

# **MANUFACTURE OF SULPHURIC ACID BY DCDA PROCESS**

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

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**DEPARTMENT OF CHEMICAL ENGINEERING**  
**SCHOOL OF BIO AND CHEMICAL ENGINEERING**

## **SATHYABAMA**

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

**Accredited with Grade "A" by NAAC**  
**JEPPIAR NAGAR, RAJIV GANDHI SALAI, CHENNAI-600 119**

**APRIL 2021**



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**DEPARTMENT OF CHEMICAL ENGINEERING**

**BONAFIDE CERTIFICATE**

This is to certify that this Project Report is the bonafide work of **SWETHA D A (37190040)**, **KARPAGAM D (37190024)** who carried out the project entitled "**MANUFACTURE OF SULPHURIC ACID BY DCDA PROCESS**" under my supervision from February, 2020 to April, 2021.

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Assistant Professor.

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Submitted for Viva voce Examination held on \_\_\_\_\_

**Internal Examiner**

**External Examiner**

## DECLARATION

I, **SWETHA D A (37190040)** hereby declare that the Project Report entitled “**MANUFACTURE OF SULPHURIC ACID BY DCDA PROCESS**” done by me under the guidance of **Dr.D.VENKATESAN M.TECH.,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: 22.03.2021**

**PLACE: Chennai**



**SIGNATURE OF THE CANDIDATE**

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## **ACKNOWLEDGEMENT**

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I convey my thanks to **Dr.S.Sathish., M.E.,Ph.D.**, and **Dr.D.PRABU., M.S.,Ph.D** Head of the department, Department of Chemical Engineering for providing me necessary support and details at the right time during the progressive reviews.

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## ABSTRACT

This project includes the detailed designing of a four stage adiabatic catalytic bed reactor and design of equipment and cost for a sulphuric acid plant of capacity 10000 TPA (tons per annum). The feed is at 1atm and 3000°C. There are two main processes for manufacture of sulphuric acid namely the chamber process and the contact process. Chamber process produced sulphuric acid of concentration less than 80 %.The major disadvantage includes the limitations in throughput, quality and concentration of the acid produced. Compared to chamber process, contact process is used in many industries.

The contact process increases yield and reduces stack emission of unconverted SO<sub>2</sub>. Conversions using a single absorption contact process were about 97-98%. While in the current double absorption flow process conversions are as high as 99.7% The catalyst used here is vanadium pent-oxide (V<sub>2</sub>O<sub>5</sub>)

This project mainly comprise of the basic parts of the sulphuric acid manufacturing plant, the equipment and the catalyst used, flow of materials in and out of the equipment, material and energy balances, heat duty of the heat exchangers.

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# CHAPTER 1

## INTRODUCTION

Sulphuric acid is a highly corrosive strong mineral acid. It is a colourless liquid which is soluble in water.

Sulphuric acid is an important chemical and it is used in many large scale industries. It is mainly used in the phosphate fertilizer industry.

Applications: Petroleum refining, Plastic and fibers, Detergents, Explosives

The changes made in the methodology of manufacturing sulphuric acid result mainly in the reduction of the emission of sulphur compounds to air and the reduction of harmful waste.

It started with the Lead Chamber process, followed by the contact process with Single Conversion Single Absorption (SCSA) and Double Conversion Double Absorption Process (DCDA). The contact process is used to produce sulphuric acid in higher concentration compared to the lead chamber process and also improvement in conversion and absorption stages are being introduced from time to time to increase conversion and absorption efficiencies, which result in the reduction of emissions.

## CHAPTER 2

### LITERATURE SURVEY

Although sulphuric acid is now one of the most widely used chemicals, it was probably little known before 16<sup>th</sup> century. It was prepared by Johann Van Helmont by destructive distillation of green vitriol (ferrous sulphate) and by burning sulphur.

In 17<sup>th</sup> century, the German-Dutch chemist Johann Glauber prepared sulphuric acid by burning sulphur together with saltpeter (potassium nitrate,  $\text{KNO}_3$ ) in the presence of steam. As saltpeter decomposes, it oxidizes the sulphur to  $\text{SO}_3$ , which combines with water to produce sulphuric acid. In 1736, Joshua Ward, a London pharmacist, used this method to begin the first large-scale production of sulphuric acid.

In 1746 in Birmingham, John Roebuck adapted this method to produce sulphuric acid in lead lined chambers, which were stronger, less expensive, and could be made larger than the previously used glass containers. Sulphuric acid created by John Roebuck's process approached a 65% concentration.

After several refinements, this method, developed into lead chamber process or "chamber process". Later refinements to the lead chamber process by French chemist Joseph Louis Gay-Lussac and British chemist John Glover improved concentration to 78%. However, the manufacture of some dyes and other chemical processes require a more concentrated product. Throughout the 18th century, this could only be made by dry distilling minerals in a technique similar to the original alchemical processes. Pyrite (iron disulphide,  $\text{FeS}_2$ ) was heated in air to yield  $\text{FeSO}_4$ , which was oxidized by further heating in air to form  $\text{Fe}_2(\text{SO}_4)_3$ , when heated to 480 °C, decomposed to iron(III) oxide and sulphur trioxide, which could be passed through water to yield sulphuric acid in any concentration. However, the expense of this process prevented the large-scale use of concentrated sulphuric acid.

In 1831, British vinegar merchant Peregrine Phillips patented the contact process, which was a far more economical process for producing sulphur trioxide and concentrated sulphuric acid. In the current flow process, achievable conversions are as high as 99.7%

## CHAPTER 3

### AIM AND SCOPE

#### 3.1 AIM

To design a plant for manufacturing of sulphuric acid by DCDA Process

#### 3.2 OBJECTIVES

1. Maximum yield of  $\text{H}_2\text{SO}_4$  production
2. Process require less raw material
3. Reaction time is less
4. Product achieved – 99.7%

## CHAPTER 4

### MATERIALS AND METHODS

#### 4.1 PROCESS SELECTION

Manufacture of sulphuric acid is of two methods

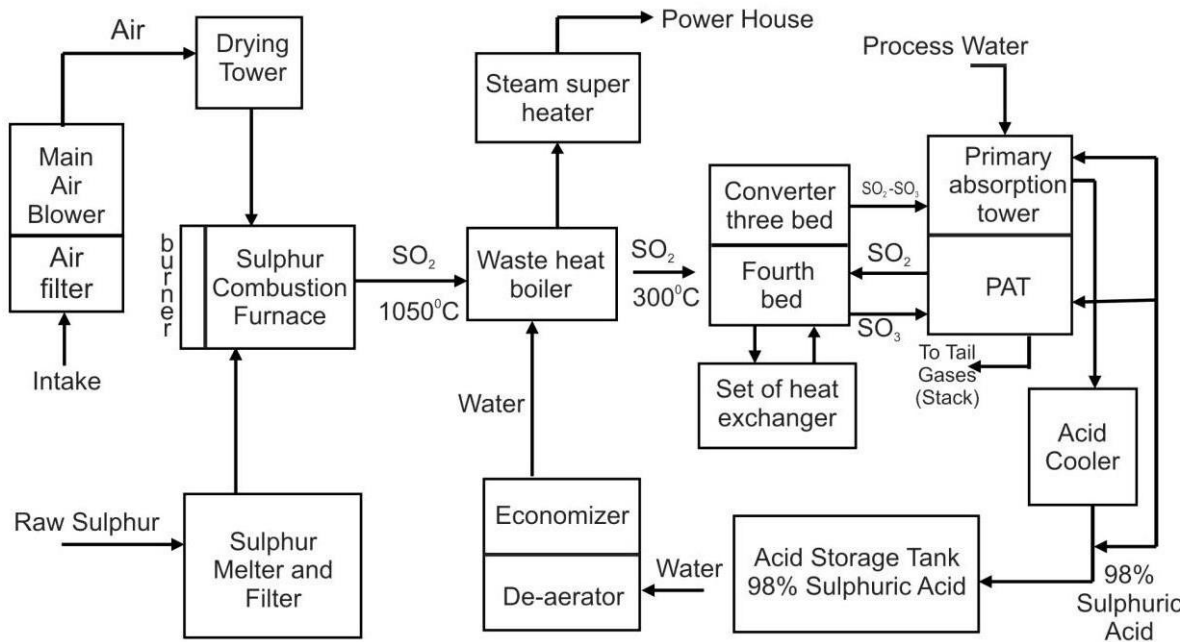
- Chamber Process
- Contact Process
  - (i) SCSA (Single contact Single absorption) process
  - (ii) DCDA (Double contact Double absorption ) process

Chamber process or lead chamber process is the oldest method and it is replaced by contact process. Contact process is used to produce sulphuric acid in the high concentration. Whereas chamber process produces sulphuric acid in less concentration. For manufacture of sulphuric acid in industry contact process is widely used.

#### CONTACT PROCESS

The primary raw material for producing intermediate product of sulphur dioxide is elemental sulphur. This can either be made by burning of sulphur in an excess of air or by heating sulphide ores like pyrite in an excess of air. In either case, an excess of air is used so that the sulphur dioxide produced is already mixed with oxygen for the next stage.

