PROCESS OPTIMIZATION FOR OLEUM PRODUCTION

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CERTIFICATE

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DEDICATION

This dissertation is dedicated to Allah Almighty for the guidance, strength and intellect, to our parents for their continuous support and encouragement and our friends without whose moral support, this could not have been possible.

ACKNOWLEDGEMENT

We would like to express our gratitude to the Project Supervisor- Dr Abdul Qadeer Malik and the Co Supervisor- Dr Erum Pervaiz for their continuous guidance, support and efficient supervision. Their timely feedback and coordination enabled us to put forward a project that is elaborate and accurate as per the required standards.

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ABSTRACT

Sulphuric Acid production and its optimization analysis construct an element of strength for global usage and facilities. It contributes to the provision of a number of different products and worldwide utility sectors. Its significance in flourishing a country's economy cannot be ignored. A close look into the insights of the subject chemical's production as well as imports/exports also explicitly depicts its immense importance and the role its plant may play for the betterment of industrial sector in a developing country like ours. Molten sulphur is the main constituent in the process and its adequate supply remains a vital factor here. The process route opted for is the Double Contact Double Absorption Process with Adiabatic Catalytic Converter for SO_2 to SO_3 conversion.

Optimization Studies have been elaborately incorporated to make the plant fully practical, cost effective, efficient and Green working facility. Maximum energy requirements of the industry are met by the plant's own expanse through Coolers. Complete and detailed design is annexed alongside a thoroughly integrated economic analysis and Simulation model to ensure that no aspect of the plant interposes any prospective ceasure or hazard.

The total capacity promised by the forwarded model is 4410 kg/hr of concentrated Sulphuric Acid. This model serves a dual functional purpose i.e. formation of both Oleum and Sulphuric Acid as per the requirement.

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CHAPTER 1

Introduction

1.1 Background

1.1.1 Global Market Demand

Sulfuric acid is one of the world's highest-production chemical produced. It is one of the parameters that denote economic stability of a country. The worldwide demand for sulfuric acid had been predicted to rise at a rate of 2% annually for 5 years since 2017, with China being its largest consumer (37%) as shown in Figure 1.1.

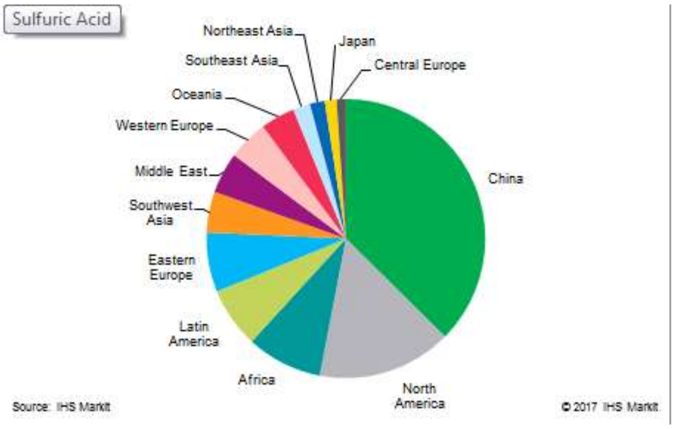


Figure 1-World consumption of sulfuric acid

1.1.2 Sulfuric Acid Production in Pakistan

The monthly produce of sulfuric acid in Pakistan was estimated to be around 4000 tons in August 2018 which was an increase from the production capacity in July 2018. Figure 1.2 shows the monthly production of sulphuric acid in thousand tons in Pakistan as reported by Pakistan Bureau of Statistics.

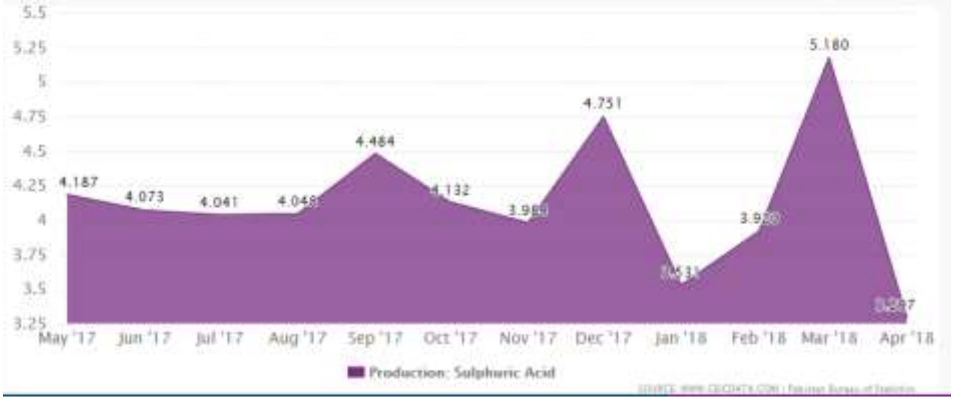


Figure 2-Sulfuric acid production in Pakistan

The change in annual production of sulfuric acid in Pakistan is given below.

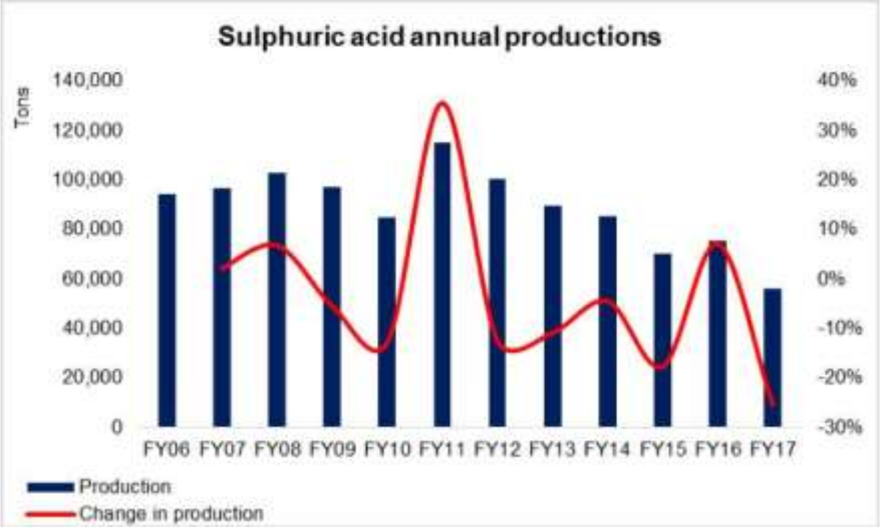


Figure 3-Sulfuric acid annual production in Pakistan

The production capacity of Pakistan has seen fluctuations which can be owed to its economic instability over the years. However, Pakistan is considered to be self-sufficient in production of sulfuric acid, with excess acid exported as well.

1.1.3 Problem Statement

The project put forward was a complex insightful design demanding detailed literature review and analytic investigation of all processes for Sulphuric Acid Production to come up with a fully workable design. Variant process routes had to be compared and looked into comprehensively to decide one that best suited the provided requirements and objectives.

The plant had to be a dual functional facility for both Sulphuric Acid and Oleum Production as per requirements. Moreover, it had to be green, optimized and efficient as a process. Additionally, a complete economic analysis of the plant had to be performed thus proving it to be cost effective and profitable on the whole.

1.1.4 Purpose of the study

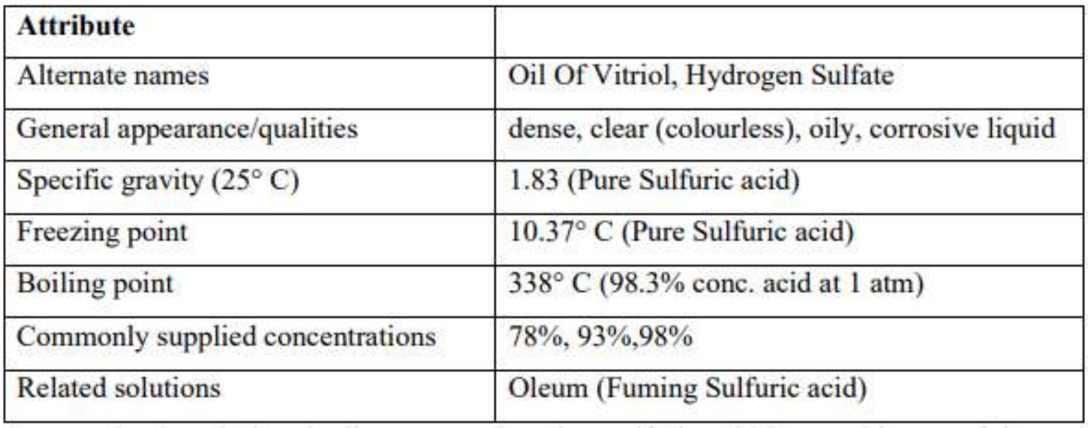
The purpose of this study is to Design and Optimize an operational industrial facility for Oleum Production with efficient workability and maximum chemical purity as well as conversion efficiency.

1.1.5 Organization of the study

For the efficient conversion of SO_2 to SO_3, the suggested method involves installation and design of an adiabatic catalytic packed bed reactor with inter-stage cooling to keep the temperature uniform throughout the reaction vessel. Moreover, as a consequence of presence of coolers in between beds, catalyst deactivation is avoided, thus saving running cost of the plant and making the process economically feasible. The chosen catalyst is Vanadium Pentaoxide (V_2O_5) which is not only low on the plant investment but also allows the equilibrium reaction to proceed more smoothly and in a better fashion. The subject reaction is an exothermic one and requires a pressure of around 1-2 atm. The final outlet of reactor i.e. that of bed 4 and that of bed 3 go to the Main Absorption Tower and Intermediate Absorption Tower respectively where absorption of SO_3 into Sulphuric Acid takes place to form Oleum. These absorbers are also packed bed reactors governing an exothermic reaction. The reactor itself and the absorbers along with all other equipment before and after the reactor-absorber assembly were elaborately designed and checked for their economic and environmental feasibility. This was completed by analytical calculations and verified by simulation done in Aspen HYSYS.

1.1.6 Physical properties of sulfuric acid

Table 1-Physical property data for sulfuric acid



Attribute	
Alternate names	Oil Of Vitriol, Hydrogen Sulfate
General appearance/qualities	dense, clear (colourless), oily, corrosive liquid
Specific gravity (25° C)	1.83 (Pure Sulfuric acid)
Freezing point	10.37° C (Pure Sulfuric acid)
Boiling point	338° C (98.3% conc. acid at 1 atm)
Commonly supplied concentrations	78%, 93%,98%
Related solutions	Oleum (Fuming Sulfuric acid)

1.1.7 Applications of sulphuric acid

It finds its major use in the production of fertilizers especially phosphate fertilizers accounting for over 60% consumption of sulfuric acid produced worldwide. According to a 2017 report, 17% of sulfuric acid is used in approximately 20 chemical processes such as processing and metallurgical applications worldwide.

Other uses include manufacturing of pigments, fertilizers, dyes, explosives, drugs, inorganic salts, acids and detergents as well as petroleum products. It is also employed as the major electrolyte in Lead Storage Batteries and as a strong dehydrating agent.

- Production of Phosphoric acid:

This acid is produced using the wet process (major source of consumption of our subject acid) and is subsequently used in producing phosphate fertilizers

- Iron and steel manufacturing:

Sulfuric acid is used to remove iron oxide flakes, scaling or any other layers formed due to oxidation from rolled sheets. To avoid costly purchase of fresh acid, most steel manufacturing units have a regeneration unit for spent sulfuric acid.

- Manufacture of Aluminum compounds:

Sulfuric acid produces gelatinous aluminum compounds which are beneficial in setting of the pulp fibers to produce hard paper surface. Sulfuric acid is also used in manufacturing other coagulants such as aluminum hydroxide.

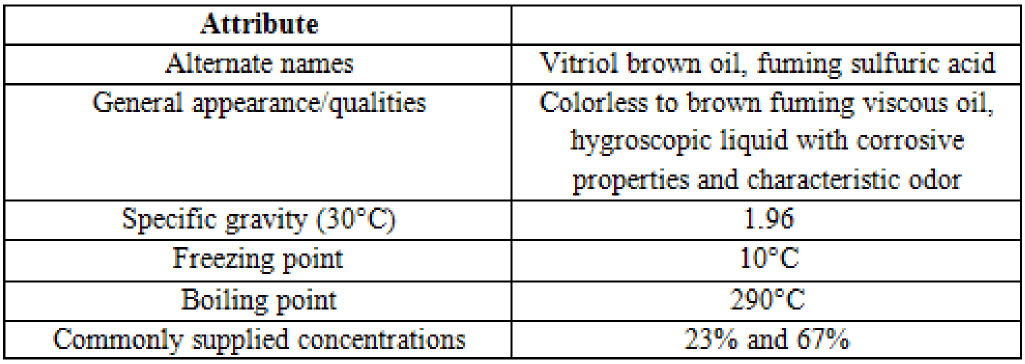
- Manufacture of Ammonium Sulfate:

It is a vital fertilizer used in agriculture industry and its mass production is through a reaction of ammonia with sulfuric acid.

$$2NH_3 + H_2SO_4 \rightarrow (NH_4)_2SO_4$$

1.1.8 Physical properties of oleum

Table 2-Physical properties of oleum



Attribute	
Alternate names	Vitriol brown oil, fuming sulfuric acid
General appearance/qualities	Colorless to brown fuming viscous oil, hygroscopic liquid with corrosive properties and characteristic odor
Specific gravity (30°C)	1.96
Freezing point	10°C
Boiling point	290°C
Commonly supplied concentrations	23% and 67%

1.1.9 Applications of oleum

Oleum is considered a vital intermediate in Sulphuric Acid manufacture process owing to its high value of enthalpy of hydration.

It has several uses such as:

- Transportation Intermediate:

It is commonly employed for use as a form of transporting Sulphuric Acid products via freight cars etc. due to the fact that a few of its compositions are solid at room temperature and hence propose a convenient and safer way to transport.

- Research in Organic Chemistry

Being a conclusively hard reagent, it is excessively used for research purposes in organic chemistry sector.

- Manufacture of Explosives:

Oleum is frequently used in the manufacture of variant explosives.

- Strong Dehydrating Agent

On being poured onto powdered glucose, it draws all water inherent constituents from within via an exothermic reaction just leaving carbon black behind as a pure solid.

CHAPTER 2

Literature Review and Process Description

2.1 Plant Site Selection and Layout:

Plants for Acid production are located well across the industrial world majorly close to the products' usage points such as in near vicinity of copper ore leach plants, fertilizer plants and refineries for petroleum. A vital reason for this is the lower transportation cost of elemental Sulphur as compared to Sulfuric Acid.

The selection of the plant plays a very significant role in many aspects of the plant processing as well as the business and the safety concerns handling. It makes sure of the safety of both the environment and humans. Location should be chosen in a way that the delivery of materials and product distribution can be accomplished with ease.

Produced acid from the plant can exceed consumption in local market and thus majority is to be exported. Plant location should be close to harbors for reduction in transportation cost of the product acid. For this case, Karachi, Gwadar and Bin Qasim areas are well-suited candidate areas if proximity to a harbor is the concern.

However, a clear priority must be delivered to the fact of raw material supply refinery and the location for plant being close to it. This hence reduces the cost for raw material, predominantly Sulphur here, which is in molten state. Long Distances can excessively complicate the matter else wise.

The plant layout diagram after the selection of the site can then be drawn on rough basis to get an idea. (Kumareswarn, 2013)

2.2 Existing Routes for Sulfuric Acid Production

After a detailed literature review and keen consideration of all options available as well as occurrences expected, two main processes are observed to be present for Sulphuric Acid and Oleum Production across the globe.

1. Lead Chamber Process
2. Contact Process

Certain variations are introduced in both these processes depending on the requirements and capacity of the plant.

2.2.1 General Overview

For an apt comparison between the two processes, a general overview of the two is given below to get a better understanding and which eventually leads to the selection of a process that meets the requirements and needs of SRC in terms of feasibility, economics and all other aspects.

2.2.1.1 Lead Chamber Process

This is a rather obsolete process, used for the production of sulphuric acid in chambers that are lined with Lead, mainly because of its corrosion resistant properties. This lead lining does not only contain the sulphuric acid being produced but it is also responsible for making the Wooden chambers waterproof.

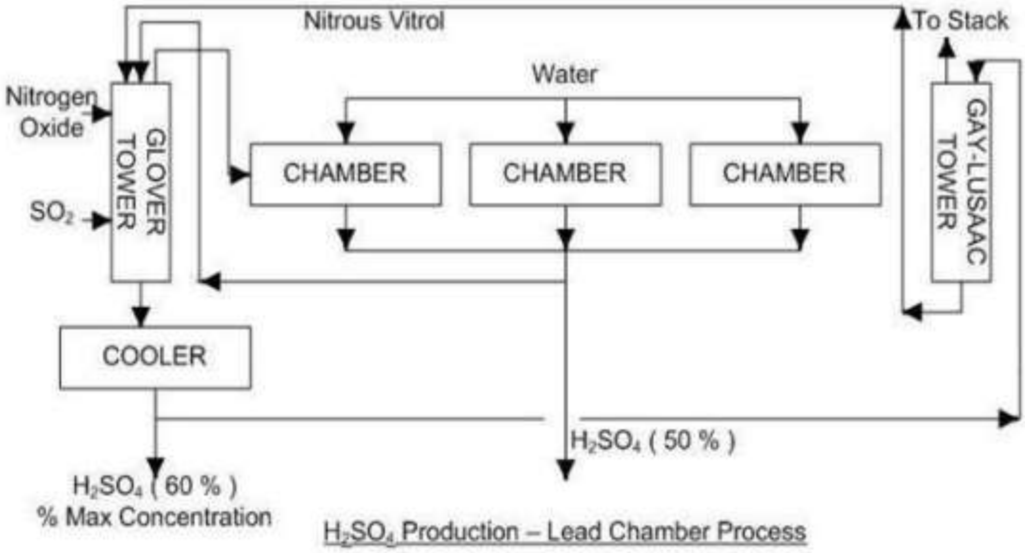


Figure 4-Lead Chamber Process-Process Flow Diagram

Steps involved:

- Burning of elemental sulphur or sulphur ore to produce SO_2

$$S + O_2 \rightarrow SO_2$$

- KNO_3 is burnt in the KNO_3 burner to produce nitrogen oxides.

- The products from both burners are filtered and sent to the three main towers one by one firstly in the glower tower followed by the Lead chamber and finally entering the Gay-Lussac Tower.
- In glower tower, Nitrogen oxides and Sulphur dioxide flow counter currently where NO_2 is absorbed in SO_2.

$$NO_2 \ + \ SO_2 \ + H_2O \rightarrow HOSO_2NO(OH)$$

- The gas from the glower tower is sent to the Lead chambers where the reaction mixture will go through a series of reactions. Due to the complexity of this process, chambers can range from three to twelve. Condensation of the acid produced occurs on the walls as well as the floors of this particular chamber. The final step taking place is as follows:

$$2HOSO_2ONO + H_2O \ \rightarrow HNO_2 + H_2SO_4$$

- The gases are then moved onto the tower called Gay-Lussac. This reactor removes any traces of acid gases. These waste gases must be treated before being purged to the atmosphere.

Some improvements have been introduced in the lead chamber from time to time, which included the introduction of packed towers, but the major process remained the same.

2.2.1.2 Contact Process

The main steps taking place in this process are as follows with optimization within each step to bring about better results.

- Sulphur Burning in the Presence of Oxygen present in Air.

$$S + \ O_2 \ \rightarrow \ SO_2$$

- Production of Sulphur Trioxide by reacting Sulphur Dioxide with oxygen in the presence of a catalyst and optimum operating conditions.

$$2SO_2 \ + O_2 \ \rightarrow \ 2\,SO_3$$

- The next step is the absorption of Sulphur Trioxide in the solvent, which is Sulphuric Acid to produce Oleum.

$$SO_3 + H_2SO_4 \rightarrow H_2S_2O_7$$

- Water is then added to Oleum to produce Conc. Sulphuric Acid.

$$H_2S_2O_7 + H_2O \rightarrow 2H_2SO_4$$

2.3 Comparison between Lead Chamber and Contact Process

2.3.1 Economic Parameters

2.3.1.1 Product concentration and Quality

Final product obtained from lead chamber process has a final concentration of 65 – 78% whereas that in contact process has a minimum concentration of 98% due to the catalytic activity of the catalyst present in the reactor. Other disadvantages of the Lead Chamber process include a smaller throughput as well as lowered quality of the Sulphuric Acid produced due to the presence of impurities. Contact process, is thus a better option in terms of both quality and concentration of the product.

2.3.1.2 Operating Cost

A lower pressure (1-2 atm) is required in the contact process, at these pressures which are only marginally greater than the atmospheric conditions, a sufficient target yield is obtained and the equipment cost is lower than the expected cost.

2.3.1.3 Energy Consumption

In terms of this parameter, both processes need the same amount of energy for the same amount of Sulphur burnt for the production of Sulphur Dioxide for further usage. However, in the lead chamber process additional burners are required to burn KNO_3 so it can be said that the Lead chamber process has a greater energy requirement. Additionally, those 3-12 chambers would also need a greater amount of energy as compared to a smaller amount of converters and absorbers in the Contact process

2.3.1.4 Power Consumption

Contact process is much more compact and requires less power requirement in comparison to the Lead Chamber process that has large gas and liquid conveying systems requiring a huge amount of air blowers and pumps owing to larger power consumption.

2.3.2 Safety Aspects

Emission of oxides of Sulphur is common in both processes due to the burning of Sulphur. However, in the chamber process additional nitrogen oxides are also emitted due to burning of KNO_3. It can be concluded that the contact process has a lesser impact on the physical environment as compared to the obsolete lead chamber process.

Table 3-Comparison between Lead Chamber and Contact Process

Aspect	Contact Process	Lead Chamber Process
Product Quality	Better Quality of Sulphuric Acid with lesser impurities	Lower Quality of Sulphuric Acid with many impurities.
Product Concentration	Min 98% conc. H_2SO_4	Max 78% conc. H_2SO_4
Pressures	Low Pressures	Comparatively higher Pressures
Energy consumption	Low energy consumption	Higher energy consumption due to additional chambers and burners
Power Consumption	Lower consumption of power	High consumption of power due to more air blowers and pumps
Operating Cost	Lower cost of operation	Higher Cost of operation
Environmental impact	Lower impact on environment	Higher impact on environment

On the basis of the aforementioned parameters, contact process is considered as the more efficient and economically feasible process.

2.4 Process Selection

An elaborate process selection, therefore, requires identification of the minute variation within the existing contact process. Based on these variations, the contact process can be further categorized into the following three types:

- Wet Contact Process
- Single Contact Single Absorption
- Double Contact Double Absorption

Table 4-Comparison between types of contact process

Process Type	Advantages	Disadvantages
Single Contact Single Absorption (SCSA)	• There are less heat losses due to absence of second absorption column	• High amount of SO_2 is discharged in flue gas due to absence of one absorption column • Impure SO_2 can poison the catalyst • It can only be used if the SO_2 content is less (3-10%) • The conversion of SO_2 is only 98%
Double Contact Double Absorption (DCDA)	• Higher conc. of sulphuric acid • Higher yield of sulphuric acid can be obtained • Most of SO_2 and SO_3 is converted (99.8%) due to second absorption column. Hence waste gases are less toxic.	• Any impurities in SO_2 can poison the catalyst

Wet Contact Process (WCP)	• Due to high sulfur recovery, heat recovery and low cooling water consumption, it is a more cost-efficient process	• The concentration of sulphuric acid produced is low • Sulfuric acid mist generation renders it as less suitable process

2.5 Process Description

2.5.1 Sulphur Burner

Assuming that maximum purity of sulphur is obtained from the sulphur burner; two different methods for the production of Sulphur Dioxide from Sulphur can be employed. Similarity between the two is the requirement of very large heat content for sulphur melting.

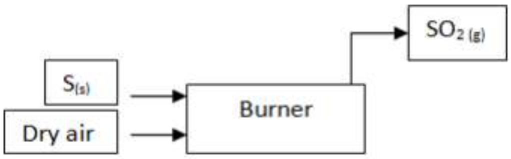


Figure 2.2 Burner Block Flow Diagram-1st method

First Method: Melting of Sulphur can be done directly in the sulphur burner along with air and with the application of heat. In this particular method melting of Sulphur and its reaction with air to produce Sulphur Dioxide will be taking place in the same space. However, this method has certain disadvantages including reduced efficiency along with hindrance in the operation leading to a temperature drop, and it may also lead to other hazards and finally cease in operation functionality.

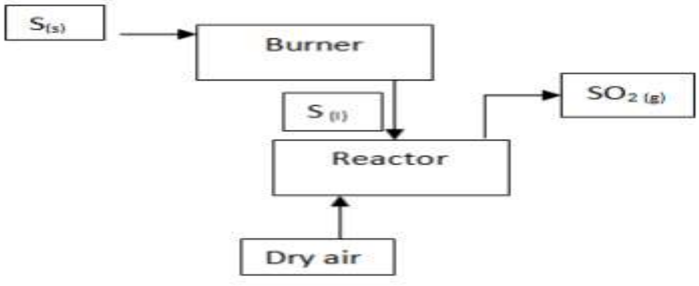


Figure 2.3 Burner Block Flow Diagram-2nd method

Second method: The performance of Sulphur melting occurs separately while excluding any contact of air, which causes oxidation of Sulphur Dioxide. Any forsake of required value of efficiency is very unlikely in the setup proposed hereby.

2.5.2 Air Dryer

One unit supplies the needed route wherever it may be needed. Drying and filtering of the air is a must as any contamination in the air supply due to the presence of moisture of any sort or dust particles may lead to a noticeable lowering of efficiency of conversion and later on a very huge operation cost for utilities in addition to any excessive price paid for replacing the catalyst.

Dehydrating agent used:

Concentrated sulphuric acid may be used for the drying of air in theory. However, oleum as a dehydrating agent is preferred in our proposed process. Firstly, because of its better ability to absorb moisture. And, secondly because using the final product i.e. concentrated Sulphuric Acid is not appropriate for parts where direct contact in the upstream processes is required. So, the dehydrating agent chosen is oleum which is recycled from within the plant via the oleum header.

2.5.3 Gas (SO_2) dryer

At this stage, existing routes deviate from one another. Air oxidation of molten Sulphur produces Sulphur Dioxide, which may be inherent of ash particles and some water content. The removal of which is important, this helps to assure that the operation proceeds without any hindrance, and the reactor operation is improved due to absence of water vapor. A close to perfect value for efficiency is desirable and needed. For gas cleaning an electrostatic precipitator maybe used to due to its higher efficiency and because it can operate at high temperatures. The dehydrating agent chosen in this part is also oleum because of the same reasons that have been mentioned in the air dryer.

2.5.4 Catalyst Aided Converter

This is a catalytic reactor, more generally, a packed bed reactor with four beds of catalyst. Since an exothermic reaction is taking place in the reactor, the temperature of products is assumed to increase as the exit each bed. Intermittent cooling arrangement is made between the beds for this purpose and this task is fulfilled by coolers in between. A temperature of 450 Degrees Celsius is ensured for each bed with an overall pressure of about 1 to 2 atm.

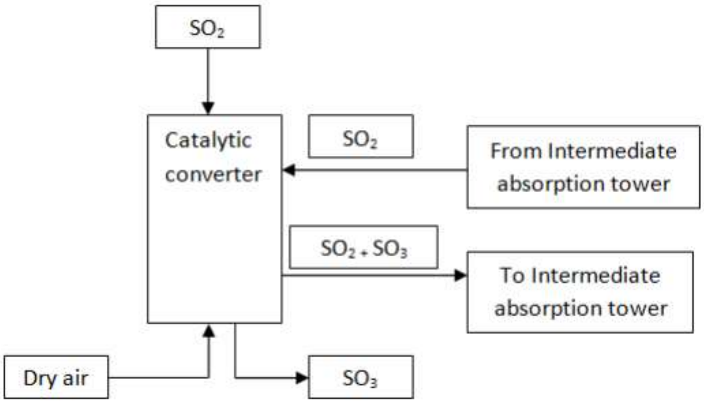


Figure 2.4 Reactor Block Flow Diagram

2.5.4.1 Choice of catalyst

The catalyst plays a very important role in converting Sulphur Dioxide to Sulphur Trioxide taking place in the catalytic reactor. The two catalysts considered are platinum and Vanadium pentaoxide.

