(H_{saturator}\right)^{0.802} \cdot F_c\right] \cdot ER$$

$PC_{saturator} = 5.768 \times 10^6\ Naira$

Purchased Cost of the Mixer:

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the tank:

$D_{mixer} := 12.164\ ft$

Height of the tank:

$H_{mixer} := 14.656\ ft$

$$PC_{mixer} := \frac{M\Omega S}{280}\left[101.9 \cdot \left(D_{mixer}\right)^{1.066} \cdot \left(H_{mixer}\right)^{0.802} \cdot F_c\right] \cdot ER$$

$PC_{mixer} = 6.447 \times 10^6\ Naira$

Purchased Cost of the Precipitator:

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the tank:

$D_{precipitator} := 15.414\ ft$

Height of the tank:

$H_{precipitator} := 18.572\ ft$

$$PC_{precipitator} := \frac{M\Omega S}{280}\left[101.9 \cdot \left(D_{precipitator}\right)^{1.066} \cdot \left(H_{precipitator}\right)^{0.802} \cdot F_c\right] \cdot ER$$

$PC_{precipitator} = 1.003 \times 10^{7}$ Naira

Purchased Cost of the Filter Tank

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the tank:

$D_{filter\_tank} := 13.646\ ft$

Height of the tank:

$H_{filter\_tank} := 16.44ft$

$$PC_{filter\_tank} := \frac{M\Omega S}{280}\left[101.9\cdot\left(D_{filter\_tank}\right)^{1.066}\cdot\left(H_{filter\_tank}\right)^{0.802}\cdot F_c\right]\cdot ER$$

$PC_{filter\_tank} = 7.991 \times 10^6$ Naira

Purchased Cost of the Ion Exchanger:

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the Exchanger:

$D_{exchanger} := 10.234$ ft

Height of the tank:

$H_{exchanger} := 12.33$ ft

$$PC_{exchanger} := \frac{M\Omega S}{280}\left[101.9\cdot\left(D_{exchanger}\right)^{1.066}\cdot\left(H_{exchanger}\right)^{0.802}\cdot F_c\right]\cdot ER$$

$PC_{exchanger} = 4.669 \times 10^6$ Naira

Purchased Cost of the Second Mixer:

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the mixer:

$D_{mixer\_2} := 12.895$

Height of the tank:

$H_{mixer\_2} := 15.536\,\mathrm{ft}$

$$PC_{mixer\_2} := \frac{M\Omega S}{280}\left[101.9\cdot\left(D_{mixer\_2}\right)^{1.066}\cdot\left(H_{mixer\_2}\right)^{0.802}\cdot F_c\right]\cdot ER$$

$PC_{mixer\_2} = 7.189 \times 10^{6}\,\mathrm{Naira}$

Purchased Cost of the electrolytic cell:

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the electrolytic cell:

$D_{cell} := 11.482\,\mathrm{ft}$

Height of the tank:

$H_{cell} := 13.834\,\mathrm{ft}$

$$PC_{electrolytic\_cell} := \frac{M\Omega S}{280}\left[101.9\cdot\left(D_{cell}\right)^{1.066}\cdot\left(H_{cell}\right)^{0.802}\cdot F_c\right]\cdot ER$$

$PC_{electrolytic\_cell} = 5.788 \times 10^{6}\,\mathrm{Naira}$

Purchased Cost of the Decomposer

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the tank:

$D_{decomposer} := 11.855\,ft$

Height of the tank:

$H_{decomposer} := 14.284\,ft$

$$PC_{decomposer} := \frac{M\Omega S}{280}\left[101.9\cdot\left(D_{decomposer}\right)^{1.066}\cdot\left(H_{decomposer}\right)^{0.802}\cdot F_c\right]\cdot ER$$

$PC_{decomposer} = 6.145 \times 10^{6}\,\text{Naira}$

Purchased Cost of the Seperator Tank

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the tank:

$D_{seperator} := 10.43\ ft$

Height of the tank:

$H_{seperator} := 12.566ft$

$$PC_{seperator} := \frac{M\Omega S}{280}\left[101.9\cdot\left(D_{seperator}\right)^{1.066}\cdot\left(H_{seperator}\right)^{0.802}\cdot F_c\right]\cdot ER$$

$PC_{seperator} = 4.837 \times 10^{6}\,\text{Naira}$

Purchased Cost of the Storage Tank

Purchase cost PC is given by the relation below

Marshal and Swift index is:

$M\Omega S := 110$

The material of construction selected is Monnel steel:

$F_m := 6.34 \qquad F_p := 1.35$

$F_c := 1 + F_m + F_p$

$F_c = 8.69$

Diameter of the tank:

$D_{storage\_tank} := 6.816\ \text{ft}$

Height of the tank:

$H_{storage\_tank} := 8.212\ \text{ft}$

$$PC_{storage\_tank} := \frac{M\Omega S}{280}\left[101.9\cdot\left(D_{storage\_tank}\right)^{1.066}\cdot\left(H_{storage\_tank}\right)^{0.802}\cdot F_c\right]\cdot ER$$