Contact process can be further subdivided into the double contact double absorption process (DCDA) and single contact single absorption (SCSA). In DCDA Process  $\text{SO}_3$  is removed from the gas stream after 3<sup>rd</sup> bed, which shifts the equilibrium and increases the rate of the forward  $\text{SO}_2$  to  $\text{SO}_3$  reaction resulting in higher overall conversion and reduces stack emission of unconverted  $\text{SO}_2$ . Conversions using single absorption contact process were typically about 97-98%. While in the current double absorption flow process, achievable conversions are as high as 99.7%.



**Figure 4.1 : Process Flow Sheet**

## 4.2 PROCEDURE

The main steps involved in DCDA process are as below:

- Melting solid Sulphur with steam coils, followed by filtration or settling of impurities to obtain clean sulphur containing less than 10 mg/l of ash.
- Burning the molten Sulphur with air to produce gas-containing  $\text{SO}_2$ .
- Cooling the hot gas in Waste Heat Boiler System to produce superheated or saturated steam at conditions fixed, as per requirements.
- Catalytic oxidation of  $\text{SO}_2$  to  $\text{SO}_3$  in three consecutive passes of converter containing  $\text{V}_2\text{O}_5$  catalyst with intercooling of gas in between. The exothermic heat of reaction is utilized to produce steam in Waste Heat Boiler system and to reheat the gases going to pass IV from the intermediate absorber



# CHAPTER 5

## PERFORMANCE ANALYSIS

### 5.1 MATERIAL BALANCE

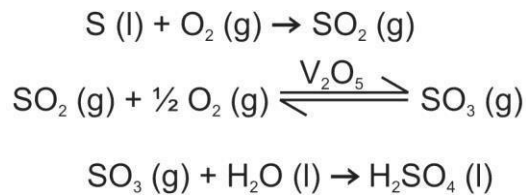
Assumptions

1. Complete burning of Sulphur in the burner.
2. 99.8% conversion of Sulphur-dioxide to Sulphur-trioxide in the reactor.
3. Overall absorption of Sulphur-trioxide in the process is 100%.
4. 40% excess oxygen is provided.
5. Humidity of entering air is 65% at 300C.

### Calculations

Basis: 10000 TPA H<sub>2</sub>SO<sub>4</sub> plant Purity: 98 % pure acid

Reactions:



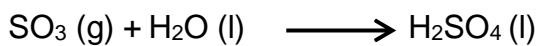
10000 TPA H<sub>2</sub>SO<sub>4</sub> = (10000 x 10<sup>3</sup>) / (325x24) = 1282.05 kg/hr

98% pure acid produced = 1282.05 x 0.98 = 1256.40 kg/hr

No. of moles of acid produced = 1256.40 / 98 = 12.82 kmol/hr

(Molecular weight of H<sub>2</sub>SO<sub>4</sub>=98)

Overall absorption of acid = 100 %



Therefore, by stoichiometry,

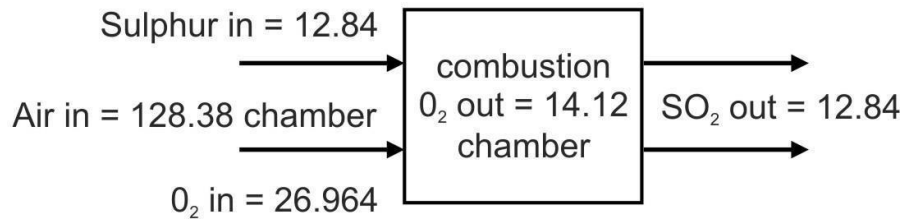
SO<sub>3</sub> required = 12.82 kmol/hr

Overall conversion of SO<sub>2</sub> to SO<sub>3</sub> = 99.8 % (Assumption)

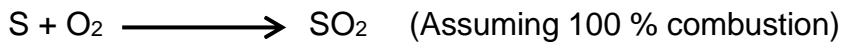
Let SO<sub>2</sub> formed be *x* kmol/hr

$0.998 \cdot x = 12.82$        $\text{SO}_2 \text{ required} = x = 12.82/0.998 = 12.84 \text{ kmol/hr}$   
 $\text{O}_2 \text{ required} = 12.84 \times 0.5 = 6.42 \text{ kmol/hr}$

### 5.1.1 Combustion Chamber Balance



**Figure 5.1 : Combustion Chamber**



S required = 12.84 kmol/hr = 410.88 kg/hr

$\text{O}_2$  required = 12.84 x 1 = 12.84 kmol/hr

Total  $\text{O}_2$  required = 12.84 + 6.42 = 19.26 kmol/hr

$\text{O}_2$  is taken in 40% excess

$\text{O}_2$  in the combustion chamber = 19.26 x 1.4 = 26.964 kmol/hr

(Dry air contains 21%  $\text{O}_2$ )

Dry air in = 26.964/0.21 = 128.38 kmol/hr

(Molecular weight of air=29)

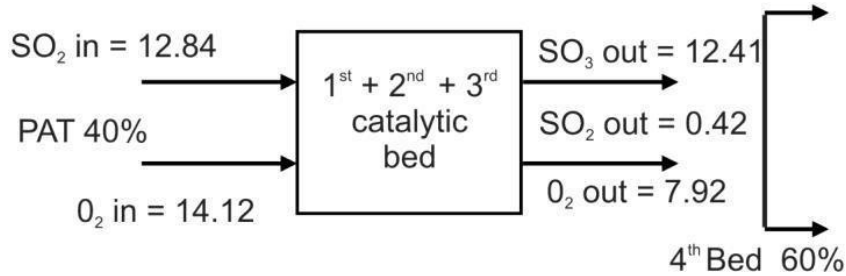
Dry air in = 128.38 x 29 = 3723.02 kg/hr

**TABLE 5.1: COMBUSTION CHAMBER**

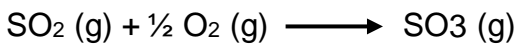
Component	Inlet (kmol)	Inlet ( kg)	Outlet (kmol)	Outlet(kg)
S	12.84	410.88	-	-
$\text{O}_2$	26.96	862.72	14.12	451.84
$\text{N}_2$	3298.306	92352.56	3298.306	92352.56
$\text{SO}_3$	-	-	12.84	821.76
TOTAL	2438.106	93627.168	3325.266	93627.168

Material Balance MASS IN = MASS OUT

### 5.1.2 Overall Balance over First Three Catalytic Beds



**Figure 5.2: Overall Balance Over First Three Catalytic Bed**



(Conversion = 96.7 %)  $\text{SO}_2$  in = 12.84 kmol

$\text{O}_2$  in = 14.12 kmol

$\text{SO}_2$  reacted =  $12.84 \times 0.967 = 12.426$  kmol/hr

$\text{O}_2$  reacted =  $12.41 \times 0.5 = 6.20$  kmol/hr

$\text{SO}_3$  formed = 12.416 kmol/hr

$\text{SO}_2$  out =  $\text{SO}_2$  in –  $\text{SO}_2$  reacted

$$= 12.84 - 12.416$$

$$= 0.424 \text{ kmol/hr}$$

$\text{O}_2$  out =  $\text{O}_2$  in –  $\text{O}_2$  reacted

$$= 14.12 - 6.20$$

$$= 7.92 \text{ kmol/hr}$$

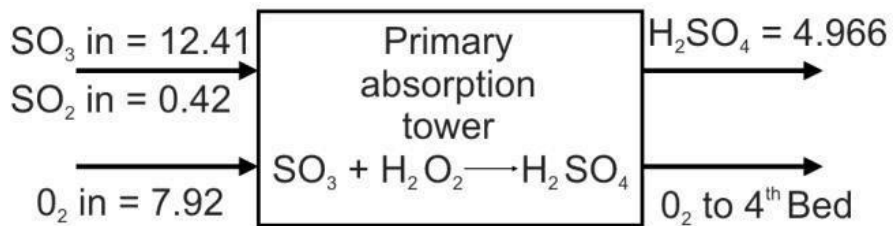
**TABLE 5.2: OVERALL BALANCE OVER FIRST THREE CATALYTIC BED**

Component	Inlet (kmol)	Mole %	Inlet ( kg)	Outlet (kmol)	Outlet (kg)
SO <sub>2</sub>	12.84	10	821.76	0.424	26.88
O <sub>2</sub>	14.12	11	451.84	7.92	253.44
N <sub>2</sub>	3298.306	79	92352.56	3298.306	92352.56
SO <sub>3</sub>	-	-	-	12.416	993.2
TOTAL	3325.266	100	93627.16	-	93627.16

MASS IN = MASS OUT

After the third stage, 40 % of product goes to economizer and then to the Inter pass

### 5.1.3 Primary Absorber



**Figure 5.3 : Primary Absorber**

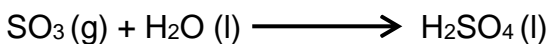
(Assuming 100 % absorption)

$$\text{SO}_2 \text{ in} = 0.4 \times 0.42 = 0.168 \text{ kmol/hr}$$

$$\text{SO}_3 \text{ in} = 0.4 \times 12.416 = 4.9664 \text{ kmol/hr}$$

$$\text{O}_2 \text{ in} = 0.4 \times 7.92 = 3.168 \text{ kmol/hr}$$

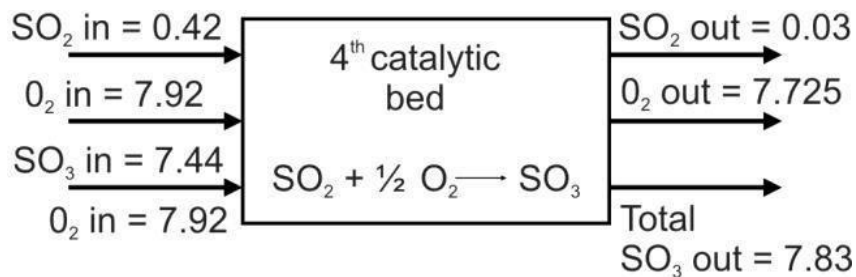
$$\text{N}_2 \text{ in} = 0.4 \times 3298.306 = 1319.3224 \text{ kmol/hr}$$



$\text{H}_2\text{SO}_4$  formed = 4.966 kmol/hr

(Sulphur dioxide, Oxygen and Nitrogen are recycled from inter pass absorber to the 4<sup>th</sup> stage of reactor).

