

IMPLEMENTATION & VALIDATION OF ZERO LIQUID DISCHARGE SYSTEM

*Submitted in partial fulfillment of the requirements for the award of
Bachelor of Technology degree in Chemical Engineering*

by

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SATHYABAMA

**INSTITUTE OF SCIENCE AND TECHNOLOGY
(DEEMED TO BE UNIVERSITY)**

Accredited with Grade "A" by NAAC

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BONAFIDE CERTIFICATE

This is to certify that this Project Report is the bonafide work of **AJITH KUMAR G S (Reg.No.38690001)** and **KUMARAN T (Reg.No. 38690014)** who carried out this Project entitled **IMPLEMENTATION & VALIDATION OF ZERO LIQUID DISCHARGE SYSTEM** under my supervision from June 2021 to October 2021.

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ABSTRACT

Zero liquid discharge (ZLD) refers to a treatment process in which the plant discharges no liquid effluent into surface waters, in effect completely eliminating the environmental pollution associated with treatment. Apart from this benefit, a ZLD process also makes effective use of wastewater treatment, recycling, and reuse, thereby contributing to water conservation through reduced intake of fresh water. This chapter discusses the challenges of ZLD in a variety of process plants, such as bulk drug manufacturers, textile factories, and chemical and fertilizer plants. Creating zero discharge solutions can be accomplished by concentrating the effluent using various techniques, including membrane-based and multiple effect evaporation-based systems, and recovery and recycling of water.

DECLARATION

We, **AJITH KUMAR. G.S (Reg.No.38690001)** and **KUMARAN T (Reg.No.38690006)**, hereby declare that this Project Report entitled **IMPLEMENTATION & VALIDATION OF ZERO LIQUID DISCHARGE SYSTEM** done by us under the guidance of **Professor Dr. S. SATHISH, M.E., Ph.D** at **Sathyabama Institute of Science and Technology** is submitted in partial fulfillment of the requirements for the award of Bachelor of Technology degree in Chemical Engineering.

DATE:

PLACE:

SIGNATURE OF THE CANDIDATES

1. AJITH KUMAR G S

2. KUMARAN T

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CHAPTER 1

1. INTRODUCTION

Zero liquid discharge (ZLD) is an advance treatment process in which all effluent is purified and recycled; therefore, leaving zero discharge at the end of treatment cycle. The methods that include ultrafiltration, reverse osmosis, evaporation, crystallization. This process that is beneficial to industrial as well as the environment because it saves money and no effluent, or discharge, is left over. ZLD systems employ the most advanced effluent treatment technologies to purify and recycle virtually all the wastewater produced.

1.1 FIELD OF APPLICATION

Most industries use water in different processes in some way. However, the wastewater effluent is widely varying in different industries. One of the main issues to designing a ZLD system is how to describe the waste stream. The contamination of the water the flow rate and the purity demand of the water are other factors that are essential when designing a ZLD-system. Because the waste stream is impossible to design a general ZLD-system. Every wastewater system is unique and must be custom made each time.

Typical waste streams that produce large volumes of wastewater include cooling tower blowdown, gas scrubber blowdown, ion-exchange regeneration effluent and rinses, plant washdown and rainwater runoff, and process wastes.

- Silicone manufacturing industries
- Pharma industry
- Paint manufacturing industries
- Food and cosmetics industries

1.2 ZLD-SYSTEM

For over 40 years vapor compression evaporation has been the most useful technology to achieve zero liquid discharge. Evaporation recovered about 96 % of an effluent as distillate for reuse. Effluent brine can then be reduced to

solids in a crystallizer device. However, evaporation alone can be an expensive when flow rates are considerable. To solve this problem is to integrate membrane processes with evaporation. These technologies are often combined to provide complete ZLD-systems. The most common membrane processes used so far are reverse osmosis (RO) and electrodialysis reversal (EDR). By combining these technologies with evaporation and crystallization ZLD-systems have become less expensive.

- pH treatment and adjustment
- Process of degasification
- Mixing and separating bed
- Oil and water separator
- Treatment of neutralization
- Oxidation process of UV and sodium hypochlorite
- Dissolved air flotation (DAF)
- Process of carbon adsorption

1.3 DESIGNING OF ZLD SYSTEM

As mentioned before, the wastewater stream is difficult when designing a ZLD-system. It is important with a realistic estimate of composition, feed chemistry and flow rate. A poorly described waste stream will likely lead to a design which is far from its optimum. The system will either be too large and expensive or too small to the required separation. The selection of the wastewater flow rates the size and therefore the initial capital cost of the ZLD-system. For plants, waste stream compositions can be measured directly, preferably on different occasions to characterize a range of compositions. Depending on the process, they may change occasionally, and it is of great importance that one has this in consideration.

1.4 ZLD TREATMENT LEVELS

- Preliminary level
- Primary level
- Secondary level
- Tertiary level

1.4.1 Preliminary Level

Purpose: The process of physical separation of large impurities like used filter cloth, plastics, wood logs, paper.

Screening: A screen with openings of uniform size is used to remove large solids. Generally maximum 10mm is used.

Sedimentation: Effluent water treatment using gravity to remove the suspended solids from effluent water.

Clarification: Used for separation of solids from fluids.

1.4.2 Primary Level

pH adjustment and Control:

- To control the pH in the treatment process to make pH neutral.
- **For acidic effluent wastes** (low pH): NaOH.
- **For alkali effluent wastes** (high pH): H_2SO_4 .

Chemical coagulation and Flocculation

Coagulation to collecting the small solid particles dispersed in a liquid into a larger mass. Chemical coagulants like $\text{Al}_2(\text{SO}_4)_3$ are added to effluent improve the attraction among fine particles so that come together and form large particles called flocs. A chemical flocculent usually a polyelectrolyte the flocculation process together particles to form larger flocs, which settle out more quickly. Flocculation by gentle mixing which causes the particles to collide.

Multiple Effect Evaporator

A multiple effect evaporator is a heat exchanger in liquid is boiled to give a vapor, so that it is in low pressure steam generator. The single effect evaporator uses rather 1 kg of steam to evaporate almost 1 kg of water. To treat an evaporator as allow pressure boiler.

1.4.3 Secondary Level

Methods: Biological process and chemical treatments are involved in this level.

Biological Unit Process

- To remove and reduce the concentration of organic and inorganic compounds.
- Biological treatment process all is based around microorganisms and bacteria.

Aerobic Processes

Aerobic process take place in the presence of oxygen to utilizes those microorganisms (aerobes), which are used in molecular and free oxygen to organic impurities i.e. convert them in to carbon dioxide, water and biomass.

1.4.4 Tertiary Level

Mechanism : Removes remaining inorganic compounds, and substances, such as the nitrogen and phosphorus. Bacteria, viruses and parasites, which are harmful to public health, are also removed at this stage.

CHAPTER 2

2. PLANT OPERATIONS

2.1 PLANT OPERATIONS FOR HPS

2.1.1 Equalizing Tank

Effluent from the collection tank comes to the equalization tank in wastewater treatment. The main function is to act as buffer. To collect the incoming raw effluent that comes at widely fluctuating rates and average flow rate. During the peak hours ETP comes at high flow rate.

2.1.2 Neutralizing Tank

A salt is compound which from the neutralization of an acid and a base. The word "neutralization" is used because the acid and base properties of H^+ and OH^- are neutralized. The reaction, H^+ and OH^- combine to form HOH or H_2O or water molecules. A neutralization is a double replacement reaction. A salt is product of an acid-base reaction.