Platinum has two main disadvantages, firstly the cost of acquisition is very high and thus it increases the capital and operating cost by a huge extent. The increase in operating cost is due to the regular maintenance needed.

On the other hand, vanadium pentaoxide is a better candidate for the catalytic activity-taking place in the reactor beds. Vanadium pentaoxide has many advantages including its lower tendency of being poisoned due to contamination Secondly; it has a lower cost of acquisition due to a lower capital and operating cost (This is because it does not need very frequent replacements due to reduced tendency of poisoning).

Hence, the catalyst chosen for our process is vanadium pentaoxide because of its economic feasibility. It is also important to state that while conducting the literature review, this catalyst has always been preferred as a better option which also clearly shows the merits of vanadium pentaoxide over all other catalysts that are used for this conversion in the reactor.

2.5.5 Intermediate Absorption tower (IAT)

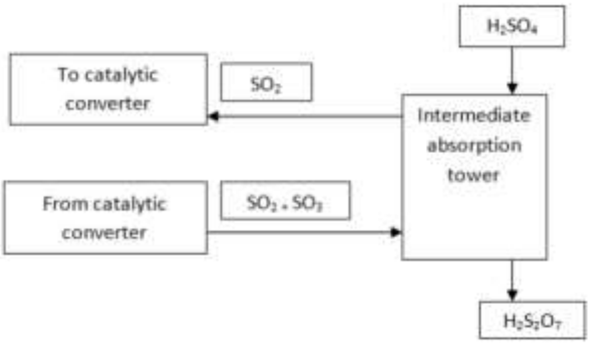


Figure 2.5 IAT Block Flow Diagram

Right after the catalytic reactor, the intermediate absorption tower has been inserted in our proposed process flow. The main reaction-taking place in this intermediate absorption tower is the chemical absorption of Sulphur Trioxide gas in the solvent which is concentrated sulphuric Acid to form oleum which is then sent to the oleum header. The sulphuric acid for this absorption is being recycled from the storage tank.

2.5.6 Main Absorption Tower (MAT)

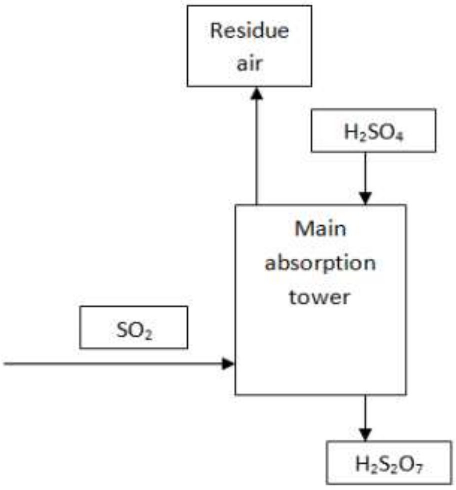


Figure 2.6 MAT Block Flow Diagram

The remaining absorption of Sulphur Trioxide gases in our solvent occurs in the main absorption tank. The exothermic absorption-taking place in this tower might need a heat exchanger for maintenance of the temperature at an optimum value for considerable absorption to take place. This is also done for the prevention of excessive sulphuric mist formation which will be sent to the scrubber for removal.

2.5.7 Scrubber (Flue gas treatment)

Gas coming from the main absorption tower contains Sulphur oxides as well as the sulphuric mists formed, this scrubber is required to remove these oxides that are harmful for the environment in order to make sure that the process that we are proposing is clean and green.

2.5.8 Oleum Tank

Oleum produced in both intermediate and absorption tower needs a tank to hold it. This storage/holding tank can also be the starting point for the recycle of Oleum back to the air dryer and Sulphur Dioxide dryer where is being used as a dehydrating agent as mentioned earlier. It is also a point for the distribution of Oleum to the dilution tank and for the extraction of oleum if needed. Since, the plant that is being proposed is dual functional, it means that sometimes oleum will have to be extracted as a product. This extraction of oleum as a product for further use will be done from this oleum tank.

2.5.9 Dilution Tank

The oleum coming from the oleum tank is converted to Sulphuric Acid which is also one of our final products, in this Dilution tank. Mass flow rate of processed water entering the dilution tank depends on the amount of oleum that needs to be converted. These dilution tanks must have agitators or mixers to facilitate and ensure homogenization as well as proper mixing of both oleum and processed water to form our product, Sulphuric Acid which is sent to the storage tank where part of it is extracted as a final product while the remaining Sulphuric Acid is recycled back to the absorption towers.

2.6 Process Flow Sheet

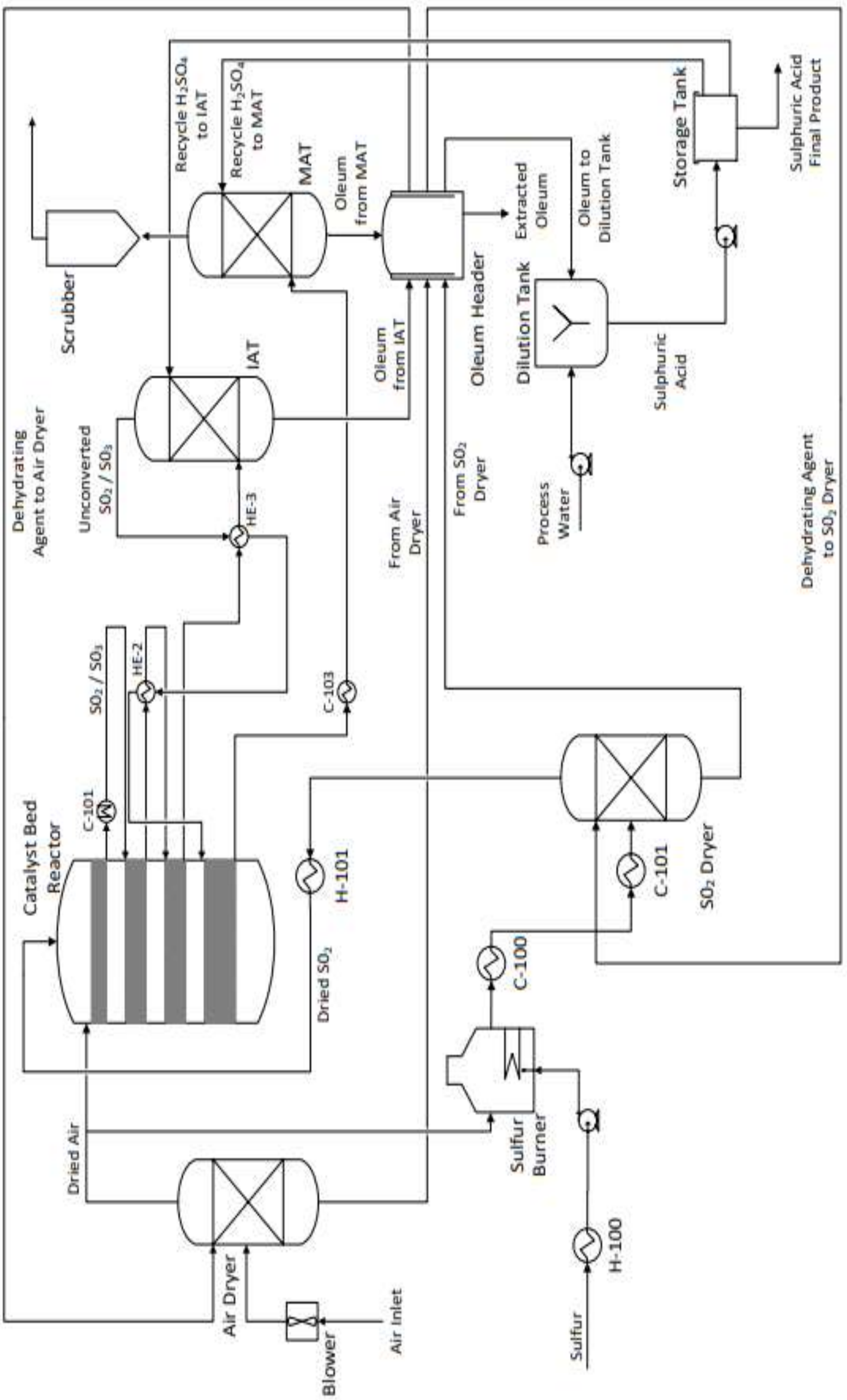


Figure 5-Process Flow sheet

CHAPTER 3

Material Balance

The material balance of this plant was organized and prepared for the operation under steady state conditions (Accumulation was assumed to be zero) with a basis of 1 hour.

The basic material balance equation is as follows:

Mass in + Mass generated = Mass out + Mass consumed + Accumulation

For the preparation of this balance, a set of variables or parameters were recognized. Allocation of provisions was done based on their necessity in equipment. These provisions along with the overall provisions helped us in coming up with realistic figures for the overall balance of our Sulphuric Acid and Oleum plant. (Watson, 1954)

3.1 Physical Properties

The only physical property that was required for preparing the material balance of our plant were the molecular weights of the materials being used and produced during plant operation. These molecular weights are important for the conversion of mass flows to molar flows or vice versa. The molecular weights have been mentioned below in Table 3.1

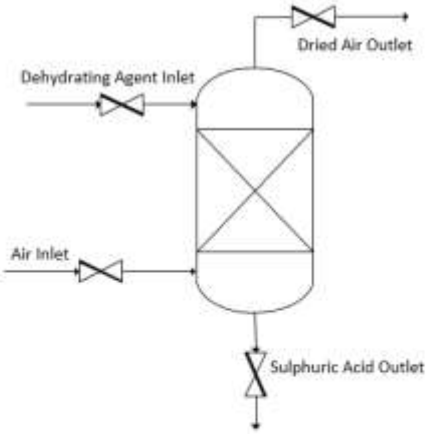
Table 5-Molecular weights of components

Species	Molecular weights
Sulphur	32.0
Sulphur Dioxide	64.1
Sulphur Trioxide	80.1
Oleum	178
Sulphuric Acid	98.0
Water	18.0

Hydrogen Sulphide	34.1
Oxygen	31.9
Nitrogen	28.0
Carbon Dioxide	44.0

3.2 Equipment wise Material Balance

3.2.1 Air dryer



The air entering the plant needs to be dried because if H_2O is a part of the combustion gas, there is a very high likelihood of its combination with SO_3 to form H_2SO_4 mist, which will obviously hinder the downstream process. This drying will be done in a packed tower with oleum as the dehydrating agent. The following reaction will take place in the tower for the removal of H_2O:

$$H_2O\ (l)\ +\ H_2S_2O_7(aq) \rightarrow 2H_2SO_4\ (aq)$$

Assumptions

- Humidity considered in order to estimate the content of water vapor in the inlet stream is 0.027125 kg (H_2O/dry air)
- The air is properly filtered before entering the air dryer or blower for efficient operation.

Table 6-Air Inlet flow sheet

Air inlet	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.75	0.72	0.74
	O_2	2915	91.09	0.20	0.18
	CO_2	667	15.16	0.05	0.03
	H_2O	376.6	20.92	0.03	0.04
Total		14255.6	494.93	1.00	1.00

Table 7-Dehydrating agent inlet flow sheet

Dehydrating Agent Inlet	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	3282.9	18.4	0.93	0.89
	H_2SO_4	228.5	2.33	0.07	0.11
Total		3511.4	20.8	1.00	1.00

According to the balanced chemical equation mentioned above, the following material balance is obtained:

Table 8-Dried Air Outlet flow sheet

Dried Air	**Components**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.8	0.74	0.78
	O_2	2914	91.1	0.21	0.19
	CO_2	667	15.2	0.05	0.03
Total		13878	473.9	1.00	1.00

Table 9-H_2SO_4 Outlet flow sheet

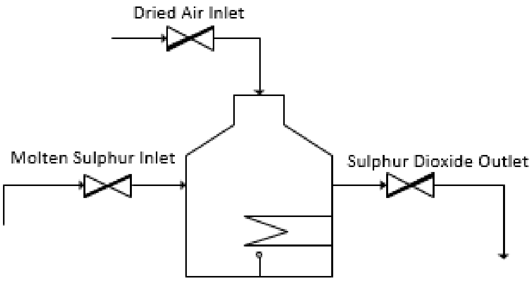
H_2SO_4 Outlet	Component	Mass flows (kg/hr)	Molar flow (kgmol/hr)	Composition (wt/wt %)	Composition (mol/mol %)
	$H_2S_2O_7$	0.00	0.00	0.00	0.00
	H_2SO_4	3843.5	39.2	1.00	1.00
Total		3843.5	39.2	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this equipment remained conserved.

In	Out
17767.1	17721.5

3.2.2 Sulphur Burner

Moving onto the next equipment, which is the sulphur burner. Combustion of molten sulphur to SO_2 takes place in the sulphur furnace according to the following reaction:



$$S(l) + O_2(g) \rightarrow SO_2(g)$$

Assumptions

- Conversion in the sulphur burner is assumed to be 100 %.
- The N_2 gas in the inlet air stream is not making any reactions in the furnace.

Table 10-Molten sulphur inlet flow sheet

Molten Sulfur	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	S	1443	45.1	0.99	0.99
	H_2S	14.60	0.43	0.01	0.01
Total		1457.6	45.5	1.00	1.00

Table 11-Dried Air inlet flow sheet

Dried Air	**Component**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	7568	270.3	0.74	0.78
	O_2	2142	66.9	0.21	0.19
	CO_2	489	11.1	0.05	0.03
Total		10199	348.3	1.00	1.00

According to the balanced chemical equation shown above, the material balance on the outlet streams is as follows:

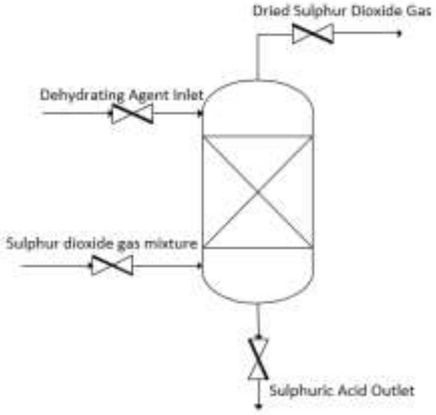
Table 12-SO_2 Outlet flow sheet

SO_2 gas	**Component**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	7568	270.3	0.65	0.77
	CO_2	489	11.1	0.04	0.03
	O_2	700.1	21.9	0.06	0.06
	SO_2	2884	45.1	0.25	0.13
	H_2O	17.7	0.98	0.00	0.00
	H_2S	14.6	0.43	0.00	0.00
Total		11673.4	349.7	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this equipment remained conserved.

In	Out
11656.6	11673

3.2.3 SO_2 Dryer



This is an absorption tower, which helps in the removal of water content in the SO_2 stream coming from the Sulphur burner. This is necessary, because if this stream enters the reactor, it may eventually lead to process failure due to the poisoning of the catalyst by H2O. Just like the air dryer, the dehydrating agent in this case is also Oleum that is being recycled from the plant. The following reaction will be taking place.

$$H_2O\ (l)\ +\ H_2S_2O_7\ (aq) \rightarrow 2H_2SO_4(aq)$$

Assumptions

- The conversion in SO2 dryer is assumed 100 %.

Table 13-SO_2 inlet flow sheet

SO_2 gas	Components	Mass flows (kg/hr)	Molar flow (kgmol/hr)	Composition (wt/wt %)	Composition (mol/mol %)
	N_2	7568	270.3	0.65	0.77
	CO_2	489	11.1	0.04	0.03
	O_2	700.1	21.9	0.06	0.06
	SO_2	2884	45.1	0.25	0.13
	H_2O	6.91	0.98	0.00	0.00
	H_2S	14.6	0.43	0.00	0.00
Total		11662.6	349.8	1.00	1.00

Table 14-Drying Agent Inlet flow sheet

Drying Agent	**Components**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	321.9	1.81	0.82	0.72
	H_2SO_4	69.1	0.71	0.18	0.28
Total		391.1	2.51	1.00	1.00

According to the balanced chemical equation shown above, the material balance on the outlet streams is as follows:

Table 15-Dried SO_2 gas Outlet flow sheet

Dried SO_2 gas	**Component**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	7568	270.3	0.65	0.77
	CO_2	489.9	11.1	0.04	0.03
	O_2	701.8	21.9	0.06	0.06
	SO_2	2884	45.1	0.25	0.13
	H_2O	0.00	0.00	0.00	0.00
	H_2S	14.6	0.43	0.00	0.00
Total		11658.4	348.8	1.00	1.00

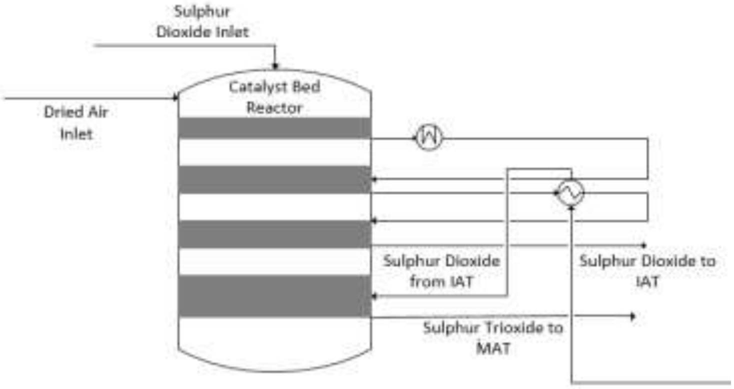
Table 16-H_2SO_4 Outlet flow sheet

H_2SO_4 outlet	**Components**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	195.5	1.99	0.58	0.71
	$H_2S_2O_7$	144.3	0.81	0.42	0.29
Total		339.8	2.81	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this equipment remained conserved.

In	Out
12054	11998

3.2.4 Reactor



In the design of the reactor, we have four packed beds; the catalyst being used is V_2O_5. The air that was previously filtered and cleaned enters the reactor where it reacts with the SO_2 gas mixture to produce SO_3. The first three beds will be considered in series whereas the fourth bed is required to convert the SO_2 that is being recovered from IAT. Intermediate heat exchangers are present between the beds to maintain an optimum temperature for an efficient conversion of reactants to products. We will be dealing with all these reactor beds one by one. The chemical reaction-taking place on all four beds is as follows:

$$2SO_2(g) + O_2(g) \rightarrow 2SO_3(g)$$

Assumptions:

- It is assumed that only conversion of SO_2 is taking place in these reactor beds since all other gases available in the reactor are thermally stable in these operating conditions.
- Overall conversion on each bed is as follows:
  1st bed = 63%
  2nd bed = 84%
  3rd bed = 93%
  4th bed =99.7%

However, conversion on each bed except the first bed is different from the overall conversion and is calculated as follows:

<u>For 2nd bed</u>

Overall conversion = 0.84

Overall Moles of SO_2 entering the reactor = 45.06

Moles of SO_2 Entering 2nd bed = 16.67 mol

Moles of SO_2 Leaving = 45.06 – (45.06 x 0.84) = 7.22 mol

Individual conversion on 2nd bed = (16.67 – 7.22 / 16.67) x 100 = **<u>56.7 %</u>**

<u>For 3rd bed</u>

Overall conversion = 0.93

Overall Moles of SO_2 entering the reactor = 45.06

Moles of SO_2 Entering 3rd bed = 7.22 mol

Moles of SO_2 Leaving = 45.06 – (45.06 x 0.93) = 3.16 mol

Individual conversion on 3rd bed = (7.22 – 3.16 / 7.22) x 100 = **<u>56.3 %</u>**

<u>For 4th bed</u>

Overall conversion = 0.997

Overall Moles of SO_2 entering the reactor = 45.06

Moles of SO_2 Entering 4th bed = 3.16 mol

Moles of SO_2 Leaving = 45.06 – (45.06 x 0.997) = 0.14 mol

Individual conversion on 3rd bed = (3.16 – 0.14 / 3.16) x 100 = **<u>95.5 %</u>**

We will now be looking at the material balance of each reactor bed one by one in the following tables:

3.2.4.1 Reactor bed 1

Table 17-Air inlet flow sheet

Air Inlet	**Component**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	2728	97.4	0.74	0.78
	O_2	772.5	24.1	0.21	0.19
	CO_2	176.0	4.00	0.05	0.03
Total		3676.5	125.6	1.00	1.00

Table 18-Dried SO2 inlet flow sheet

Dried SO_2 gas inlet	**Component**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	7568	270.3	0.65	0.77
	CO_2	489.9	11.1	0.04	0.03
	O_2	701.8	21.9	0.06	0.06
	SO_2	2884	45.1	0.25	0.13
	H_2O	0.00	0.00	0.00	0.00
	H_2S	14.6	0.43	0.00	0.00
Total		11658.4	348.8	1.00	1.00

Table 19-SO_2/SO_3 gas Outlet flow sheet

Partially converted SO_2/SO_3	**Component**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.8	0.67	0.80
	O_2	1020.5	31.9	0.07	0.07
	SO_2	1067.1	16.7	0.07	0.04
	CO_2	666.0	15.1	0.04	0.03

	SO_3	2273.5	28.4	0.15	0.06
	H_2S	14.59	0.43	0.00	0.00
Total		15339	460.30	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this bed remained conserved.

In	Out
15334.86	15338.68

3.2.4.2 Reactor bed 2

Table 20-SO_2/SO_3 inlet flow sheet

SO_2/SO_3 from 1st bed	Component	Mass flow (kg/hr)	Molar flow (kmol/hr)	Composition (wt/wt %)	Composition (mol/mol %)
	N_2	10297	367.8	0.67	0.81
	O_2	1020.5	31.9	0.07	0.07
	SO_2	1067.1	16.7	0.07	0.04
	CO_2	665.9	15.1	0.04	0.03
	SO_3	2273.5	23.2	0.15	0.05
	H_2S	14.6	0.43	0.00	0.00
Total		15339	455.1	1.00	1.00

Table 21-SO_2/SO_3 Outlet flow sheet

SO_2/SO_3 Outlet	**Component**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.8	0.67	0.81
	O_2	869.4	27.2	0.06	0.06
	SO_2	462.1	7.22	0.03	0.02
	CO_2	666	15.14	0.04	0.03
	SO_3	3029	37.87	0.20	0.08
	H_2S	14.6	0.43	0.00	0.00
Total		15339	455.6	1.00	1.00

In	**Out**
15339	15339

3.2.4.3 Reactor bed 3

Table 22-SO_2/SO_3 Outlet flow sheet

SO_2/SO_3 from 2nd bed	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.8	0.67	0.81
	O_2	869.4	27.2	0.06	0.06
	SO_2	462.1	7.22	0.03	0.02
	CO_2	665.9	15.1	0.04	0.03
	SO_3	3029.6	37.9	0.20	0.08
	H_2S	14.6	0.43	0.00	0.00
Total		15339	455.6	1.00	1.00

Table 23-SO_2/SO_3 Outlet flow sheet

SO_2/ SO_3	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.7	0.67	0.81
	O_2	804.5	25.1	0.05	0.06
	SO_2	201.9	3.16	0.01	0.01
	CO_2	666	15.1	0.04	0.03
	SO_3	3355	41.9	0.22	0.09
	H_2S	14.6	0.43	0.00	0.00
Total		15338	453.5	1.00	1.00

In	**Out**
15338	15338

3.2.4.4 Reactor bed 4

Table 24-Unconverted SO_2/SO_3 Outlet flow sheet

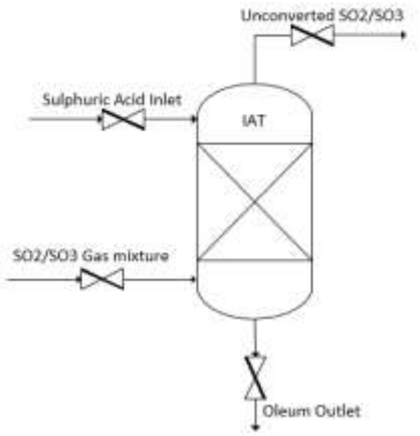
From IAT	**Components**	**Mass flows (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.8	0.84	0.88
	O_2	804.8	25.2	0.07	0.06
	SO_2	201.9	3.15	0.02	0.01
	CO_2	666.2	15.1	0.05	0.04
	SO_3	315	3.94	0.03	0.01
	H_2S	14.6	0.43	0.00	0.00
Total		12299	415.6	1.00	1.00

Table 25-SO_2/SO_3 Outlet flow sheet

To MAT	Components	Mass flow (kg/hr)	Molar flow (kgmol/hr)	Composition (wt/wt %)	Composition (mol/mol %)
	N_2	10297	367.8	0.84	0.89
	O_2	756.9	23.6	0.06	0.06
	SO_2	8.68	0.14	0.00	0.00
	CO_2	666.2	15.1	0.05	0.04
	SO_3	556.7	6.96	0.05	0.02
	H_2S	14.6	0.43	0.00	0.00
Total		12300	414.1	1.00	1.00

In	Out
12299	12300

3.2.5 Intermediate Absorption Tower (IAT)



The main purpose of IAT is the removal of SO_3 that is present in the stream coming from the first three reactor beds. The solvent used for this purpose is concentrated sulphuric acid that this being recycled. The chemical reaction taking place is as follows:

$$H_2SO_4(aq) \ + \ SO_3\ (g) \rightarrow H_2S_2O_7(aq)$$

Assumptions

- Considered ideal with 100 % efficiency
- No excess sulphuric acid provided

Table 26-SO_2/SO_3 Inlet flow sheet

3rd bed outlet	**Component**	**Mass flows (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.75	0.67	0.81
	O_2	804.5	25.14	0.05	0.06
	SO_2	201.9	3.16	0.01	0.01
	CO_2	666	15.1	0.04	0.03
	SO_3	3355	41.9	0.22	0.09
	H_2S	14.6	0.43	0.00	0.00
Total		15338	453.5	1.00	1.00

Table 27-H_2 SO_4 inlet flow sheet

H_2SO_4 Inlet	**Components**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	3650	37.2	1.00	0.99
	H_2O	4.93	0.27	0.00	0.01
Total		3654.9	37.5	1.00	1.00

According to the balanced chemical equation shown above, the material balance on the outlet streams is as follows:

Table 28-SO_2/SO_3 Outlet flow sheet

To 4th bed	**Components**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.8	0.84	0.88
	O_2	804.8	25.2	0.07	0.06
	SO_2	201.9	3.15	0.02	0.01
	CO_2	666.2	15.1	0.05	0.04

	SO_3	315	3.94	0.03	0.01
	H_2S	14.6	0.43	0.00	0.00
Total		12299	415.6	1.00	1.00

Table 29-Oleum Outlet flow sheet

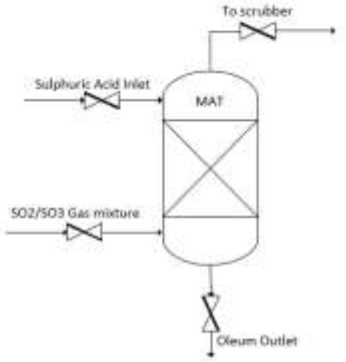
Oleum	**Components**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	6756	37.9	1.00	1.00
Total		6756	37.9	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this equipment remained conserved.