#### 5.1.4 Fourth Catalytic Bed



**Figure 5.4: Fourth Catalytic Bed**

Conversion = 3.1 % (Overall = 99.8 %)

Gases in to the fourth stage of reactor constitute 60 % of product from third stage and gases recycled from Inter pass absorber.

$\text{SO}_2$  in = Total  $\text{SO}_2$  out from the 3<sup>rd</sup> bed = 0.42 kmol/hr

$\text{O}_2$  in = Total  $\text{O}_2$  out from the 3<sup>rd</sup> bed = 7.92 kmol/hr

$\text{N}_2$  in = 3298.30 kmol/hr

$\text{SO}_3$  in = 60% of  $\text{SO}_3$  out from the 3<sup>rd</sup> bed

$$= 0.6 * 12.41 = 7.44 \text{ kmol/hr}$$

$\text{SO}_2(\text{g}) + \frac{1}{2} \text{O}_2(\text{g}) \longrightarrow \text{SO}_3(\text{g})$  (Overall Conversion = 99.8 %)

$\text{SO}_2$  reacted up to 4<sup>th</sup> bed = 99.8 % of  $\text{SO}_2$  entering the 1<sup>st</sup> bed

$$= 0.998 * 12.84 = 12.81 \text{ kmol/hr}$$

$\text{SO}_2$  out = Initial  $\text{SO}_2$  – Total  $\text{SO}_2$  reacted

$$= 12.84 - 12.81 = 0.036 \text{ kmol/hr}$$

Therefore  $\text{SO}_2$  reacted in the 4<sup>th</sup> bed

$$= \text{SO}_2 \text{ in} - \text{SO}_2 \text{ out} = 0.42 - 0.036 = 0.39 \text{ kmol/hr}$$

$$\text{O}_2 \text{ reacted} = 0.39 * 0.5 = 0.19 \text{ kmol/hr}$$

$$\text{SO}_3 \text{ formed} = 0.39 \text{ kmol/hr}$$

$$\text{O}_2 \text{ out} = \text{O}_2 \text{ in} - \text{O}_2 \text{ reacted}$$

$$= 7.92 - 0.195 = 7.725 \text{ kmol/hr}$$

$$\text{Total SO}_3 \text{ outlet} = \text{SO}_3 \text{ inlet from the 3}^{\text{rd}} \text{ bed} + \text{SO}_3 \text{ formed in the 4}^{\text{th}} \text{ bed}$$

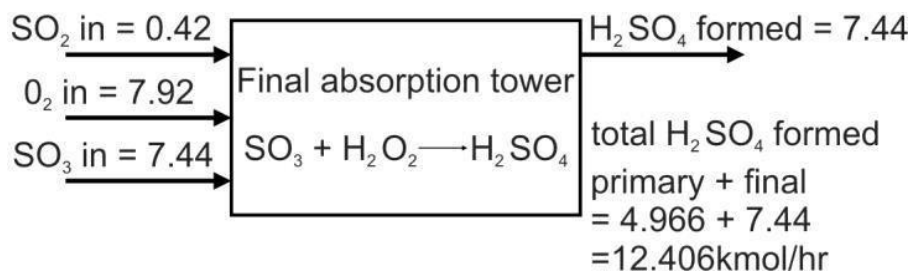
$$= 7.44 + 0.39 = 7.83 \text{ kmol/hr}$$

**TABLE 5.3: FOURTH CATALYTIC BED**

Component	Inlet (kmol)	Inlet (kg)	Outlet (kmol)	Outlet (kg)
SO <sub>2</sub>	0.42	26.88	0.03	1.92
O <sub>2</sub>	7.92	253.44	7.725	247.2
N <sub>2</sub>	3298.30	92,353.56	3298.30	92,353.56
SO <sub>3</sub>	7.44	595.2	7.83	626.4
Total	3314.08	93,229.08	3313.885	93229.08

MASS IN = MASS OUT

### 5.1.5 Final Absorption Tower



**Figure 5.5 : Final Absorption Tower**

## 5.2 ENERGY BALANCE

The energy or heat coming into a process in the inlet material plus any net energy added to the process is equal to the energy leaving in the

$$\Sigma H_R + (\Delta H^\circ_{298}) + q = \Sigma H_P$$

Where  $\Sigma H_R$  is the sum of enthalpies of all materials entering the reaction

process  $\Delta H^\circ_{298}$  = standard heat of reaction at 298K AND 101.32kPa

Q= net energy or heat added to the system.

$\Sigma H_P$ = sum of enthalpies of all leaving materials referred to the standard reference state at 298 K

Equations for calculating net enthalpy change for catalyst bed Heat capacity is given by the equation

$$C_p^\circ = a + bT + cT^2 + \dots$$

$$\Delta H^\circ_{RT} = T \int^{298} \Sigma (n_i C_{p,i}^\circ)_{\text{reactants}} dT + \Delta H^\circ_{298} + \int_{298}^T \Sigma (n_i C_{p,i}^\circ)_{\text{products}} dT \text{ where, } \Delta H^\circ_{RT} \text{ is}$$

the standard heat of reaction at temperature T (K)

$$\Delta H^\circ_{RT} = \Delta H^\circ_{298} + \int_{298}^T (\Sigma (n_i C_{p,i}^\circ)_{\text{Products}} - \Sigma (n_i C_{p,i}^\circ)_{\text{Reactants}}) dT \dots (4)$$

Enthalpy change, with 298 K as the reference temperature, can be calculated from the formula,

$$\Sigma H = \Delta H^\circ_{RT} - \Delta H^\circ_{298} = (\Sigma a_i n_i)(T - 298) + ((\Sigma b_i n_i)/2)(T^2 - 298^2) + \dots (5)$$

Similarly, Enthalpy change between T<sub>1</sub> K and T<sub>2</sub> K, can be calculated from the formula

$$\Sigma H = (\Sigma a_i n_i)(T_1 - T_2) + ((\Sigma b_i n_i)/2)(T_1^2 - T_2^2) \dots (6)$$

Standard heat of reaction at 298 K ( $\Delta H^\circ_{298}$ ) from heat of formation is given by

$$\Delta H^\circ_{298} = [\Sigma (n_i \Delta H^\circ_{f,i})]_{\text{Products}} - [\Sigma (n_i \Delta H^\circ_{f,i})]_{\text{Reactants}} \dots (7)$$

where  $[\Sigma (n_i \Delta H^\circ_{f,i})]_{\text{products}}$  = standard heat of formation of products and

$[\Sigma (n_i \Delta H^\circ_{f,i})]_{\text{reactants}}$  = standard heat of formation of reactants Net heat to be

removed from the catalytic bed ,

$$q = \sum H_P - \sum H_R + \Delta H^0_{298} \text{ DATA GIVEN}$$

Inlet and effluent temperature

First stage : Inlet temperature =410°C

Effluent temperature =603°C

Second stage : Inlet temperature =439°C

Effluent temperature =506°C

Third stage : Inlet temperature =433°C

Effluent temperature =455°C

Fourth stage : Inlet temperature =428°C

Effluent temperature=436°C

Heat capacities:

$$N_2 = 29.5909 - 5.14 \times 10^{-3}T$$

$$O_2 = 26.0257 + 11.755 \times 10^{-3}T$$

$$SO_2 = 24.7706 + 62.9481 \times 10^{-3}T$$

$$SO_3 = 22.0376 + 121.624 \times 10^{-3}T$$

CALCULATION

$$\Delta a = 22.036 - 24.771 - 0.5(26.026) = -15.748$$

$$\Delta b \cdot 10^3 = 121.624 - 62.948 - 0.5(11.755) = 52.799$$

$$\Delta c \cdot 10^6 = -91.867 - (-44.258) - 0.5(-2.343) = -46.438$$

$$\Delta d \cdot 10^9 = 24.369 - 11.122 - 0.5(-0.562) = 13.258$$

From eqn 7

$$\Delta H^0_{298} = -395720 - (-296810)$$

$$= -98910 \text{ kJ/kmol } SO_2 \text{ reacted}$$



Substituting the value  $\Delta H^\circ$  in eqn 6 we get

$$\Delta H^\circ = -96178 - 15.748T + (26.4 \times 10^{-3})T^2 - (15.48 \times 10^{-6})T^3 + (3.382 \times 10^{-9})T^4 \dots$$