$PC_{storage\_tank} = 2.185 \times 10^6\ \text{Naira}$

Total Purchased Cost of equipments

$$PC_{equip} := PC_{saturator} + PC_{mixer} + PC_{precipitator} + PC_{filter\_tank} + PC_{exchanger} \cdots \\ + PC_{mixer\_2} + PC_{electrolytic\_cell} + PC_{decomposer} + PC_{seperator} + PC_{storage\_tank}$$

$PC_{equip} = 6.105 \times 10^7\ \text{Naira}$

3.4.3 ECONOMIC ESTIMATION OF TOTAL CAPITAL INVESTIMENT

The total capital investiment can be given by

$T_{inv} = TFC + WC + TLC$

where, TFC =total fixed cost

WC=working capital

TLC=total land cost

3.4.3.1 Total fixed cost

The factorial method can be used with the relationship below

Total physical plant cost (Direct cost)

$PPC := 10PC_{equip}$

$PPC = 6.105 \times 10^{8}$ Naira

Fixed Capital cost (Indirect cost)

$TFC := 5PPC$

$TFC = 3.053 \times 10^{9}$ Naira

Working capital

$WC := 2 \cdot TFC$

$WC = 6.105 \times 10^{9}$ Naira

Total land cost

$TLC := 0.2 \cdot TFC$

$TLC = 6.105 \times 10^{8}$ Naira

$T_{inv} := TLC + WC + TFC$

$T_{inv} = 9.768 \times 10^{9}$ Naira

3.4.3.2 Operating cost

This is divided into Fixed and Variable operating cost

Let the plant life be 15yrs $n := 15yr$

3.4.3.2.1 Fixed operating cost

Direct labour cost

$Lb_{c} := 0.06TFC$

$Lb_{c} = 1.832 \times 10^{8}$ Naira

Plant maintenace and repairs

$M_c := 0.08TFC$

$M_c = 2.442 \times 10^8$ Naira

Insurance

$Ins_c := 0.01TFC$

$Ins_c = 3.053 \times 10^7$ Naira

Local Taxes

$Ltx_c := 0.02TFC$

$Ltx_c = 6.105 \times 10^7$ Naira

Royalties and licences fee

$Roy_c := 0.01 \cdot TFC$

$Roy_c = 3.053 \times 10^7$ Naira

Laboratory cost

$Lab_c := 0.2Lb_c$

$Lab_c = 3.663 \times 10^7$ Naira

Supervision

$S_c := 0.2Lb_c$

$S_c = 3.663 \times 10^7$ Naira

Plant overhead cost

$POH_c := 0.5Lb_c$

$POH_c = 9.158 \times 10^7$ Naira

GENERAL EXPENSES

i administrative

$ADM_c := 0.25Lb_c$

$ADM_c = 4.579 \times 10^7$ Naira

ii. Research and development

$RAD_c := 0.015 \cdot TFC$

$RAD_c = 4.579 \times 10^7 \text{Naira}$

Total fixed operating cost

$TFO_c := RAD_c + ADM_c + POH_c + Lab_c + Roy_c + Ltx_c + Ins_c + M_c + Lb_c + S_c$

$TFO_c = 8.059 \times 10^8 \text{Naira}$

Annual fixed operating cost

$$TFO_{annum} := \frac{TFO_c}{n}$$

$$TFO_{annum} = 5.373 \times 10^7 \frac{\text{Naira}}{\text{yr}}$$

3.4.3.2.2 Variable operating cost

Cost of raw materials CRMc

$$Salt\_{cost} := \left[(35283.899) \cdot \frac{kg}{hr}\right] \cdot \frac{0.35ER}{kg}$$

$$Salt\_{cost} = 1.852 \times 10^6 \frac{\text{Naira}}{\text{hr}}$$

$$Raw\_Brine_{cost} := \left(35283.899 \cdot \frac{kg}{hr}\right) \cdot \frac{0.45ER}{kg}$$

$$Raw\_Brine_{cost} = 2.382 \times 10^6 \frac{\text{Naira}}{\text{hr}}$$

$$Na2CO3_{cost} := \left(191976.6503 \cdot \frac{kg}{hr}\right) \cdot \frac{0.15ER}{kg}$$

$$Na2CO3_{cost} = 4.319 \times 10^6 \frac{\text{Naira}}{\text{hr}}$$

$$NaOH_{cost} := \left(48301.025 \cdot \frac{kg}{hr}\right) \cdot \frac{0.5ER}{kg}$$

$$NaOH_{cost} = 3.623 \times 10^6 \frac{\text{Naira}}{\text{hr}}$$

$$HCl_{cost} := \left(8814.937 \cdot \frac{kg}{hr}\right) \cdot \frac{0.06ER}{kg}$$

$$HCl_{cost} = 7.933 \times 10^4 \frac{\text{Naira}}{\text{hr}}$$

$$H2O_{cost} := \left(25023.5536 \cdot \frac{kg}{hr}\right) \cdot \frac{0.009ER}{kg}$$

$$H2O_{cost} = 3.378 \times 10^{4} \frac{Naira}{hr}$$

$$CRM_c := Salt\_{cost} + H2O_{cost} + Raw\_Brine_{cost} + Na2CO3_{cost} + NaOH_{cost} + H2O_{cost} + HCl_{cost}$$