In	**Out**
18993.6	19055.5

3.2.6 Main Absorption Tower (MAT)

MAT is needed to remove the remaining SO_3 that was not absorbed in IAT and additional SO_3 that was produced in the fourth bed of the reactor. It is also attached to a scrubber that is responsible for the prevention of oxides of sulphur into the atmosphere. The solvent used for absorption is again concentrated Sulphuric Acid and the chemical reaction is as follows:



$$H_2SO_4(aq) + SO_3(g) \rightarrow H_2S_2O_7(aq)$$

Assumptions:

- Conversion efficiency is assumed to be 100%
- All the gases (inert and harmful) get removed in the MAT
- No excess Concentrated Sulphuric Acid is introduced

Table 30-SO_2/SO_3 Inlet flow sheet

SO_2/SO_3 from 4th bed	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.8	0.84	0.89
	O_2	756.9	23.7	0.06	0.06
	SO_2	8.68	0.14	0.00	0.00
	CO_2	666.2	15.1	0.05	0.04
	SO_3	556.7	6.96	0.05	0.02
	H_2S	14.6	0.43	0.00	0.00
Total		12300	414.1	1.00	1.00

Table 31-H_2SO_4 Inlet flow sheet

H_2SO_4 Inlet	**Components**	**Mass flows (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	894.7	9.13	1.00	0.99
	H_2O	1.21	0.07	0.00	0.01
Total		895.9	9.20	1.00	1.00

The material balance on the outlet streams is as follows:

Table 32-To Scrubber Outlet flow sheet

To Scrubber	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	N_2	10297	367.75	0.88	0.90
	O_2	756.7	23.65	0.06	0.06
	SO_2	8.73	0.14	0.00	0.00
	CO_2	666.1	15.14	0.06	0.04
	H_2O	11.9	0.67	0.00	0.00
	H_2S	14.6	0.43	0.00	0.00
Total		11755	407.8	1.00	1.00

Table 33-Oleum Outlet flow sheet

Oleum	**Component**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	1238.7	6.96	0.85	0.76
	H_2SO_4	212.6	2.17	0.15	0.24
Total		1451.3	9.13	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this equipment remained conserved.

In	**Out**
13196	13206.4

3.2.7 Oleum Header

Oleum Header is merely a distributor that distributes oleum to the necessary locations (Air dryer, SO_2 dryer, dilution tank) and receives oleum from different location, where it is being produced i.e. from IAT and MAT. There are no reactions taking place in the oleum header.

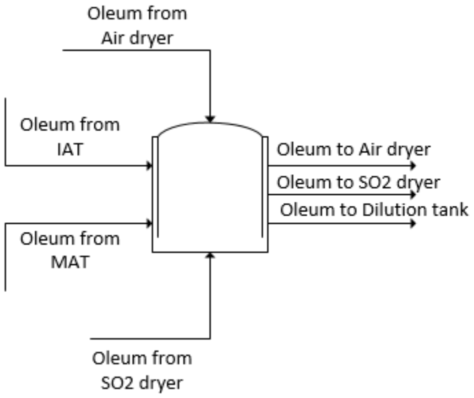


Table 34-Oleum from IAT Inlet flow sheet

Oleum (IAT)	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	6756	37.9	1.00	1.00
Total		6756	37.9	1.00	1.00

Table 35-Oleum from MAT Inlet flow sheet

Oleum (MAT)	**Component**	**Mass flow(kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	1238.7	6.96	0.85	0.76
	H_2SO_4	212.6	2.17	0.15	0.24
Total		1451.3	9.13	1.00	1.00

Table 36-Oleum from Air Dryer Inlet flow sheet

Oleum (Air Drier)	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	0.00	0.00	0.00	0.00
	H_2SO_4	3843.5	39.2	1.00	1.00
Total		3843.5	39.2	1.00	1.00

Table 37-Oleum from SO_2 Dryer Inlet flow sheet

Oleum from SO_2 Dryer	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	195.5	1.99	0.58	0.71
	$H_2S_2O_7$	144.3	0.81	0.42	0.29
Total		339.8	2.81	1.00	1.00

Table 38-Oleum to Air Dryer Outlet flow sheet

Oleum to Air Dryer	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	3282.9	18.4	0.93	0.89
	H_2SO_4	228.5	2.33	0.07	0.11
Total		3511.5	20.8	1.00	1.00

Table 39-Oleum to SO_2 Dryer Outlet flow sheet

Oleum to SO_2 Dryer	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	321.9	1.81	0.82	0.72
	H_2SO_4	69.1	0.71	0.18	0.28
Total		391.1	2.51	1.00	1.00

Table 40-Oleum to Dilution Tank Outlet flow sheet

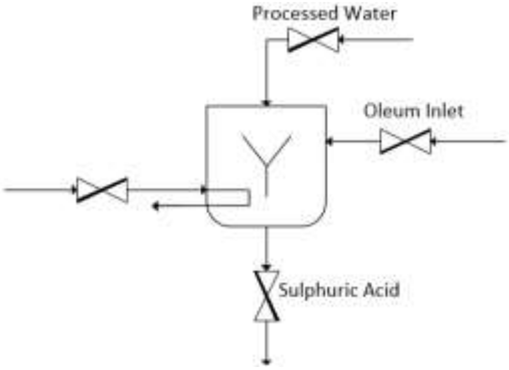
Oleum to dilution Tank	**Components**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	4534	25.5	0.53	0.38
	H_2SO_4	3953	40.3	0.46	0.60
	H_2O	20.6	1.14	0.00	0.02
Total		8508	66.95	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this equipment remained conserved.

In	**Out**
12390.5	12410.1

3.2.8 Dilution Tank

In the dilution tank, water is added to oleum to produce concentrated sulphuric acid. The chemical reaction for this equipment is as follows:



$$H_2O(l) + H_2S_2O_7(aq) \rightarrow 2H_2SO_4(aq)$$

Assumption:

- Conversion is assumed 100%.

Table 41-Oleum from Oleum header Inlet flow sheet

Oleum from Oleum Header	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	$H_2S_2O_7$	4534	25.5	0.53	0.38
	H_2SO_4	3953	40.3	0.46	0.60
	H_2O	20.6	1.14	0.00	0.02
Total		8507.6	66.95	1.00	1.00

Table 42-Water Inlet flow sheet

Water inlet	**Components**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2O	450	25.0	1.00	1.00
Total		450	25.0	1.00	1.00

According to the balanced chemical equation shown above, the material balance on the outlet streams is as follows:

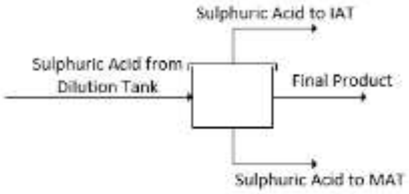
Table 43-H_2 SO_4 Outlet flow sheet

H_2SO_4 outlet	**Component**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	8946	91.3	1.00	0.99
	H_2O	12.07	0.67	0.00	0.01
Total		8958	91.9	1.00	1.00

Adding both the inlet and outlet streams gives us the following result, proving that the materials in this equipment remained conserved.

In	Out
8957.6	8958.1

3.2.9 H_2SO_4 Storage Tank



H_2SO_4 storage tank is only used to keep the sulphuric acid produced to be extracted as a final product or recycled to various locations in the plant where it is needed.

Table 44-H_2SO_4 Inlet flow sheet

H_2SO_4 Inlet	**Components**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	8946	91.3	1.00	0.99
	H_2O	12.1	0.67	0.00	0.01
Total		8958.07	91.96	1.00	1.00

Table 45-H_2SO_4 Outlet flow sheet

H_2SO_4 to IAT	**Component**	**Mass flow (kg/hr)**	**Molar flow (kgmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	3650	37.24	0.998	0.99
	H_2O	4.93	0.27	0.002	0.01
Total		3654.93	37.52	1.00	1.00

Table 46-H_2SO_4 to MAT Outlet flow sheet

H_2SO_4 to MAT	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	894.7	9.13	0.998	0.99
	H_2O	1.21	0.07	0.002	0.01
Total		895.91	9.20	1.00	1.00

Table 47-H_2SO_4 Extracted flow sheet

H_2SO_4 extracted	**Component**	**Mass flow (kg/hr)**	**Molar flow (kmol/hr)**	**Composition (wt/wt %)**	**Composition (mol/mol %)**
	H_2SO_4	4401.8	44.9	0.998	0.99
	H_2O	5.95	0.33	0.002	0.01
Total		4407	45.3	1.00	1.00

Adding both the inlet and outlet streams proves mass conservation.

In	**Out**
8958.1	8958.5

Table 48-Overall Material Balance

	Air dryer			Sulfur Burner				Catalytic Reactor				IAT			MAT				Dilution Tank		
Component	Moist air	Oleum inlet	H_2SO_4 outlet	Dried Air	Molten sulfur	Dried Air	SO_2 gas	Dried Air	SO_2/SO_3 from IAT	SO_2/ SO_3 to MAT	SO_2/SO_3 from 3rd bed	H_2SO_4 inlet	Oleum	SO_2/SO_3 gas to 4th bed	SO_2/SO_3 from 4th bed	H_2SO_4 inlet	Oleum	To scrubber	Oleum from Absorbers	Water	H_2SO_4
O_2	2915	0	0	2915	0	2142	700.1	772.5	804.8	756.9	804.5	0	0	804.8	756.9	0	0	756.7	0	0	0
N_2	10297	0	0	10297	0	7568	7568	2728	10297	10297	10297	0	0	10297	10297	0	0	10297	0	0	0
CO_2	667	0	0	667	0	489	489	176	666.2	666.23	666	0	0	666.2	666.2	0	0	666.1	0	0	0
H_2O	376.6	0	0	0	0	0	17.66	0	0	0	0	0	0	0	0	1.21	0	11.98	20.6	450	12.07
H_2S	0	0	0	0	14.6	0	14.6	0	14.59	14.59	14.59	0	0	14.59	14.59	0	0	14.59	0	0	0
S	0	0	0	0	1443	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO_2	0	0	0	0	0	0	2884	0	201.8	8.68	201.9	0	0	201.8	8.68	0	0	8.73	0	0	0
SO_3	0	0	0	0	0	0	0	0	315	556.7	3354.7	0	0	315	556.8	0	0	0	0	0	0
H_2SO_4	0	228.5	3843.5	0	0	0	0	0	0	0	0	3650	0	0	0	894.7	212.6	0	3953	0	8946
$H_2S_2O_7$	0	3282	0	0	0	0	0	0	0	0	0	4.93	6756	0	0	0	1238.7	0	4534	0	0
Total (kg/hr)	14255	3510	3843.5	13879	1457.6	10199	11673	3676.5	12299.39	12300.1	15338.69	3654.9	6756	12299.39	12300.17	895.91	1451.3	11755.1	8507.6	450	8958.07
	17766		17722.5		11656.6		11673	27649.25		27638.79		18993.6	19055.39		13196.08		13206.4		8957.6		8958.07

CHAPTER 4

Energy Balance

Basis = 1hr **Pressure = 101.3 kPa (unless changed)**

The energies are calculated using the formula:

$$Q = mCp\Delta T$$

Where Q is the energy required in MJ,

m is the mass flow in kg,

Cp is specific heat capacity in kJ/kg.°C, and

ΔT is temperature difference between the stream temperature and reference temperature (25°C) (Cengel, 2014)

4.1 Sulfur melter

The solid sulfur is sent to a melter to be converted into its molten form for safer and efficient conversion to SO_2 gas in sulfur burner.

Table 49-Energy calculation for sulfur melter

Parameter	**Solid sulfur**	**Molten sulfur**
Pressure (kPa)	110	101.3
Temperature (^{0}C)	25	150
Cp (kJ/kg.^{0}C)	1.59	1.33
Mass (kg/hr)	1458	1458
Q (MJ)	0	242.5

4.2 Pump (Sulfur)

The power-driven pump provides a head for the feed to flow. Isentropic compression increases the pressure of the incoming stream of sulfur from 101.3 kPa to 110 kPa. The efficiency of the pump is taken to be 75%.

4.2.1 Specifications and equations required for pump calculations

Volume expansivity β (°C $^{-1}$) = 0.00075

ΔP (kPa) = 8.67

$$Isentropic\ Work\ (Ws) = V * \Delta P\ (1 - \beta T)$$

$$Power(P) = \frac{Wa}{t}$$

$$Actual\ Work\ (Wa) = m * (\frac{Ws}{\eta})$$

Table 50-Energy calculation for pump

Parameter	Value
Feed mass flow rate (kg/hr)	1457.6
Feed volume flow rate (m^3/hr)	0.693
Specific Volume (m^3/kg)	0.000475
Isentropic Work/Ws (kJ/kg)	0.0037
Actual Work/Wa (kJ)	0.0020
Power (kW)	0.002

4.3 Air Dryer

The air dryer is used to remove moisture content from the air using dehydrating agent oleum. Since it is an exothermic process, the dried air is at a higher temperature than inlet air.

Reaction: $H_2O + H_2S_2O_7 \rightarrow 2H_2SO_4$

Table 51-Energy calculation for Air Dryer

Parameter	Air	Oleum from Oleum Header	Dried Air	H_2SO_4 to Oleum Header
Temperature (^{0}C)	30	172.7	45	45
Mass flow rate (kg/hr)	14256	3511.5	13880	3843.5

Cp (kJ/kg.^{0}C)	1.05	0.73	1.01	1.44
dH (MJ)	74.9	378.5	280.3	110.8
Total (MJ)	453.4		391.1	
Energy released (MJ)	62.3			

4.4 Sulfur Burner

The dried air and molten sulfur react in the sulfur burner to produce SO_2. Since the reaction is highly exothermic, the product is at a very high temperature. The energy difference between the inlet and outlet is the energy provided by the fuel to carry out the reaction.

Reaction: $S + O_2 \rightarrow SO_2$

Table 52-Energy calculation for sulfur burner

Parameter	**Air**	**Molten Sulfur**	**SO_2 gas**
Temperature (^{0}C)	45	150	2014
Mass Flow rates (kg/hr)	10199	1458	11673
Cp (kJ/kg.^{0}C)	1.01	1.33	1.2
dH (MJ)	206	242.4	27861
Total	448.4		27861
Energy required (MJ)	27412		

4.5 SO_2 Dryer

The SO_2 product gas from sulfur burner is sent to the SO_2 dryer so that water content is completely removed from the stream before it enters the reactor. Due to the exothermic reaction, products are at a higher temperature.

Table 53-Energy calculation for SO_2 dryer

Parameter	**SO_2 from furnace**	**Oleum from Oleum Header**	**Dried SO_2 to Reactor**	**H_2SO_4 to Oleum Header**
Temperature (^{0}C)	70	143	184.6	184.6
Mass flow rate (kg/hr)	11663	391.1	11658.4	339.8
Cp (kJ/kg.^{0}C)	0.934	0.712	0.959	1.29
dH (MJ)	990	532.9	1483	69.9
Total (MJ)	1522.9		1552.8	
Energy released (MJ)	0.1			

4.6 Reactor Bed 1

The dried SO_2 gas enters the packed bed catalytic reactor at Bed 1. V_2O_5 catalyst increases the rate of reaction between SO_2 and dried air and the optimum temperature of 450°C allows for efficient conversion of SO_2 to SO_3. 63% conversion of SO_2 to SO_3 is achieved.

Reaction: $2SO_2 + O_2 \rightarrow 2SO_3$

\

Table 54-Energy calculation for Reactor Bed 1

Parameter	**Dried Air**	**SO_2 from SO_2 Dryer**	**SO_2/SO_3 gas**
Temperature (^{0}C)	45	450	525.8
Mass flow rates (kg/hr)	3677	11658	15338
Cp (kJ/kg.^{0}C)	1.01	1.03	1.07
dH (MJ)	74.2	7188.6	7253.9
Total (MJ)	7262.8		7253.9
Energy released (MJ)	8.90		

4.7 Reactor Bed 2

The same reaction occurs here as Bed 2. Since the temperature of the product increases to 525°C, it is cooled back to the optimum temperature for reaction in Bed 2. 84% conversion is achieved here.

Table 55-Energy calculation for Reactor Bed 2

Parameter	**SO_3 from 1st bed**	**SO_3 from 2nd bed**
Temperature (^{0}C)	450	507.1
Mass flow rates (kg/hr)	15338.68	15338.7
Cp (KJ/kg.^{0}C)	1.05	1.07
dH (MJ)	6838.4	6835
Total (MJ)	6838.4	6835
Energy released (MJ)	3.7	

4.8 Heat Exchanger 2

The Bed 2 outlet is cooled in Heat Exchanger 2 to further convert in Bed 3.

Table 56-Energy calculation for Heat Exchanger 2

Parameter	**Tube Side In (SO_2/SO_3 from Bed 2)**	**Shell Side In (From IAT)**	**Tube Side Out (SO_2/SO_3 to Bed 3)**	**Shell Side Out (SO_2/SO_3 to Bed 4)**
Temperature (^{0}C)	507.1	444.5	460	501.4
Mass flow rate (kg/hr)	15338	12299	15338	12299
Cp (kJ/kg.^{0}C)	1.065	1.09	1.053	1.104
dH (MJ)	7875	5624	7026	6469
Total (MJ)	13499.45		13494.9	
Energy required (MJ)	4.55			

4.9 Reactor Bed 3

The cooled stream from Heat Exchanger 2 reacts with air and achieves a conversion of 93%. The temperature rise is less compared to the first two beds because of decreasing SO_2 concentration in each bed.

Table 57-Energy calculation for Reactor Bed 3

Parameter	**Reactant (From IAT)**	**Product (To IAT)**
Temperature (^{0}C)	460	490
Mass flow rates (kg/hr)	15338	15338
Cp (KJ/kg.^{0}C)	1.053	1.062
dH (MJ)	7526	7524.7
Total (MJ)	7526	7524.7
Energy released (MJ)	1.30	

4.10 IAT (Intermediate Absorption Tower)

The SO_2/SO_3 product stream from Bed 3 is sent to IAT where it reacts with concentrated sulphuric acid to form oleum. Most of the SO_3 is utilized in this absorber. The unconverted gases are sent to Bed 4 to achieve further conversion.

Table 58-Energy calculation for IAT

Parameter	**SO_3 from 3rd bed)**	**H_2SO_4 from Storage Tank**	**Unconverted SO_2 to bed 4**	**Oleum to Oleum Header**
Temperature (^{0}C)	290	45	200	200
Mass flow rates (kg/hr)	15338	3655	12299	6756
Cp (KJ/kg.^{0}C)	1.006	1.43	1.032	0.968
dH (MJ)	4089	105	2221	1145
Total (MJ)	4194		3366	
Energy released (MJ)	828			

4.11 Reactor Bed 4

The unconverted SO_2 from IAT achieves a maximum conversion of (99.7%).

Table 59-Energy calculation for Reactor Bed 4

Parameter	**SO_2/SO_3 From IAT**	**SO_2/SO_3 To MAT**
Temperature (^{0}C)	501.4	523.2
Mass flow rates (kg/hr)	12299	12299
Cp (KJ/kg.^{0}C)	1.104	1.11
dH (MJ)	6469	6469
Total (MJ)	6468.8	6469
Energy released (MJ)	0.2	

4.12 MAT (Main Absorption Tower)

The remaining SO_3 from IAT is consumed here so that minimum amount of contaminants enters the scrubber.

Table 60-Energy calculation for MAT

Parameter	**SO_2/SO_3 from 4th bed**	**H_2SO_4 from Storage Tank**	**To Scrubber**	**Oleum to Oleum Header**
Temperature (^0C)	200	45	87	87
Mass flow rate (kg/hr)	12299	896	11755	1451
Cp (KJ/kg.^0C)	1.032	1.43	1.03	0.543
dH (MJ)	1821	25.6	949.9	649
Total (MJ)	1847		1599	
Energy released (MJ)	248			

4.13 Dilution Tank

A major fraction of oleum from IAT and MAT reacts with water to form concentrated sulphuric acid.

Table 61-Energy calculation for Dilution Tank

Parameter	**Oleum from Oleum Header**	**Process Water**	**H_2SO_4 to Storage Tank**
Temperature (^0C)	92.4	70	45
Mass flow rate (kg/hr)	8508	450	8958
Cp (KJ/kg.^0C)	0.920	4.18	1.43
dH (MJ)	528	84.7	256
Total (MJ)	612		256
Energy released (MJ)	356		

CHAPTER 5

Equipment Design

The equipment design for all our components in the process flow diagram is described in detail below. (Coulson, 2005)

5.1 Air Dryer

In the air dryer, the water content in the incoming air is removed via an absorption reaction with the dehydrating agent, oleum. For this reason, the design of the air dryer is similar to that of an absorption column.

For our case, the calculations for chemical absorption have been carried out. Onda's method is used to calculate the gas and liquid film mass-transfer coefficients which are subsequently used in combination with reaction kinetics to determine the vessel height. (Zein, 2014)

The first step of the design is to determine the specifications of the column.

5.1.1 Calculation of Cross-sectional Area of Column

The physical property data required for the calculation of area are given below.

Table 5.1 Physical property data for calculating cross sectional area of column

Flow rate of gas inlet / G (kg/sec)	3.96
Flow rate of solvent inlet /L (kg/sec)	0.975
Temperature of gas inlet (°C)	30
Temperature of solvent inlet (°C)	173.3
Pressure (bar)	1.013
Average Molecular Weight of gases	28.8
Density of gas mixture (kg/m3)	1.62
Density of solvent at $45^{0}C$ (kg/m3)	1042
Viscosity of solvent (Cp)	0.0279

Viscosity of solvent (kg/m.sec)	0.0000279
Viscosity of gaseous mixture (Cp)	0.0183
Viscosity of gaseous mixture (kg/m.sec)	0.0000183

Using Figure 11.44 from Coulson Richardson Volume 6, we calculate the value of K_4 which is the gas mass flow rate per unit cross sectional area. The figure used generalized pressure drop correlation for calculation of area and diameter.

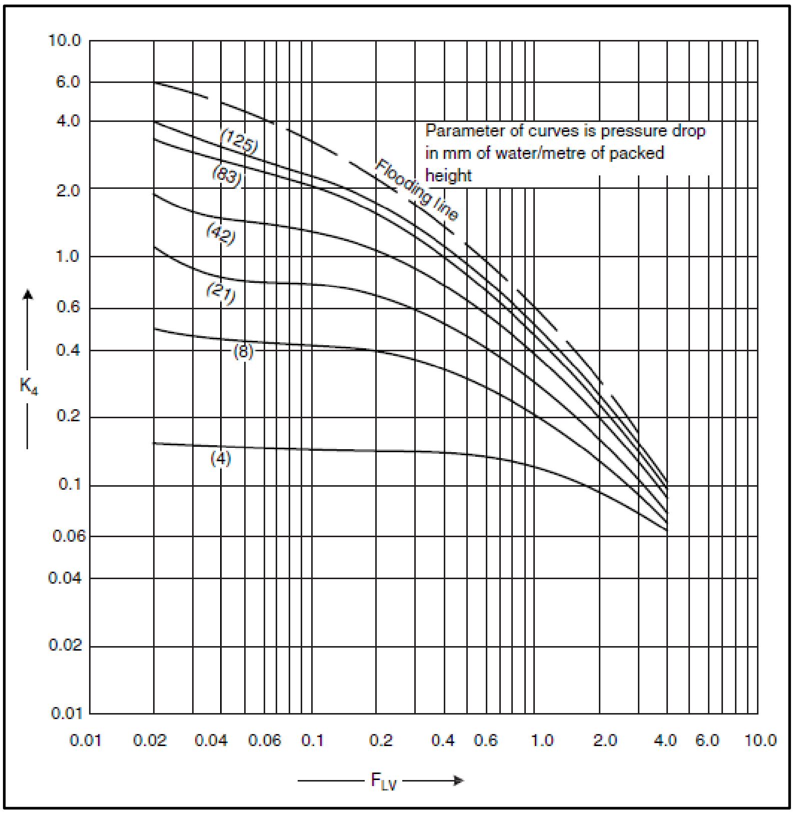


Figure 6-Generalized pressure drop correlation for calculation of area

We have assumed the pressure drop to be 21 mm of water/meter of packed height as per the recommended design values for absorbers and strippers (15 to 50 mm of water/meter of packed height).

$$F_{LV} = \frac{L_w^*}{V_w^*}\sqrt{\frac{\rho_v}{\rho_l}}$$

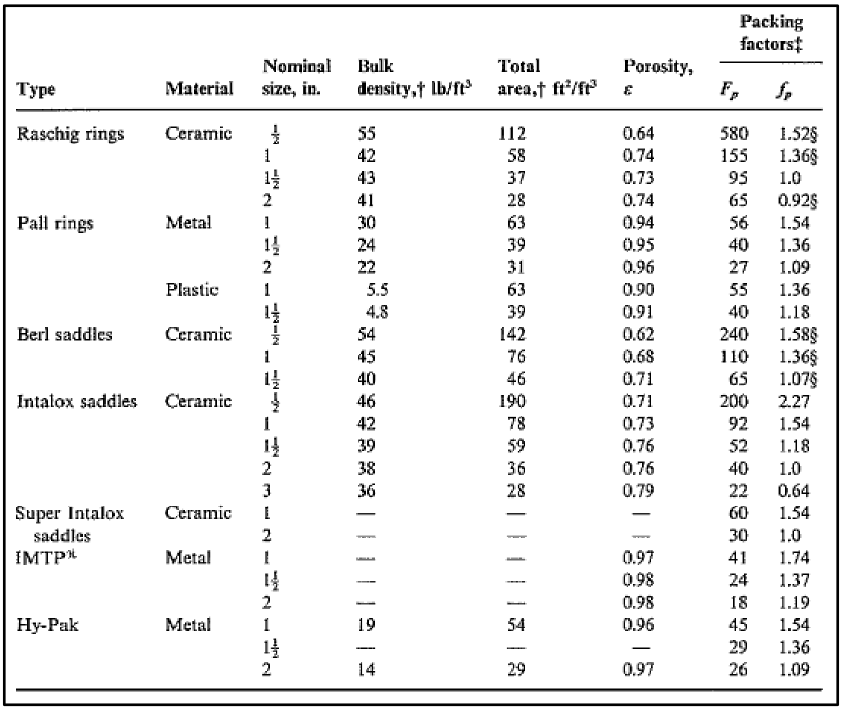
K4=1.98

5.1.1.1 Calculation of mass flow rate per unit area (G*).

The packing material chosen for air dryer is Intalox Saddle Ceramic. Due to presence of corrosive materials like oleum, ceramic is the most suitable choice. Most of the moisture from the air is removed in the air dryer. An efficient liquid distribution is required for maximum removal of moisture. Hence the most effective packing that can be used is Intalox Saddles.

Using Table 22.1 from the book Unit Operations of Chemical Engineering, we calculate the packing factor F_P. (McCabe, 1993)

We assumed the nominal size for the packing to be 1.5 in.