Calculation of heat duty required in each bed

### 5.2.1 First Catalytic Bed

**TABLE 5.4 : HEAT CAPACITY EQUATION CONSTANTS FOR INCOMING GAS MIXTURE**

Component	$n_i$ (kmol/hr)	$a_i$	$a_i n_i$	$b_i$	$b_i n_i \times 10^{-3}$
SO <sub>2</sub>	12.84	24.7706	318.054	62.9481	808.253
O <sub>2</sub>	14.12	26.0257	367.482	11.7551	165.982
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	—	22.0376	—	121.624	—
TOTAL	3345.266	—	98285.379	—	-15982.355

Enthalpy of incoming gas mixture at 683K over 298K can be calculating by substituting  $T=298$  in eqn (5)

$$\begin{aligned} \Sigma H_R &= (\Sigma a_i n_i)(T_1 - T_2) + ((\Sigma b_i n_i)/2)(T_1^2 - T_2^2) \\ &= 98285.379(683 - 298) + (-15982.355 \times 10^{-3}/2)(683^2 - 298^2) \\ &= 37839870.9 + (-7.991)(377685) \\ &= 37839870.9 - 3018080.8 = 34821790.1 \text{ kJ/hr} \end{aligned}$$

$$\begin{aligned} 1 \text{ kW} &= 3600 \text{ kJ/hr} \\ &= 34821790.1 \text{ kJ/hr} \times 1 \text{ kW} \div 3600 \text{ kJ/hr} \\ &= 9672.719 \text{ kW} \end{aligned}$$

**TABLE 5.5 : HEAT CAPACITY EQUATION CONSTANTS FOR  
OUTGOING GAS MIXTURE**

<b>Component</b>	<b>n<sub>i</sub> (kmol/hr)</b>	<b>a<sub>i</sub></b>	<b>a<sub>i</sub>n<sub>i</sub></b>	<b>b<sub>i</sub></b>	<b>b<sub>i</sub>n<sub>i</sub>*10<sup>-3</sup></b>
SO <sub>2</sub>	3.339	24.7706	82.709	62.9481	210.18
O <sub>2</sub>	9.37	26.0257	243.86	11.7551	110.144
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	9.501	22.0376	209.37	121.624	1155.54
<b>TOTAL</b>	<b>3321.92</b>	<b>–</b>	<b>98135.78</b>	<b>–</b>	<b>-15480.72</b>

Enthalpy of outgoing gas mixture at T=876K over T=298K can be calculated by substituting T=876 K in eqn (5)

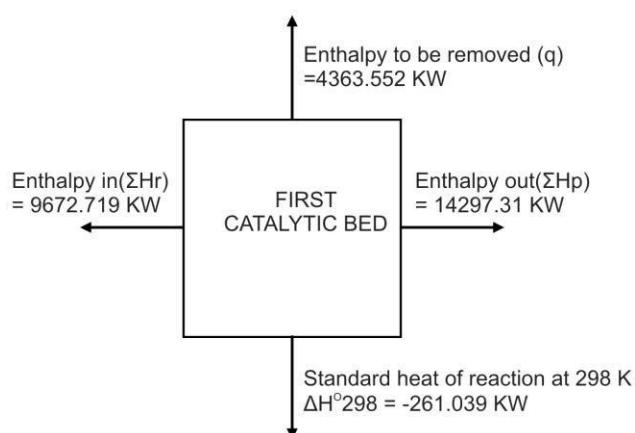
$$\begin{aligned}\Sigma H_P &= 98170.75(876-298) + ((-153480.72 \times 10^{-3}/2)(876^2-298^2)) \\ &= 56742693.5 + (-7.74)(678592) \\ &= 56742693.5 - 5252147.28 \\ &= 51470333.52 \text{ kJ/hr}\end{aligned}$$

$$\begin{aligned}\Sigma H_P &= 51470333.52/3600 \\ &= 14297.31 \text{ kW}\end{aligned}$$

Total heat of reaction at 298K

$$\begin{aligned}\Delta H_{298}^0 &= -98910 \text{ kJ/kmol} \\ &= (12.84 - 3.339)(-98910/3600) \\ &= 9.501(-27.475) \\ &= -261.039 \text{ kW}\end{aligned}$$

$$\begin{aligned}\text{Net enthalpy change } q &= \Sigma H_P - \Sigma H_R + \Delta H_{298}^0 \\ &= 14297.31 - 9672.719 - 261.039 \\ &= 4363.55 \text{ kW}\end{aligned}$$



**Figure 5.6 : Enthalpy Balance Over First Catalytic Bed**

## 5.2.2 SECOND CATALYTIC BED

**TABLE 5.6 : HEAT CAPACITY EQUATION CONSTANT FOR INCOMING GAS MIXTURE**

Component	N <sub>i</sub> (kmol / hr)	a <sub>i</sub>	a <sub>i</sub> n <sub>i</sub>	b <sub>i</sub>	b <sub>i</sub> n <sub>i</sub> *10 <sup>-3</sup>
SO <sub>2</sub>	3.339	24.7706	82.709	62.948	210.18
O <sub>2</sub>	9.37	26.0547	243.86	11.7551	110.144
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	9.501	22.0376	209.37	121.624	1155.54
TOTAL	3321.92	_	98135.78	_	-15480.72

Enthalpy of incoming gas mixture at 712 K over 298 K sub T = 712 In eqn (5)

$$\Sigma H_R = 98135.78 (712 - 298) + \frac{(-15391.84 \times 10^{-3})}{2} (718^2 - 298^2)$$

$$= 40628212.92 + (-7.74) (418140)$$

$$= 46642690.5 - 3236403.6$$

$$= 37391809.32 \text{ kJ/hr}$$

$$= 37391809.32 / 3600$$

$$= 10386.61 \text{ kW}$$

**TABLE 5.8: HEAT CAPACITY EQUATION CONSTANT FOR  
OUTGOING GAS MIXTURE**

Component	$n_i$	$a_i$	$n_i a_i$	$b_i$	$b_i n_i \cdot 10^{-3}$
SO <sub>2</sub>	0.976	24.7706	24.176	62.9481	61.437
O <sub>2</sub>	8.188	26.0257	213.09	11.7551	96.250
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	11.864	22.0376	261.45	121.624	1442.94
TOTAL	3319.33	—	98098.55	—	-15355.96

Enthalpy of outgoing gas mixture at 779 K over 298 K can be calculated by substituting  $T = 779$  K in eqn(5)

$$\Sigma H_P = 98098.55 (779 - 298) + ((-15355.96 \times 10^{-3})/2) (779^2 - 298^2)$$

$$= 47185402.55 + (-7.677)(518037)$$

$$= 41785402.55 - 3976970.04$$

$$= 43208432.51 \text{ kJ/hr}$$

$$\Sigma H_P = 12002.342 \text{ kW}$$

Total heat of reaction at 298 K

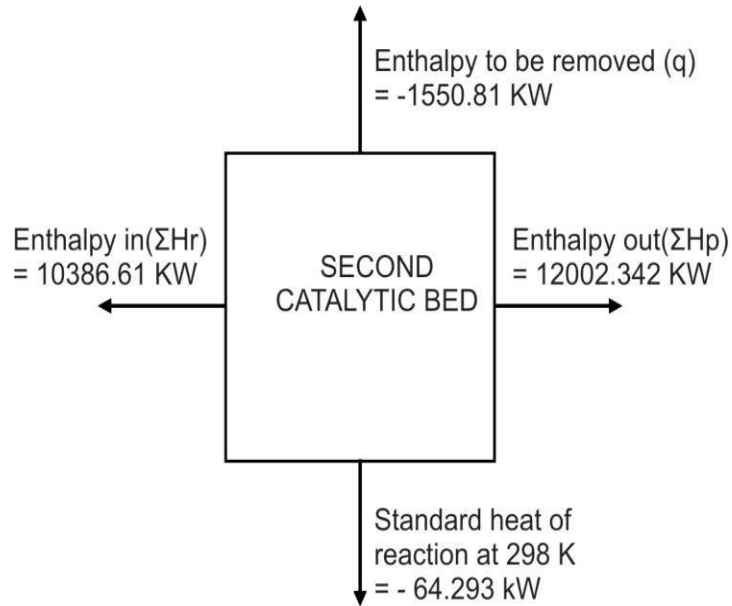
$$\Delta H_{298}^0 = -98910 \text{ kJ/kmol SO}_2 \text{ reacted}$$

$$= (3.339 - 0.976) (-98910/3600)$$

$$= 2.363 (-27.475) = -64.923 \text{ kW}$$

$$\text{Net enthalpy change } q = \Sigma H_P - \Sigma H_R + \Delta H_{298}^0$$

$$= 12002.342 - 10386.61 - 64.923 = -1550.81 \text{ kW}$$



**Figure 5.7 : Enthalpy Balance Over Second Catalytic Bed**

**5.2.3 THIRD CATALYTIC BED**

**TABLE 5.8 : HEAT CAPACITY EQUATION CONSTANT FOR INCOMING GAS MIXTURES**

Component	$n_i$ (kmol / hr)	$a_i$	$a_i n_i$	$b_i$	$b_i n_i * 10^{-3}$
SO <sub>2</sub>	0.976	24.7706	24.176	62.9481	61.437
O <sub>2</sub>	8.188	26.0257	213.09	11.7551	96.250
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	11.864	22.0376	261.45	121.624	1442.94
TOTAL	3319.33	—	98098.55	—	-15355.96