$$CRM_c = 1.232 \times 10^{7} \frac{Naira}{hr}$$

$$ACRM := (CRM_c) \cdot \left(7992 \frac{hr}{yr}\right)$$

$$ACRM = 9.849 \times 10^{10} \frac{Naira}{yr}$$

Miscellaneuos

$$Ms_c := 0.05Lb_c$$

$$Ms_c = 9.158 \times 10^{6} Naira$$

Utilities cost

$$Ut_c := 0.02Lb_c$$

$$Ut_c = 3.663 \times 10^{6} Naira$$

packaging

$$PA_c := 0.001Lb_c \cdot Naira$$

$$PA_c = 1.832 \times 10^{5} Naira$$

Total variable operating cost

$$TVO_c := PA_c + Ms_c + ACRM + Ut_c$$

$$TVO_c = 9.85 \times 10^{10} \ Naira$$

Annual variable operating cost

$$TVO_{annum} := \frac{TVO_c}{n}$$

$$TVO_{annum} = 6.567 \times 10^{9} \frac{Naira}{yr}$$

Total annual operating cost

$$AOC_c := TFO_{annum} + TVO_{annum}$$

$$AOC_c = 6.62 \times 10^{9} \frac{Naira}{yr}$$

3.4.3.3 Profit analysis

Annal sales of Product or Revenue (A_{sp})

Sales of NaOH

$$NaOH_{sales} := \left(41667 \cdot \frac{kg}{hr}\right) \cdot \frac{0.35 \cdot ER}{kg}$$

$$NaOH_{sales} = 2.188 \times 10^{6} \frac{Naira}{hr}$$

Total Sales

$$Tot_{sales} := NaOH_{sales}$$

$$Tot_{sales} = 2.188 \times 10^{6} \frac{Naira}{hr}$$

Annual sales of product

$$As_{p} := Tot_{sales} \cdot \left(7992 \frac{hr}{yr}\right)$$

$$As_{p} = 1.748 \times 10^{10} \frac{Naira}{yr}$$

Profit before tax (PBT)

$$PBT := As_{p} - AOC_{c}$$

$$PBT = 1.086 \times 10^{10} \frac{Naira}{yr}$$

Annual depreciation (Depr)

let S = Salvage Value after n years of the plant. Assume plant life of 15 yrs

Then $S := 10 \cdot 10^{8} Naira$ $n := 15$

$$Depr := \frac{T_{inv} - S}{n}$$

$$Depr = 5.846 \times 10^{8} \frac{Naira}{yr}$$

Tax payable (TP)

Assume tax ratio of 70% and depreciation is tax allowable, hence $Tax\_ratio := 0.6$

$$TP := (PBT - Depr) \cdot Tax\_ratio$$

$$TP = 6.167 \times 10^{9} \frac{Naira}{yr}$$

Profit after tax (PAT)

$$PAT := PBT - TP$$

$$PAT = 4.696 \times 10^{9}\,\frac{Naira}{yr}$$

Net income (NIN)

$$NIN := PAT + Depr$$

$$NIN = 5.28 \times 10^{9}\,\frac{Naira}{yr}$$

Pay back period (PBP)

$$PBP := \frac{T_{inv}}{NIN}$$

$$PBP = 1.85\,yr$$

Rate of Return (ROR)

$$ROR := \frac{NIN + Depr}{T_{inv}} \cdot 100$$

$$ROR = 60.039\,\%$$

3.6 PLANT LOCATION AND SITE SELECTION, AND PLANT LAYOUT

3.6.1 Plant Location and Site Selection

The geographical location of the final plant can have strong influence on the success of the industrial venture. Considerable care must be exercised in selecting the plant site, and many different factors must be considered. Primarily the plant must be located where the minimum cost of production and distribution can be obtained but, other factors such as room for expansion and safe living conditions for plant operation as well as the surrounding community are also important. The location of the plant can also have a crucial effect on the profitability of a project. The choice of the final site should first be based on a complete survey of the advantages and disadvantages of various geographical areas and ultimately, on the advantages and disadvantages of the available real estate. The various principal factors that must be considered while selecting a suitable plant site, are briefly discussed in this section. The factors to be considered are:

1. Raw material availability.
2. Location (with respect to the marketing area.)
3. Availability of suitable land.
4. Transport facilities.
5. Availability of labors.
6. Availability of utilities (Water, Electricity).
7. Environmental impact and effluent disposal.
8. Local community considerations.
9. Climate.
10. Political strategic considerations.
11. Taxations and legal restrictions

3.6.1.1 Raw materials availability

The source of raw materials is one of the most important factors influencing the selection of a plant site. This is particularly true for the spent dry cell recovery plant because large volumes of spent dry cells are used in the production process which will result in the reduction of the transportation and storage charges. Attention should be given to the purchased price of the raw materials, distance from the source of supply, flight and transportation expenses, availability and reliability of supply, purity of raw materials and storage requirements.

3.6.1.2 Market location

The location of markets or intermediate distribution centers affects the cost of product distribution and time required for shipping. Proximity to the major markets is an important consideration in the selection of the plant site, because the buyer usually find it advantageous to purchase from near-by sources.