Type	Material	Nominal size, in.	Bulk density,† lb/ft³	Total area,† ft²/ft³	Porosity, ε	Packing factors‡	
						F_p	f_p
Raschig rings	Ceramic	$\frac{1}{2}$	55	112	0.64	580	1.52§
		1	42	58	0.74	155	1.36§
		$1\frac{1}{2}$	43	37	0.73	95	1.0
		2	41	28	0.74	65	0.92§
Pall rings	Metal	1	30	63	0.94	56	1.54
		$1\frac{1}{2}$	24	39	0.95	40	1.36
		2	22	31	0.96	27	1.09
	Plastic	1	5.5	63	0.90	55	1.36
		$1\frac{1}{2}$	4.8	39	0.91	40	1.18
Berl saddles	Ceramic	$\frac{1}{2}$	54	142	0.62	240	1.58§
		1	45	76	0.68	110	1.36§
		$1\frac{1}{2}$	40	46	0.71	65	1.07§
Intalox saddles	Ceramic	$\frac{1}{2}$	46	190	0.71	200	2.27
		1	42	78	0.73	92	1.54
		$1\frac{1}{2}$	39	59	0.76	52	1.18
		2	38	36	0.76	40	1.0
		3	36	28	0.79	22	0.64
Super Intalox saddles	Ceramic	1	—	—	—	60	1.54
		2	—	—	—	30	1.0
IMTP®	Metal	1	—	—	0.97	41	1.74
		$1\frac{1}{2}$	—	—	0.98	24	1.37
		2	—	—	0.98	18	1.19
Hy-Pak	Metal	1	19	54	0.96	45	1.54
		$1\frac{1}{2}$	—	—	—	29	1.36
		2	14	29	0.97	26	1.09

Figure 7-Characteristics of tower packing

The packing factor F_P corresponding to our chosen packing specifications is 52.

$$G^* = \sqrt{\frac{K_4 * \rho_v * (\rho_l - \rho_v)}{13.1 * F_P * (\frac{\mu_L}{\rho_L})^{0.1}}}$$

$$G^* = \sqrt{\frac{1.98 * 1.62 * (1042 - 1.62)}{13.1 * 52 * \left(\frac{0.00002794}{1024}\right)^{0.1}}}$$

$\mathrm{G}^{*}\left(\frac{\mathrm{kg}}{\mathrm{m2}}.\mathrm{sec}\right) = 5.29$

The formula used for calculation of area is

$$Area(m^2) = \frac{G}{G^*}$$

Area = 0.748 m^2

5.1.2 Calculation of diameter of column

The formula used to calculate the diameter of the column is

$$Diameter\ of\ the\ column\ (m) = \sqrt{\frac{4 * Area}{3.142}}$$

Diameter = 0.976 m

5.1.3 Calculation of wetted area of packing (a_w)

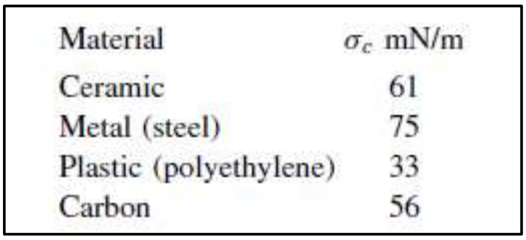
The data required for the calculation of a_w is given below.

Table 5.2 Physical property data for calculation of wetted area of packing

L_w* (kg/m^2.sec)	1.30
a (ft^2/ft^3)	59
a (m^2/m^3)	193.6
Critical surface tension-σ_c (N/m)	0.061
Liquid surface tension- σ_l (N/m)	0.00876
Viscosity of solvent (kg/m.sec)	0.00002794
Density of solvent (kg/m^3)	1042

$$a_w = 1 - exp\left(-1.45 * \left(\frac{\sigma_c}{\sigma_l}\right)^{0.75} * \left(\frac{L_w^*}{a * \mu_l}\right)^{0.1} * \left(\frac{L_w^2 * a}{\rho_l^2 * g}\right)^{-0.05} * \left(\frac{L_w^2}{\rho_l * \sigma_l * a}\right)^{0.2}\right)$$

σ_c is taken from Coulson Richardson Volume 6.



Material	σ_c mN/m
Ceramic	61
Metal (steel)	75
Plastic (polyethylene)	33
Carbon	56

Figure 8-Critical surface tension data for different packing material

σ_l , density of solvent and viscosity of solvent are taken from our HYSYS model.

Effective interfacial area for mass transfer (a) is taken from Table 22.1 of Unit Operations of Chemical Engineering.

$$a_w = 1 - exp\left(-1.45 * \left(\frac{0.061}{0.00876}\right)^{0.75} * \left(\frac{1.3}{193.6 * 0.00002794}\right)^{0.1} * \left(\frac{1.3^2 * 193.6}{1042^2 * 9.81}\right)^{-0.05} * \left(\frac{1.3^2}{1042 * 0.00876 * 193.6}\right)^{0.2}\right)$$

a_w (m^2/m^3) = 191.4

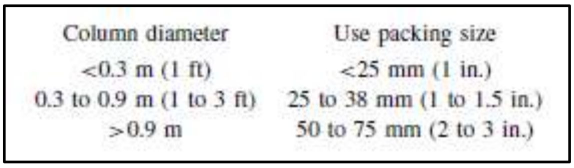
5.1.4 Calculation of liquid film mass transfer coefficient K_L

The data required for the calculation of K_L is given below.

Table 5.3 Physical property data for calculation of K_L

ρ_l (kg/m^3)	1042
μ_l (kg/m.sec)	0.0000279
Lw* (kg/m^2.sec)	1.30
aw (m^2/m^3)	191.4
Diffusivity/ DL	0.000000109
Packing size /d_p (m)	0.075
a (m^2/m^3)	193.6

The packing size is taken from the recommended size ranges corresponding to the diameter of the column in Coulson Richardson Volume 6.



Column diameter	Use packing size
<0.3 m (1 ft)	<25 mm (1 in.)
0.3 to 0.9 m (1 to 3 ft)	25 to 38 mm (1 to 1.5 in.)
>0.9 m	50 to 75 mm (2 to 3 in.)

Figure 9-Recommended packing size for a particular range of diameter

The value of D_L is taken from literature. The formula to calculate K_L is:

$$K_L = 0.0051 * \left(\frac{L_w^*}{a_w * \mu_l}\right)^{\frac{2}{3}} * \left(\frac{\mu_l}{\rho_l * D_L}\right)^{\frac{-1}{2}} * \left(a * d_p\right)^{0.4} * \left(\frac{\mu_l * g}{\rho_l}\right)^{\frac{1}{3}}$$

$$K_L\left(\frac{m}{s}\right) = 0.0051 * \left(\frac{1.30}{191.44 * 0.0000279}\right)^{\frac{2}{3}} * \left(\frac{0.0000279}{1042 * 0.000000109}\right)^{\frac{-1}{2}}$$
$$* (193.58 * 0.075)^{0.4} * \left(\frac{0.0000279 * 9.81}{1042}\right)^{\frac{1}{3}}$$

K_L = 0.0079 m/s

5.1.5 Calculation of gas film mass transfer coefficient K_g

The physical property data needed for its calculation is given below.

Table 5.4 Physical property data for calculation of K_g

R (J/mol.K)	0.0802
Temperature of gas inlet (^{0}C)	30
a (m^2/m^3)	193.6
Diffusivity of gas (m^2/sec)	0.00000012
K_5	5.23
V_w (kg/m^2.sec)	5.29
Viscosity of gas (kg/m.sec)	0.000018
Density of gas (kg/m^3)	1.62
d_p	0.075

K_5 is a constant that has a value of 5.23 for packing sizes above 15mm as per the standard range given in Coulson Richardson Volume 6.

The physical properties of the gas are taken from our HYSYS model.

Diffusivity of gas is taken from literature.

The formula used to calculate K_g is

$$K_g = \left(\frac{a * D_g}{R * T_g}\right) * K_5 * \left(\frac{V_w^*}{a * \mu_g}\right)^{0.7} * \left(\frac{\mu_g}{\rho_g * D_g}\right)^{\frac{1}{3}} * \left(a * d_p\right)^{-2}$$

$$K_g = \left(\frac{193.58 * 0.00000012}{0.0802 * 30}\right) * 5.23 * \left(\frac{5.29}{193.58 * 0.000018}\right)^{0.7} * \left(\frac{0.000018}{1.62 * 0.00000012}\right)^{\frac{1}{3}} * (193.58 * 0.075)^{-2}$$

$K_g = 0.00018\ (\frac{kmol}{m2}.sec.bar)$

5.1.6 Calculation of Rate of Reaction

The data required for the calculation of reaction rate is given below.

Table 5.5 Physical property data for calculation of rate of reaction

Molar flow rate of oleum (kgmol/hr)	20.8	Molar flow rate of air (kgmol/hr)	494
Vol. flow rate of oleum (m^3/hr)	1042	Vol. flow rate of air (m^3/hr)	8803
C_{ao} (mol/m^3)	19.9	C_{bo} (mol/m^3)	56.2
K_L (m/s)	0.0079	K_g (kmol/m^2.sec.bar)	0.00018
D_L (m^2/s)	0.000000109	D_g (m^2/s)	0.00000012

$$P_a = P * mol.frac.of\ water\ in\ gas$$

P=100000 atm

P_a = 4220

The formula used to calculate the rate of reaction is

$$-r_a'' = K_g * P_a$$

$-r_a'' = 0.0075\ \frac{mol}{s.m^2}$

To decide the regime where the reaction will take, the value of K_LC_b is calculated.

$K_LC_b = 0.443$

Since K_LC_b is greater than K_gP_a, an instantaneous reaction occurs in the first regime.

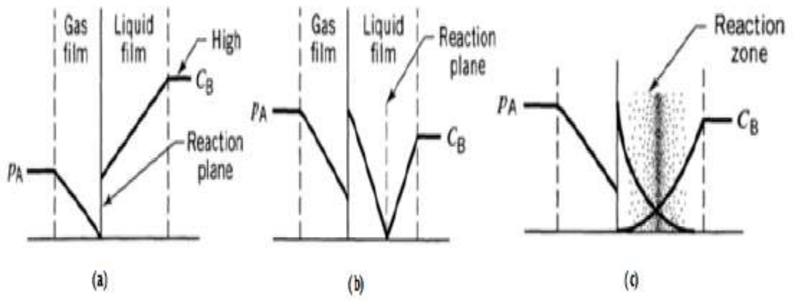


Fig. 4. Concentration profile by two film theory for (a) instantaneous reaction with high silver ion concentration (first regime), (b) instantaneous reaction with low silver ion concentration (second regime), and (c) fast reaction with low silver ion

Figure 10-Concentration profile by two film theory

5.1.7 Calculation of height of column

The data required for the calculation is given below.

Table 5.6 Physical property data for calculation of height of column

G' (molar flow of inert)/kgmol/hr	115
L' (molar flow of inert in liquid)/kgmol/hr	2332
Mol. Frac. of water in gas inlet	0.0422
Mol. Frac. of water in gas outlet	0.0028
Cross sectional area (m^2)	0.748
a (m^2/m^3)	193.6
$1/-r_a''$	132.5

$$Height\ of\ packing = \left(\frac{G'}{a * a_c}\right) * \left(x_{g1} - x_{g2}\right) * \frac{1}{-r_a''}$$

$$Height\ of\ packing = \left(\frac{115}{193.58 * 0.748}\right) * (0.0422 - 0.0028) * 132.49$$

Height of packing = 4.14 m

25% provision is given for the spaces above and below the packed region of the column. Hence the total height of the column is:

$$Height\ of\ column = Height\ of\ packing + 0.25 * Height\ of\ packing$$

Height of column = 5.18 m

5.1.8 Calculation of Volume of Packing

$$Volume = \frac{\pi * d^2 * h}{4}$$

D = 0.976m

H = 4.14m

Volume = 3.1 m^3

Table 62-Specification Summary of Air Dryer

Air Dryer Specification Sheet	
Cross-sectional area of column (m^2)	0.748
Diameter of column (m)	0.976
a_W (m^2/m^3)	191.4
D_P (m)	0.075
Pressure Drop (mm of H_2O/m of packed height)	21
K_L (m/s)	0.0079
K_g (kmol/m^2.sec.bar)	0.00018
$-r_a''$ (mol/m^2.sec)	0.0075
Height of packing (m)	4.14
Height of column (m)	5.18
Volume of packing (m^3)	3.1

5.2 SO_2 Dryer

Table 63-Specification Summary of SO_2 Dryer

SO_2 Dryer Specification Sheet	
Cross-sectional area of column (m^2)	0.898
Diameter of column (m)	1.07
a_w (m^2/m^3)	73.6
Packing material	Raschig Rings
Pressure Drop (mm of H_2O/m of packed height)	42
K_L (m/s)	0.000959
K_g (kmol/m^2.sec.bar)	0.000319
$-r_a''$ (mol/m^2.sec)	0.000904
Height of packing (m)	3.13
Height of column (m)	3.91
Volume of packing (m^3)	2.81

5.3 Intermediate Absorption Tower (IAT)

Maximum absorption of SO_3 takes place in this absorption tower via chemical reaction with sulphuric acid. Hence the dimensions of this column are significantly larger than MAT for complete absorption. (Henley, 2005)

Table 64-Specification Summary of Intermediate Absorption Tower (IAT)

IAT Specification Sheet	
Cross-sectional area of column (m^2)	1.09
Diameter of column (m)	1.18
a_w (m^2/m^3)	43.0
Packing Material	2 in. Intalox Saddle Ceramic

Pressure Drop (mm of H_2O/m of packed height)	42
K_L (m/s)	0.0025
K_g (kmol/m^2.sec.bar)	0.000138
$-r_a''$ (mol/m^2.sec)	0.0129
Height of packing (m)	6.27
Height of column (m)	7.84
Volume of packing (m^3)	6.84

5.4 Main Absorption Tower (MAT)

In this column, the remaining SO_3 is absorbed in sulphuric acid to form oleum. The main purpose of this absorption tower is to reduce the amount of toxic gases in the flue gases before they are discharged into the atmosphere.

Table 65-Specification Summary of Main Absorption Tower (MAT)

MAT Specification Sheet	
Cross-sectional area of column (m^2)	0.54
Diameter of column (m)	0.84
a_w (m^2/m^3)	40.23
D_P (m)	0.025
Packing material	1.5 in. Berl Saddle Ceramic
Pressure Drop (mm of H_2O/m of packed height)	42
K_L (m/s)	0.000369
K_g (kmol/m^2.sec.bar)	0.00000317
$-r_a''$ (mol/m^2.sec)	0.00539
Height of packing (m)	4.25
Height of column (m)	5.31

Volume of packing (m^3)	2.33

5.5 Heat Exchanger - 2

For the design of this heat exchanger that is being used to cool the stream coming out of the second bed of the reactor, we will be starting the calculation with some basic assumptions in order to have a basis for our elaborate design process. The assumptions taken are based on actual data that we have been able to obtain with the help of both our material and energy balance as well as from literature. The type of Heat exchanger being chosen is depending on heat transfer and temperature difference that needs to be attained for our operation and process. (Perry, 2007)

Heat exchanger type: Shell and tube heat exchanger

Tube bundle and shell arrangement pattern: Pull through floating head type
Tube side inlet stream is basically a gas mixture that is coming from the reactor at a relatively high temperature and needs to be cooled down in order to reach the optimum temperature required in the reactor. This will be done with the help of two heat exchangers; this one is thus the first of those two heat exchangers. The reason why this gas mixture is being sent through the tubes is to prevent heat loss to the environment. The second reason being the concern for the pressure drop - streams are to be chosen as such, that the mass velocities result in a value for a lower and acceptable pressure drop on both, the shell and the tube side.

The cold gas mixture coming from the absorber through the second heat exchanger is entering the Shell side. Again, the reason for sending the cold fluid through the shell side is firstly because there are greater chances of heat loss in the shell side and secondly because we had to ensure that the pressure loss stays at a minimum.

Design Parameters

1. The value of UA achieved in our heat exchanger should be greater than or equal to the value of UA required.
2. The pressure-drop in both, the shell and tube side, should not exceed 10 psi.

5.5.1 Tube side data

Subject stream: Gas mixture from reactor (SO_2/SO_3)

Table 5.10 Physical property data for Tube Side

Inlet temperature (t_1) / °C	507.10
Outlet temperature (t_2) / °C	460.00
Mass flow rate (kg/hr)	15338.70
Mass flow rate (kg/s)	4.26

5.5.2 Shell side data

Subject stream: Gas mixture from Absorber (SO_2/SO_3)

Table 5.11 Physical property data for Shell Side

Inlet temperature (T_1) / °C	444.5
Outlet temperature (T_2) / °C	501.4
Mass flow rate (kg/hr)	12299
Mass flow rate (kg/s)	3.42

5.5.3 LMTD

Following the basic LMTD formula,

$$LMTD = \frac{\Delta t_1 - \Delta t_2}{\ln(\Delta t_1 / \Delta t_2)}$$

To be able to calculate the final value of LMTD, it is first necessary to find the terminal differences between the temperatures

Cold terminal difference (Δt_1) = $t_2 - T_1$ = 460 – 444.5 = 15.5 °C

Hot terminal difference (Δt_2) = $t_1 - T_2$ = 507.1 – 501.4 = 5.7 °C

And hence,

$$LMTD = \frac{15.5 - 5.7}{\ln(^{15.5}/_{5.7})} = 9.80\ ^\circ\mathrm{C}$$

5.5.4 Corrected LMTD

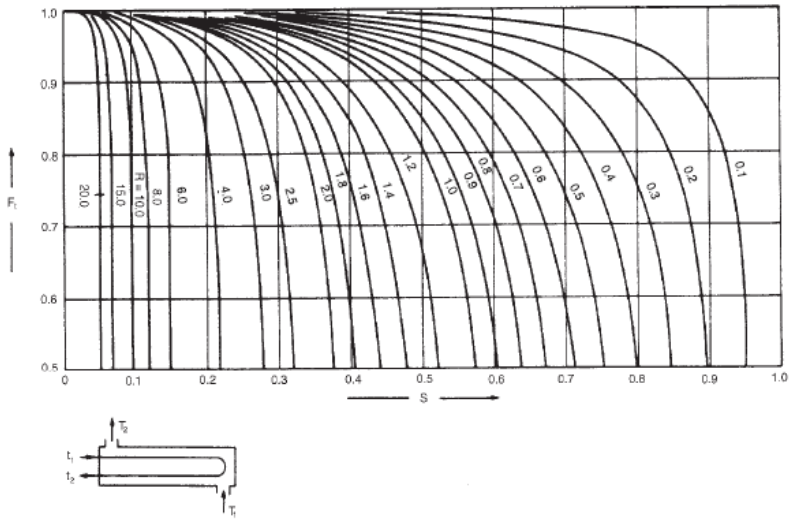
Now, if we talk about the LMTD correction factor (F_T), this factor is used to bring into account the actual or true temperature difference. This is basically needed because a heat exchanger is never completely counter current or co-current. Sometimes, it is a combination of both. And now we introduce the LMTD correction factor to account for the true temperature difference. This is done because the heat exchanger is not truly and solely either counter-current or co-current. It is a combination of both. The correction factor (F_T) depends on the geometry of the heat exchanger as well as the temperature of the inlets and outlets. Hence, we find the factors R & S, which will help define the correction factor (F_T) and the true LMTD.

$$R = \frac{T_1 - T_2}{t_2 - t_1} \qquad S = \frac{t_2 - t_1}{T_1 - t_1}$$

$$R = \frac{15 - 53.2}{65 - 78} = 1.21 \qquad S = \frac{460 - 507.1}{444.5 - 507.1} = 0.23$$

Using these values of R & S, we find the value of the F_T from the graph correlation

Table 66-Temperature Correction Factor: One Shell Pass; Two Or More Even Tube Passes



Using the values of R & S, we will be finding out a value for the correction factor, which will eventually help in finding the true value of the mean temperature difference.

$$F_T = 0.98$$

$$\Delta T_m = F_T \times LMTD = 0.98 \times 9.80 = 9.60\ ^\circ\mathrm{C}$$

5.5.5 Required UA

The value for the duty of the heat exchanger was already calculated and quoted in the energy balance chapter, and it was as follows.

Table 5.12 Data for UA calculation

Duty (Q) / (kJ / hr)	787543
ΔT_m / °C	9.60

And now, using the equation for basic heat transfer, we will be finding out the UA required in order for to design the heat exchanger.

$$Q = UA\Delta T_m$$

$$UA = {}^{Q}/_{T} = {}^{787543.8}/_{9.6} = 22786.8 = 2.3 \times 10^4\ {}^{W}/_{^\circ\mathrm{C}}$$

This is the value of the required UA, one of our design parameters is attaining a value of UA that is greater than or equal to this specific value.

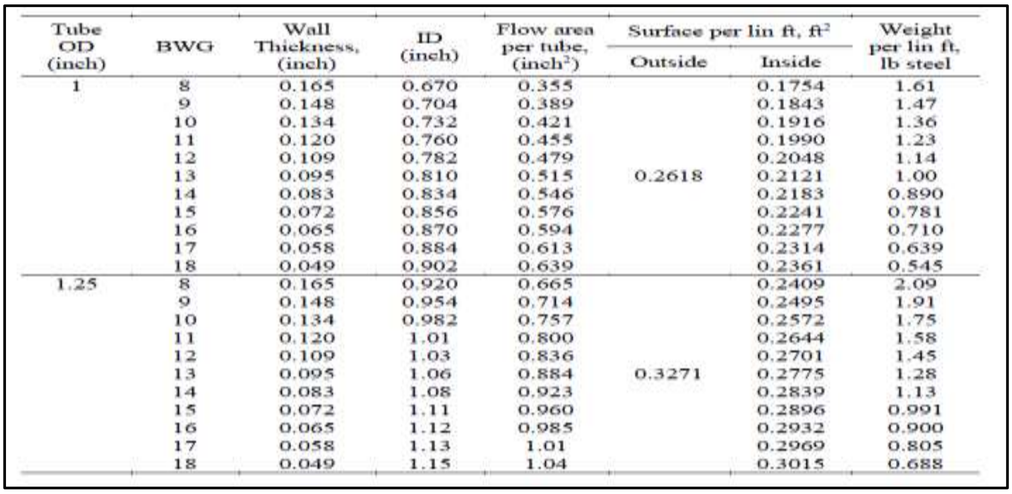
5.5.6 Tube Dimensions

Assumptions:

Table 5.13 Specification Sheet for Heat Exchanger Tubes

Number of tube passes (N)	2
Length of tube / m	6
Outer Diameter (OD) / m	0.032
BWG	10

Using this data as our basis, we perform further calculations in order to find out the other specifications. For the tube ID and the flow area the following data table is referred to:



Tube OD (inch)	BWG	Wall Thickness, (inch)	ID (inch)	Flow area per tube, (inch^2)	Surface per lin ft, ft^2		Weight per lin ft, lb steel
					Outside	Inside	
1	8	0.165	0.670	0.355	0.2618	0.1754	1.61
	9	0.148	0.704	0.389		0.1843	1.47
	10	0.134	0.732	0.421		0.1916	1.36
	11	0.120	0.760	0.455		0.1990	1.23
	12	0.109	0.782	0.479		0.2048	1.14
	13	0.095	0.810	0.515		0.2121	1.00
	14	0.083	0.834	0.546		0.2183	0.890
	15	0.072	0.856	0.576		0.2241	0.781
	16	0.065	0.870	0.594		0.2277	0.710
	17	0.058	0.884	0.613		0.2314	0.639
	18	0.049	0.902	0.639		0.2361	0.545
1.25	8	0.165	0.920	0.665	0.3271	0.2409	2.09
	9	0.148	0.954	0.714		0.2495	1.91
	10	0.134	0.982	0.757		0.2572	1.75
	11	0.120	1.01	0.800		0.2644	1.58
	12	0.109	1.03	0.836		0.2701	1.45
	13	0.095	1.06	0.884		0.2775	1.28
	14	0.083	1.08	0.923		0.2839	1.13
	15	0.072	1.11	0.960		0.2896	0.991
	16	0.065	1.12	0.985		0.2932	0.900
	17	0.058	1.13	1.01		0.2969	0.805
	18	0.049	1.15	1.04		0.3015	0.688

Figure 11-Heat Exchanger Tube Data

From this table the following data is found using the outer diameter and BWG assumed in the table above:

Table 5.14 Data from the Standard Ranges for Tube Specifications

Internal diameter (ID) / m	0.025
Flow area per tube / m^2	0.000488

For the number of tubes, we have to initially assume a value for U (the overall heat transfer coefficient), which can be found out from literature. The range of U found from literature allowed a range from 30 – 120 W/m^2°C when dealing with gases on both inlet and outlet streams. Hence, with the help of hit and trial, we were able to assume a value, which was most suitable for our design.

Table 5.15 Estimated Ranges for Value of Overall Heat Transfer Coefficient U

Allowable range of U (W/m^2°C)	30 - 120
Assumed value of U (W/m^2°C)	110

Using the value of UA as 22,786 as mentioned above, the required area is calculated as follows:

$$A = \frac{UA}{U} = \frac{22786}{110} = 207.15\ m^2$$

For the tube dimensions mentioned above, the area is calculated using the following formula:

$$Area\ per\ tube = \pi DL = \pi \times 0.032 \times 6 = 0.599\ m^2$$

The number of tubes required for this particular area would then be,

$$Number\ of\ tubes = \frac{Total\ area\ required}{Area\ of\ one\ tube} = \frac{207.15}{0.599} = 346.10 \approx 347\ tubes$$

The pitch chosen for this type of heat exchanger is triangular because of the higher rates of heat transfer provided by a triangular pitch along with the ease with which it can be cleaned. The tube pitch would be in accordance with the KERN method would be,

$$p_t = 1.25 d_o = 1.25\ \times 0.032 = 0.0397\ m$$

In addition, the tube clearance would then be calculated using the following formula,

$$Clearance = p_t - d_o = 0.0397 - 0.032 = 0.0079\, m$$

The tube dimensions so far are summarized in the table below.

Table 5.16 Tube Specification Summary (Heat Exchanger)

Tube Specification Summary	
Specification Title	**Specification Value**
Number of passes (N_p)	2
Tube length (L)	6 m
Outer Diameter (d_o)	0.032 m
Inner Diameter (d_i)	0.025 m
BWG	10
Number of tubes (N_t)	347
Flow area per tube	0.000488 m^2
Surface area per tube	0.599 m^2
Tube pitch (p_t)	Triangular & 0.0397 m
Tube clearance	0.0079 m

5.5.7 Shell Dimensions

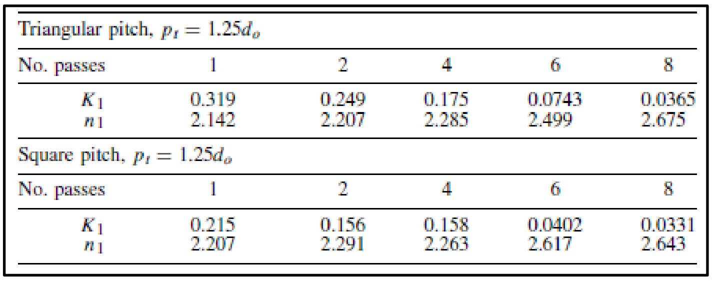
The following assumptions have been made for the calculation of shell dimensions:

Table 5.17 Shell Specifications

Number of shell passes	1
Baffle cut percentage (%)	25

In order to calculate the shell inside diameter (D_i), there are some parameters that need to be calculated namely the shell clearance and the bundle diameter.

For the bundle diameter (D_b) the following table is used to obtain the data required:



Triangular pitch, $p_t = 1.25d_o$					
No. passes	1	2	4	6	8
K_1	0.319	0.249	0.175	0.0743	0.0365
n_1	2.142	2.207	2.285	2.499	2.675
Square pitch, $p_t = 1.25d_o$					
No. passes	1	2	4	6	8
K_1	0.215	0.156	0.158	0.0402	0.0331
n_1	2.207	2.291	2.263	2.617	2.643

Figure 12-Bundle Diameter Data

By looking at the table and the assumptions taken above i.e. triangular pitch and 2 tube passes, the values derived from the above figure are 0.249 for K and 2.207 for N. Using these values the bundle diameter (D_b) is found out using the following formula:

$$D_b = d_o \left(\frac{N_t}{K_I}\right)^{1/n_1} = 0.032 \times \left(\frac{269}{0.249}\right)^{1/2.207} = 0.843\ m$$

For the shell bundle clearance, we refer to the following graph.