Enthalpy of incoming gas mixture at 706 K over 298 K can be calculated by substituting T = 706 K in eqn (5)

$$\begin{aligned}\Sigma H_R &= 98098.55 (706 - 298) + ((-15355.96 * 10^{-3})/2) \\ &= 40024208.4 + (-7.677) (409632) \\ &= 40024208.4 - 3144744.86 \\ &= 36879463.5 \text{ kJ/hr}\end{aligned}$$

$$\Sigma H_R = 10244.29 \text{ kW}$$

**TABLE 5.9 : HEAT CAPACITY EQUATION CONSTANT FOR OUTGOING GAS MIXTURE**

Component	$n_i$ (kmol / hr)	$a_i$	$a_i n_i$	$b_i$	$b_i n_i * 10^{-3}$
SO <sub>2</sub>	0.42	24.7706	10.403	62.9481	26.689
O <sub>2</sub>	7.92	26.0257	206.12	11.7551	93.100
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	7.44	23.0376	163.959	121.624	904.882
TOTAL	3314.06	—	97980.32	—	-15931.91

Enthalpy of outgoing gas mixture at 718 K over 298 K can be calculated by substituting T= 718 K in eqn (5)

$$\begin{aligned}\Sigma H_P &= 97980.32 (718 - 298) + ((-15931.91 * 10^{-3})/2)(701^2 - 298^2) \\ &= 41151734.4 + (-7.965)(426720) \\ &= 41151734.4 - 3398824.8 \\ &= 37752909.6 \text{ KJ/hr} \\ &= 10486.91 \text{ kW}\end{aligned}$$

Total heat of reaction at 298 K

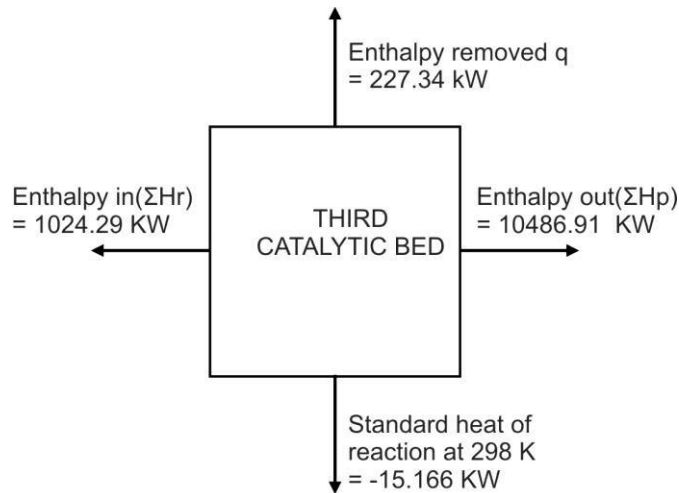
$$\Delta H_{298}^0 = -98910 \text{ kJ/kmol SO}_2 \text{ reacted}$$

$$= (0.976 - 0.42) (-98910/3600)$$

$$= (0.556)(-27.475) = -15.276$$

Net enthalpy change  $q = \Sigma H_P - \Sigma H_R + \Delta H^\circ_{298}$

$$= 10486.91 - 10244.29 - 15.276 = 227.34 \text{ kW}$$



**Figure 5.8 : Enthalpy Balance Over Third Catalytic Bed**

#### 5.2.4 Fourth Catalytic Bed

**TABLE 5.10 : HEAT CAPACITY EQUATION CONSTANT FOR INCOMING GAS MIXTURE**

Component	$n_i$ (kmol / hr)	$a_i$	$a_i n_i$	$b_i$	$b_i n_i \cdot 10^{-3}$
SO <sub>2</sub>	0.42	24.7706	10.403	62.9481	26.689
O <sub>2</sub>	7.92	26.0257	206.12	11.7551	93.100
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	7.44	22.0376	163.959	121.624	904.882
TOTAL	3314.08	—	97980.32	—	-15931.91

T= 701 K in eqn (5)

$$\begin{aligned}\Sigma H_R &= 97980.32 ( 701 - 298) + ((15031.91 * 10^{-3})/2)( 701^2 - 298^2) \\ &= 39486068.96 + ( -7.96)(402597) \\ &= 39486068.96 - 3204672.12 \\ &= 36281396.84 \text{ kJ/hr} \\ &= 10078.16 \text{ kW}\end{aligned}$$

**TABLE 5.11 : HEAT CAPACITY CONSTANT EQUATION FOR  
OUTGOING GAS MIXTURE**

Component	$n_i$ (kmol / hr)	$a_i$	$a_i n_i$	$b_i$	$b_i n_i * 10^{-3}$
SO <sub>2</sub>	0.03	24.7706	0.743	62.9481	1.888
O <sub>2</sub>	7.725	26.0257	247.2	11.7551	90.808
N <sub>2</sub>	3298.306	29.5909	97599.843	-5.141	-16956.59
SO <sub>3</sub>	7.83	22.0376	172.554	121.624	952.31
TOTAL	3313.89	_	98020.34	_	-15911.58

T=709 K in eqn(5)

$$\begin{aligned}\Sigma H_P &= 98020.34 ( 709 - 298) + (( - 15911.58 * 10^{-3})/2) ( 709^2 - 298^2) \\ &= 40286359.74 + ( - 7.95) ( 413877) \\ &= 40286359.74 - 3290322.15 \\ &= 36996037.5 \text{ kJ/hr} \\ &= 10,276.67 \text{ kW}\end{aligned}$$

Total heat of reaction at 298 K

$$\begin{aligned}\Delta H_{298} &= -98910 \text{ kJ/kmol SO}_2 \text{ reacted} \\ &= (0.42 - 0.03 ) (-98910/3600) \\ &= 0.39( -27.47)\end{aligned}$$

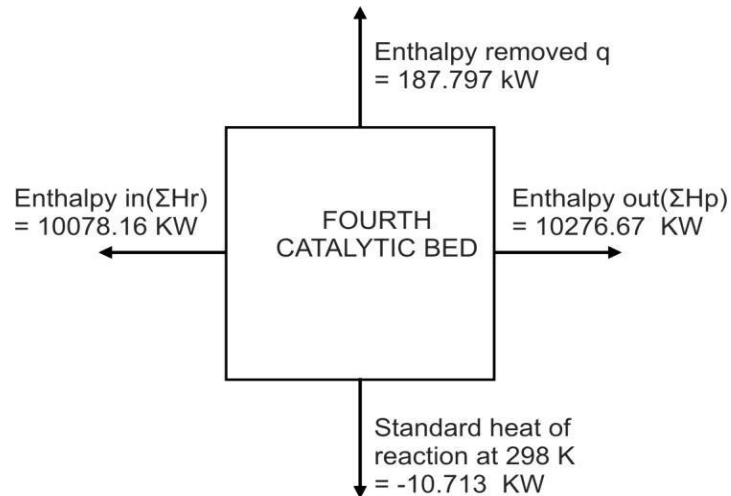


$$= -10.713 \text{ kW}$$

$$\text{Net enthalpy change } q = \Sigma H_P - \Sigma H_R + \Delta H^{0}_{298}$$

$$= 10,276.67 - 10078.16 - 10.713$$

$$= 187.797 \text{ kW}$$



**Figure 5.9 : Enthalpy Balance Over Fourth Catalytic Bed**

### 5.2.5 CALCULATION OF HEAT LOAD FOR HEAT EXCHANGER

Heat load required b/w first and second catalytic bed At the outlet of first catalytic bed

$$\Sigma a_i n_i = 98,135.78$$

$$\Sigma b_i n_i = -15480.72 * 10^{-3}$$

First stage outlet temp = 876 K

Second stage inlet temp = 712 K

Temp change = 876 to 712 K

Heat load is calculated by

$$\Delta H_1 = 98135.78 ( 876 - 712 ) + ((-15480.72 * 10^{-3})/2)$$

$$= 16094267.92 + (-7.74) (260432)$$

$$= 16094267.92 - 2015743.68$$

$$= 14078524.24 \text{ kJ/hr}$$

$$= 3910.70 \text{ kW}$$

Heat load required b/w second and third catalytic bed At the outlet of second catalytic bed

$$\Sigma a_i n_i = 98098.55$$

$$\Sigma b_i n_i = -15355.96 * 10^{-3}$$

$$\text{Second stage outlet temp} = 506^\circ\text{C} = 779 \text{ K}$$

$$\text{Third stage inlet temp} = 433^\circ\text{C} = 706 \text{ K}$$

$$\text{Temp change} = 779 \text{ K to } 706 \text{ K}$$

Heat load is calculated by substituting  $T_1 = 779 \text{ K}$   $T_2 = 706 \text{ K}$

$$\begin{aligned}\Delta H_2 &= 98098.55(779 - 706) + ((-15355.96 * 10^{-3})/2) (779^2 - 706^2) \\ &= 7161194.15 + (-7.67) (108405) \\ &= 6329727.8 \text{ kJ/hr} \\ &= 1758.25 \text{ kW}\end{aligned}$$