3.6.1.3 Availability of suitable land

The characteristics of the land at the proposed plant site should be examined carefully. The topography of the tract of land structure must be considered, since either or both may have a pronounced effect on the construction costs. The cost of the land is important, as well as local building costs and living conditions. Future changes may make it desirable or necessary to expand the plant facilities. The land should be ideally flat, well drained and have load-bearing characteristics. A full site evaluation should be made to determine the need for piling or other special foundations

3.6.1.4 Transport

The transport of materials and products to and from plant will be an overriding consideration in site selection. If practicable, a site should be selected so that it is close to at least two major forms of transport: road, rail, waterway or a seaport. Road transport is being increasingly used, and is suitable for local distribution from a central warehouse. Rail transport will be cheaper for the long-distance transport. If possible the plant site should have access to all three types of transportation. There is usually need for convenient rail and air transportation facilities between the plant and the main company head quarters, and the effective transportation facilities for the plant personnel are necessary.

3.6.1.5 Availability of labours

Labors will be needed for construction of the plant and its operation. Skilled construction workers will usually be brought in from outside the site, but there should be an adequate pool of unskilled labors available locally; and labors suitable for training to operate the plant. Skilled tradesmen will be needed for plant maintenance. Local trade union customs and restrictive practices will have to be considered when assessing the availability and suitability of the labors for recruitment and training.

3.6.1.6 Availability of utilities

The word "utilities" is generally used for the ancillary services needed in the operation of any production process. These services will normally be supplied from a central facility and includes Water, Fuel and Electricity which are briefly described as follows:

(i) Water: - The water is required for large industrial as well as general purposes, starting with water for cooling, washing, steam generation and as a raw material in the. The plant therefore must be located where a dependable water supply is available namely lakes, rivers, wells, seas. If the water supply shows seasonal fluctuations, it's desirable to construct a reservoir or to drill several standby wells. The temperature, mineral content, slit and sand content, bacteriological content, and cost for supply and purification treatment must also be considered when choosing a water supply. Demineralized water, from which all the minerals have been removed is used where pure water is needed for the process use, in boiler feed. Natural and forced draft cooling towers are generally used to provide the cooling water required on site.

(ii) Electricity: - Power and steam requirements are high in most industrial plants and fuel is ordinarily required to supply these utilities. Power, fuel and steam are required for running the various equipments like generators, motors, turbines, plant lightings and general use and thus be considered as one major factor is choice of plant site.

3.6.1.7 Environmental impact and effluent disposal

Facilities must be provided for the effective disposal of the effluent without any public nuisance. In choosing a plant site, the permissible tolerance levels for various effluents should be considered and attention should be given to potential requirements for additional waste treatment facilities. As all industrial processes produce waste products, full consideration must be given to the difficulties and cost of their disposal. The disposal of toxic and harmful effluents will be covered by local regulations, and the appropriate authorities must be consulted during the initial site survey to determine the standards that must be met.

3.6.1.8 Local community considerations

The proposed plant must fit in with and be acceptable to the local community. Full consideration must be given to the safe location of the plant so that it does not impose a significant additional risk to the community.

3.6.1.9 Climate

Adverse climatic conditions at site will increase costs. Extremes of low temperatures will require the provision of additional insulation and special heating for equipment and piping. Similarly, excessive humidity and hot temperatures pose serious problems and must be considered for selecting a site for the plant. Stronger structures will be needed at locations subject to high wind loads or earthquakes.

3.6.1.10 Political and strategic considerations

Capital grants, tax concessions, and other inducements are often given by governments to direct new investment to preferred locations; such as areas of high unemployment. The availability of such grants can be the overriding consideration in site selection.

3.6.1.11Taxation and legal restrictions

State and local tax rates on property income, unemployment insurance, and similar items vary from one location to another. Similarly, local regulations on zoning, building codes, nuisance aspects and others facilities can have a major influence on the final choice of the plant site.

3.6.2 Plant Layout

After the flow process diagrams are completed and before detailed piping, structural and electrical design can begin, the layout of process units in a plant and the equipment within these process unit must be planned. This layout can play an important part in determining construction and manufacturing costs, and thus must be planned carefully with attention being given to future problems that may arise. Thus the economic construction and efficient operation of a process unit will depend on how well the plant and equipment specified on the process flow sheet is laid out. The principal factors that are considered are listed below:

1. Economic considerations: construction and operating costs.
2. Process requirements.
3. Convenience of operation.
4. Convenience of maintenance.
5. Health and Safety considerations.
6. Future plant expansion.
7. Modular construction.
8. Waste disposal requirements

3.6.2.1 Economic considerations: operating costs

The cost of construction can be minimized by adopting a layout that gives the shortest run of connecting pipe between equipment, and least amount of structural steel work. However, this will not necessarily be the best arrangement for operation and maintenance.

3.6.2.2 Process requirements

An example of the need to take into account process consideration is the need to elevate the base of columns to provide the necessary net positive suction head to a pump.

3.6.2.3 Convenience of operation

Equipment that needs to have frequent attention should be located convenient to the control room. Valves, sample points, and instruments should be located at convenient positions and heights. Sufficient working space and headroom must be provided to allow easy access to equipment.

3.6.2.4 Convenience of maintenance

Heat exchangers need to be sited so that the tube bundles can be easily withdrawn for cleaning and tube replacement. Vessels that require frequent replacement of catalyst or packing should be located on the out side of buildings. Equipment that requires dismantling for maintenance, such as compressors and large pumps, should be places under cover.