2.1.3 Coagulation Tank

Coagulant chemicals with charges opposite those of the suspended solids are added to the water to neutralize the negative charges on non-settle able solids (such as clay and color-producing organic substances).

Once the charge is neutralized, the small suspended particles are capable of sticking together. These slightly larger particles are called micro flocs and are not visible to the naked eye.

2.1.4 Flocculation Tank

Flocculation, a gentle mixing stage, increases the particle size from submicroscopic micro floc to visible suspended particles. Micro floc particles collide, causing them to bond to produce larger, visible flocs called pin flocs. Floc size continues to build with additional collisions and interaction with added inorganic polymers (coagulant) or organic polymers.

2.1.5 Clarifier

Clarifiers are settling tanks built with mechanical means for continuous removal of solids being deposited by sedimentation. A clarifier is generally used to remove solid particulates or suspended solids from liquid for clarification and (or) thickening. Concentrated impurities, discharged from the bottom of the tank are known as sludge, while the particles that float to the surface of the liquid are called scum.

2.2 MULTIPLE EFFECT EVAPORATOR

An evaporator is essentially a heat exchanger in which a liquid is boiled to give a vapor, so that it is also simultaneously, a low pressure steam generator. The single effect evaporator uses rather more than 1 kg of steam to evaporate almost 1 kg of water.

There are three types of feeding are:

- Forward feed
- Backward feed
- Parallel feed

2.2.1 Forward Feed Arrangement

In this arrangement the feed and steam introduced in the first effect and pressure in the first effect is highest and pressure in last effect is minimum, so transfer of feed from one effect to another can be done without pump.

2.2.2 Backward Feed Arrangement

In this arrangement feed is introduced in last effect and steam is introduced in first effect. For transfer of feed, it requires pump since the flow is from low pressure to high pressure. Concentrated liquid is obtained in first effect.

2.2.3 Parallel Feed Arrangement

In this arrangement feed is introduced in intermediate effect flows in forward feed to the end of the series and is then pumped back to the first effect for final concentration. This permits the final evaporation to be done at the highest temperature.

2.3 PLANT OPERATIONS OF SPENT ACID STREAM

2.3.1 Continuous Stirred Tank Reactor

The continuous stirred-tank reactor (CSTR), also known as vat- or backmix reactor, is a common ideal reactor type in chemical engineering. A CSTR often refers to a model used to estimate the key unit operation variables when using a continuous agitated-tank reactor to reach a specified output. The behavior of a CSTR is often approximated or modeled by that of a Continuous Ideally Stirred-Tank Reactor (CSTR).

2.3.2 Agitated Thin Film Dryer

The typical Agitated Thin Film Dryer consists of a tubular heat transfer area with an external heating jacket and a fast-revolving, inner rotor with flexible or rigid wiper elements. The driving speed is adapted to the product being handled, its particular specifications and task.

2.3.3 Condensor

In systems involving heat transfer, a condenser is a device or unit used to condense a substance from its gaseous to its liquid state, by cooling it. In so doing, the latent heat is given up by the substance and transferred to the surrounding environment.

2.4 OVERALL FLOW SHEET

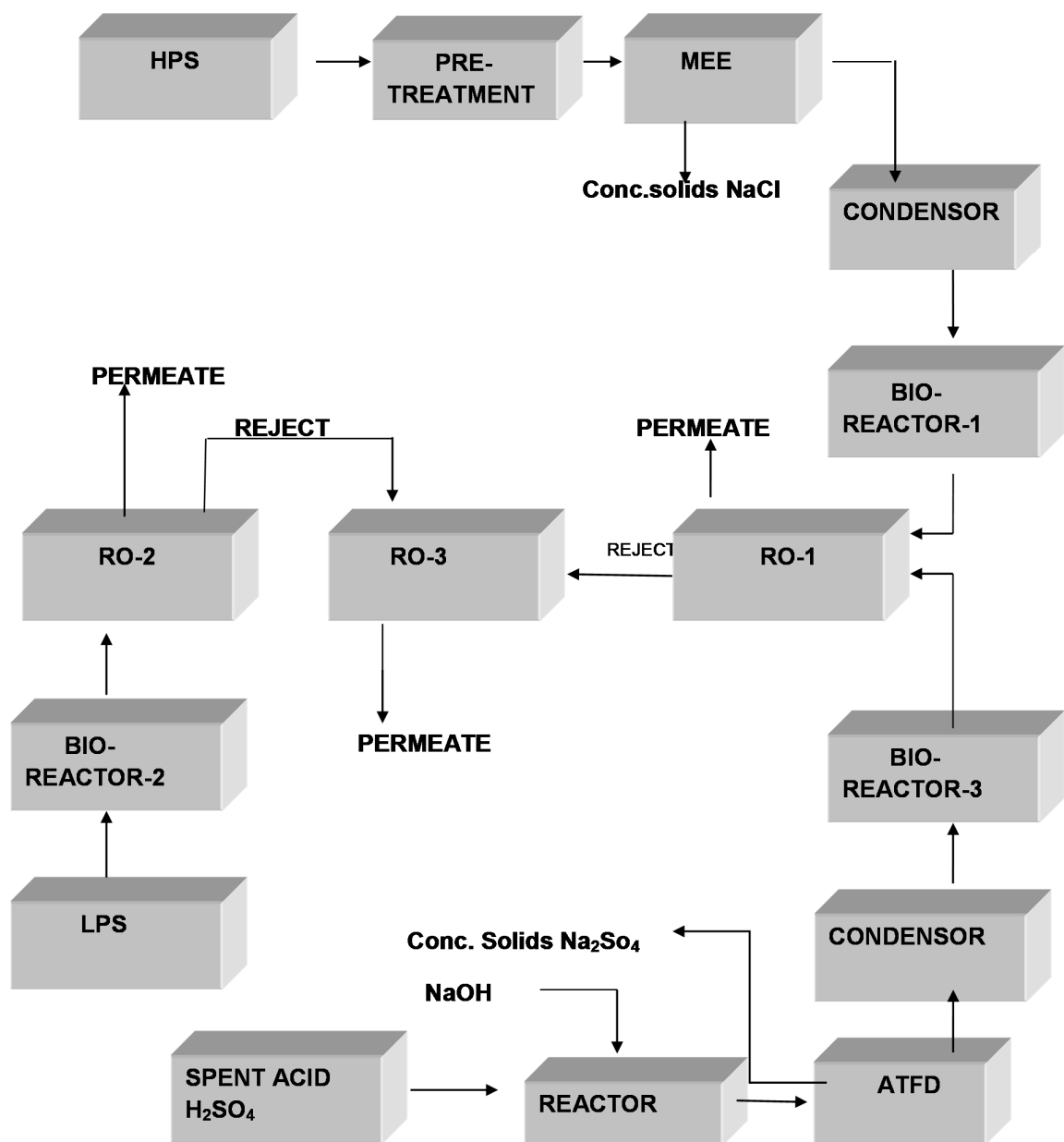


Figure 2.4: Overall Flow Sheet

2.5 PROCESS DESCRIPTION

2.5.1 Process Description For Hps

The HPS means high pollutant steam. It contains acetic acid, NaCl etc and the HPS is poured into the equalising tank to control the fluctuation of the feed and the feed is stored for 3-4 hrs to become a homogenous mixture. And then it is pumped to the neutralising tank in this tank the PH value is adjusted by adding a alkali solution called NaOH. The PH value is adjusted to 4to 7.5. and it is pumped to the coagulation and flocculation tank. the alum as the coagulant and the poly as the flocculant is added to get colloidal substances and is pumped to the clarifier. In clarifier, the chemical degradants sludge can remove, and the remaining solution is pumped to the multiple effect evaporator. The forward feed arrangement evaporator is handled in this process. The water vapour is move into the condenser and the water is pumped to the Bio-reactor in this reactor micro-organism are removed. And it is pumped into the RO plant. After the RO plant process. The water becomes to the purpose of utilities.