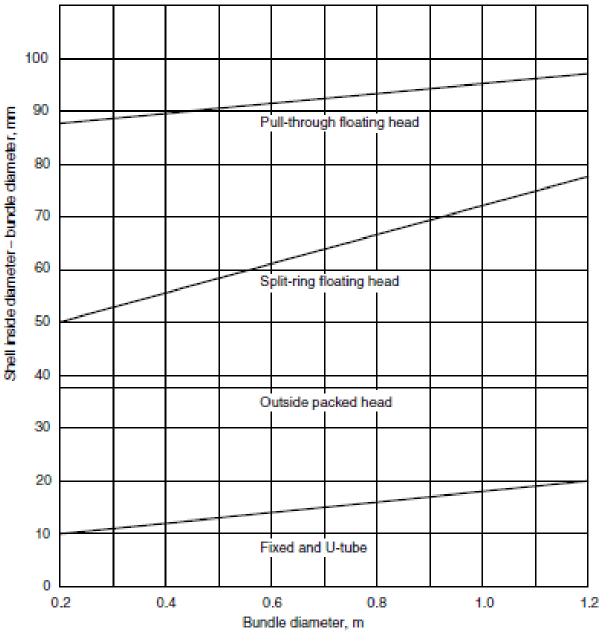


Figure 13-Shell Bundle Clearance

As mentioned at the start of our design, the arrangement used was pull through floating head, which when used along with the bundle diameter, gives the value for the clearance from the graph as follow

Shell bundle clearance = 0.095 m

Now, the shell inside diameter will be calculated using the following formula and values calculated above:

$$D_s = shell\ bundle\ clearance + bundle\ diameter = 0.095 + 0.843 = 0.938\ m$$

For the baffle spacing, by rule

$$\left(\frac{1}{5}\right) D_s < Baffle\ spacing\ (L_s)\ < D_s$$

So, we chose an intermediate baffle spacing factor i.e. 0.5.

$$Baffle\ spacing, L_s = 0.92 \times D_s = 0.92 \times 0.938 = 0.863\ m$$

And then,

$$Number\ of\ baffles = \frac{L}{L_s} - 1 = \frac{6}{0.863} - 1 = 6$$

A summary of the shell specifications has been mentioned in the table below:

Table 5.18 Shell Specification Summary (Heat Exchanger 2)

Shell Specification Summary	
Specification Title	**Specification Value**
Number of passes	1
Bundle diameter (D_b)	0.843 m
Shell bundle clearance	0.095 m
Shell inside diameter (D_s)	0.938 m
Baffle spacing (L_s)	0.863 m
Number of baffles	6
Baffle Cut	25 % cut baffles

5.5.8 Tube Side Coefficients

The stream flowing through the tube side is a mixture of gases, and the calculation being done is based on the physical properties of this particular gas mixture at an average value of temperature between the inlets and outlets. The tube wall temperature variation is taken to be negligible in this case as neither the temperature difference is high enough and nor the viscosity plays that important a role.

For the mean temperature,

$$Mean\ temperature\ (tube\ side) = \frac{t_1 + t_2}{2} = \frac{507.1 + 460}{2} = 483.55\ °\mathrm{C}$$

Physical property data for the gas mixture at 483.55 °C is as follows.

Table 5.19 Tube side physical property data (Heat Exchanger 2)

Property Tag	**Property Value**
Density of gas from reactor (Ρ) / (Kg/m^3)	0.5272
Viscosity of Solution (μ) / (kg/m.sec)	0.00003529
Heat Capacity (C$_p$) / (J/kg.°C)	1013
Thermal Conductivity (K$_f$) / (W/m.°C)	0.2245

For the value for h$_i$ (the inside heat transfer coefficient), the following steps are to be followed.

$$Tubes\ per\ pass = \frac{Total\ tubes}{Number\ of\ passes} = \frac{347}{2} = 173\ tubes\ per\ pass$$

$$Total\ flow\ area = Tubes\ per\ pass\ \times Flow\ area\ per\ tube = 173 \times 0.000488$$
$$= 0.0845\ m^2$$

$$Mass\ velocity = \frac{Mass\ flow\ rate\ (per\ second)}{Total\ flow\ area} = \frac{4.26}{0.0845} = 50.42\ {}^{kg}/_{s.m^2}$$

$$Reynold's\ number\ (Re) = \frac{Mass\ velocity\ \times tube\ inner\ diameter}{Viscosity\ of\ solution}$$

$$= \frac{50.42 \times 0.025}{0.00003529} = 35633.6$$

The value for the Heat transfer factor (j_h) can be found from the figure below.

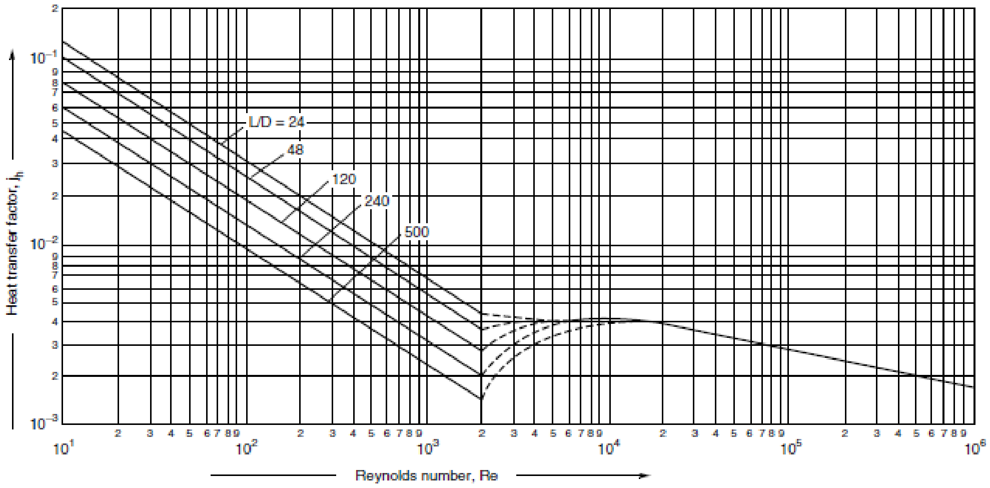


Figure 14-Tube Side Heat Transfer Factor jH

From the figure, for Re = 35,634

$$j_h = 0.0035$$

Now,

$$h_i = \left(\frac{k_f}{d_i}\right) j_h Re \left(\frac{C_p \mu}{k_f}\right)^{1/3} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

$$h_i = \left(\frac{0.2245}{0.025}\right) \times 0.0035 \times 33634 \times \left(\frac{1013 \times 0.00003529}{0.2245}\right)^{1/3} (1)^{0.14}$$

$$h_i = 608.81\ W/m^2 °C$$

And now to refer the value of the coefficient to the outer surface of the tubes,

$$h_{io} = h_i \times \frac{d_i}{d_o} = 608.81 \times \frac{0.025}{0.032} = 478.3\ W/m^2 °C$$

5.5.10 Shell Side Coefficients

The stream flowing on the shell side is the gas mixture coming from the absorber and another heat exchanger, and calculation being done is based on the physical properties of this particular gas mixture at an average value of temperature between the inlets and outlets. The tube wall temperature variation is taken to be negligible in this case as neither the temperature difference is high enough and nor the viscosity plays that important a role.

For the mean temperature,

$$Mean\ temperature\ (shell\ side) = \frac{T_1 + T_2}{2} = \frac{444.5 + 501.4}{2} = 472.95\ °C$$

Physical property data for cooling water at 472.95 °C is as follows.

Table 5.20 Shell Side Physical Property Data (Heat Exchanger 2)

Property Tag	Property Value
Density of Solution (P) / (kg/m^3)	0.561
Viscosity of Solution (μ) / (kg/m.sec)	0.0000338
Heat Capacity (C$_p$) / (J/kg.°C)	1404
Thermal Conductivity (K$_f$) (W/m.°C)	0.0507

For the value for h$_o$ (the outside heat transfer coefficient), the following steps are to be followed.

$$Shell\ equivalent\ diamater, d_e = \left(\frac{1.10}{d_o}\right)(p_t^2 - 0.917 d_0^2)$$

$$d_e = \left(\frac{1.10}{0.032}\right)(0.0397^2 - 0.917(0.032)^2)$$

$$d_e = 0.0225\ m$$

$$Cross\ flow\ area, A_s = \frac{D_s \times clearance \times L_s}{p_t} = \frac{0.938 \times 0.0079 \times 0.863}{0.0397}$$

$$A_s = 0.162\ m^2$$

$$Mass\ velocity, G_s = \frac{Mass\ flow\ rate\ (per\ second)}{Cross\ flow\ area} = \frac{3.42}{0.162} = 21{,}10\ {}^{kg}/_{s.m^2}$$

$$Reynold's\ number\ (Re) = \frac{G_s\ \times d_e}{\mu} = \frac{21.10 \times 0.0225}{0.00003378} = 14083.33$$

The value for the Heat transfer factor (j_h) can be found from the figure below.

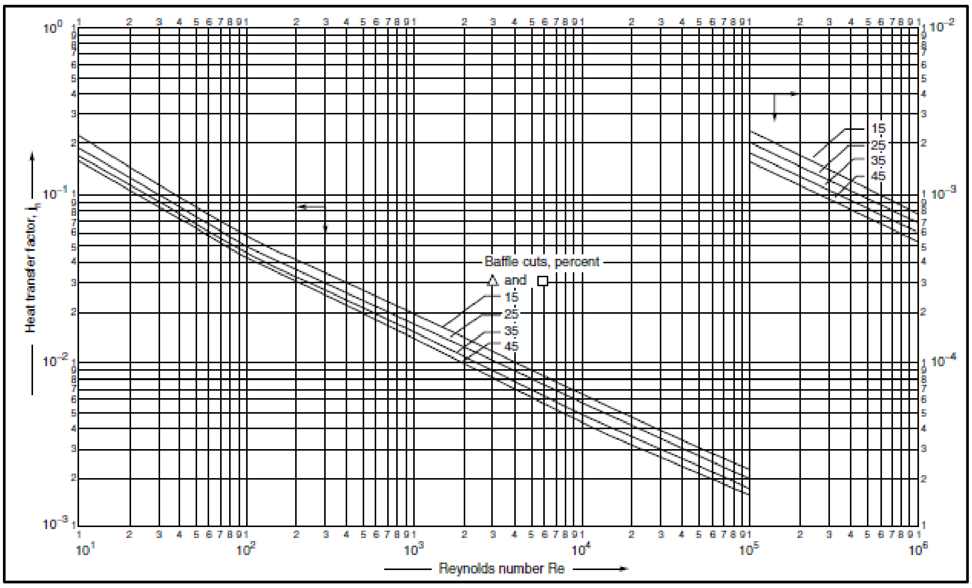


Figure 15-Shell-Side Heat-Transfer Factors, Segmental Baffles

From the figure, for Re = 14083 and 25 baffle cut percent

$$j_h = 0.005$$

Now,

$$h_o = \left(\frac{k_f}{d_e}\right) j_h Re \left(\frac{C_p \mu}{k_f}\right)^{1/3} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

$$h_o = \left(\frac{0.0507}{0.0225}\right) \times 0.005 \times 14083 \times \left(\frac{1404 \times 0.00003378}{0.0507}\right)^{1/3} (1)^{0.14}$$

$$h_o = 155\ W/m^2{}^{\circ}\mathrm{C}$$

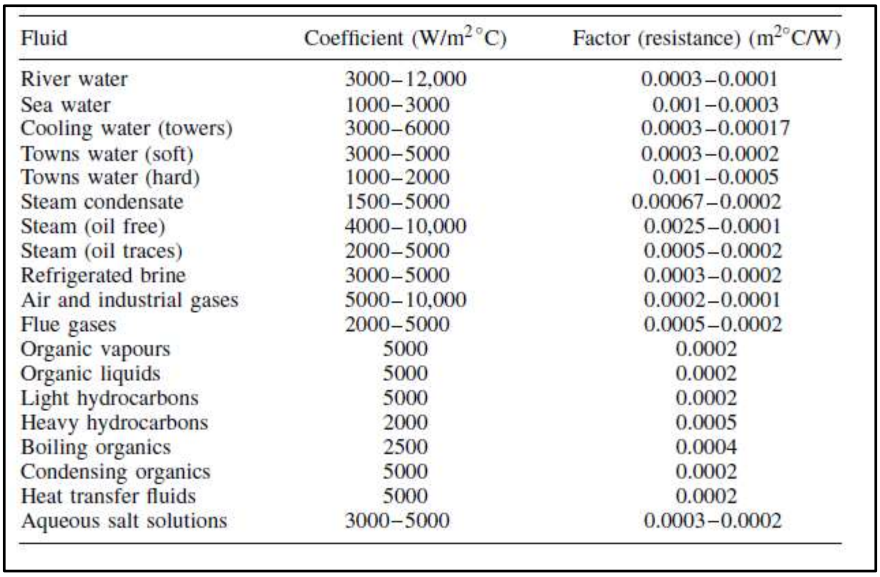
5.5.11 Overall Coefficient

Some basic data relevant to the fouling factor coefficient for the tube and shell side streams can be found form the figure below.

Thermal conductivity of cupro-nickel alloys, k_w = 50 W/m°C

Shell side fouling coefficient (organic vapors), h_{od} = 5000 W/m^2°C

Tube side fouling coefficient (organic vapors), h_{id} = 5000 W/m^2°C



Fluid	Coefficient (W/m^2°C)	Factor (resistance) (m^2°C/W)
River water	3000–12,000	0.0003–0.0001
Sea water	1000–3000	0.001–0.0003
Cooling water (towers)	3000–6000	0.0003–0.00017
Towns water (soft)	3000–5000	0.0003–0.0002
Towns water (hard)	1000–2000	0.001–0.0005
Steam condensate	1500–5000	0.00067–0.0002
Steam (oil free)	4000–10,000	0.0025–0.0001
Steam (oil traces)	2000–5000	0.0005–0.0002
Refrigerated brine	3000–5000	0.0003–0.0002
Air and industrial gases	5000–10,000	0.0002–0.0001
Flue gases	2000–5000	0.0005–0.0002
Organic vapours	5000	0.0002
Organic liquids	5000	0.0002
Light hydrocarbons	5000	0.0002
Heavy hydrocarbons	2000	0.0005
Boiling organics	2500	0.0004
Condensing organics	5000	0.0002
Heat transfer fluids	5000	0.0002
Aqueous salt solutions	3000–5000	0.0003–0.0002

Figure 16-Fouling Factor (Coefficients)

And now to refer the value of the coefficient to the outer surface of the tubes,

$$h_{iod} = h_{id} \times \frac{d_i}{d_o} = 5000 \times \frac{0.025}{0.032} = 3906 \frac{W}{m^2} ℃$$

And now to finally calculate the overall coefficient (U$_o$) based on the outside area of the tube,

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o ln\left(\frac{d_o}{d_i}\right)}{2k_w} + \frac{1}{h_{io}} + \frac{1}{h_{iod}}$$

$$= \frac{1}{154.8} + \frac{1}{5000} + \frac{0.032 \times ln\left(\frac{0.032}{0.025}\right)}{2 \times 50} + \frac{1}{478.3} + \frac{1}{3906}$$

$$\frac{1}{U_o} = 0.00903$$

$$U_o = 110.8\, W/m^2 ℃$$

This obtained value of U$_o$, is above the initially assumed value of the coefficient which was equal to 110 W/m^2℃

Additionally, in order to calculate the value for the UA delivered by the heat exchanger is as follows:

- U_o achieved = 110.8 W/m^2°C
- Surface area per tube = 0.599 m^2
- UA required for suitable performance (from previous calculation) = 22786 W/°C

$$Total\ heating\ area = area\ per\ tube\ \times\ tubes = 0.599 \times 347 = 207.2\ m^2$$

$$UA\ achieved =\ U_o \times total\ heating\ area = 110.79 \times 207.15 = 22951\ W/°C$$

Hence as the UA achieved > UA required; we have been able to satisfy the first design parameter.

5.5.12 Tube Side Pressure Drop

For our tube side, the pressure drop calculation proceeds as follows.

For the number of tube passes, N_p = 2 and Re = 35,633 the friction factor j_f from the graph below is 0.0035.

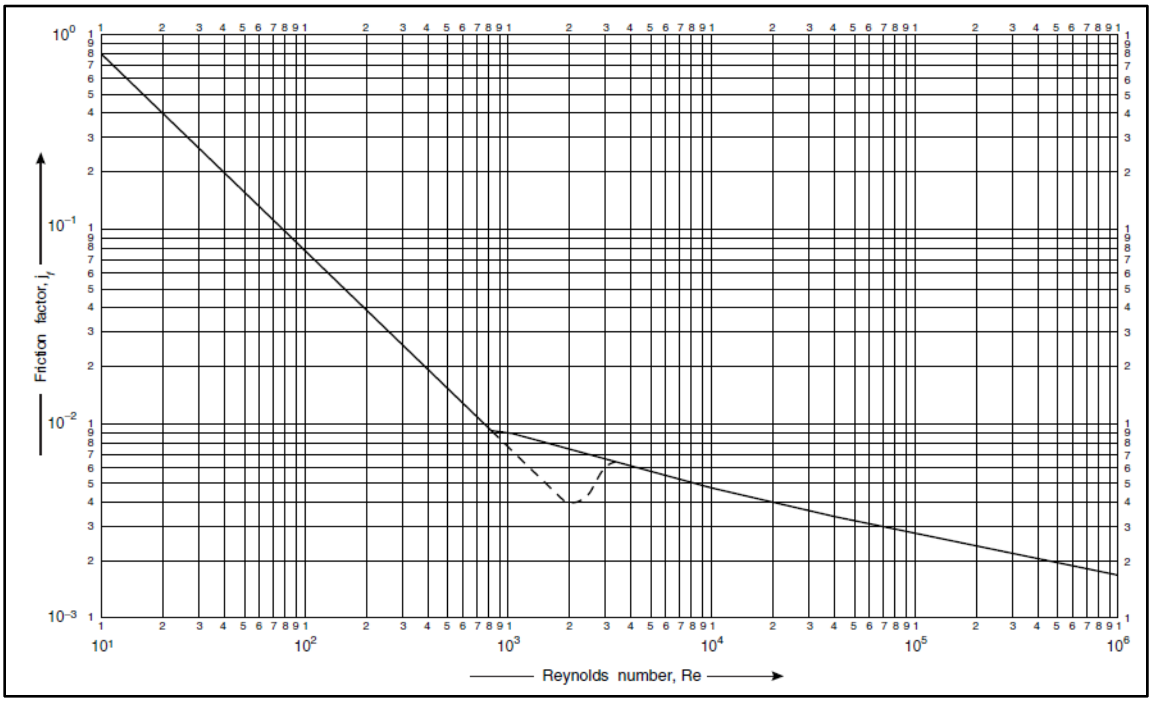


Figure 17-Tube Side Friction Factors

For linear velocity,

$$u_t = \frac{mass\ velocity}{density\ of\ gas\ mixture} = \frac{50.4}{0.527} = 95.63\ m/s$$

For the tube side pressure drop,

$$\Delta P_t = N_p\left[8j_f\left(\frac{L}{d_i}\right)\left(\frac{\mu}{\mu_w}\right)^{-m} + 2.5\right]\left(\frac{\rho u_t^2}{2}\right)$$

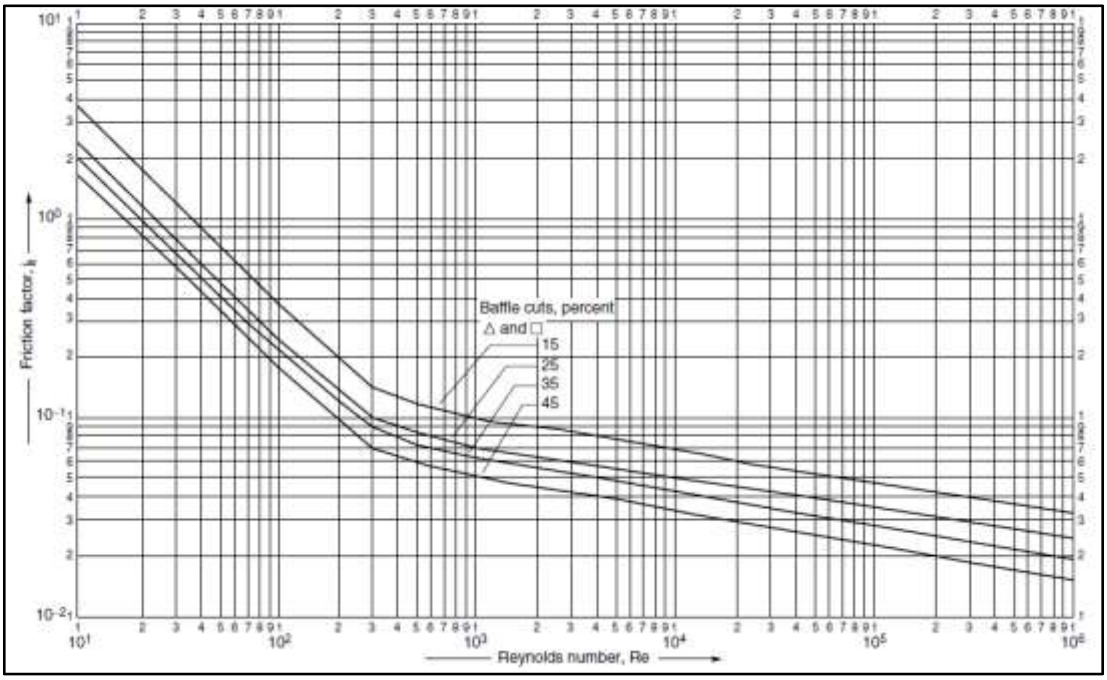
$$= 2 \times \left[8 \times 0.0035 \times \left(\frac{6}{0.025}\right) + 2.5\right]\left(\frac{0.5272 \times (95.62)^2}{2}\right)$$

$$\Delta P_t = 44525.7\ \frac{N}{m^2}$$

5.5.13 Shell Side Pressure Drop

For our shell side, the pressure drop calculation proceeds as follows.

For a Re = 14083, the friction factor j_f from the graph below is 0.048.



For linear velocity,

$$u_s = \frac{mass\ velocity}{density\ of\ gas\ mixture} = \frac{21.1}{0.561} = 37.6\ m/s$$

For the shell side pressure drop,

$$\Delta P_s = 8j_f\left(\frac{D_s}{d_e}\right)\left(\frac{L}{L_s}\right)\left(\frac{\rho u_s^2}{2}\right) = 8 \times 41.60 \times \left(\frac{0.809}{0.0225}\right)\left(\frac{6}{0.863}\right)\left(\frac{0.5611 \times (37.6)^2}{2}\right)$$

$$\Delta P_s = 44081.2\ \frac{N}{m^2} = 44.081\ kPa = 6.39\ psi$$

So, as both the shell side and tube side have a pressure drop less than 10 psi, the heat exchanger is suitable for the operation and with that we have our second design variable satisfied as well.

Summary: The Proposed Design

Table 67-Heat Exchanger -2 Specification Summary

Heat Exchanger – 2			
Heat exchanger type	Shell & Tube HE	**TEMA designation**	Pull through, floating head
Exchanger duty	787543 kJ/hr	Heat transfer area	207.2 m^2
U_o required	110 W/m^2°C	U_o achieved	110.8 W/m^2°C
U_oA required	22786 W/°C	U_oA achieved	22951 W/°C
Tube Side		**Shell Side**	
Subject stream	Gas mixture	Subject stream	Gas mixture
Construction material	Carbon steel	Construction material	Carbon steel
Flow rate	15,338 kg/hr	Flow rate	12,299 kg/hr
Inlet temperature	507.1 °C	Inlet temperature	444.5 °C
Outlet temperature	460 °C	Outlet temperature	501.4°C
Number of passes	2	Number of passes	1
Number of tubes	347	Tube bundle diameter	0.843 m
Length of tubes	6 m	Shell bundle clearance	0.095 m
Outer diameter	0.032 m	Inner diameter	0.938 m
BWG	10	Equivalent diameter	0.0225 m
Inner diameter	0.025 m	Baffle spacing	0.863 m

Flow area per tube	0.000488 m^2	Number of baffles	6
Tube pitch	0.0397 m Triangular	Baffle cut	25 % cut
Tube clearance	0.0079 m	Cross flow area	0.162 m^2
(h_{io})	478 W/m^2°C	*(h_o)*	155 W/m^2°C
Fouling coefficient	5000 W/m^2°C	Fouling coefficient	5000 W/m^2°C
Pressure drop	6.460 psi	Pressure drop	6.39 psi

Table 68-Heat Exchanger -3 Specification Summary

Heat Exchanger – 3			
Heat exchanger type	Shell & Tube HE	**TEMA designation**	Split ring, floating head
Exchanger duty	3510271 kJ/hr	Heat transfer area	150.9 m^2
U_o required	110 W/m^2°C	U_o achieved	110.98 W/m^2°C
U_oA required	16606 W/°C	U_oA achieved	16754 W/°C
Tube Side		**Shell Side**	
Subject stream	Gas mixture	Subject stream	Gas mixture
Construction material	Carbon steel	Construction material	Carbon steel
Flow rate	15,338 kg/hr	Flow rate	12,299 kg/hr
Inlet temperature	490 °C	Inlet temperature	200°C
Outlet temperature	290 °C	Outlet temperature	444.5°C
Number of passes	2	Number of passes	1
Number of tubes	189	Tube bundle diameter	0.641 m

Length of tubes	8 m	Shell bundle clearance	0.093 m
Outer diameter	0.032 m	Inner diameter	0.734 m
BWG	10	Equivalent diameter	0.0225 m
Inner diameter	0.025 m	Baffle spacing	0.697 m
Flow area per tube	0.000570 m^2	Number of baffles	10
Tube pitch	0.0397 m Triangular	Baffle cut	25 % cut
Tube clearance	0.0079 m	Cross flow area	0.102 m^2
(h_{io})	659 $W/m^2°C$	(h_o)	145 $W/m^2°C$
Fouling coefficient	4500 $W/m^2°C$	Fouling coefficient	4500 $W/m^2°C$
Pressure drop	8.950 psi	Pressure drop	9.24 psi

5.7 Packed Bed Catalytic Reactor

The reactor chosen for our process i.e. Contact process is Packed Bed Catalytic Reactor. The successive bed arrangement allows for maximum conversion of SO_2 to SO_3 with intermediate cooling in between the beds. (Moulijin, 2013) The reaction is exothermic and hence the temperature of the gases increases on each bed. If the streams are not brought back to the optimum temperature of 450°C, the equilibrium will shift towards the reactant side which results in less conversion.

Now we will discuss the elaborate design process of our reactor. (Levenspiel, Chemical Reaction Engineering, 1999)

5.7.1 Calculation of equilibrium constant K_P

The value of equilibrium constant is calculated to aid us in finding out the rate of equilibrium reaction. (Wikipedia, 2012)

Table 5.23 Temperature Conversion Data

Kelvin to Rankine conversion	1 R= 1.8K
Temperature in K	723
Temperature in R	1301.4

The formula used to calculate the value of K_P is

$$K_P = \exp\left(\frac{42311}{R * T}\right) - 11.24$$

Where R= 1.986 Btu/lbmol.R

$$K_P = \exp\left(\frac{42311}{1.986 * 1301.4}\right) - 11.24$$

$K_P = 169.1 \text{ atm}^{-0.5}$

5.7.2 Calculation of rate constant k

The formula used is

$$k = \exp\left[\left(\frac{-176008}{T}\right) - 110.1 \ln T + 912.8\right]$$

k = 6.18 e-06 lbmol/lb cat.sec.atm

5.7.3 Calculation of rate of reaction

The physical property data required is:

Table 5.24 Pressure data for rate equation

P (atm)	2
P_{SO2}	0.54
P_{SO3}	0.486
P_{O2}	0.974

The rate equation is:

$$r_{SO_2} = k * \left(\frac{P_{SO_2}}{P_{SO_3}}\right)^{0.5} * \left[P_{O_2} - \left(\frac{P_{SO_3}}{K_P * P_{SO_2}}\right)^2\right]$$

$$r_{SO_2} = 6.18E^{-06} * \left(\frac{0.54}{0.486}\right)^{0.5} * \left[0.974 - \left(\frac{0.486}{169.1 * 0.54}\right)^2\right]$$

r_{SO_2} = 6.35 e-06 lbmol/lb cat.sec

5.7.4 Calculation of total weight of catalyst

Feed of SO_2 inlet = 0.129 lbmol/sec

The formula for weight of catalyst is

$$W = \left(\frac{F_{ao}}{-r_a}\right) * \int_0^x dx$$

$$W\ (lb) = \left(\frac{0.129}{6.35e - 06}\right) * (0.997 - 0)$$

W = 20259 lb = 9177 kg

5.7.5 Calculation of Volume of Catalyst and Volume of Bed

Density of V_2O_5 catalyst = 1281.5 kg/m^3

$$Volume\ of\ catalyst = \frac{Mass\ of\ catalyst}{Density\ of\ catalyst}$$

Volume of catalyst = 7.17 m^3

Assumption:

Void fraction = 0.4

$$Volume\ of\ bed = \frac{Volume\ of\ catalyst}{1 - void\ fraction}$$

Volume of bed = 11.9 m^3

5.7.6 Calculation of Reactor Specifications

$$Diameter\ of\ bed = \left(\frac{2V}{\pi}\right)^{\frac{1}{3}}$$

Diameter = 1.95 m

$$Height\ of\ bed = \left(\frac{4V}{\pi * D^2}\right)$$

Height of bed = 3.91 m

$$Height\ of\ reactor = Height\ of\ bed + 0.25 * Height\ of\ bed$$

Height of reactor = 4.49 m

$$Volume\ of\ reactor = \left(\frac{\pi * D^2}{4}\right) * H$$

Volume of reactor = 13.5 m^3

5.7.7 Calculation of catalyst and bed specifications

As the concentration of SO_2 decreases down each bed, the weight of catalyst required on each bed will change and hence the rate of reaction will change.