3.6.2.5 Health and safety considerations

Blast walls may be needed to isolate potentially hazardous equipment, and confine the effects of an explosion. At least two escape routes for operators must be provided from each level in process buildings.

3.6.2.6 Future plant expansion

Equipment should be located so that it can be conveniently tied in with any future expansion of the process. Space should be left on pipe alleys for future needs, and service pipes over-sized to allow for future requirements.

12.2.7 Modular construction

In recent years there has been a move to assemble sections of plant at the plant manufacturer's site. These modules will include the equipment, structural steel, piping and instrumentation. The modules are then transported to the plant site, by road or sea.

The advantages of modular construction are:

1. Improved quality control.
2. Reduced construction cost.
3. Less need for skilled labors on site.

The disadvantages of modular construction are:

1. Higher design costs & more structural steel work.
2. More flanged constructions and possible problems with assembly on site.

3.6.2.8 Waste disposal requirements

In a plant layout, the permissible tolerance levels for various waste should be considered and attention should be given to potential requirements for additional waste treatment facilities. As all industrial processes produce waste products, full consideration must be given to the difficulties and cost of their disposal. The disposal of toxic and harmful effluents will be covered by local regulations, and the appropriate authorities must be consulted during the initial site survey to determine the standards that must be met.

3.7 PROCESS CONTROL AND INSTRUMENTATION

3.7.1 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).

3.7.2 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.

1. **Production rate:** To achieve the design producer output
2. **Product quality:** To maintain the product composition within the specific quality standards.
3. **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.

a. **Flow rates/feed ratio control:** a feed ratio controller is aim at maintaining a complete conversion to the desired products. This is necessary to avoid excess built up in the reactor.

b. **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.

c. **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

CHAPTER FOUR

4.0 RESULTS AND DISCUSSION OF RESULT

4.1 RESULTS

4.1.1 SUMMARY OF MATERIAL BALANCE

Table 4.1: Material Balance around the Saturator Tank (Unit 1)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 1		Stream 2		Stream 3	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	35283.73	603.76	285.77	4.89	35568.34	608.63
Water	25023.47	1388.65	0	0	25023.47	1388.65
Brine	0	0	683.16	11.69	683.16	11.69
Na_2CO_3	0	0	0	0	0	0
$CaCO_3$	0	0	0	0	0	0
$Mg(OH)_2$	0	0	0	0	0	0
Chlorate	0	0	32132.41	301.88	32132.41	301.88
Hydrogen	0	0	0	0	0	0
NaOH	0	0	0	0	0	0
Chlorine	0	0	0	0	0	0
HCl	0	0	0	0	0	0
Hg	0	0	0	0	0	0
Total	60307.2	1992.42	33101.34	318.44	93408.54	2310.86
	Total Input (Stream 1 + Stream 2)				**Total Output** Stream 3	
	(kg/hr)		(kmol/hr)		(kg/hr)	(kmol/hr)
Overall Total	**93408.54**		**2310.86**		**93408.54**	**2310.86**

Table 4.2: Material Balance around the Mixer (Unit 2)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 4		Stream 5		Stream 36	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	0	0	35575.93	608.76	35575.93	608.63
Water	0	0	25023.47	1388.65	25023.47	1388.65
Brine	35283.73	603.76	683.16	11.69	35967.48	615.46
Na_2CO_3	0	0	0	0	0	0
$CaCO_3$	0	0	0	0	0	0
$Mg(OH)_2$	0	0	0	0	0	0
Chlorate	0	0	0	0	0	0
Hydrogen	0	0	0	0	0	0
NaOH	0	0	0	0	0	0
Chlorine	0	0	0	0	0	0
HCl	0	0	0	0	0	0
Hg	0	0	0	0	0	0
Total	35283.73	603.76	61282.57	2008.98	96566.88	2612.74
	Total Input (Stream 4 + Stream 5)				**Total Output** Stream 6	
	(kg/hr)		(kmol/hr)		(kg/hr)	(kmol/hr)
Overall Total	**96566.30**		**2612.74**		**96566.30**	**2612.74**

Table 4.3: Material Balance around the Precipitator (Unit 3)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 7		Stream 8		Stream 9	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	0	0	35575.93	608.63	35575.93	608.63
Water	0	0	25023.47	1388.65	25023.47	1388.65
Brine	0	0	35967.48	615.46	35967.48	615.46
Na_2CO_3	191976.82	1811.29	0	0	191976.82	1811.29
$CaCO_3$	0	0	0	0	0	0
$Mg(OH)_2$	0	0	0	0	0	0
Chlorate	0	0	0	0	0	0
Hydrogen	0	0	0	0	0	0
NaOH	48301.2	1207.53	0	0	48301.2	1207.53
Chlorine	0	0	0	0	0	0
HCl	0	0	0	0	0	0
Hg	0	0	0	0	0	0
Total	240278.02	3018.81	96566.88	2612.74	336844.9	5631.56
	Total Input (Stream 7 + Stream 8)				**Total Output** Stream 9	
	(kg/hr)		(kmol/hr)		(kg/hr)	(kmol/hr)
Overall Total	**336844.9**		**5631.56**		**336844.9**	**5631.56**