2.5.2 Process Description for Spent Acid Treatment

The sulphuric acid is the spent acid. The spent acid is poured into the gas lined reactor and the alkali solution is added to the spent acid to reduce the PH level of the spent acid. The solution is stirred in the reactor for 30 minutes and the temperature is maintained at 60-70°C. The PH is maintained in 7- 7.5. The solution is pumped into the Agitated thin film dryer. The water and the sodium sulphate is separated at temperature 60-105°C. The water is passed to the Bio-reactor in this process the micro-organism are killed. And the water is pumped into the RO plant and then it can be reused.

CHAPTER 3

3 MATERIAL BALANCE

3.1 *Material Balance For Hps Process*

✓ For Equalising Tank

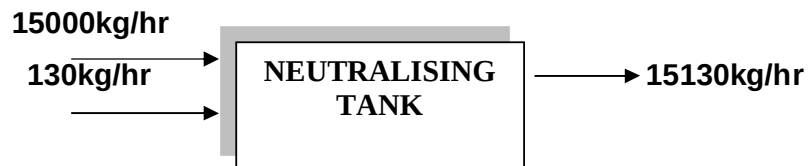


Inlet of HPS feed = 15,000kg/hr

Outlet of HPS feed = 15,000kg/hr

Inlet feed = outlet feed, it does not contain any chemical reaction. Only it can control the fluctuation.

✓ For Neutralising Tank



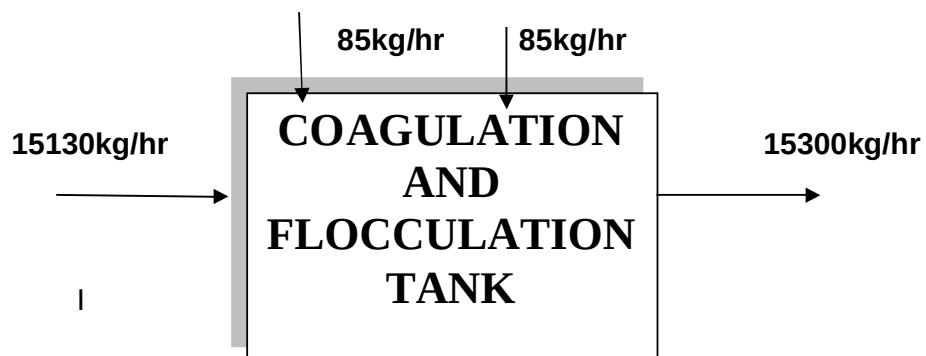
Inlet of HPS feed = 15,000kg/hr

Inlet of NaOH = 130kg/hr

Outlet of aqueous NaCl solution = 15,130kg/hr

Inlet feed = outlet feed, it can adjust the PH level.

✓ For Coagulation and Flocculation Tank



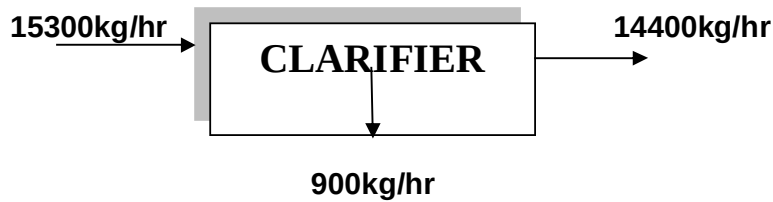
Inlet of aqueous NaCl solution = 15,130kg/hr

Inlet of Alum = 85 kg/hr

Inlet of poly = 85kg/hr

Outlet of semi solid NaCl solution = 15,300kg/hr

✓ For Clarifier



Inlet of semi solid NaCl solution = 15,300kg/hr

Outlet of sludge (chemical degradants) = 900kg/hr

Outlet of NaCl solution = 14,400kg/hr

For Multiple Effect Evaporator

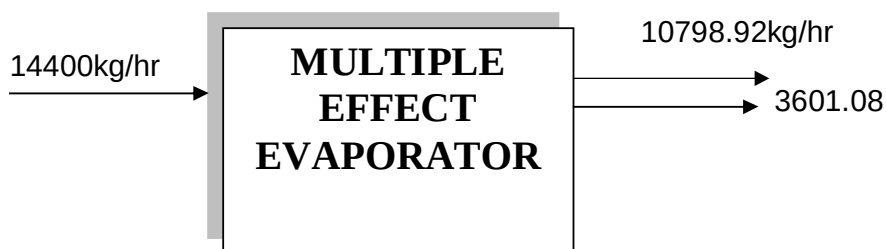


Table 3.1 : MATERIAL BALANCE FOR HPS PROCESS

CALANDRIA	INLET Kg/hr	CONC. SOLIDS Kg/hr	VAPOUR Kg/hr	OUTLET Kg/hr
1	14400	11095.2	3304.8	14400
2	11095.2	7516.8	3578.4	11095.2
3	7516.8	3601.08	3915.72	7516.8

OVERALL MATERIAL BALANCE FOR MEE

Inlet feed of NaCl solution = 14,400kg/hr

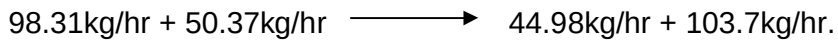
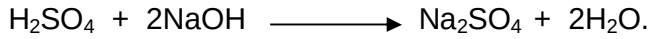
Outlet of water vapour = 10798.92kg/hr

Outlet feed of conc. Solids = 3601.08kg/hr

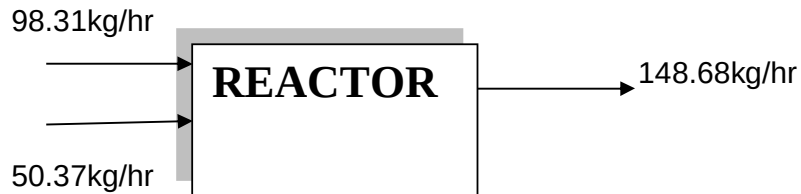
3.2 MATERIAL BALANCE FOR SPENT ACID PROCESS

OVERALL MATERIAL BALANCE FOR SPENT ACID

REACTION :



➤ For Reactor



Inlet feed of sulphuric acid = 98.31kg/hr

Inlet feed of sodium hydroxide = 50.37kg/hr

Outlet of sodium sulphate and water = 148.68kg/hr

➤ For Agitated Thin Film Dryer



Inlet feed of sodium sulphate and water = 148.68kg/hr

Outlet of sodium sulphate = 44.98kg/hr

Outlet of water vapour = 103.7kg/hr

CHAPTER 4

4 ENERGY BALANCE

4.1 ENERGY BALANCE FOR HPS PROCESS

➤ For Multiple Effect Evaporator

Forward Flow Calculations:

Mass flow rate of feed = 14,400kg/hr = 4kg/s

Feed inlet temperature = 30°C

Initial solid content = 10%

Final solid content = 40%

Saturated dry steam temperature = 2.05bar = 205kN/m²

Specific heat of feed = 4.184KJ/Kg. k

Pressure of 3rd effect evaporator = 13kN/m² = 0.13bar

From steam table at 0.13 boiling point of water = 325k

Temperature of dry saturated steam at 205kN/m² = 394 k

Temperature difference $\Sigma \Delta T = (394-325) = 69k$

Area = $A_1=A_2=A_3=65m^2$

Temperature difference across each evaporator

$\Delta T_1=18K$ $\Delta T_2=17K$ $\Delta T_3=34K$

Latent heat are given by $\lambda_0, \lambda_1, \lambda_2, \lambda_3$

For steam:

$T_0 = 394k$ $\lambda_0=2200kJ/kg$

$T_1=376k$ $\lambda_1=2249kJ/kg$

$T_2=359k$ $\lambda_2=2293kJ/kg$

$T_3=325k$ $\lambda_3=2377kJ/kg$

Values obtained from steam table

4.2 ENERGY BALANCE

CALANDRIA -1:

Mass flow rate of feed = 14400 kg/hr = 4kg/s

Enthalpy of feed = 89.494

Steam rate $D_0 = 1.5 \text{ kg/s}$

Latent heat of vaporisation $\lambda_0 = 2200 \text{ kJ/kg}$

Outlet of vapour $D_1 = 0.918 \text{ kg / s}$

Latent heat of vaporisation $\lambda_1 = 2249 \text{ kJ/kg}$

Outlet of Concentrated solids of Nacl = 3.082 kg / s

Enthalpy of concentrated solids Nacl = 517

$$M_f h_f + D_0 \lambda_0 = D_1 \lambda_1 + M_{f1} h_{f1}$$

$$4(89.494) + 1.5(2200) = 0.918(2249) + 3.082(517)$$

$$357.976 + 3300 = 2064.582 + 1593.394$$

$$3657.439 = 3657.976$$

$$Q_1 = Q_2$$

CALANDRIA - 2:

Inlet of vapour $D_1 = 0.918 \text{ kg / s}$

Latent heat of vaporisation $\lambda_1 = 2249 \text{ kJ/kg}$

Inlet of Concentrated solids of Nacl = 3.082 kg / s

Enthalpy of concentrated solids Nacl = 517

Outlet of vapour $D_2 = 0.994 \text{ kg/s}$

Latent heat of vaporisation $\lambda_2 = 2293 \text{ kJ/kg}$

Outlet of Concentrated solids of Nacl = 2.088 kg / s

Enthalpy of concentrated solids Nacl = 517

$$D_1 \lambda_1 + M_{f1} h_{f1} = M_{ph2} h_{f2} + D_2 \lambda_2$$

$$(3.081)(517) + (0.918)(2249) = (0.994)(2293) + (2.088)(517)$$

$$1592.877 + 2064.582 = 2578.436 + 1079.493$$

$$3657.459 = 3657.932$$

$$Q_2 = Q_3$$

CALANDRIA - 3:

Inlet of vapour $D_2 = 0.994 \text{ kg/s}$

Latent heat of vaporisation $\lambda_2 = 2293 \text{ kJ/kg}$

Inlet of Concentrated solids of Nacl = 2.088 kg / s

Enthalpy of concentrated solids Nacl = 517

Outlet of vapour $D_3 = 1.0877 \text{ kg/s}$

Latent heat of vaporisation $\lambda_3 = 2377 \text{ kJ/kg}$

Outlet of Concentrated solids of Nacl = 1.003 kg / s

Enthalpy of concentrated solids Nacl = 517

$$M_{ph} h_{f2} + D_2 \lambda_2 + D_3 \lambda_3 + M_{f3} h_{f3}$$

$$(0.994)(2293) + (2.088)(517) = (1.0877)(2377) + (1.0003)(517)$$

$$2578.436 + 1079.496 = 3140.78 + 517.1551$$

$$3657.936 = 3657.936$$

$$Q_3 = Q_4$$

Steam Economy = kg vaporized / kg steam input

$$E = 3/1.5$$

$$= 2 \text{ Kg/s}$$

The heat transfer coefficient are then:

$$U_1 = D_0 \lambda_0 / A_1 \Delta T_1$$

$$= (1.5 \times 2200) / (65 \times 18)$$

$$= 2.82 \text{ kJ/m}^2 \text{s K}$$

$$U_2 = D_1 \lambda_1 / A_2 \Delta T_2$$

$$= (0.918 \times 2249) / (65 \times 17)$$

$$= 1.86 \text{ kJ/m}^2 \text{s K}$$

$$U_3 = D_3 \lambda_3 / A_3 \Delta T_3$$

$$= (0.994 \times 2293) / (65 \times 34)$$

$$= 1.03 \text{ kJ/m}^2 \text{s K}$$

4.3 ENERGY BALANCE FOR REACTOR

INPUT

Inlet of sulphuric acid $M_1 = 98.31\text{kg/hr.}$

Specific heat of sulphuric acid $C_p = 1.34\text{kJ/kg. k}$

Inlet of sodium hydroxide $M_2 = 58.37\text{kg/hr.}$

Specific heat of sodium hydroxide $C_p = 3.471\text{kJ/kg. k}$

Inlet temperature = 30°C

$$Q_1 = M_1 C_p \Delta T + M_2 C_p \Delta T$$

$$Q_1 = 98.31(1.34) (303-298) + 58.37(3.471) (303-298)$$

$$Q_1 = 658.677 + 1013.01135$$

$$Q_1 = 1671.68835\text{kJ/hr.}$$

OUTPUT

Outlet of sodium sulphate $M_3 = 44.98\text{kg/hr.}$

Specific heat of sodium sulphate = 3.77kJ/kg k

Outlet of water $M_4 = 103.7\text{kg/hr.}$

Specific heat of water $C_p = 4.18\text{kJ/kg. k}$

Outlet temperature = 60°C

Inlet temperature = 30°C

$$Q_2 = M_2 C_p \Delta T + M_3 C_p \Delta T$$

$$Q_2 = 44.98(3.77) (333-303) + 103.7(4.18) (333-303)$$

$$Q_2 = 4277.598 + 13003.98$$

$$Q_2 = 17281.578\text{kJ/hr.}$$

$$\begin{aligned}\text{Heat load} &= \text{Heat input} + \text{Heat of reaction} = \text{Heat output} \\ &= Q_1 + Q_R = Q_2 \\ &= 1671.68835 + 15609.8896 = 17281.578 \text{ KJ/hr.}\end{aligned}$$

4.4 AGITATED THIN FILM DRYER

Mass flow rate of feed	= 148.68 Kg/hr.
Specific heat capacity of feed	= 1 kcal/°C
Inlet temperature of feed	= 60°C
Outlet temperature of feed	= 100°C
Steam flow rate	= 103.7 kg/hr
Latent heat of vaporization	= 540 kcal/kg

CALCULATIONS

$$\begin{aligned}Q &= mc_p \Delta T + m\lambda \\ Q &= 148.68 \times 1(100-60) + 103.7 \times 540 \\ Q &= 148.68 \times 40 + 55998 \\ Q &= 5947.2 + 55998 \\ Q &= 61945.2 \text{ kcal/hr}\end{aligned}$$

Steam consumption in ATFD:

$$\begin{aligned}\text{Steam consumption} &= Q/\lambda \\ &= 61945.2/540 \\ &= 114.7 \text{ kg/hr}\end{aligned}$$