The diameter will remain throughout the length of the reactor. However the height of each bed will also vary due to different weight of catalyst on each bed. (Donobam, 1984)

5.7.7.1 1st Bed

Table 5.25 Pressure Data for Rate Equation

$P1_{O2}$	1.01
$P1_{SO2}$	0.988
$P1_{SO3}$	0.002

$$r_{SO_2} = k * \left(\frac{P_{SO_2}}{P_{SO_3}}\right)^{0.5} * \left[P_{O_2} - \left(\frac{P_{SO_3}}{K_P * P_{SO_2}}\right)^2\right]$$

$$r_{SO_2} = 6.18E - 06 * \left(\frac{0.988}{0.002}\right)^{0.5} * \left[1.01 - \left(\frac{0.002}{169.1 * 0.988}\right)^2\right]$$

r_{SO_2}= 0.000139 lbmol/lb cat.sec = 0.000139 kgmol/kg cat.sec

Table 5.26 Data for use in rate equation

SO2 molar flow rate (kgmol/hr)	66.9
Air Molar Flow rate (kgmol/hr)	24.1
Fao (kgmol/hr)	91
Conversion	0.63

Table 69-Specification sheet for 1st bed

W1 (kg)	115
Volume of catalyst (m^3)	0.0895
Volume of bed (m^3)	0.149
Diameter (m)	1.95
Height of bed (m)	0.0498

5.7.7.2 2nd Bed

Table 5.28 Pressure data for Rate Equation

$P2_{O2}$	0.880
$P2_{SO2}$	0.460
$P2_{SO3}$	0.646

r_{SO_2}= 4.59 e-06 kgmol/kg cat.sec

Table 5.29 Data required for rate equation

SO_2 molar flow rate (kgmol/hr)	71.8
SO_2 molar flow rate (kgmol/sec)	0.0199
Conversion	0.567

Table 70-Specification sheet for 2nd Bed

W2 (kg)	2463
Volume of catalyst (m3)	1.92
Volume of bed (m3)	3.20
Diameter (m)	1.95
Height (m)	1.07

5.7.7.3 3rd bed

Table 5.31 Pressure data for Rate Equation

$P3_{O2}$	0.752
$P3_{SO2}$	0.198
$P3_{SO3}$	1.05

r_{SO_2}= 2.02 e-06 kgmol/kg cat.sec

Table 5.32 Data required for rate equation

SO_2 molar flow rate (kgmol/hr)	72.3
SO_2 molar flow rate (kgmol/sec)	0.0201
Conversion	0.563

Table 71-Specification sheet for 3rd bed

W3 (kg)	5607
Volume of catalyst (m3)	4.38
Volume of bed (m3)	7.29
Diameter (m)	1.95
Height (m)	2.43

5.7.7.4 4th bed

Table 5.34 Pressure Data for Rate Equation

$P4_{O2}$	1.56
$P4_{SO2}$	0.196
$P4_{SO3}$	0.244

r_{SO_2} = 8.64 e-06 kgmol/kg cat.sec

Table 5.35 Data required for use in rate equation

SO_2 molar flow rate (kgmol/hr)	32.2
SO_2 molar flow rate (kgmol/sec)	0.00896
Conversion	0.957

Table 72-Specification sheet for 4th bed

W4 (kg)	991.6
Volume of catalyst (m3)	0.774
Volume of bed (m3)	1.29
Diameter (m)	1.95
Height (m)	0.430

Table 73-Specification sheet for the combined dimensions of all beds

Reactor bed specification sheet	
Total height of beds (m)	3.88
Volume of bed (m^3)	11.9
Total weight of catalyst (kg)	9177

5.8 Sulphuric Acid Storage Tank

The sulphuric acid storage tank is required for the storage of the Sulphuric acid to be extracted as a final product or to be recycled back to the initial locations. In order to estimate the cost of the storage tank correctly and aptly, it is necessary to find out certain design specifications with the help of calculations shown below.

5.8.1 Calculation of volume of tank

In order to calculate the volume of the tank, we need to know the volumetric flow rate (V_F) of the Sulphuric Acid that is being collected in the storage tank; the following physical properties are needed for the calculation of volumetric flow rate (V_F):

Table 5.38 Physical properties of H_2SO_4 entering storage tank

Mass flow rate (kg/hr)	8958
Density of H_2SO_4 (kg/m^3)	1781

$$V_F = \frac{Mass\ flow\ rate}{Density\ of\ H_2SO_4}$$

$$V_F = \frac{8958}{1781}$$

Volumetric flow rate = 5.03 m^3/hr

Instead of taking the volume of the tank (V_T) equal to the volumetric flow rate (V_F), a safety clearance (S_F) is included to make sure that the size of the tank is safe from overflowing or other hazards. The value of safety clearance (S_F) is taken to be 0.05 from literature.

$$V_T = V_F\ (1 + \ S_F)$$

$$V_T = 5.03\ (1 + 0.05)$$

Volume of tank = 5.28 m^3

5.8.2 Calculation of the Diameter of tank

The calculation of the internal diameter (ID) is done using the volume of the tank (V_T) using a simple geometric formula:

$$ID = \sqrt[3]{\frac{2\ x\ V_T}{\pi}}$$

$$ID = \sqrt[3]{\frac{2\ x\ 5.28}{\pi}}$$

Internal Diameter = 1.49 m

Now, for the calculation of the outer diameter (OD), wall thickness (T_M) is incorporated in the formula, which is taken to be 0.006 to account for the difference between the internal and outer diameters:

$$OD = ID + (2\ x\ T_M)$$

$$OD = 1.49 + (2\ x\ 0.006)$$

Outer Diameter = 1.50 m

5.8.3 Calculation of the Height of tank

In order to calculate the height of the tank, volumetric flow rate is first used for the calculation of height of the liquid (H_L).

$$H_L = \sqrt[3]{\frac{4\ x\ V_F\ x\ 4}{\pi}}$$

$$H_L = \sqrt[3]{\frac{4\ x\ 5.03\ x\ 4}{\pi}}$$

Height of the liquid (H_L) = 2.92 m

The pressure head (ΔP) created due to this height can be calculated using a very simple formula mentioned below:

$$\Delta P = \frac{\rho\ x\ g\ x\ H_L}{1000}$$

$$\Delta P = \frac{1781\ x\ 9.81\ x\ 2.92}{1000}$$

Pressure head = 50.9 kN/m^2

For the total pressure (T_P), atmospheric pressure needs to be added to the pressure head:

$$T_P = \ \Delta P + 101.3$$

$$T_P = \ 50.9 + 101.3$$

Total pressure = 152.25 kN/m^2

Height of the tank (H_T) is again calculated, using a very simple ratio where the formula is as follows:

$$H_T = 2\ x\ OD$$

$$H_T = 2\ x\ 1.50$$

Height of the tank = 3.00m

Table 74-Specification Summary of Storage Tank

Storage Tank Specification Sheet	
Volumetric flow of liquid (m^3/hr)	5.03
Volume of the tank (m^3)	5.28
Internal Diameter (m)	1.49
Out Diameter (m)	1.50
Height of liquid (m)	2.92
Height of Tank (m)	3.00
Total Pressure (kN/m^2)	152

5.9 Oleum Header

This equipment is needed and designed to collect the oleum coming from different locations in the plant and is distributed to other locations as well as extracted as a product. It may also be processed further to make Sulphuric Acid in the dilution Tank. The method of calculation of its design is exactly like the storage tank.

Table 75-Specification Summary of Oleum Header

Oleum Header Specification Sheet	
Volumetric flow of liquid (m^3/hr)	6.75
Volume of the tank (m^3)	7.09
Internal Diameter (m)	1.64
Out Diameter (m)	1.66
Height of liquid (m)	3.21
Height of Tank (m)	3.31
Total Pressure (kN/m^2)	159

5.10 Scrubber

The scrubber is installed after MAT to further remove the SO_X from the flue gases before discharging them into the atmosphere. The design is based on physical absorption process.

5.10.1 Calculation of diameter

Table 5.41 Physical property Data for Diameter of Scrubber

Volumetric flow rate (m3/sec)	3.35
Gas velocity/u_g (m/s)	1.50

According to literature, the value of gas velocity should ideally be between 0.3 and 2.3 m/s.

$$Diameter = \sqrt{\left(\frac{4 * V}{\pi * gas\ velocity}\right)}$$

Diameter = 1.685 m

5.10.2 Calculation of Specifications of Spray section

Table 5.42 Physical property Data for Specifications of Spray Section

Gas flow rate Gm (kgmol/sec)	0.113
Mol. frac. Of SO2/y1 (gas inlet)	0.00033
Mol. frac. Of SO2/y2 (gas outlet)	0.0001

$$N_{OG} = \frac{y1 - y2}{y2}$$

N_{OG} = 2.3

K_g = 0.0409 kmol/m^3.sec (from literature)

$$Volume = \frac{N_{OG} * G_M}{K_g}$$

Volume = 6.37 m^3

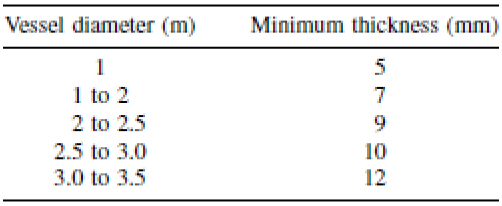
$$Height = \frac{4 * V}{\pi * D^2}$$

Height = 2.85 m

$$Area = \pi * r^2 + 2 * \pi * r * h$$

R=D/2=0.843 m

Area = 17.3 m^2



Vessel diameter (m)	Minimum thickness (mm)
1	5
1 to 2	7
2 to 2.5	9
2.5 to 3.0	10
3.0 to 3.5	12

5.10.3 Mechanical Properties of Spray Tower

Table 76-Mechanical properties of Spray Tower

Mechanical Properties of Spray Section	
Material of construction	Carbon Steel
Operating pressure (N/m2)	100000
Design pressure (N/m2)	10000
Joint factor	1.0
Stress factor-Carbon Steel (N/m2)	125000000
Wall thickness-diameter (mm)	0.0793
Corrosion allowance-Carbon Steel(mm)	2
Total wall thickness (mm)	2.08

5.11 Dilution Tank

The oleum produced in the absorption towers is finally sent to the dilution tank where it reacts with water to produce Sulphuric Acid. Our dilution tank is a continuously stirred tank reactor because firstly the output composition is identical to the composition of the material inside the reactor and because the temperature and pH control is relatively easy in the CSTR. Hence, the design calculations are based on the design equation of a CSTR. The reaction-taking place in the dilution tank is as follows:

$$H_2S_2O_7 + H_2O \rightarrow 2\ H_2SO_4$$

5.11.1 Calculation of Volume of the Dilution Tank

The physical property data required for the calculation of volume are given in the table below.

Table 5.41 Physical property data for calculating volume of the reactor

Molar flow rate of Oleum / F_{A0} (kgmol/hr)	66.9
Molar flow rate of Water / F_{B0} (kgmol/hr)	25.0
Volumetric flow rate of Oleum / V_{A0} (m^3/hr)	4.27
Volumetric flow rate of water / V_{B0} (m^3/hr)	0.27
Conversion / X	0.99
K / (m^3/kgmolsec)	6×10^{-5}

In order to move forward with our calculations, the first thing that is needed for the design calculation is the concentration of both oleum (C_{A0}) and water (C_{B0}) with a simple formula:

$$C_{A0} = \frac{F_{A0}}{V_{A0}}$$

Concentration of Oleum = 15.68 kgmol/m

$$C_{B0} = \frac{F_{B0}}{V_{B0}}$$

Concentration of water = 92.70 kgmol/m^3

In order to calculate the volume of our dilution tank, we first need to find the value of r_A, for this purpose the order of the reaction is taken as 2 because the reaction between oleum and water is considered to be an elementary reaction and hence, the value of r_A is calculated using the following formula: (Levenspiel, Chemical Reaction Engineering, 1999)

$$r_A = k * C_{AO}(1 - X_A) * (C_{B0} - (C_{A0} * X_A))$$

$$r_A = 0.00006 \ * 15.7 * (1 - 0.99)(92.7 - (15.7 * 0.99))$$

$r_A = 7.26 \times 10^{-4}$ kgmol/m^3sec

Now, that we have the value of r_A the next step only involves plugging the values in the design equation to calculate the volume (V_{DT}) of the dilution tank:

$$V_{DT} = \frac{(F_{AO} + F_{BO}) * X_A}{r_A}$$

$$V_{DT} = \frac{(66.9 + 25) * 0.99}{0.000726}$$

Volume of the tank = 34.84 m^3

5.11.2 Calculation of diameter of tank

Using the volume of the tank, the diameter (D_{DT}) of the dilution tank can be found out using a simple geometric formula, which is mentioned below:

$$D_{DT} = \sqrt[3]{\frac{4 * V_{DT}}{2\pi}}$$

$$D_{DT} = \sqrt[3]{\frac{4 * 34.8}{2\pi}}$$

Diameter of the dilution tank = 2.81m

5.11.3 Height of the column

In the industrial plant, a suitable L/D ratio = 2 is considered a good design, so in order to calculate the height the following equation is used:

$$H_{Dt} = 2 * D_{DT}$$

$$H_{Dt} = 2 * 2.81$$

Height of the tank = 5.61m

Table 77-Specification Summary of Dilution Tank

Dilution tank Specification Sheet	
r_A (kgmol/m^3sec)	7.26 x 10^{-4}
Volume of Dilution Tank (m)	34.8
Diameter of Dilution Tank (m)	2.81
Height of Dilution Tank (m)	5.61

CHAPTER 6

Simulation

The simulation model for our plant was developed for the purpose of reaching the most optimum conditions that ensured high yield of our final products i.e. Sulphuric Acid and Oleum and enable us to design a GREEN process.

6.1 General Process Description

The moisture is removed from the air in the Air Dryer using Oleum as a dehydrating agent. A fraction of the dried air then reacts with molten sulfur in Sulfur Burner to produce SO_2 gas. This SO_2 gas mixture is sent to an SO_2 dryer to ensure maximum removal of moisture content before it enters the packed bad catalytic reactor. The high temperature gas from the burner is cooled to the optimum reactor temperature of 450°C. In our simulation model, the four beds of our reactor can be seen as four separate convertors due to limitations in Aspen HYSYS. The remaining fraction of the air from Air Dryer reacts with SO_2 gas to form SO_3. The catalyst used for this oxidation reaction is V_2O_5 (vanadium pentaoxide). Since the reaction is exothermic, it results in an increase in temperature of the outlet streams. Heat exchangers are used to cool down the temperature of the streams leaving the reactor. This is crucial as the SO_3 produced in the reactor has to react with sulphuric acid to form oleum and absorption reactions are favorable at low temperatures. After maximum conversion of SO_3 to oleum is achieved using an intermediate and a main absorption column, the oleum produced is split into fractions for use throughout the plant as a dehydrating agent. The unconverted gases in the second absorber are sent to a scrubber to bring down the amount of the toxic gases to an allowable limit before releasing them into the atmosphere. Part of the oleum extracted from oleum header is sent to the dilution tank to react with water and form concentrated sulphuric acid. (Sultana, 2011)

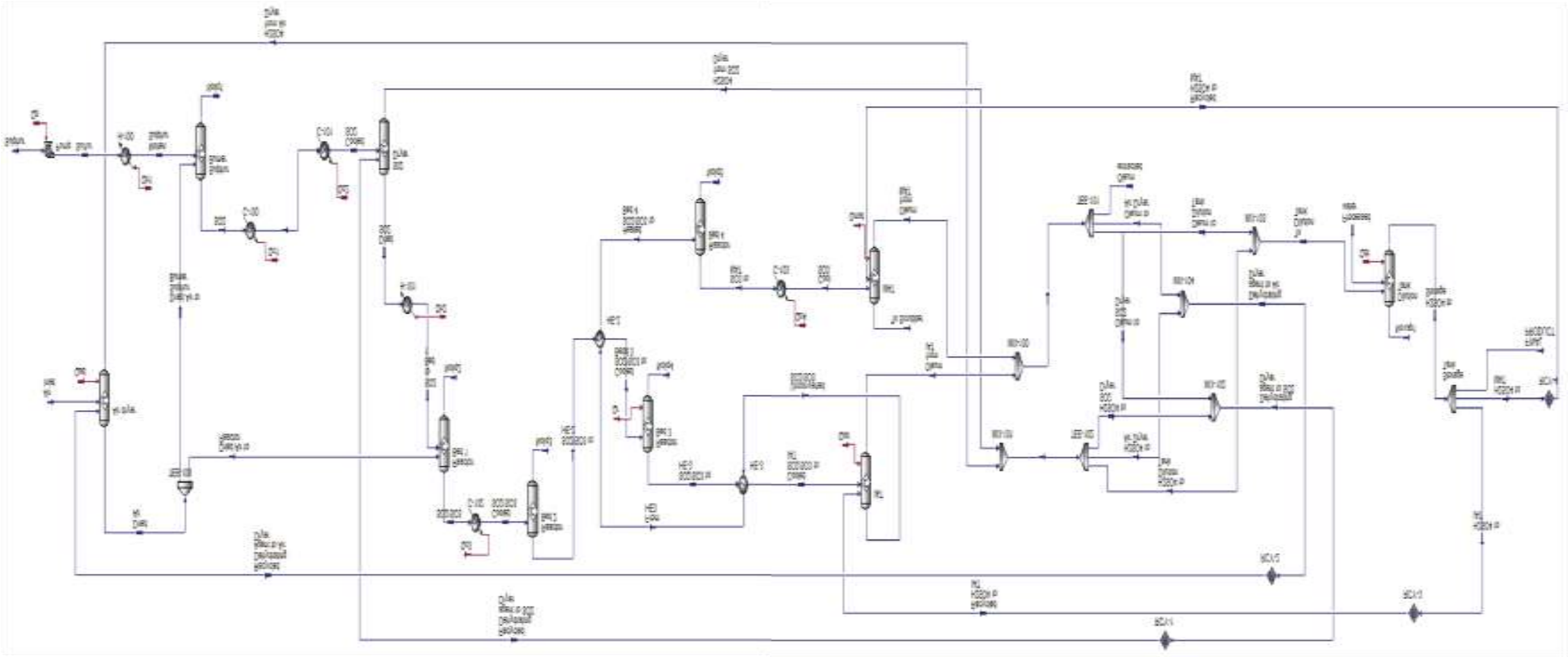


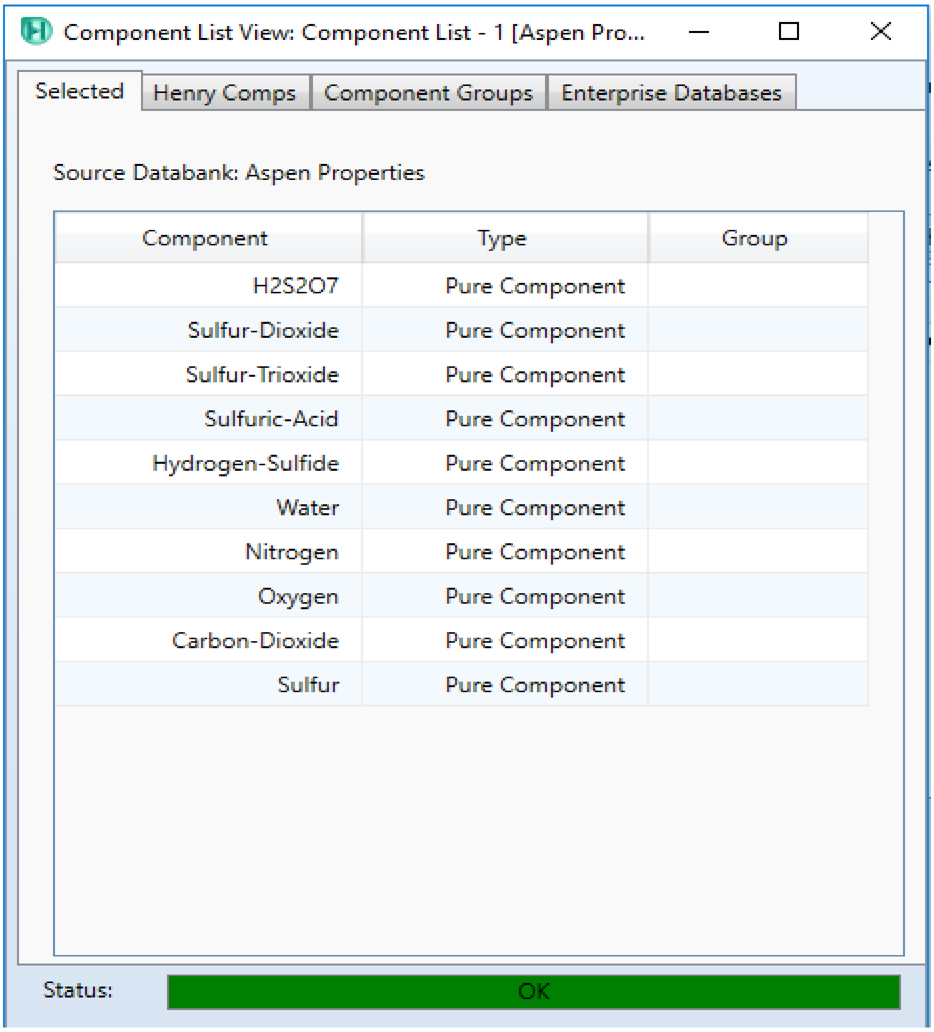
Figure 19-Simulation Model

6.2 Process Simulation

6.2.1 Components

The components were extracted from the Aspen properties as oleum was not included in HYSYS properties. (Magazine, 2010)

Following is the list of components that were used in our simulation model.



Source Databank: Aspen Properties

Component	Type	Group
H2S2O7	Pure Component	
Sulfur-Dioxide	Pure Component	
Sulfur-Trioxide	Pure Component	
Sulfuric-Acid	Pure Component	
Hydrogen-Sulfide	Pure Component	
Water	Pure Component	
Nitrogen	Pure Component	
Oxygen	Pure Component	
Carbon-Dioxide	Pure Component	
Sulfur	Pure Component	

Status: OK

Figure 20-Component list for simulation model

6.2.2 Fluid Package

Fluid Package	Component List	Property Package
Basis-1	Component List - 1 [Aspen Properties Databanks]	Aspen Properties (Electrolyte NRTL)

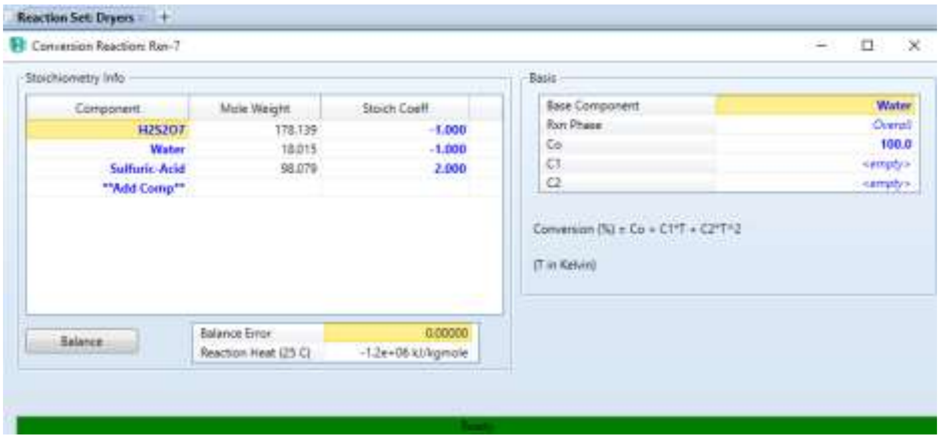
Figure 21-Fluid Package

Reactions

6.2.3.1 Air Dryer/SO_2 Dryer

Table 6.1 Reaction equation for Dryers

Specification type	**Stoichiometric equation**
Conversion	$H_2S_2O7 + H2O \rightarrow 2H2SO4$



Component	Mole Weight	Stoich Coeff
H2S2O7	178.139	-1.000
Water	18.015	-1.000
Sulfuric-Acid	98.079	2.000
Add Comp		

Base Component	Water
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

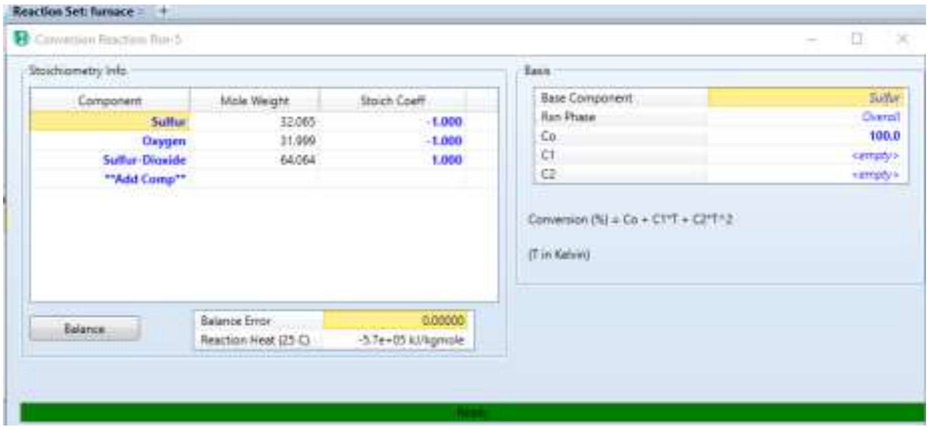
Balance Error	0.00000
Reaction Heat (25 C)	-1.2e+06 kJ/kgmole

Figure 22-Reaction Set for the Dryers

6.2.3.2 Sulfur Burner

Table 6.2 Reaction equation for Sulfur Burner

Specification type	**Stoichiometric equation**
Conversion	$S + O_2 \rightarrow SO_2$



Component	Mole Weight	Stoich Coeff
Sulfur	32.065	-1.000
Oxygen	31.999	-1.000
Sulfur-Dioxide	64.064	1.000
Add Comp		