Table 4.4: Material Balance around the Filter Tank (Unit 4)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 10		Stream 11		Stream 12	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	0	0	35575.93	608.63	35575.93	608.63
Water	0	0	25023.47	1388.65	25023.47	1388.65
Brine	0	0	35967.48	615.46	35967.48	615.46
Na_2CO_3	0	0	0	0	0	0
$CaCO_3$	181290.20	1811.29	0	0	181290.20	1811.29
$Mg(OH)_2$	70422.67	1207.53	0	0	70422.67	1207.53
Chlorate	0	0	0	0	0	0
Hydrogen	0	0	0	0	0	0
NaOH	0	0	0	0	0	0
Chlorine	0	0	0	0	0	0
HCl	0	0	0	0	0	0
Hg	0	0	0	0	0	0
Total	251712.87	3018.81	96566.88	2612.74	348279.75	5631.56
	Total Input (Stream 10 + Stream 11)				**Total Output** Stream 12	
	(kg/hr)		(kmol/hr)		(kg/hr)	(kmol/hr)
Overall Total	**348279.75**		**5631.56**		**348279.75**	**5631.56**

Table 4.5: Material Balance around the Ion Exchange Purificator (Unit 5)

	INPUT STREAMS		**OUTPUT STREAMS**	
	Stream 13		Stream 14	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	35575.93	608.63	35575.93	608.63
Water	25023.47	1388.65	25023.47	1388.65
Brine	35967:48	615.46	35967.48	615.46
Na_2CO_3	0	0	0	0
$CaCO_3$	36257.52	362.26	36257.52	362.26
$Mg(OH)_2$	14084.77	241.51	14084.77	241.51
Chlorate	0	0	0	0
Hydrogen	0	0	0	0
NaOH	0	0	0	0
Chlorine	0	0	0	0
HCl	0	0	0	0
Hg	0	0	0	0
Total	146909.17	3216.51	146909.17	3216.51

	Total Input Stream 13		**Total Output** Stream 14	
	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Overall Total	**146909.17**	**3216.51**	**146909.17**	**3216.51**

Table 4.6: Material Balance around the Second Mixer (Unit 6)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 15		Stream 16		Stream 17	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	0	0	35575.93	608.63	35575.93	608.63
Water	0	0	25023.47	1388.65	25023.47	1388.65
Brine	0	0	35967.48	615.46	35967.48	615.46
Na_2CO_3	0	0	0	0	0	0
$CaCO_3$	0	0	0	0	0	0
$Mg(OH)_2$	0	0	0	0	0	0
Chlorate	0	0	0	0	0	0
Hydrogen	0	0	0	0	0	0
NaOH	0	0	0	0	0	0
Chlorine	0	0	0	0	0	0
HCl	8815.12	241.51	0	0	8815.12	241.51
Hg	0	0	0	0	0	0
Total	8815.12	241.51	96566.88	2612.74	105382	2854.25
	Total Input (Stream 15 + Stream 16)				**Total Output** Stream 17	
	(kg/hr)		(kmol/hr)		(kg/hr)	(kmol/hr)
Overall Total	**105382**		**2854.25**		**105382**	**2854.25**

Table 4.7: Material Balance around the Electrolytic Tank (Unit 7)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 18		Stream 19		Stream 20	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	28455.02	486.91	0	0	28455.02	486.91
Water	17516.52	972.06	0	0	17516.52	972.06
Brine	34168.70	584.68	0	0	34168.70	584.68
Na_2CO_3	0	0	0	0	0	0
$CaCO_3$	0	0	0	0	0	0
$Mg(OH)_2$	0	0	0	0	0	0
Chlorate	32132.41	301.88	0	0	32132.41	301.88
Hydrogen	0	0	0	0	0	0
NaOH	0	0	0	0	0	0
Chlorine	0	0	0	0	0	0
HCl	0	0	0	0	0	0
Hg	0	0	60554.11	301.88	60554.11	301.88
Total	112272.65	2345.53	60554.11	301.88	172826.76	2647.41
	Total Input (Stream 18 + Stream 19)				**Total Output** Stream 20	
	(kg/hr)		(kmol/hr)		(kg/hr)	(kmol/hr)
Overall Total	**172826.76**		**2647.41**		**172826.76**	**2647.41**

Table 4.8: Material Balance around the Amalgam Decomposer (Unit 8)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 21		Stream 22		Stream 23	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	0	0	0	0	0	0
Water	0	0	17516.52	972.06	17516.52	972.06
Brine	0	0	0	0	0	0
Na_2CO_3	0	0	0	0	0	0
$CaCO_3$	0	0	0	0	0	0
$Mg(OH)_2$	0	0	0	0	0	0
Chlorate	0	0	0	0	0	0
Hydrogen	608.59	301.88	0	0	608.59	301.88
NaOH	0	0	0	0	0	0
Chlorine	21405.10	301.88	0	0	21405.10	301.88
HCl	0	0	0	0	0	0
Hg	121108.22	603.76	0	0	121108.22	603.76
Total	143121.81	1207.53	17516.52	972.06	160638.33	2179.58
	Total Input (Stream 21 + Stream 22)				**Total Output** Stream 23	
	(kg/hr)		(kmol/hr)		(kg/hr)	(kmol/hr)
Overall Total	**160638.33**		**2179.58**		**160638.33**	**2179.58**

Table 4.9: Material Balance around the Separation Tank (Unit 9)