Steam economy:

$$\begin{aligned}\text{Steam economy} &= \text{water evaporated} / \text{steam consumed} \\ &= 103.7 / 114.7 \\ &= 0.904 \text{ kg/hr}\end{aligned}$$

4.5 CONDENSER

Shell side = water vapour

Tube side = cooling water

Temp. For shell = 103/80

Temp. For tube = 30/36

Mass of the feed, m = 103.7 kg/hr

Latent heat of condensate = 540 kcal/kg

Condensed product, $m^{\circ} = 103 \text{ kg/hr}$

Specific heat, $C_p = 1 \text{ Kcal/kg } ^{\circ}\text{C}$

For shell side:

$$Q = mc_p \Delta T + m^{\circ} \lambda$$

$$Q = 103.7 \times 1 \times (103 - 80) + 103.7 \times 540$$

$$Q = 59740 \text{ kcal/hr}$$

For Tube Side:

Specific heat of water: 4.189 kJ/hr

$$Q = M C_p \Delta T$$

$$M = Q / C_p \Delta T$$

$$M = 249952.16 / (4.189 \times (36 - 30))$$

$$M = 9944 \text{ Kg/hr.}$$

CHAPTER 5

5. DESIGN FOR EQUIPMENT

5.1 DESIGN OF MULTIPLE EFFECT EVAPORATOR

5.1.1 Multiple Effect Evaporator

Forward Flow Calculations:

Mass flow rate of feed = 14,400kg/hr = 4kg/s

Feed inlet temperature = 30°C

Initial solid content = 10%

Final solid content = 40%

Steam input = 1.5kg/s

Saturated dry steam temperature = 2.05bar = 205KN/m²

Specific heat of feed = 4.184KJ/Kg. k

Pressure of 3rd effect evaporator = 13KN/m² = 0.13bar

From steam table at 0.13 boiling point of water = 325k

Temperature of dry saturated steam at 205KN/m² = 394 k

Temperature difference $\Sigma\Delta T = (394-325) = 69k$

Area = $A_1=A_2=A_3=65m^2$

Temperature difference across each evaporator

$\Delta T_1=18K$ $\Delta T_2=17K$ $\Delta T_3=34K$

Latent heat are given by $\lambda_0, \lambda_1, \lambda_2, \lambda_3$

For Steam :

$$\lambda_0 = 394 \text{ k} \quad \lambda_0 = 2200 \text{ kJ/kg}$$

$$\lambda_1 = 376 \text{ k} \quad \lambda_1 = 2249 \text{ kJ/kg}$$

$$\lambda_2 = 359 \text{ k} \quad \lambda_2 = 2293 \text{ kJ/kg}$$

$$\lambda_3 = 325 \text{ k} \quad \lambda_3 = 2377 \text{ kJ/kg}$$

Values obtained from steam table

ENERGY CALCULATIONS

Effect 1:

$$D_0 \lambda_0 = m_f c_p (T_1 - T_f) + D_1 \lambda_1$$

$$2200 D_0 = 4 \times 4.18 (376 - 303) + 2249 D_1$$

$$D_0 = 0.5548 + 1.022 D_1 \longrightarrow 1$$

Effect 2:

$$D_2 \lambda_2 = d_1 \lambda_1 + (m_f - d_1) c_p (T_1 - T_2)$$

$$D_2 (2293) = 2249 D_1 + (4 - D_1) 4.184 (376 - 354)$$

$$D_2 (2293) = 2249 D_1 + (4 - D_1) 71.128$$

$$D_2 = 0.948 D_1 + 0.124 \longrightarrow 2$$

Effect 3 :

$$D_3 \lambda_3 = D_2 \lambda_2 + (m_f - D_1 - D_2) 4.184 (359 - 325)$$

$$D_3 (2377) = 2293 D_2 + (4 - D_1 - D_2) 142.256$$

$$D_3 = 0.0965 D_2 + 0.0598 (4 - D_1 - D_2)$$

$$= 0.0965 D_2 + 0.240 - 0.0598 D_1 - 0.0598 D_2$$

$$D_3 = 0.905 D_2 - 0.0598 D_1 + 0.240 \longrightarrow 3$$

$$D_1 + D_2 + D_3 = 3 \text{ kg/s}$$

Substituting the values of D_2 and D_3

$$D_3 + 0.948 D_1 + 0.124 + 0.80 D_1 + 0.352 = 3$$

$$2.748 D_1 = 2.524$$

$$\mathbf{D_1 = 0.918 \text{ kg/s}}$$

$$D_2 = 0.948 D_1 + 0.124$$

$$= 0.994 \text{ kg/s}$$

$$D_3 = 0.9052 (0.9048 D_1 + 0.124) - 0.05981 D_1 + 0.240$$

$$= 0.798 D_1 + 0.352$$

$$D_3 = 1.0877 \text{ kg/s}$$

$$D_0 = 0.5548 + 1.022 D_1$$

$$D_0 = 0.5548 + 1.022 \times 0.918$$

$$D_0 = 1.50$$

Steam Economy = kg vaporized / kg steam input

$$E = 3/1.5$$

$$= 2$$

The heat transfer coefficient is then:

$$U_1 = D_0 \lambda_0 / A_1 \Delta T_1$$

$$= (1.5 \times 2200) / (65 \times 18)$$

$$= 2.82 \text{ kJ/m}^2\text{s K}$$

$$U_2 = D_1 \lambda_1 / A_2 \Delta T_2$$

$$= (0.918 \times 2249) / (65 \times 17)$$

$$= 1.86 \text{ kJ/m}^2\text{s K}$$

$$U_3 = D_3 \lambda_3 / A_3 \Delta T_3$$

$$= (0.994 \times 2293) / (65 \times 34)$$

$$= 1.03 \text{ kJ/m}^2\text{s K}$$

Let us select 50mm outer diameter, 40mm inner diameter tubes each of 2m arranged on 75mm square pitch.

N = Number of tubes in the pitch

$$\text{So, } \pi D_0 L N = 65 \text{ m}^2$$

$$N = 65 / \pi D_0 L$$

$$N = 208$$

Down take area, A_D :

$$A_D = \pi/4 D_i^2 N \times 0.75$$

$$= 0.1960 \text{ m}^2$$

Annular area

$$= NP_T^2$$

$$= 208 \times 0.075^2$$

$$= 1.17 \text{m}^2$$

Evaporator diameter:

$$\text{Down take area} = 0.1960 \text{m}^2$$

$$\text{Annular area} = 1.17 \text{m}^2$$

$$\text{Total area} = 1.366 \text{m}^2$$

$$\text{Area of body} = 1.1(1.366) = 1.5026 \text{m}^2$$

$$\text{Diameter} = 0.7513 \text{m}$$

$$\text{Height} = 3L = 6 \text{m}$$

5.2 DESIGN OF CONDENSOR

TABLE 5.2: DESIGN OF CONDENSOR

	Unit	Shell Side	Tube Side
Fluid flow rate	Kg/hr	103.7	
Fluid temperature	°C	103/60	36/30
Viscosity	Cp	0.8	0.69
Density	Kg/m ³	960	993
Thermal conductivity	W/m°C	0.68	0.62
Fouling coefficient	W/m ² °C	5000	7400
Pressure drop allowable	Bar	1.0	0.5