Base Component	Sulfur
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

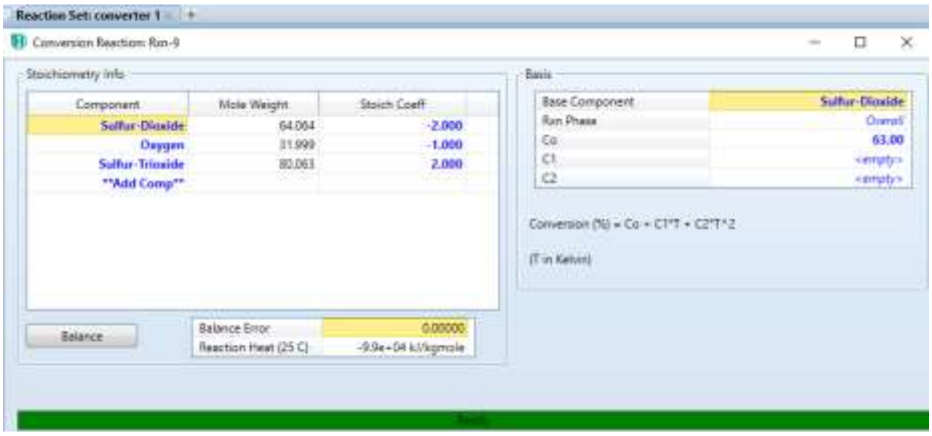
Balance Error	0.00000
Reaction Heat (25 C)	-5.7e+05 kJ/kgmole

Figure 23-Reaction Set for Sulfur Burner

6.2.3.3 Packed Bed Reactor (Bed 1)

Table 6.3 Reaction equation for Packed Bed Reactor

Specification type	**Stoichiometric equation**
Conversion	$SO_2 + 0.5O_2 \rightarrow SO_3$



Component	Mole Weight	Stoich Coeff
Sulfur-Dioxide	64.064	-2.000
Oxygen	31.999	-1.000
Sulfur-Trioxide	80.063	2.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	-9.9e+04 kJ/kgmole

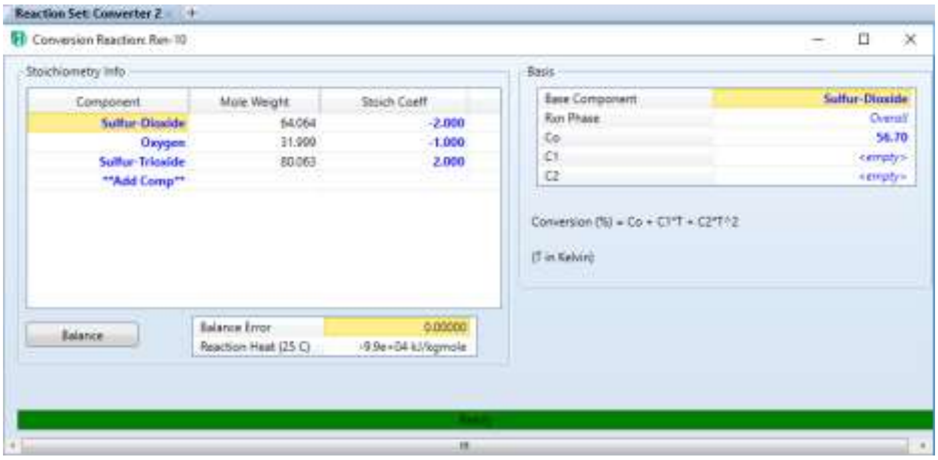
Base Component	Sulfur-Dioxide
Rxn Phase	Overall
Co	63.00
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 24-Reaction Set for Packed Bed Reactor Bed 1

6.2.3.4 Packed Bed Reactor (Bed 2)



Component	Mole Weight	Stoich Coeff
Sulfur-Dioxide	64.064	-2.000
Oxygen	31.999	-1.000
Sulfur-Trioxide	80.063	2.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	-9.9e+04 kJ/kgmole

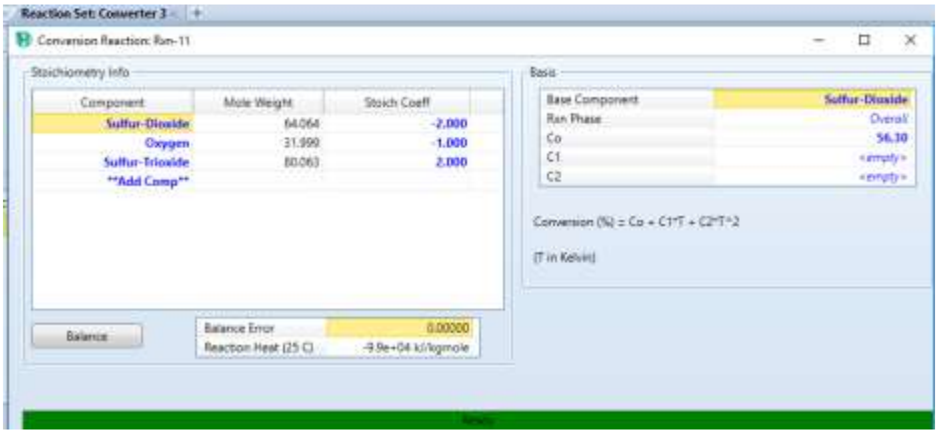
Base Component	Sulfur-Dioxide
Rxn Phase	Overall
Co	56.70
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 25-Reaction Set for Packed Bed Reactor Bed 2

6.2.3.5 Packed Bed Reactor (Bed 3)



Component	Mole Weight	Stoich Coeff
Sulfur-Dioxide	64.064	-2.000
Oxygen	31.999	-1.000
Sulfur-Trioxide	80.063	2.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	-9.9e+04 kJ/kgmole

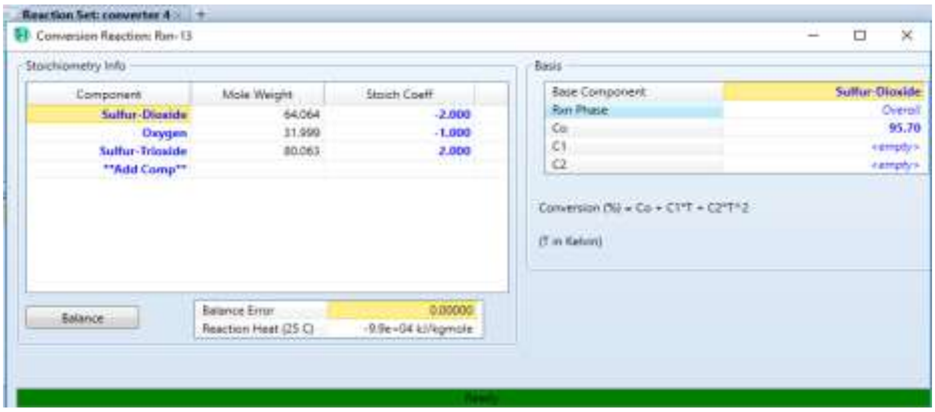
Base Component	Sulfur-Dioxide
Rxn Phase	Overall
Co	56.30
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 26-Reaction Set for Packed Bed Reactor Bed 3

6.2.3.6 Packed Bed Reactor (Bed 4)



Component	Mole Weight	Stoich Coeff
Sulfur-Dioxide	64.064	-2.000
Oxygen	31.999	-1.000
Sulfur-Trioxide	80.063	2.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	-9.9e+04 kJ/kgmole

Base Component	Sulfur-Dioxide
Rxn Phase	Overall
Co	95.70
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

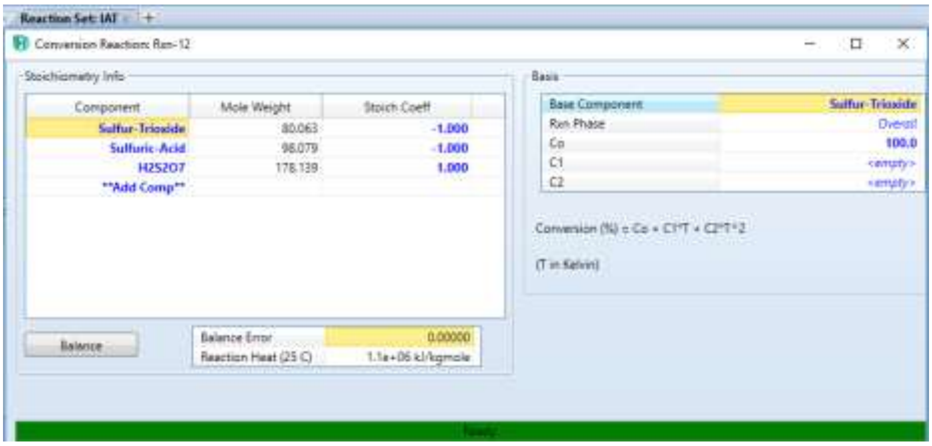
(T in Kelvin)

Figure 27-Reaction Set for Packed Bed Reactor Bed 4

6.2.3.7 IAT (Intermediate Absorption Tower)

Table 6.4 Reaction Equation for IAT

Specification type	**Stoichiometric equation**
Conversion	$SO_3 + H_2SO_4 \rightarrow H_2S_2O7$



Stoichiometry Info

Component	Mole Weight	Stoich Coeff
Sulfur-Trioxide	80.063	-1.000
Sulfuric-Acid	98.079	-1.000
H2S2O7	178.139	1.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	1.1e+06 kJ/kgmole

Basis

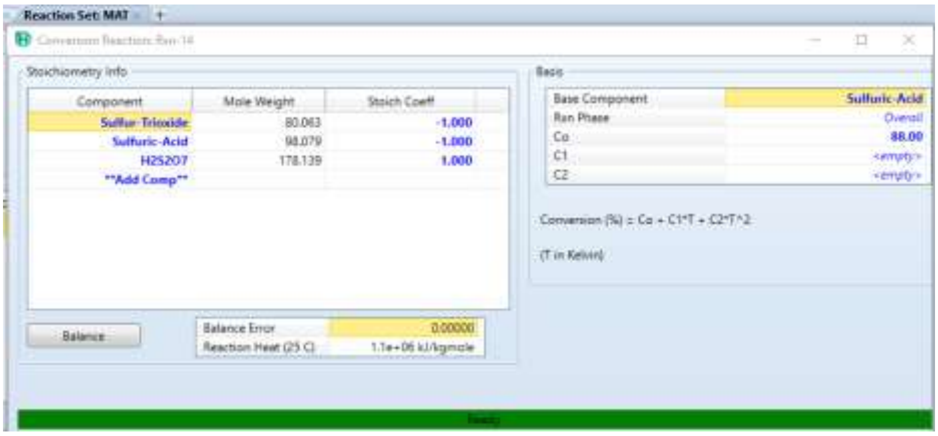
Base Component	Sulfur-Trioxide
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 28-Reaction Set for IAT

6.2.3.8 MAT (Main Absorption Tower)



Stoichiometry Info

Component	Mole Weight	Stoich Coeff
Sulfur-Trioxide	80.063	-1.000
Sulfuric-Acid	98.079	-1.000
H2S2O7	178.139	1.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	1.1e+06 kJ/kgmole

Basis

Base Component	Sulfuric-Acid
Rxn Phase	Overall
Co	88.00
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

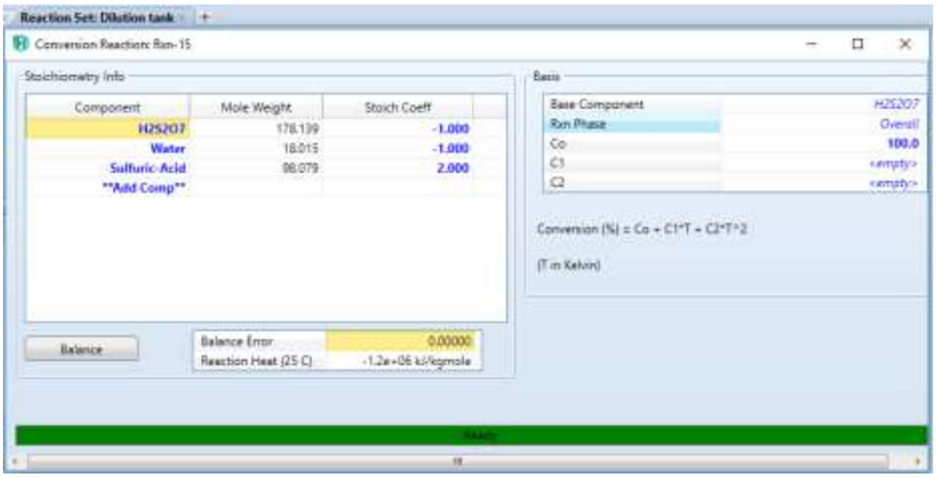
(T in Kelvin)

Figure 29-Reaction Set for MAT

6.2.3.9 Dilution Tank

Table 6.5 Reaction Equation for Dilution Tank

Specification type	**Stoichiometric equation**
Conversion	$H_2S_2O_7 + H_2O \rightarrow 2H_2SO_4$



Component	Mole Weight	Stoich Coeff
H2S2O7	178.139	-1.000
Water	18.015	-1.000
Sulfuric Acid	98.079	2.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	-1.2e+06 kJ/kgmole

Base Component	H2S2O7
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

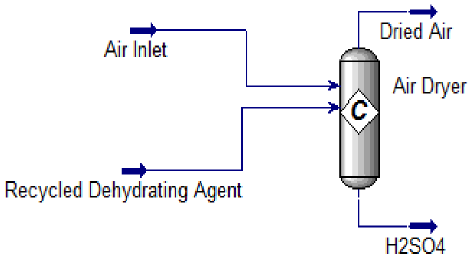
(T in Kelvin)

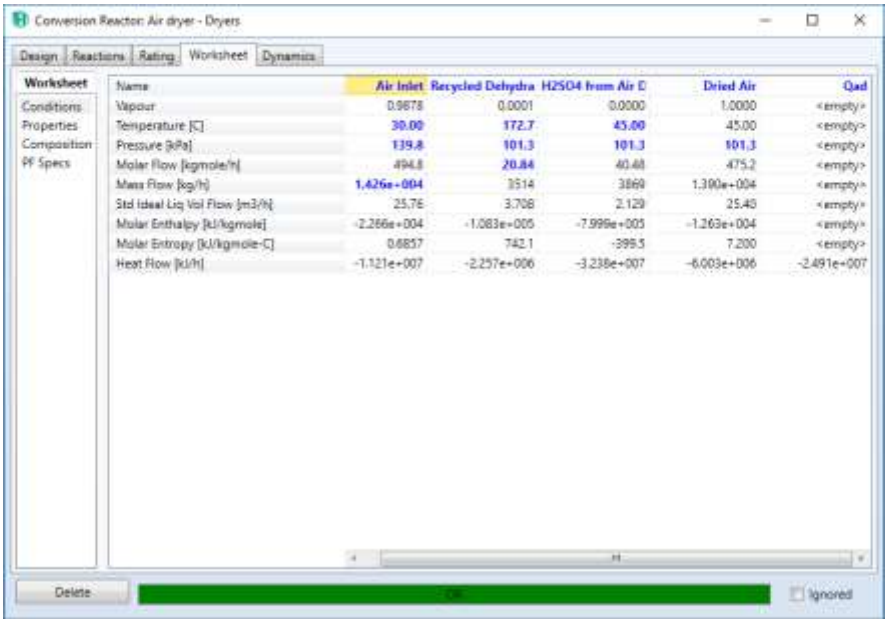
Figure 30-Reaction Set for Dilution Tank

6.2.4 Process Equipment

6.2.4.1 Air Dryer

The moisture content in the air is removed in the air dryer using oleum as a dehydrating agent. The chemical absorption reaction between water and oleum is exothermic and forms sulphuric acid as a product. The temperature of the air dryer is kept below 70°C.

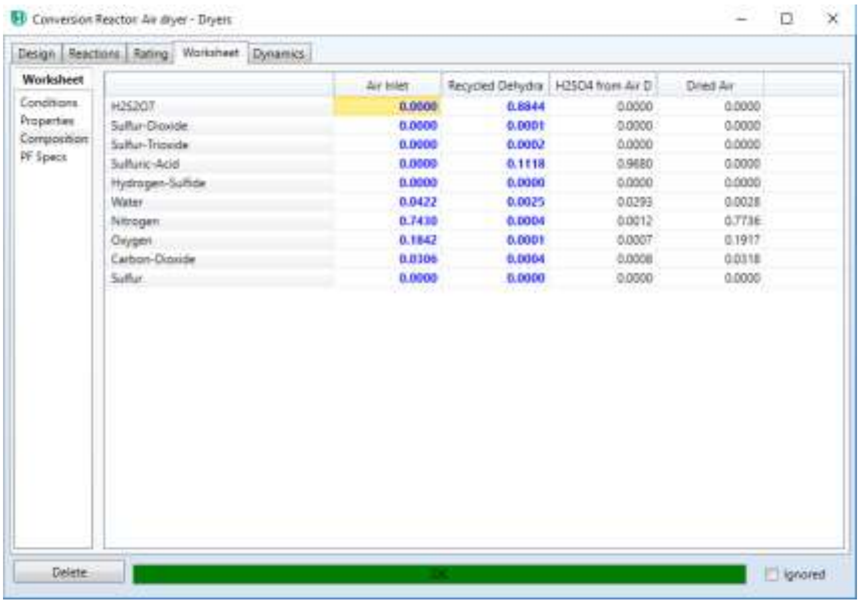




Name	Air Inlet	Recycled Dehydra	H2SO4 from Air D	Dried Air	Qad
Vapour	0.9878	0.0001	0.0000	1.0000	<empty>
Temperature [C]	30.00	172.7	45.00	45.00	<empty>
Pressure [kPa]	139.8	101.3	101.3	101.3	<empty>
Molar Flow [kgmole/h]	494.8	20.84	40.48	475.2	<empty>
Mass Flow [kg/h]	1.426e+004	3514	3869	1.390e+004	<empty>
Std Ideal Liq Vol Flow [m3/h]	25.76	3.708	2.129	25.40	<empty>
Molar Enthalpy [kJ/kgmole]	-2.266e+004	-1.083e+005	-7.999e+005	-1.263e+004	<empty>
Molar Entropy [kJ/kgmole-C]	0.6857	742.1	-399.5	7.200	<empty>
Heat Flow [kJ/h]	-1.121e+007	-2.257e+006	-3.238e+007	-6.003e+006	-2.491e+007

Figure 31-Air Dryer Conditions Worksheet

Most of the moisture content from the air is removed in the air dryer.

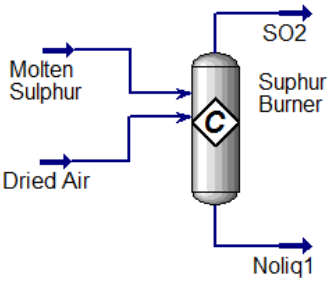


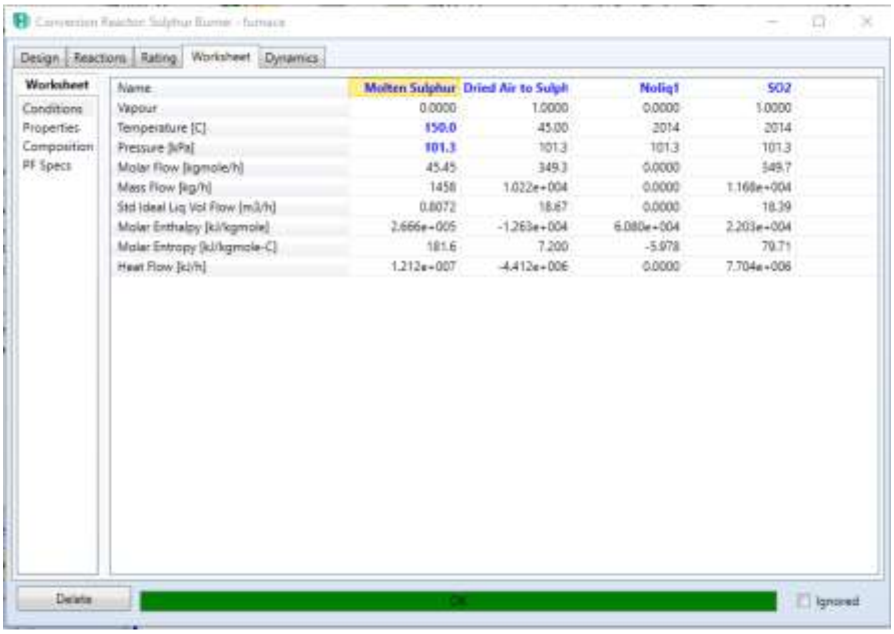
	Air Inlet	Recycled Dehydra	H2SO4 from Air D	Dried Air
H2S2O7	0.0000	0.8844	0.0000	0.0000
Sulfur-Dioxide	0.0000	0.0001	0.0000	0.0000
Sulfur-Trioxide	0.0000	0.0002	0.0000	0.0000
Sulfuric-Acid	0.0000	0.1118	0.9680	0.0000
Hydrogen-Sulfide	0.0000	0.0000	0.0000	0.0000
Water	0.0422	0.0025	0.0293	0.0028
Nitrogen	0.7430	0.0004	0.0012	0.7736
Oxygen	0.1842	0.0001	0.0007	0.1917
Carbon-Dioxide	0.0306	0.0004	0.0008	0.0318
Sulfur	0.0000	0.0000	0.0000	0.0000

Figure 32-Air Dryer Streams Compositions

6.2.4.2 Sulphur Burner

The dried air from the air dryer reacts with molten sulfur in the sulfur burner to form SO_2 gas. The reaction is highly exothermic and the temperature of the burner increases to a very high value.

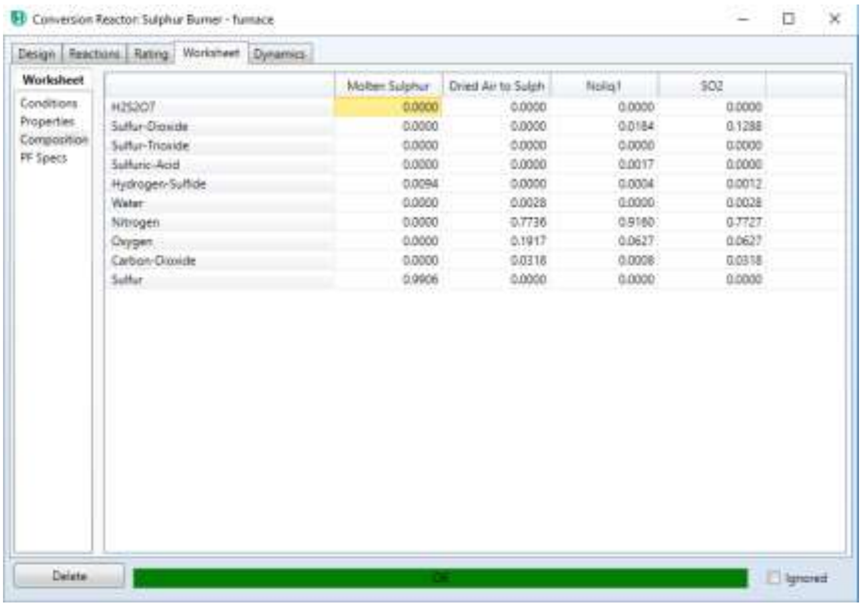




Name	Molten Sulphur	Dried Air to Sulph	Noliq1	SO2
Vapour	0.0000	1.0000	0.0000	1.0000
Temperature [C]	150.0	45.00	2014	2014
Pressure [kPa]	101.3	101.3	101.3	101.3
Molar Flow [kgmole/h]	45.45	349.3	0.0000	349.7
Mass Flow [kg/h]	1458	1.022e+004	0.0000	1.168e+004
Std Ideal Liq Vol Flow [m3/h]	0.8072	18.67	0.0000	18.39
Molar Enthalpy [kJ/kgmole]	2.666e+005	-1.263e+004	6.080e+004	2.203e+004
Molar Entropy [kJ/kgmole-C]	181.6	7.200	-5.978	79.71
Heat Flow [kJ/h]	1.212e+007	-4.412e+006	0.0000	7.704e+006

Figure 33-Sulphur burner Conditions Worksheet

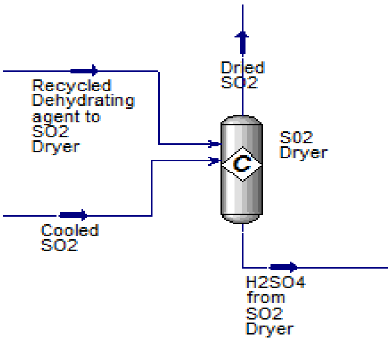
The composition of SO_2 in our product gas is 12.88 mol% which meets the general limit of SO_2 (11-15%) expected to form in this reaction.



	Molten Sulphur	Dried Air to Sulph	Noliq1	SO2
H2S2O7	0.0000	0.0000	0.0000	0.0000
Sulfur-Dioxide	0.0000	0.0000	0.0184	0.1288
Sulfur-Trioxide	0.0000	0.0000	0.0000	0.0000
Sulfuric-Acid	0.0000	0.0000	0.0017	0.0000
Hydrogen-Sulfide	0.0094	0.0000	0.0004	0.0012
Water	0.0000	0.0028	0.0000	0.0028
Nitrogen	0.0000	0.7736	0.9160	0.7727
Oxygen	0.0000	0.1917	0.0627	0.0627
Carbon-Dioxide	0.0000	0.0318	0.0008	0.0318
Sulfur	0.9906	0.0000	0.0000	0.0000

Figure 34-Sulfur Burner Streams Compositions

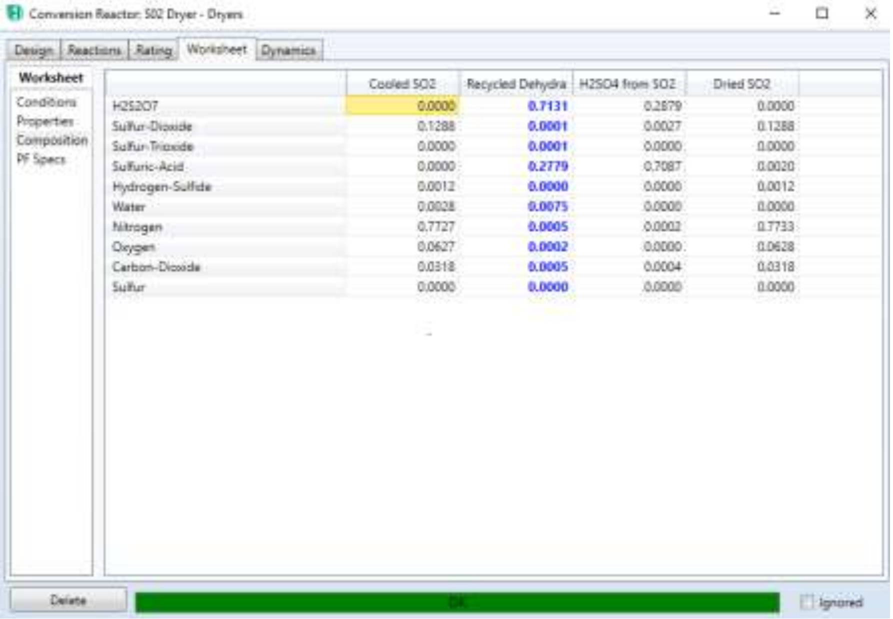
6.2.4.3 SO_2 Dryer



The SO_2 gas from the sulfur burner is cooled and sent to the SO_2 dryer to completely remove any moisture content left in the stream. This is to ensure that at high temperatures of the reactor, the water in the gas stream does not react with oxides of sulphur to form acid.