	INPUT STREAMS				**OUTPUT STREAMS**	
	Stream 24		Stream 25		Stream 26	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	0	0	0	0	0	0
Water	0	0	17516.52	972.06	17516.52	972.06
Brine	0	0	0	0	0	0
Na_2CO_3	0	0	0	0	0	0
$CaCO_3$	0	0	0	0	0	0
$Mg(OH)_2$	0	0	0	0	0	0
Chlorate	0	0	0	0	0	0
Hydrogen	0	0	608.59	301.88	608.59	301.88
NaOH	24150.4	603.76	0	0	24150.4	603.76
Chlorine	0	0	21405.10	301.88	21405.10	301.88
HCl	0	0	0	0	0	0
Hg	0	0	60554.11	301.88	60554.11	301.88
Total	24150.4	603.76	100084.32	1877.70	124234.72	2481.47

	Total Input (Stream 24 + Stream 25)		**Total Output** Stream 26	
	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Overall Total	**124234.72**	**2481.47**	**124234.72**	**2481.47**

Table 4.10: Material Balance around the NaOH Storage Tank (Unit 10)

	INPUT STREAMS		**OUTPUT STREAMS**	
	Stream 27		Stream 28	
Components	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Salt	0	0	0	0
Water	17516.52	972.06	17516.52	972.06
Brine	0	0	0	0
Na_2CO_3	0	0	0	0
$CaCO_3$	0	0	0	0
$Mg(OH)_2$	0	0	0	0
Chlorate	0	0	0	0
Hydrogen	0	0	0	0
NaOH	24150.4	603.76	24150.4	603.76
Chlorine	0	0	0	0
HCl	0	0	0	0
Hg	0	0	0	0
Total	41666.95	1575.82	41666.95	1575.82
	Total Input Stream 27		**Total Output** Stream 28	
	(kg/hr)	(kmol/hr)	(kg/hr)	(kmol/hr)
Overall Total	**41666.95**	**1575.82**	**41666.95**	**1575.82**

4.1.2 SUMMARY OF EQUIPMENT SPECIFICATION

TABLE 4.11 Result of Capacities or Sizes of Various Equipment

SERIAL NO.	EQUIPMENT	SIZE OR CAPACITY (m^3)
1	Saturator	33.478
2	Mixer	40.031
3	Precipitator	81.454
4	Filter Tank	56.507
5	Ion Exchange Purificator	23.835
6	Second Mixer	47.688
7	Electrolytic Tank	33.668
8	Amalgam Decomposer	37.059
9	Separator	25.235
10	Storage Tank	7.042

4.1.3 SUMMARY OF ECONOMIC ANALYSIS

TABLE 4.12 Result of Cost Estimation of Equipment

SERIAL NO.	EQUIPMENT	COST OF EQUIPMENT (₦)
1	Saturator	₦5768000.00
2	Mixer	₦6447000.00
3	Precipitator	₦10030000.00
4	Filter Tank	₦7991000.00
5	Ion Exchange Purificator	₦4669000.00
6	Second Mixer	₦7189000.00
7	Electrolytic Tank	₦5788000.00
8	Amalgam Decomposer	₦6145000.00
9	Separator	₦4837000.00
10	Storage Tank	₦2185000.00
11	Total Cost of Equipment	₦61050000.00

12	Fixed Capital Cost FCC	₦3053000000.00
13	Working Capital Cost WCC	₦6105000000.00

PROFIT ANALYSIS

TABLE 4.13 Result of Cost Estimation of Equipment

SERIAL NO.	EQUIPMENT	COST OF EQUIPMENT (₦)
1	Annual Sales of Product	₦17480000000.00
2	Profit Before Tax	₦10860000000.00
3	Profit After Tax	₦4696000000.00
4	Net Income	₦5280000000.00
5	Pay Back Period	1.85 yrs
6	Rate of Return	60.039 %

4.2 DISCUSSION OF RESULTS

The results presented above gives the summary of the material balance calculation, the summary of the equipment specification and a detail summary of the cost analysis. The table showing the sizes or specification of the different equipment used in the process is been obtained from their calculated values of the material and energy balances. The results of the sizes can then be used to estimate the diameter and heights of the different equipment to give a detail dimensioning of the equipments.

The results obtained for the summary of material balances across each equipment was presented in both mass flow and molar flow. It was seen that there is a net flow balance between the input material into a unit and the output of the unit both in the molar and mass flow units. It is only in the reactor where a chemical reaction takes place that there will be a change between the molar flow and mass flow of components entering or leaving a unit because of the differences in the molecular weights. Therefore across each unit, the total input is equal to the total output.

The results of the material balance were used to calculate the amount of heat leaving or entering a unit. These will aid in the selection of the particular type of material to be used during construction. The capacity of the different type of equipment used was also determined from the amount of material entering the unit. This capacity was used to dimension the equipments which enabled us to know the actual dimension of the equipment. Different equipment has different sizes depending on the amount of material entering the unit, for example the capacity of the saturator tank was calculated to be 33.48 m^3 while the capacity of mixer is 40 m^3. The determination of the sizes will assist in estimation of the cost of equipment which invariably will affect the cost analysis of the process plant. The equipments were properly dimensioned so as to achieve a better economy analysis determined from the purchased cost of equipment.