Shell side = water vapour

Tube side = cooling water

Temp. For shell = 103/80

Temp. For tube = 30/36

Mass of the feed, $m = 103.7 \text{ kg/hr}$

Latent heat of condensate, $\lambda = 540 \text{ kcal/kg}$

Condensed product, $m^\circ = 103 \text{ kg/hr}$

Specific heat, $C_p = 1 \text{ Kcal/kg } ^\circ\text{C}$

$$Q = mc_p \Delta T + m^\circ \lambda$$

$$Q = 103.7 \times 1 \times (103 - 80) + 103.7 \times 540$$

$$Q = 59740 \text{ kcal/hr}$$

$$Q = 2449952.16 \text{ kJ/hr}$$

For tube side:

Specific heat of water: 4.189 kJ/hr

$$Q = M C_p \Delta T$$

$$M = Q / C_p \Delta T$$

$$M = 249952.16 / (4.189 \times (36 - 30))$$

$$M = 9944 \text{ Kg/hr}$$

5.2.1 LMTD

Vapor : 103°C to 80°C

Cooling water: 36°C to 30°C

$$\Delta T_{lm} = (T_2 - t_1) - (T_1 - t_2) / \ln((T_2 - t_1) / (T_1 - t_2))$$

$$\Delta T_{lm} = (103 - 30) - (80 - 36) / \ln((103 - 30) / (80 - 36))$$

$$\Delta T_{lm} = 29 / 0.506$$

$$\Delta T_{lm} = 57.3^\circ\text{C}$$

For, overall heat transfer coefficient can be assumed

$$U_0 = 300 \text{ W/m}^2\text{ } ^\circ\text{C}$$

5.2.2 Trial Area

$$A_0 = Q / U_0 \Delta T_m$$

$$A_0 = 2449952.16 / 300 \times 57.3^\circ\text{C}$$

$$A_0 = 4 \text{ m}^2$$

5.2.3 Number of Tubes Required

Tube dimensions = 20mm OD x 18mm ID x 4.88m long

Tube pitch = 25mm square pitch

Effective tube length = tube length – (2 x tube thickness)

$$= 4.88 - (2 \times 0.02)$$

$$= 4.83 \text{ m}$$

No of tubes = trial area / surface area

$$= \text{Area} / \pi \times \text{OD} \times L$$

$$= 4 / 3.14 \times 0.02 \times 4.83$$

$$= 13 \text{ tubes}$$

5.2.4 Film Coefficient – Shell Side

$$h_c = 0.95 \times K_L \times ((\rho_L \times (\rho_L - \rho_G) \times g) / (\mu_L \times \tau_L))^{(1/3)} \times (Nr)^{-(1/6)}$$

$$\tau_L = W_C / L \times (N_T)^{2/3}$$

$$\tau_L = 1.56 / 4.88 \times 5.528$$

$$\tau_L = 1.56 / 17.857$$

$$\tau_L = 0.057 \text{ Kg/ms}$$

$$N_r = (2/3) \times (L_b/P_T)$$

$$N_T = C(L_b/P_T)^2$$

$$13 = 0.86(L_b/0.025)^2$$

$$(13/0.86) = (L_b/0.025)^2$$

$$8.139^{1/2} = L_b/0.025$$

$$2.853 = L_b/0.025$$

$$L_b = 2.853 \times 0.025$$

$$L_b = 0.066$$

$$N_r = (2/3) \times (L_b/P_T)$$

$$N_r = 0.66 \times 2.64$$

$$N_r = 1.759$$

$$h_c = 0.95 \times K_L \times ((\rho_L \times (\rho_L - \rho_G) \times g) / (\mu_L \times T_L))^{(1/3)} \times (N_r)^{-(1/6)}$$

$$h_c = 0.95 \times 0.68 \times ((960 \times (960 - 4) \times 9.81) / (0.8 \times 0.057))^{(1/3)} \times (1.759)^{-(1/6)}$$

$$h_c = 0.95 \times 0.68 \times 505.74 \times 0.910$$

$$h_c = 340 \text{ W/m}^2\text{°C}$$

5.2.5 Film Coefficient-Tube Side

$$h_i D_i / k = 0.023 \times \text{Re}^{0.8} \times \text{Pr}^{0.33} \times (\mu / \mu_w)^{0.14}$$

5.2.6 Flow Area Per Pass

$$A = \pi/4 \times D_i^2 \times (N_T/N_P)$$

$$A = \pi/4 \times 0.018^2 \times 6.5$$

$$A = 0.00330 \text{ m}^2$$

MASS VELOCITY:

$$G_t = M_c / A_t$$

$$G_t = 2.76 / 0.00330$$

$$G_t = 840 \text{ kg/m}^2\text{s}$$

REYNOLDS NUMBER:

$$Re = D_i G_t / \mu$$

$$Re = 0.018 \times 840 / 0.69 \times 10^{-3}$$

$$Re = 21913.04$$

PRANDTL NUMBER:

$$Pr = C_p \mu / k$$

$$Pr = 4.184 \times 0.69 / 0.62$$

$$Pr = 4.67$$

$$\mu / \mu_w = 1$$

$$h_i D_i / k = 0.023 \times Re^{0.8} \times Pr^{0.33} \times (\mu / \mu_w)^{0.14}$$

$$h_i = 0.023 \times (21913.04)^{0.8} \times (4.67)^{0.33} \times (1)^{0.14} \times (0.62 / 0.018)$$

$$h_i = 0.023 \times 2968 \times 1.66 \times 1 \times 34.44$$

$$h_i = 39035.6 \text{ W/m}^2\text{°C}$$

$$h_{i0} = h_i \times (D_i / D_0)$$

$$h_{i0} = 39035.6 \times (18 / 20)$$

$$h_{i0} = 35132.04 \text{ W/m}^2\text{°C}$$

TUBE WALL RESISTANCE:

$$K_w = 50 \text{ W/m}^2\text{ }^{\circ}\text{C}$$

$$R_w = (D_o \ln(D_o/D_i))/2 \times K_w$$

$$R_w = (0.02 \ln(0.02/0.018))/2 \times 50$$

$$R_w = 2.11 \times 10^{-5} \text{ m}^2\text{ }^{\circ}\text{C/W}$$

FOULING RESISTANCE:

$$R_{d0} = 1/5000 \text{ m}^2\text{ }^{\circ}\text{C/W}$$

$$R_{d0} = 2 \times 10^{-4} \text{ m}^2\text{ }^{\circ}\text{C/W}$$

$$R_{di} = 1/7400 \text{ m}^2\text{ }^{\circ}\text{C/W}$$

$$R_{di0} = R_{di} \times (OD/ID)$$

$$= (1/7400) \times 0.02/0.018$$

$$R_{di0} = 1.5 \times 10^{-4} \text{ m}^2\text{ }^{\circ}\text{C/W}$$

OVERALL HEAT TRANSFER COEFFICIENT:

$$U_0 = (1/h_c) + (1/h_{i0}) + R_w + R_{di0} + R_{d0})^{-1}$$

$$U_0 = (1/340) + (1/35132.04) + (2.11 \times 10^{-5}) + (1.5 \times 10^{-4}) + (2 \times 10^{-4})^{-1}$$

$$U_0 = 299.33 \text{ W/m}^2\text{ }^{\circ}\text{C}$$

HEAT TRANSFER AREA REQUIRED:

$$A_t = Q/U_0 \Delta T_M$$

$$= 69.31/299.33 \times 57.6$$

$$= 4.02 \text{ m}^2$$

AREA AVAILABLE:

$$A_a = \pi \times O \times L_e \times N_T$$

$$A_a = 3.14 \times 0.02 \times 4.83 \times 13.5$$

$$A_a = 4.4 \text{ m}^2$$

EXCESS AREA:

$$\% \text{ Excess area} = ((\text{Area available} - \text{Area required}) / (\text{Area required})) \times 100$$

$$= ((4.4 - 4.09) / (4.09)) \times 100$$

$$= 7.5\%$$

PRESSURE DROP- (TUBE SIDE):

$$\Delta P_t = \{((f G_t^2 L N_p) / (2 \times 10^6 \times D_e S \times \phi T)) + 2.5 N_p v^2 s\}$$

$$f = 0.72 (\text{Re})^{-0.33}$$

$$f = 0.72 (21913)^{-0.33}$$

$$f = 0.0266$$

$$v = G_t / p$$

$$v = 840 / 993$$

$$v = 0.84 \text{ m/s}$$

$$\Delta P_t = \{((0.0266 \times 840^2 \times 4.88 \times 2) / (2 \times 10^6 \times 0.018 \times 1 \times 0.994)) + 2.5 \times 2 \times 0.7^2 \times 1\}$$

$$\Delta P_t = (183185.04 / 35784) + 2.45$$

$$\Delta P_t = 5.119 + 2.45$$

$$\Delta P_t = 7.569 \text{ kPa}$$

$$\Delta P_t = 0.07569 \text{ bar}$$

PRESSURE DROP- (SHELL SIDE):

$$\Delta P_s = (fG_t^2 L N_p) / (2 \times 10^6 \times D_e \times S \times 0.994)$$

$$f = 1.87(\text{Re})^{-0.2}$$

$$f = 1.87(21913)^{-0.2}$$

$$f = 0.25$$

$$N_b = L_e / D_s$$

$$N_b = 4.83 / 0.591$$

$$N_b = 8$$

$$D_e = 4[(P_T^2 - \pi/4(D_o^2)) / \pi D_o]$$

$$D_e = 4[0.025^2 - \pi/4(0.02)^2] / \pi(0.02)$$

$$D_e = 0.0198 \text{ m}$$

$$\Delta P_s = (0.25 \times 22.29^2 \times 0.591 \times (8+1)) / (2 \times 10^6 \times 0.0198 \times 1 \times 0.994)$$

$$\Delta P_s = 660.67 / 39362.4$$

$$\Delta P_s = 0.016 \text{ kPa}$$

$$\Delta P_s = 0.00016 \text{ bar}$$

5.3 DESIGN OF REACTOR

Inlet feed of sulphuric acid = 98.31 kg/hr

Inlet feed of sodium hydroxide = 50.37 kg/hr

Outlet of sodium sulphate and water = 148.68 kg/hr

INPUT:

Inlet of sulphuric acid $M_1 = 98.31 \text{ kg/hr}$

Specific heat of sulphuric acid $C_p = 1.34 \text{ kJ/kgK}$

Inlet of sodium hydroxide $M_2 = 58.37 \text{ kg/hr}$

Specific heat of sodium hydroxide $C_p = 3.471 \text{ kJ/kgK}$

Inlet temperature = 30°C

$$Q_1 = M_1 C_p \Delta T + M_2 C_p \Delta T$$

$$Q_1 = 98.31(1.34) (303-298) + 58.37(3.471)(303-298)$$

$$Q_1 = 658.677 + 1013.01135$$

$$Q_1 = 1671.68835 \text{ kJ/hr.}$$

OUTPUT:

Outlet of sodium sulphate $M_3 = 44.98 \text{ kg/hr.}$

Specific heat of sodium sulphate = 3.77 kJ/kgK

Outlet of water $M_4 = 103.7 \text{ kg/hr}$

Specific heat of water $C_p = 4.18 \text{ kJ/kgK}$

Outlet temperature = 60°C

Inlet temperature = 30°C

$$Q_2 = M_2 C_p \Delta T + M_3 C_p \Delta T$$

$$Q_2 = 44.98(3.77) (333-303) + 103.7(4.18)(333-303)$$

$$Q_2 = 4277.598 + 13003.98$$

$$Q_2 = 17281.578 \text{ kJ/hr}$$

Heat load = Heat input + Heat of reaction = Heat output

$$= Q_1 + Q_R = Q_2$$

$$= 1671.68835 + 15609.8896 = 17281.578 \text{ kJ/hr}$$

5.3.1 Volume of Reactor

Rate of reaction = 15609.8896kJ/hr

$$=4.336 \text{ kJ/s}$$

$$F(x) = rV_b$$

The reaction conversion is 40%

$$V_b = (148.68 \times 0.40)/4.336$$

$$V_b = 11.157\text{m}^3$$

$$\text{Volume} = \pi r^2 h$$

$$11.157 = \pi r^2 h$$

$$H/D \text{ ratio} = 1.6$$

$$H = 1.6D$$

$$11.157 = \pi r^2 (1.6D)$$

$$2.2207 = r^3$$

$$(2.2207)^{1/3} = r$$

$$r = 1.30\text{m}$$

Diameter of the Reactor:

$$D = 2r$$

$$D = 2 \times 1.30$$

$$D = 2.60\text{m}$$

Height of the reactor:

$$H = 1.6 \times 2.60$$

$$H = 4.16\text{m}$$

Diameter of the agitator:

$$Da/D = 1/3$$

$$Da = 0.333 \times 2.60$$

$$Da = 0.858\text{m}$$

Da = Diameter of the agitator

E = Distance of the blade from the bottom of the reactor

$$E/D = (1/3)$$

$$E = 0.333 \times 2.60$$

$$E = 0.858\text{m}$$

Blade width:

$$W/Da = 1/5$$

$$W = 0.2 \times 0.858$$

$$W = 0.1716\text{m}$$

$$Lb/Da = 1/4$$

$$Lb = 0.25 \times 0.858$$

$$Lb = 0.2145\text{m}$$

CHAPTER 6

6 PROCESS ECONOMICS

6.1 *Specifications*

Basis: 13333.33kg/day of IBU-profen(product)

Number of working days per year = 300 days

Production of IBU-profen per hour = 555.55kg/hr

Production of IBU-profen per year = $555.55 \times 300 \times 24 = 40,00,000$

Cost of soft per tray = Rs.800

So, gross sale for one year (Total income) = $40,00,000 \times 800 =$
Rs.320,00,00,000

Turnover ratio = Gross annual sales / Fixed cost investment

For a chemical industry,

Turnover = 1

So, gross annual sales investment = fixed cost investment

Therefore fixed capacity investment

Fixed capital investment cost = Direct cost + Indirect cost

6.2 DIRECT COST

It's 70% of fixed capital investment. Hence

Direct cost = $320,00,00,000 \times 0.7$
=Rs:2240000000

Direct cost consists of,

- Equipment
- Land cost
- Installation

- Piping
- Insulation
- Painting

6.3 EQUIPMENT COST

It is 34.5% of Direct Costs.

Therefore equipment cost = Rs. 2240000000×0.345

=Rs.772800000

6.4 LAND COST

It is 1% of Direct Costs.