Heat load required b/w third and fourth catalytic bed At the outlet of third catalytic bed

$$\Sigma a_i n_i = 97980.32$$

$$\Sigma b_i n_i = -15931.91 * 10^{-3}$$

$$\text{Third stage outlet temp} = 455^\circ\text{C} = 718 \text{ K}$$

$$\text{Fourth stage inlet temp} = 428^\circ\text{C} = 701 \text{ K}$$

$$\text{Temp change from } 718 \text{ to } 701 \text{ K}$$

Heat load is calculated by substituting  $T_1 = 718 \text{ K}$   $T_2 = 701 \text{ K}$

$$\begin{aligned}\Delta H_3 &= 97980.32(718 - 701) + ((-15931.91 * 10^{-3})/2) (718^2 - 701^2) \\ &= 1665665.44 + (-7.96) (24123) \\ &= 1665665.44 - 192019.08 \\ &= 1473646.36 \text{ kJ/hr} \\ &= 409.34 \text{ kW}\end{aligned}$$

## 5.3 EQUIPMENT DESIGN

### 5.3.1 Shell & Tube Heat Exchanger

Consider the reactants are flow on tube side and cooling water are on shell side

Catalysts are fill inside the tube.

Process Design:

$$\begin{aligned}\text{Consider length of tube} &= 5\text{m} \\ \text{Diameter of tube} &= 2.6 \text{ cm} \\ \text{Volume of one tube} &= \pi (d)^2 (L)^4 \\ &= \pi (2.6 \times 10^{-2})^2 (5)^4 \\ &= 2.65 \times 10^{-3} \text{ m}\end{aligned}$$

**TABLE 5.12 : PROPERTIES AT ARITHMETIC MEAN TEMPERATURE**

PROPERTIES	Shell Side (water) (30°C)	Tube Side ( SO <sub>2</sub> + H <sub>2</sub> O) (65°C)
C <sub>P</sub> ( Specific heat )	4180 kJ/kg K	4.185 kJ/kg K
μ( viscosity)	7.97 x 10 <sup>-4</sup> kg/m.s	4.33 x 10 <sup>-4</sup> kg/m.s
K ( Thermal conductivity)	0.616 W/m.K	0.659 W/m.K
ρ (density)	995.65 kg/m <sup>3</sup>	980.57 kg/m <sup>3</sup>

$$\text{No. of tube} = \frac{\text{Volume of one reactor}}{\text{Volume of one tube}}$$

$$\begin{aligned}&= \frac{2.57}{0.0265}\end{aligned}$$

No. of tube = 969 Nos.

Area of tube per pass:

$$\begin{aligned}A_{tp} &= \pi (d)^2 (N_{tp})^4 \\ &= \pi (2.6 \times 10^{-2})^2 (969)^4 \\ A_{tp} &= 0.51 \text{m}^2\end{aligned}$$

Velocity:

$$V = \frac{m}{\rho A}$$

$m = 12.98 \text{ Kg/Sec}$

$$V = \frac{12.98}{980.57 \times 0.51}$$

$V = 0.02 \text{ m/s}$

$$\begin{aligned}\text{Now, } N_{Re} &= \frac{dV\rho}{(1-\epsilon)} \\ &= (2.6 \times 10^{-2})^2 * (0.02) * (980.57) / (4.33 \times 10^{-4}) (1-0.8)\end{aligned}$$

$$N_{Re} = 1683.95$$

$$\begin{aligned}\text{Now, } A_o &= N_{tp} \times \pi \times d \times L \\ &= (969) \times 3.14 \times (2.6 \times 10^{-2}) \times (5)\end{aligned}$$

$$A_o = 396 \text{ m}^2$$

Shell diameter:

$$DS = 0.637 \sqrt{C_L/C_P [A_o (P_R^2 \times d_o) \div L]}$$

Consider the Triangular pitch

$$\begin{aligned}C_{TP} &= 0.8 \\ C_L &= 0.6\end{aligned}$$

Where  $C_{TP}$  = Tube count calculation constant

$C_L =$  Tube layout constant

Take  $P_R =$  Tube pitch ratio

$$= 1.26$$

$$D_s = 0.637 \sqrt{0.6/0.8 [395.54 \times (1.26)^2 \times (2.6 \times 10^{-2}) \div 5]^{0.5}}$$

$$D_s = 1 \text{ m}$$

Now, No. of tubes that can be accommodate

$$\begin{aligned} N_t &= 0.875 (C_{TP} / C_L) [ (D_s)^2 / (P_R)^2 (d)^2 ] \\ &= 0.875 (0.8/0.6) [ (1.103)^2 / (1.26)^2 (0.026)^2 ] \\ &= 1115 \end{aligned}$$

Shell side H.T.C :

$$h_o D_e / k = 0.36 (D_e G / \mu)^{0.5} (C_P \mu / K) (\mu_b / \mu_w)$$

For triangular pitch

$$D_e = 4 \sqrt{3/2} P_T^2 - \pi d^2 / 4 \div \pi d$$

$$\text{Pitch ratio } P_R = P_T / d$$

$$1.26 = P_T / 0.026$$

$$P_T = 0.0327 \text{ m}$$

$$D_e = 4(\sqrt{3/2} (0.0327)^2 - 3.14/4) \times (2.6 \times 10^{-2})^2 \div \pi \times 0.026$$

$$D_e = 0.018 \text{ m}$$

$$G_s = m \div AS$$

$$A_s = D_s \times C \times B \div P_T$$

Where  $A_s =$  Cross flow area of shell side  $B =$  baffle spacing

$$\begin{aligned} C &= P_T - d_o \\ &= 0.0327 - 0.026 \\ &= 0.0067 \end{aligned}$$

$$\begin{aligned} B &= 0.4 D_s \\ &= 0.4 \times 1 \\ &= 0.4 \text{ m} \end{aligned}$$

$$A_s = 1.103 \times 0.0067 \times 0.44 \div 0.0327$$

$$A_s = 0.099 \text{ m}^2$$

$$G_s = m \div A_s = 12.84 \div 0.099 = 129.69 \text{ kg/m}^2\text{sec}$$

From the above equation,

$$h_o \times 0.018 / 0.61 = 0.36 (0.018 \times 129.69 / 7.9 \times 10^{-4})^{0.5} \times (4.180 \times 10^{-3} \times 7.9 \times 10^{-4} / 7.9 \times 10^{-4} / 0.61)^{0.333}$$

$$h_o = 0.068 \text{ W/m}^2\text{K}$$

Tube side H.T.C:

$$Nu = 0.023 (NR_e)^{0.8} (Pr)^{0.4}$$

$$h_i d_i / k = 0.023 (1683.95)^{0.8} (cp\mu/k)^0$$

$$h_i \times 0.025 / 0.659 = 0.023 (1683.95)^{0.8} (4.185 \times 4.33 \times 10^{-4} / 0.659)^{0.4}$$

$$h_i = 21.8 \text{ W/m}^2\text{K}$$

Over all H.T.C

$$\begin{aligned} 1/U_o &= (1/h_o + 1/h_i) \\ &= (1/0.068 + 1/21.8) \end{aligned}$$

$$U_o = 14.77 \text{ W/m}^2\text{K}$$

Now,

Rate of H.T. is given by

$$Q = UA\Delta T$$

$$A = 396 \text{ m}^2$$

Tube side pressure drop :

Tube side pressure drop is given by

$$\Delta P / Z (g_c (\epsilon)^3 D_P \rho) \div (1-\epsilon)(G')^2 = 150 (1-\epsilon) / N_{re} + 1.75$$

Where

$\rho$  = density (kg/m<sup>3</sup>)

$\epsilon$  = effectiveness  $N_{re}$  = Reynolds no.