The results of the economic analysis were obtained from the standard cost index used in carrying out the economic analysis of a process plant. From the results obtained from the summary of economic analysis, it can be seen that different analysis were made which include: Cost estimation of equipment, Estimation of operating cost and finally profit analysis as not left out. The results shows that the plant will be very profitable with a profit after tax of ₦4.696 billion, pay back period of 1.85 yrs and rate of return of 60.039 %. It can be recommended for worth investing on because of the high rate of return and the profit involved. Based on this fact from the economic analysis, these plant can be recommended for worth investing on because of its high rate of return and the profitability of the plant as making profit is dream of all process designer.

CHAPTER FIVE

5.0 CONCLUSION AND RECOMMENDATIONS

5.1 CONCLUSION

The major objective of this design is to develop a cost effective, environmentally friendly process that takes into consideration the technological development of a viable economy and market structure. Hence a preliminary design of a plant to produce sodium hydroxide from sodium chloride has so far been designed. Based on the design, the basic equipment were properly sized and specified after carrying out detail material and energy balances around each of the units and hence the overall plant. Safety and effective waste disposal measures were considered and thus the plant can be said to be environmentally acceptable. From the economic analysis as calculated above, the equipment cost of the plant is 61.05 million naira. The profit after tax was found to be 4.696 billion naira per annum, with pay back period of approximately two years. Hence the plant designed so far can be said to be economically viable with rate of return on investment (ROI) of 60.039%.

5.2.0 RECOMMENDATION

For any intending industrialist or government the following recommendations are of paramount importance.

1. The safety of life and property should of paramount in the design implementation.
2. Experience mechanical engineer or contractor should be consulted for the construction.
3. Well trained personnel should keep proper statistic of technological parameters on order to prevent any unwanted accident.
4. For the operational services, well trained personnel should be employed.
5. The industrialist should ensure that a good maintenance culture is put in place.
6. The design should be properly supervised by personnel of adequate technical know how.
7. Expansion of the plant from small scale production to large scale production should not be delayed since there is high demand for this product.

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APPENDIX

A. SUMMARY OF THE ENERGY BALANCE CALCULATION

1.0 ENERGY BALANCE OF THE PRECIPITATOR

Total Enthalpy of Species in the Inlet Streams = $-7.474 * 10^6$ kJ/hr

Total Enthalpy of Species in the Outlet Streams = $7.105 * 10^6$ kJ/hr

Total Heat Load = $1.458 * 10^7$ kJ/hr

2.0 ENERGY BALANCE OF THE ION EXCHANGE PURIFIER

Total Enthalpy of Species in the Inlet Streams = $-9.606 * 10^6$ kJ/hr

Total Enthalpy of Species in the Outlet Streams = $-5.185 * 10^6$ kJ/hr

Total Heat Load = $4.42 * 10^6$ kJ/hr

3.0 ENERGY BALANCE OF THE SECOND MIXER

Total Enthalpy of Species in the Inlet Streams = $-5.751 * 10^6$ kJ/hr

Total Enthalpy of Species in the Outlet Streams = $-1.052 * 10^6$ kJ/hr

Total Heat Load = $-4.768 * 10^6$ kJ/hr

4.0 ENERGY BALANCE OF THE ELECTROLYTIC CELL

Total Enthalpy of Species in the Inlet Streams = $-6.971 * 10^6$ kJ/hr

Total Enthalpy of Species in the Outlet Streams = $-8.78 * 10^6$ kJ/hr

Total Heat Load = $-1.809 * 10^6$ kJ/hr

5.0 ENERGY BALANCE OF THE AMALGAM DECOMPOSER

Total Enthalpy of Species in the Inlet Streams = $3.028 * 10^{10}$ kJ/hr

Total Enthalpy of Species in the Outlet Streams = $2.461 * 10^{10}$ kJ/hr

Total Heat Load = $-5.662 * 10^9$ kJ/hr

B. SUMMARY OF THE EQUIPMENT SPECIFICATION

1.0 DESIGN OF MIXER

Total Volume = 40.031 m^3

Diameter = 12.164 ft

Height = 14.656 ft

2.0 DESIGN OF PRECIPITATOR

Total Volume = 81.454 m^3

Diameter = 15.414 ft

Height = 18.572 ft

3.0 DESIGN OF FILTER TANK

Total Volume = 56.507 m^3

Diameter = 13.646 ft

Height = 16.44 ft

4.0 DESIGN OF EXCHANGER

Total Volume = 23.835 m^3

Diameter = 10.234 ft

Height = 12.33 ft

5.0 DESIGN OF SECOND MIXER

Total Volume = 47.688 m^3

Diameter = 12.895 ft

Height = 15.536 ft

6.0 DESIGN OF ELECTROLYTIC CELL

Total Volume = 33.668 m^3

Diameter = 11.482 ft

Height = 13.834 ft

7.0 DESIGN OF DECOMPOSER

Total Volume = 37.059 m^3

Diameter = 11.855 ft

Height = 14.284 ft

8.0 DESIGN OF SEPARATOR

Total Volume = 25.235 m^3

Diameter = 10.43 ft

Height = 12.566 ft

9.0 DESIGN OF NaOH STORAGE TANK

Total Volume = 7.042 m^3

Diameter = 6.816 ft

Height = 8.212 ft