Land cost = Rs. 2240000000×0.01

=Rs. 22400000

6.5 INSTRUMENT AND INSTALLATION COST

These cost account to 20% of equipment cost,

=Rs. 772800000×0.2

=Rs.154560000

6.6 ELECTRICAL AND INSTRUMENT COST

These cost account to 29% of equipment cost,

=Rs. 772800000×0.29

=Rs.224112000

6.7 BUILDING, PROCESS AND AUXILIARY COST

This cost is 31% of equipment cost,

=Rs. 772800000×0.31

=Rs. 239568000

6.8 SERVICE FACILITIES AND YARD IMPROVEMENT

This cost is 20% of equipment cost,

=Rs. 772800000×0.2

=Rs.154560000

6.9 INDIRECT COST

These are expenses which are not including with material and labor of actual installation of complete facilities. This account for 30% of fixed capital investment

=Rs. 320,00,00,000×0.3

=Rs. 960000000

The indirect cost consist of

- Engineering supervision
- Construction expenses and construction fees
- Contingency

6.10 ENGINEERING SUPERVISION

It is 20% of Indirect Costs,

$$=Rs. 960000000 \times 0.2$$

$$=Rs. 192000000$$

6.11 CONSTRUCTION EXPENSES AND CONSTRUCTION FEES

It is 5% of Indirect costs,

$$=Rs. 960000000 \times 0.05$$

$$=Rs. 48000000$$

6.12 CONTINGENCY

It is 5% of Indirect costs,

$$=Rs. 960000000 \times 0.05$$

$$=Rs. 48000000$$

6.13 ESTIMATION OF TOTAL PRODUCTION COST

Total annual income = Rs. 320,00,00,000

Total gross earning is 15% of total income

$$= \text{Rs. } 320,00,00,000 \times 0.15$$

$$= \text{Rs. } 48,00,00,000$$

6.14 PRODUCTION COST

Production cost = Total annual income – Gross earning

$$= \text{Rs. } 320,00,00,000 - \text{Rs. } 48,00,00,000$$

$$= \text{Rs. } 272,00,00,000$$

Total production cost = Direct production cost + Fixed charges + Plant overhead

6.15 DIRECT PRODUCTION COST

It is 75% of total production cost,

$$= 272,00,00,000 \times 0.75$$

$$= \text{Rs. } 204,00,00,000$$

6.16 OPERATING LABOR COST

It is 15% of total production cost,

$$= \text{Rs. } 272,00,00,000 \times 0.15$$

$$= \text{Rs. } 40,80,00,000$$

6.17 DIRECT SUPERVISORY AND CLINICAL COST

It is 5% of total production cost,

$$= \text{Rs. } 272,00,00,000 \times 0.05 = \text{Rs. } 13,60,00,000$$

6.18 UTILITIES

It is 7% of total production cost,

$$= \text{Rs. } 272,00,00,000 \times 0.07$$

$$= \text{Rs. } 19,04,00,000$$

6.19 MAINTENANCE AND REPAIR COST

It is 8% of total production cost,

$$= \text{Rs. } 272,00,00,000 \times 0.08$$

$$= \text{Rs. } 21,76,00,000$$

6.20 PATENTS AND ROYALTIES COST

It is 15% of total production cost,

$$= \text{Rs. } 272,00,00,000 \times 0.15$$

$$= \text{Rs. } 40,80,00,000$$

6.21 FIXED CHARGES

It is 20% of total production cost
= Rs. 2720000000×0.2
= Rs. 544000000

6.22 PLANT OVERHEAD

This includes cost for following general plant and overheads payroll. Overheads packing medical services, safety and protection, restaurant, recreation, salvage, laboratories and storage facilities.

This cost is 5% of production cost
= Rs. 2720000000×0.05
= Rs. 136000000

6.23 DEPRECIATION

Depreciation for machinery is 10% of the fixed capital cost

= Rs. 320,00,00,000 ×0.1
= Rs. 320000000

Depreciation for building is 3% of the land cost

= Rs.2240000000 ×0.03
= Rs.67200000

Total depreciation = Rs. 320000000 + Rs. 67200000
= Rs. 387200000

6.24 INSURANCE

It is 1% of fixed capital cost
= Rs. 320000000 ×0.01
= Rs. 3200000

6.25 GENERAL EXPENSES

Administrative cost includes cost for executive offices, clinical wages, legal fees, office supplies and communication.

It is 5%of total production cost,
= 2720000000×0.05

$$= \text{Rs.}136000000$$

6.26 DISTRIBUTION AND SELLING COST

It includes cost for sales officers, sales men and shipping

It is 7% of total production cost,

$$= \text{Rs. } 2720000000 \times 0.07$$

$$= \text{Rs.}190400000$$

6.27 RESEARCH AND DEVELOPMENT

This cost 1% of total production cost,

$$= \text{Rs. } 2720000000 \times 0.01$$

$$= \text{Rs.}27200000$$

6.28 FINANCING

It is 2% of total production cost,

$$= \text{Rs. } 2720000000 \times 0.02$$

$$= \text{Rs.}54400000$$

6.29 GROSS EARNING COST

It is the net profit obtained after deduction of tax from goods earning

Tax payable = 30% gross profit

$$= \text{Rs. } 4800000000 \times 0.3 = \text{Rs.}1440000000$$

6.30 NET PROFIT

Net profit = Gross – tax

$$= \text{Rs. } 4800000000 - \text{Rs. } 1440000000$$

$$= \text{Rs.}3360000000$$

6.31 PAYBACK PERIOD

Payback period = $\frac{\text{Depreciated fixed capital investment}}{(\text{Annual profit} + \text{Average depreciation}) / \text{year}}$

$$= \text{Rs. } 387200000 / (\text{Rs. } 3360000000 + \text{Rs. } 672000000)$$

$$= 1.01 \text{ year}$$

Payback period = 368.65 days.

CHAPTER 7

7 ADVANTAGES AND DISADVANTAGES OF ZLD

7.1 ADVANTAGES

- To reduce the usage of fresh water in industry.

- To preserve natural environment against pollution.
- To reduce water expenditure on water acquisition.
- To clean industry effluent and recycle it for further use.
- It is cost effective.
- Easily maintained mechanical work.
- Capable of removing 98% of suspended solids.

7.2 DISADVANTAGES

- Cleaning is hassle.
- Most plants need at least three tanks.
- Temperature changes affect the tank greatly.

7.3 APPLICATIONS

- Sludges are used for gardening purposes.
- It is also used in cement industry.
- It is also used in paint industry.

CHAPTER 8

CONCLUSION

Thus, the project based on ZERO LIQUID DISCHARGE. The corresponding material balance and energy balance for the whole process was performed. The design calculation for bio-reactor and condenser have been calculated. The economic considerations of plant were analyzed. The rate of return for ZLD is 99.2%. Thus, the project covers all the aspects required for design of ZLD and could be implemented in reality.

REFERENCES

1. Kirk, R.E. and Othmer D.F., (2005) 'Encyclopedia of Chemical Technology', Seventh Edition.
2. M.Gopala Rao and Marshall Sitting, 'Shreve's Chemical Process Industries', Fifth Edition.
3. Max S. Peters and Klaus Timmerhaus, 'Plant Design and Economics for Chemical Engineers', Mc-Graw Hill Book Company, Third Edition
4. McCabe W.L., Smith J.C. and Harriott P. (1993) 'Unit Operations of Chemical Engineering', McGraw-Hill, Fifth Edition
5. R. H. Perry and Don W. Green, (1997)'Perry's Chemical Engineers' Hand Book', Mc Graw Hill International Edition, Volume – 6
6. R. K. Sinnott, Butter Worth-Heinemann., (2005) 'Coulson and Richardson's Chemical Engineering Series', Third Edition, Volume – 6
7. DWSIM Simulation software.