$$\Delta P/Z (1)(0.8)^3 (1.2 \times 10^{-3})(980.59) \div (1-0.8)(19.46)^2 = 150 (1-0.8)/1683.95 + 1.75$$

$$\Delta P/Z = 2 \text{ kN/m}^2$$

Shell side pressure drop

$$\Delta P/Z = f(GS)\{(Nb-1)\} (DS) \div 2\rho De$$

$$\phi_s f = \exp[0.57 - 0.19 \ln(\text{res})]$$

$$= \exp [0.57 - 0.19 \ln(25716)]$$

$$= 0.3997 \quad \phi_s = 1$$

$$N_b = L/B = 5/0.441 = 11$$

$$\Delta P = 0.3997 \times (129.69) \{(11.33-1)+1\} (1.103) \div 2 \times 980.57 \times 0.018 \times 1$$

$$= 18 \text{ N/m}^2 \text{ MECHANICAL DESIGN}$$

Internal pressure inside the reactor

$$P = 1.5 \text{ MPa} = 1.5 \text{ MN/m}^2$$

Design pressure  $P_d = 1.05 P$

$$= 1.05 \times 1.5 = 1.525 \text{ MN/m}^2$$

Vessel is IS: 2002-1962 class 2B vessels

So, allowable stress = 118 MN/m<sup>2</sup>

Welding joint efficiency factor  $J=0.90$

Shell design

Thickness of vessel based on internal pressure

$$t = \frac{PD}{2SE}$$

Where

P = Internal pressure

S= Allowable stress

t = wall material thickness

D = outside diameter

E = safety factor i.e joint efficiency in ASME code

$$= \frac{1.5 \times 1.103}{2 \times 118 \times 0.90} = 88 \text{ mm}$$

**TABLE 5.13 : SHELL AND TUBE HEAT EXCHANGER  
PROPERTIES**

Length of tube (l)	5m
Diameter of tube (d)	2.6 cm
Volume of one tube	0.0026 m
No.of tube (N <sub>tp</sub> )	969 nos
Area of tube per pass A <sub>tp</sub>	0.51 m <sup>2</sup>
Velocity (v)	0.02 m/sec
Reynolds no.(N <sub>Re</sub> )	1683.95
Shell diameter D <sub>s</sub>	1m
No.of tubes that can accommodate N <sub>t</sub>	1115
Shell side heat transfer coefficient h <sub>o</sub>	0.068 W/m <sup>2</sup> K
Tube side heat transfer coefficient h <sub>i</sub>	22.3 W/m <sup>2</sup> K
Overall heat transfer coefficient U	14.18 W/m <sup>2</sup> K
Rate of heat transfer area A	396 m <sup>2</sup>
Tube side pressure drop ΔP/Z	2kN/m <sup>2</sup>
Shell side pressure drop ΔP	18 N/m <sup>2</sup>
Baffle spacing	0.4 m
Cross flow area of shell side A <sub>s</sub>	0.099 m <sup>2</sup>



## 5.4 INSTRUMENTATION AND PROCESS CONTROL

Instruments are provided to monitor the various process variable .

The instrumentation can be pneumatic, hydraulic or electric. The instrumentation is required to measure temperature, pressure, flowrate, level, physical properties as density, pH, humidity, chemical composition etc

Flow Measurement :

- Types of flow meters available are orifice meter, venturimeter, pitot tube etc
- Depending on temperature and velocity condition the suitable meter is selected for measurement of flow rates and velocity.

Temperature Measuring Devices :

- Mercury in glass thermometer, bimetallic thermometer, pressure spring thermometer, thermocouples, resistance thermocouples, radiation pyrometers and optical pyrometers.
- Industrial thermocouples are competitively good as they provide large measuring range, without any error

Various pressure measuring devices are:

- U – Tube Manometer
- Differential Manometer
- Inclined Manometer
- Bourdon Tube
- Bellows
- Diaphragm valve
- Mc Leod gauge
- Pirani gauge

Liquid Level :

Liquid level is measured by direct process and indirect process.

#### **5.4.1 Valves**

It control the flow of liquid.

#### **5.4.2 Sensors**

It is used for process monitoring and controlling.

#### **5.4.3 Pumps and Pipes**

Device which is used to move the fluid, converting mechanical energy into hydraulic energy.

Pipes divided into three categories plastic pipes, cement pipe and steel pipes . Depending upon the process pipes should be selected .

#### **5.4.4 Process drawings**

- i. Block diagram
- ii. Process flow diagram (PFD)
- iii. Piping and instrumentation diagram (PID)

Block diagram

- Block diagram is the simplest form of presentation
- Each block can represent a single piece of equipment or complete stage process
- Block diagram is useful for representing a process in simplified form

Process flow diagram (PFD)

- PFD is one in which all incoming and outgoing material and utilities are shown

PFD includes

- i. Arrangement of equipments and their interconnections.
- ii. Operating condition of each stream such as temp, pressure and composition.
- iii. Heat added or removed in particular equipment.

Piping and instrumentation diagram (PID)

It show the engineering details of instruments, equipments, pipes and valves etc.

## 5.5 COST ESTIMATION

- ESTIMATION OF PRESENT PLANT COST

The cost of 10,000 TPA Sulphuric acid in 2000 was= Rs 60000000

The Chemical Engineering Plant cost index for Sulphuric acid in 2002=404

The Chemical Engineering plant cost index for Sulphuric acid in 2020 = 592

Therefore, the present plant cost can be given as

$$= \text{Rs } 60000000 * 592 / 402$$

$$= \text{Rs } 88358200$$

- DETAILED COST ANALYSIS

Direct cost = Purchased cost + Installation cost + Instrumentation and controls installation + Piping + Electrical installation + Building process and auxiliary + Service facilities + Yard improvement + Land

(1) Purchased equipment cost (15-40% of fixed capital investment).

Assume 35% of Fixed capital investment =  $0.35 * (88358200)$

$$= \text{Rs } 30925300$$

(2) Installation cost, including insulation and painting (25-55% of purchased cost)

$$\begin{aligned} \text{Assume 42\% of purchased cost} &= 0.42 \times (30925300) \\ &= \text{Rs } 12988600 \end{aligned}$$

- (3) Instrumentation and controls cost (6-30% of purchased cost) Assume 10% of purchased cost =  $0.10 \times (30925300)$

$$= \text{Rs } 3092500$$

- (4) Piping installation cost (10-80% of purchased cost)

$$\text{Assume 40\% of purchased cost} = 0.40 \times (30925300)$$

$$= \text{Rs } 12370100$$

- (5) Electrical costs (10-40% of purchased cost)

$$\text{Assume 17\% of purchased cost} = 0.17 \times (30925300)$$

$$= \text{Rs } 5257300$$

- (6) Building Process and auxiliary (10-70% of purchased cost)

$$\text{Assume 38\% of purchased cost} = 0.38 \times (30925300)$$

$$= \text{Rs } 11751600$$

- (7) Service facilities (30-80% of purchased cost)

$$\text{Assume 60\% of purchased cost} = 0.60 \times (3925300)$$

$$= \text{Rs } 18555100$$

- (8) Yard improvement (10-15% of total purchased cost) Assume 11% of purchased cost =  $0.11 \times (30925300)$

$$= \text{Rs } 3401700$$

- (9) Land (4-8% of purchased cost)

$$\text{Assume 6\% of purchased cost} = 0.06 \times (30925300)$$

$$= \text{Rs } 1855500$$

$$\text{Direct costs} = \text{Rs } 100197700$$

- Indirect costs = Engineering Supervision + Construction expenses + Contractor fee + Contingency plan

(1) Engineering supervision (15- 30% of direct cost)

$$\text{Assume 20\% of direct cost} = 0.20 \times (100197700)$$

$$= \text{Rs } 20039500$$

(2) Construction expenses and contractors feed (6-30% of direct cost)

$$\text{Assume 15\% of direct cost} = 0.15 \times (100197700)$$

$$= \text{Rs } 15029600$$

(3) Contractor fee (2-7% of direct cost)

$$\text{Assume 4\% of direct cost} = 0.04 \times (100197700)$$

$$= \text{Rs } 4008000$$

(4) Contingency costs (5-15% of direct cost)

$$\text{Assume 13\% of direct cost} = 0.13 \times (100197700)$$

$$= \text{Rs } 13025700$$

$$\text{Indirect costs} = \text{Rs } 52102800$$

- Fixed Capital Investment

$$= \text{direct costs} + \text{indirect costs}$$

$$= 100197700 + 52102800$$

$$= \text{Rs } 152300500$$

- Working Capital (10-20% of fixed capital investment)

Assume 16% of fixed capital investment

$$= 0.16 \times (152300500)$$

$$= \text{Rs } 24368000$$

- Total Capital Investment
  - = Total fixed Capital + Working Capital Investment
  - = Rs 176668500
  
- Estimation of Manufacturing Cost
  
- Fixed costs
  - 1) Depreciation
 

9% of fixed Capital investment for machinery and 5% of building value for building = Rs 14294600
  
  - 2) Local taxes (3-4% of fixed capital investment)
 

Assumed value 4% of fixed capital investment = Rs 6092000
  
  - 3) Insurance (0.4-1% of fixed capital investment)
 

Assumed value 0.8% of fixed capital investment = Rs 1218400
  
  - 4) Rent (8-12 % of the rented land and buildings) Assume 10% of building value = Rs 1175100
 

Fixed charges are 18% of Product cost =  $0.18 \times 60000000$

= Rs 10800000

Total Product cost = Rs 33581000
  
- Direct Production
  - 1) Raw Materials (10-15% of Total Product cost)
 

Assumed value 30% of Total product cost =  $0.3 \times (33580100)$

= Rs 10074030
  
  - 2) Operating labour (10-20% of total product cost)

Assumed value 15% of Total Product cost =  $0.15 \times (33580100)$   
= Rs 5037000

- 3) Direct Supervisory and Electrical labour (10- 15% of operating labour)

Assumed 12% of Operating labour =  $0.12 \times (33580100)$   
= Rs 3760900

- 4) Utilities (10-20% of total product cost)

Assumed value 13% of total product cost =  $0.13 \times (33580100)$   
= Rs 4365400

- 5) Maintenance and repairs (2-10% of Fixed Capital Investment)

Assumed value 8% of Fixed Capital Investment  
=  $0.08 \times (152300500)$   
= Rs 12184000

- 6) Operating supplies (10-20% of cost of Maintenance and repair) Assumed value 15% =  $0.15 \times (12184000)$

= Rs 1827600

- 7) Laboratory charges (0-20% of operating labour) Assumed value 15% of operating labour =  $0.15 \times (5037000)$

= Rs 755500

- 8) Patents and royalties (2-6% of Total Product cost)

Assumed value 5% of Total Product cost = Rs 1679000

Direct Production cost is sum of all the (1),(2),(3),(4),(5),(6),(7)and (8)

= Rs 39683400

- Plant Overhead costs

(50-70% of operating labour+ Direct supervisory + Maintenance cost)

Considering 60%, Overhead plant costs = Rs 12589100

Total Manufacturing Costs = Rs 85852600

- General Expenses

Includes administration expenses, distribution prices and also R & D costs.

A) Administration cost (40-60% of the Operating Labour cost)

Assuming 50%, the administration cost = Rs 2518500

B) Distribution costs (2-20 of total product cost)

Assume 10% of the product cost =  $0.10 \times (33580100)$   
= Rs 3358000

C) Research & Development cost:

Assume 3% of the production cost = Rs 1190500

General Expenses = A+B+C = Rs 7067000

- Total product cost

= Manufacture cost + general expenses  
= Rs 92919600

- Current selling price

Cost of Sulphuric acid = Rs 11/kg

Total selling price per annum

=  $(10 \times 103) \times 325 \times 11$   
=Rs 35750000/ annum

- Gross Earnings

= Total selling price – total product price  
= Rs 2169900/ annum



- Tax  
Assume the local tax rate 40%  
Tax on the gross earning =  $0.4 \times (2169900)$   
= Rs 867900
- Net Profits  
= Gross Earning – Tax  
= Rs 1302000
- Rate of return  
= Net Profit / Fixed Capital Investment  
= 18.54%
- Pay back period  
P.B.P = (Total Investment + Total Production Cost) / Profit After Taxes  
=  $(1766850 + 3968340) / 1302000$   
= Rs 4.4 Years

## 5.6 PLANT LAYOUT AND SELECTION

The selection of site to ensure the safety of both human and environment surrounded. In addition, the plant should be located in such way to get the raw materials and to distribute the finished products easily.

Sulphuric acid to be produced from the plant is expected to exceed the local market consumption hence the majority will be exported. Therefore the plant should be located near a harbour in order to reduce the transport cost of finished products .

Plant location

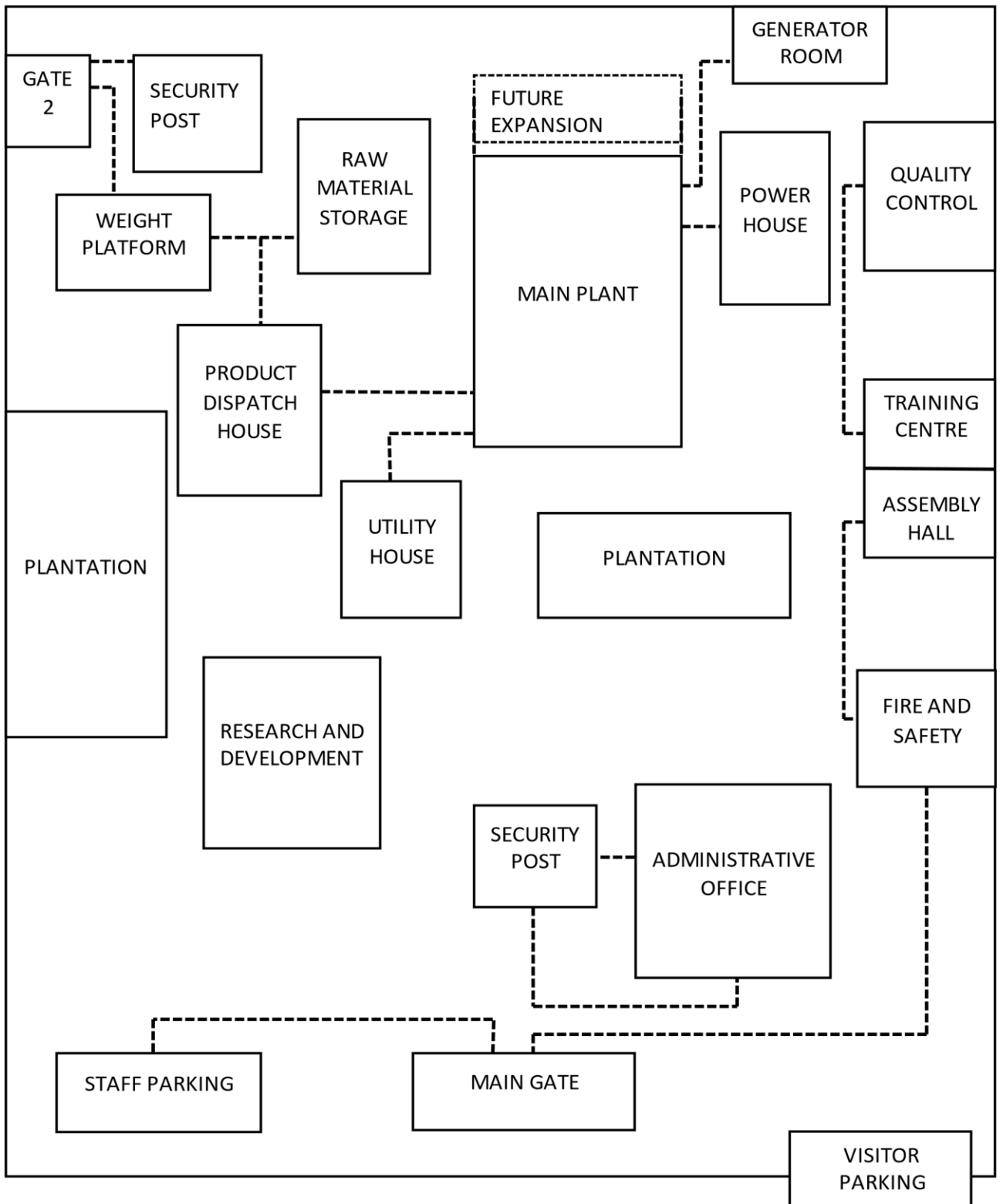
- i. Principle factors
- ii. Availability of raw materials

- iii. Nearness to potential market
  - iv. Supply of labour
  - v. Transport facilities
  - vi. Availability of utilities such as water , fuel and power
  - vii. Suitability of climate and land
  - viii. Local community consideration
  - ix. Political strategic consideration
  - x. Environmental impact and effluent disposal
- The primary objective of

#### Safe Plant Operation:

- Safe operating limits.
- To detect dangerous situations and to provide alarms
- Automatic shutdown systems.
- To prevent dangerous operating procedures.
- Production Rate and Quality:
  - To achieve the desired product output.
  - To maintain the product composition within the quality standards
- Cost : To operate at lowest production rate .

### 5.6.1 PLANT LAYOUT



**Figure 5.10 plant layout**

## 5.7 PLANT SAFETY HAZARD

Being a strong acid, mixing with water should be always exercised with caution recommended practice is to send the acid stream to water during dilution.

The dehydrating properties of concentrated acid apply even to organic materials, and so, can cause severe skin burns if made contact with skin.

Long term exposure at low concentrations with skin may dry up and redden skin while inhalation may damage teeth and enamel, and inflammation and irritation will be brought upon the respiratory system.

Personal Protective equipment in the form of respirators, breathing apparatus (for emergencies), dust tight goggles (Sulphur handling areas), safety glasses with side shields (Molten Sulphur storage areas), face shields, fire retardant clothing, heat resistant gloves (molten Sulphur storage area) are being used in the industry based on information gathered on safety aspect .

## CHAPTER 6

### SUMMARY AND CONCLUSION

The next step to the Contact Process is DCDA or Double Contact Double Absorption. In this process the product gases ( $\text{SO}_2$ ) and ( $\text{SO}_3$ ) are passed through absorption towers twice to achieve further absorption and conversion of  $\text{SO}_2$  to  $\text{SO}_3$  and production of higher grade sulphuric acid.  $\text{SO}_2$ -rich gases enter the catalytic converter, usually a tower with multiple catalyst beds, and are converted to  $\text{SO}_3$ , achieving the first stage of conversion. The exit gases from this stage contain both  $\text{SO}_2$  and  $\text{SO}_3$  which are passed through intermediate absorption towers where sulphuric acid is trickled down packed columns and  $\text{SO}_3$  reacts with water increasing the sulphuric acid concentration. Though  $\text{SO}_2$  too passes through the tower it is unreactive and comes out of the absorption tower.

This stream of gas containing  $\text{SO}_2$ , after necessary cooling is passed through the catalytic converter bed column again achieving up to 99.8% conversion of  $\text{SO}_2$  to  $\text{SO}_3$  and the gases are again passed through the final absorption column thus resulting not only achieving high conversion efficiency for  $\text{SO}_2$  but also enabling production of higher concentration of sulphuric acid.

The industrial production of sulphuric acid involves proper control of temperatures and flow rates of the gases as both the conversion efficiency and absorption are dependent on these.

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