Name	Cooled SO2	Recycled Dehydra	H2SO4 from SO2	Dried SO2
Vapour	1.0000	0.0003	0.0000	1.0000
Temperature [C]	70.00	143.3	184.6	184.6
Pressure [kPa]	101.3	101.3	101.3	101.3
Molar Flow [kgmole/h]	349.7	2.534	2.812	349.4
Mass Flow [kg/h]	1.168e+004	391.8	340.3	1.173e+004
Std Ideal Liq Vol Flow [m3/h]	18.39	0.3895	0.2647	18.41
Molar Enthalpy [kJ/kgmole]	-5.006e+004	-2.422e+005	-5.706e+005	-4.726e+004
Molar Entropy [kJ/kgmole-C]	12.18	523.0	11.36	20.87
Heat Flow [kJ/h]	-1.751e+007	-6.140e+003	-1.605e+006	-1.651e+007

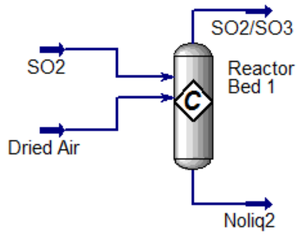
Figure 35-SO_2 Dryer Conditions Worksheet



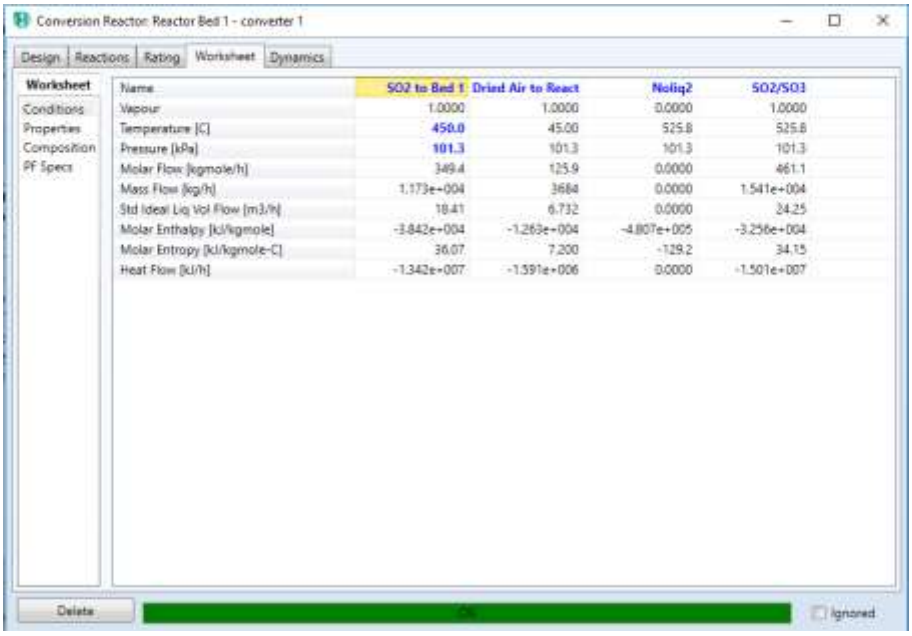
	Cooled SO2	Recycled Dehydra	H2SO4 from SO2	Dried SO2
H2S2O7	0.0000	**0.7131**	0.2879	0.0000
Sulfur-Dioxide	0.1288	**0.0001**	0.0027	0.1288
Sulfur-Trioxide	0.0000	**0.0001**	0.0000	0.0000
Sulfuric-Acid	0.0000	**0.2779**	0.7087	0.0020
Hydrogen-Sulfide	0.0012	**0.0000**	0.0000	0.0012
Water	0.0028	**0.0075**	0.0000	0.0000
Nitrogen	0.7727	**0.0005**	0.0002	0.7733
Oxygen	0.0627	**0.0002**	0.0000	0.0628
Carbon-Dioxide	0.0318	**0.0005**	0.0004	0.0318
Sulfur	0.0000	**0.0000**	0.0000	0.0000

Figure 36-SO_2 Dryer Streams Compositions

6.2.4.4 Reactor Bed 1



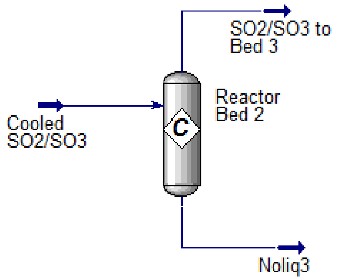
The dried SO_2 gas is heated to bring it to the reactor optimum temperature of 450°C. The V_2O_5 catalyst bed increases increase the rate of reaction. The reaction is exothermic and 63% conversion of SO_2 to SO_3 takes place.



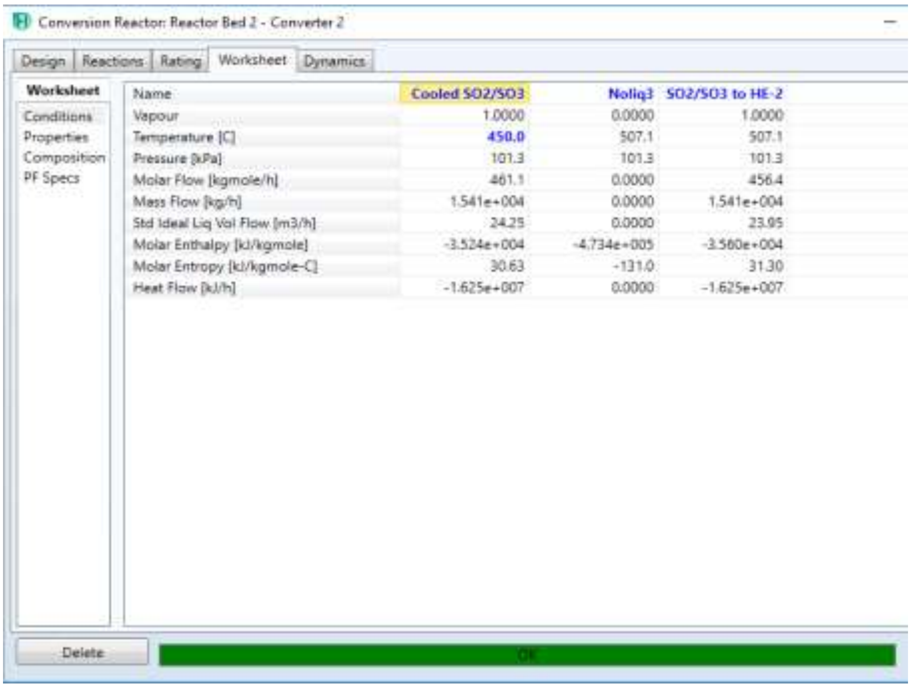
Name	**SO2 to Bed 1**	**Dried Air to React**	**Noliq2**	**SO2/SO3**
Vapour	1.0000	1.0000	0.0000	1.0000
Temperature [C]	**450.0**	45.00	525.8	525.8
Pressure [kPa]	**101.3**	101.3	101.3	101.3
Molar Flow [kgmole/h]	349.4	125.9	0.0000	461.1
Mass Flow [kg/h]	1.173e+004	3684	0.0000	1.541e+004
Std Ideal Liq Vol Flow [m3/h]	18.41	6.732	0.0000	24.25
Molar Enthalpy [kJ/kgmole]	-3.842e+004	-1.263e+004	-4.807e+005	-3.256e+004
Molar Entropy [kJ/kgmole-C]	36.07	7.200	-129.2	34.15
Heat Flow [kJ/h]	-1.342e+007	-1.591e+006	0.0000	-1.501e+007

Figure 37-Reactor Bed1 Conditions Worksheet

6.2.4.5 Reactor Bed 2



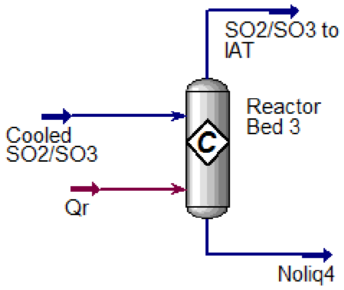
The outlet of Bed 1 is cooled and returned to Bed 2 to achieve further conversion (84%) of SO_2 to SO_3.



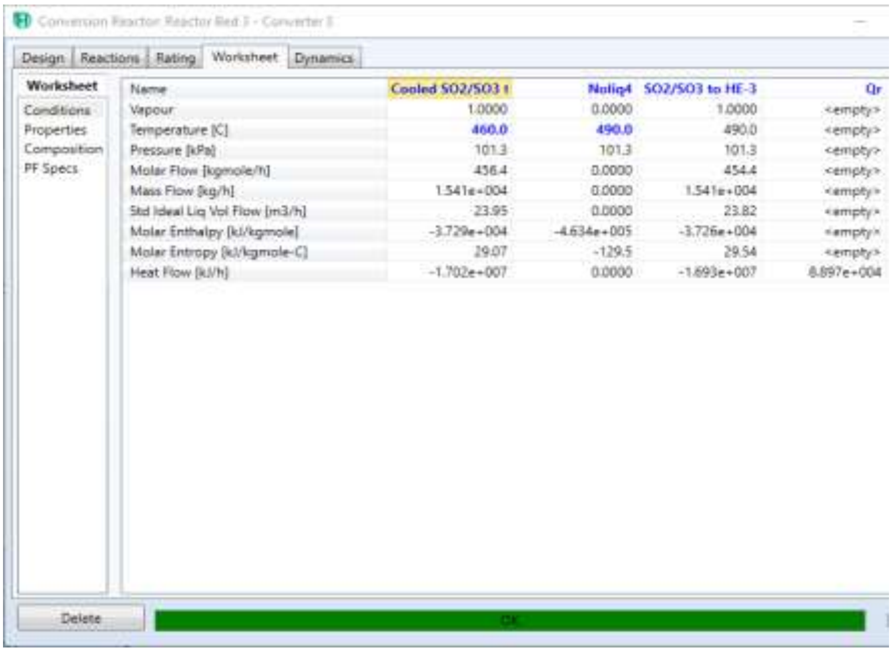
Name	Cooled SO2/SO3	Noliq3	SO2/SO3 to HE-2
Vapour	1.0000	0.0000	1.0000
Temperature [C]	450.0	507.1	507.1
Pressure [kPa]	101.3	101.3	101.3
Molar Flow [kgmole/h]	461.1	0.0000	456.4
Mass Flow [kg/h]	1.541e+004	0.0000	1.541e+004
Std Ideal Liq Vol Flow [m3/h]	24.25	0.0000	23.95
Molar Enthalpy [kJ/kgmole]	-3.524e+004	-4.734e+005	-3.560e+004
Molar Entropy [kJ/kgmole-C]	30.63	-131.0	31.30
Heat Flow [kJ/h]	-1.625e+007	0.0000	-1.625e+007

Figure 38-Reactor Bed 2 Conditions Worksheet

6.2.4.6 Reactor Bed 3



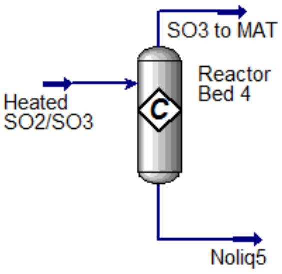
The outlet of Bed 2 is sent to the tube side of Heat Exchanger 2 to be cooled back to the optimum temperature by the unconverted gases from IAT in the shell side. The cooled gas returns to Bed 3 to achieve further conversion (93%) of SO_2 to SO_3.



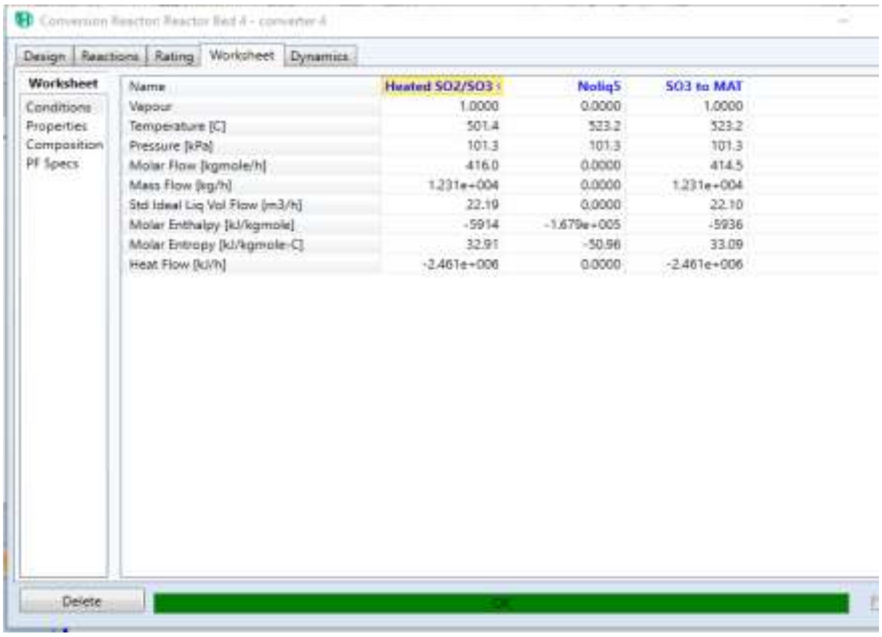
Name	Cooled SO2/SO3 t	Noliq4	SO2/SO3 to HE-3	Qr
Vapour	1.0000	0.0000	1.0000	<empty>
Temperature [C]	460.0	490.0	490.0	<empty>
Pressure [kPa]	101.3	101.3	101.3	<empty>
Molar Flow [kgmole/h]	456.4	0.0000	454.4	<empty>
Mass Flow [kg/h]	1.541e+004	0.0000	1.541e+004	<empty>
Std Ideal Liq Vol Flow [m3/h]	23.95	0.0000	23.82	<empty>
Molar Enthalpy [kJ/kgmole]	-3.729e+004	-4.634e+005	-3.726e+004	<empty>
Molar Entropy [kJ/kgmole-C]	29.07	-129.5	29.54	<empty>
Heat Flow [kJ/h]	-1.702e+007	0.0000	-1.693e+007	8.897e+004

Figure 39-Reactor Bed 3 Conditions Worksheet

6.2.4.7 Reactor Bed 4



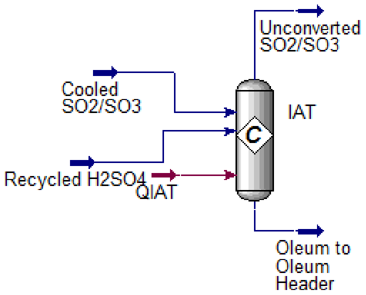
The SO_2/SO_3 gas stream from Bed 3 is sent to the Intermediate Absorption Tower (IAT). The unconverted gases from IAT return to Bed 4 after being heated by Heat Exchangers 2 and 3. In bed 4, the remaining SO_2 achieves the maximum conversion of 99.7%.



Name	Heated SO2/SO3 :	Noliq5	SO3 to MAT
Vapour	1.0000	0.0000	1.0000
Temperature [C]	501.4	523.2	523.2
Pressure [kPa]	101.3	101.3	101.3
Molar Flow [kgmole/h]	416.0	0.0000	414.5
Mass Flow [kg/h]	1.231e+004	0.0000	1.231e+004
Std Ideal Liq Vol Flow [m3/h]	22.19	0.0000	22.10
Molar Enthalpy [kJ/kgmole]	-5914	-1.679e+005	-5936
Molar Entropy [kJ/kgmole-C]	32.91	-50.96	33.09
Heat Flow [kJ/h]	-2.461e+006	0.0000	-2.461e+006

Figure 40-Reactor Bed 4 Conditions Worksheet

6.2.4.8 Intermediate Absorption Tower (IAT)

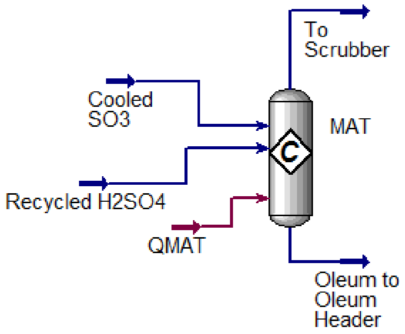


The SO_2/SO_3 gas stream from Bed 3 is cooled and enters IAT. SO_3 reacts with concentrated sulphuric acid to form oleum. The temperature of the column is not kept too high as absorption is favorable at low temperature. Most of the SO_3 is absorbed in this column.

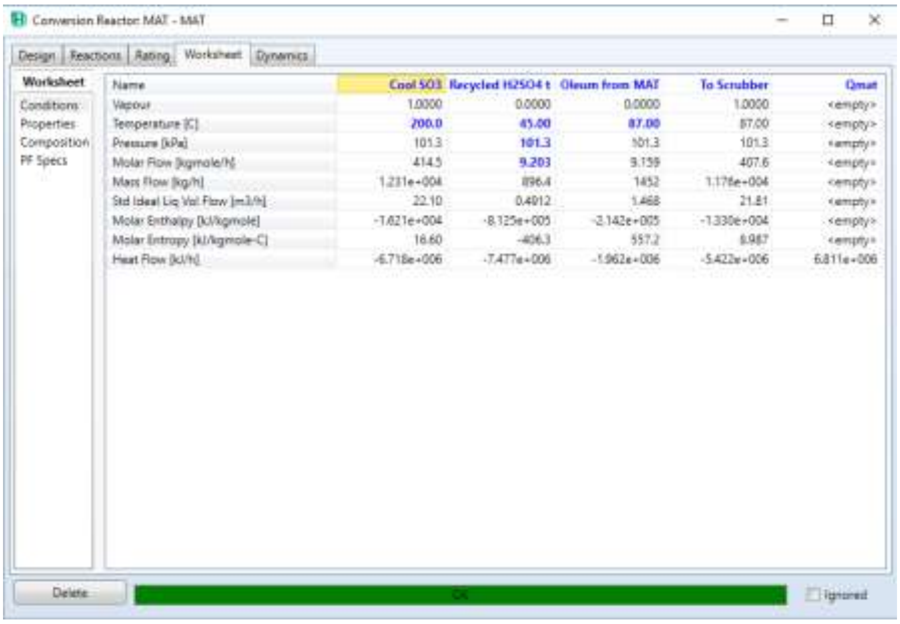
Name	Cooled SO2/SO3 1	Recycled H2SO4 1	Oleum from IAT	Unconverted SO2	Qiat
Vapour	1.0000	0.0000	0.0000	1.0000	<empty>
Temperature [C]	290.0	45.00	200.0	200.0	<empty>
Pressure [kPa]	101.3	101.3	101.3	101.3	<empty>
Molar Flow [kgmole/h]	454.4	37.55	37.97	416.0	<empty>
Mass Flow [kg/h]	1.541e+004	3658	6758	1.231e+004	<empty>
Std Ideal Liq Vol Flow [m3/h]	23.82	2.004	7.370	22.19	<empty>
Molar Enthalpy [kJ/kgmole]	-4.428e+004	-8.125e+005	-1.693e+004	-1.543e+004	<empty>
Molar Entropy [kJ/kgmole-C]	18.89	-406.3	890.7	17.40	<empty>
Heat Flow [kJ/h]	-2.012e+007	-3.051e+007	-6.426e+005	-6.419e+006	4.357e+007

Figure 41-IAT Conditions Worksheet

6.2.4.9 Main Absorber Tower (MAT)



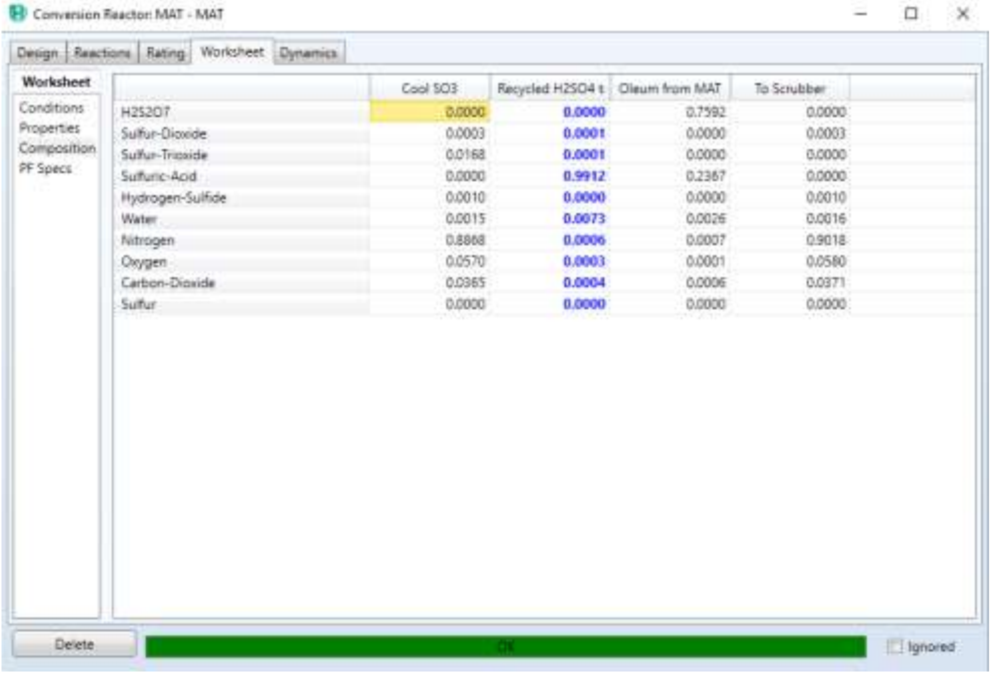
The gas stream from Bed 4 of the reactor is cooled and sent to MAT for maximum removal of SO_3 from flue gases before they are sent to the scrubber.



Name	Cool SO3	Recycled H2SO4 t	Oleum from MAT	To Scrubber	Qmat
Vapour	1.0000	0.0000	0.0000	1.0000	<empty>
Temperature [C]	200.0	45.00	87.00	87.00	<empty>
Pressure [kPa]	101.3	101.3	101.3	101.3	<empty>
Molar Flow [kgmole/h]	414.5	9.203	9.159	407.6	<empty>
Mass Flow [kg/h]	1.231e+004	896.4	1452	1.176e+004	<empty>
Std Ideal Liq Vol Flow [m3/h]	22.10	0.4912	1.468	21.81	<empty>
Molar Enthalpy [kJ/kgmole]	-1.621e+004	-8.125e+005	-2.142e+005	-1.330e+004	<empty>
Molar Entropy [kJ/kgmole-C]	16.60	-406.3	557.2	8.987	<empty>
Heat Flow [kJ/h]	-6.718e+006	-7.477e+006	-1.962e+006	-5.422e+006	6.811e+006

Figure 42-MAT Conditions Worksheet

The composition of SO_3 in MAT is decreased to zero.



	Cool SO3	Recycled H2SO4 t	Oleum from MAT	To Scrubber
H2S2O7	0.0000	0.0000	0.7592	0.0000
Sulfur-Dioxide	0.0003	0.0001	0.0000	0.0003
Sulfur-Trioxide	0.0168	0.0001	0.0000	0.0000
Sulfuric-Acid	0.0000	0.9912	0.2367	0.0000
Hydrogen-Sulfide	0.0010	0.0000	0.0000	0.0010
Water	0.0015	0.0073	0.0026	0.0016
Nitrogen	0.8868	0.0006	0.0007	0.9018
Oxygen	0.0570	0.0003	0.0001	0.0580
Carbon-Dioxide	0.0365	0.0004	0.0006	0.0371
Sulfur	0.0000	0.0000	0.0000	0.0000

Figure 43-MAT Streams Compositions

6.2.4.10 Oleum Header

The oleum produced in IAT and MAT is sent to oleum header where it is separated into fractions for use in different equipment across the plant. The ratios of oleum and sulphuric acid were adjusted according to the need of all equipment. For example, in the dryers, oleum is used as a dehydrating agent and hence the ratio of sulphuric acid in that stream can be minimized. On the other hand, the absorption columns require H_2SO_4 for the reaction. Hence oleum does not need to be sent as part of that stream.

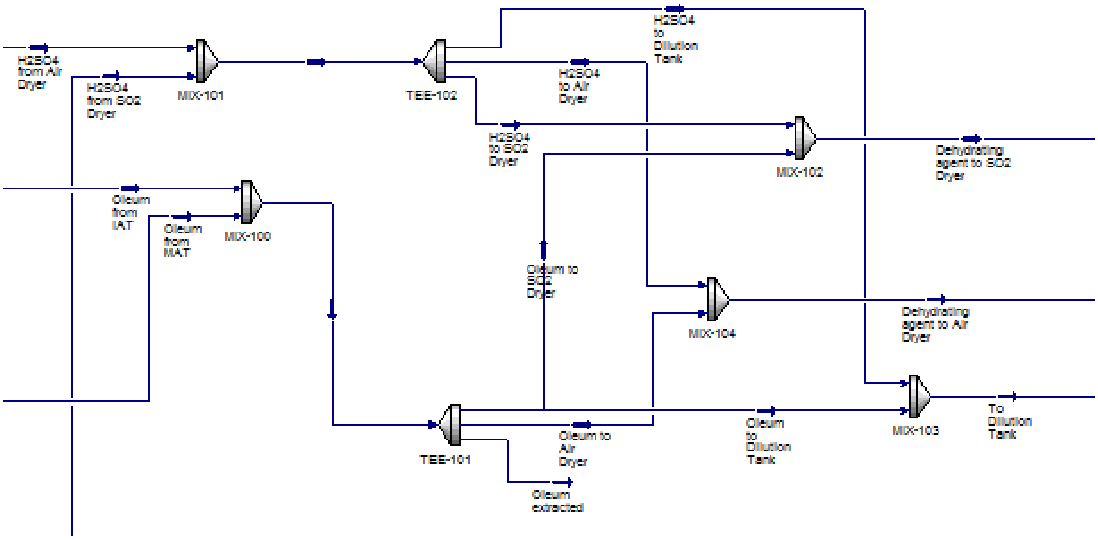
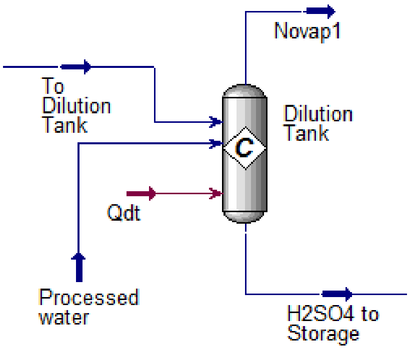
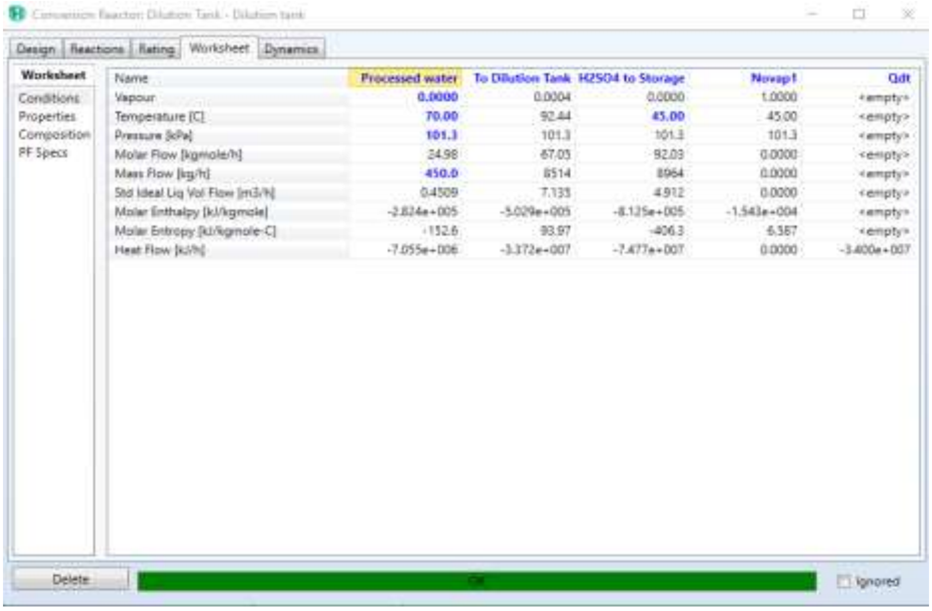


Figure 44-Oleum Header Streams Splits and Flows

6.2.4.11 Dilution Tank

After setting the split ratios for oleum in different equipment, the major part of oleum is sent to the dilution tank for reaction with water to produce sulphuric acid. Since this reaction is highly exothermic, there is a risk of acid mist formation. To prevent this, coolants are circulated in the jackets surrounding the reaction tank (CSTR).

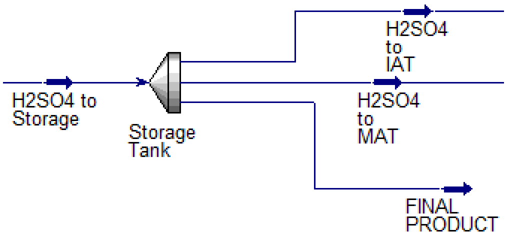




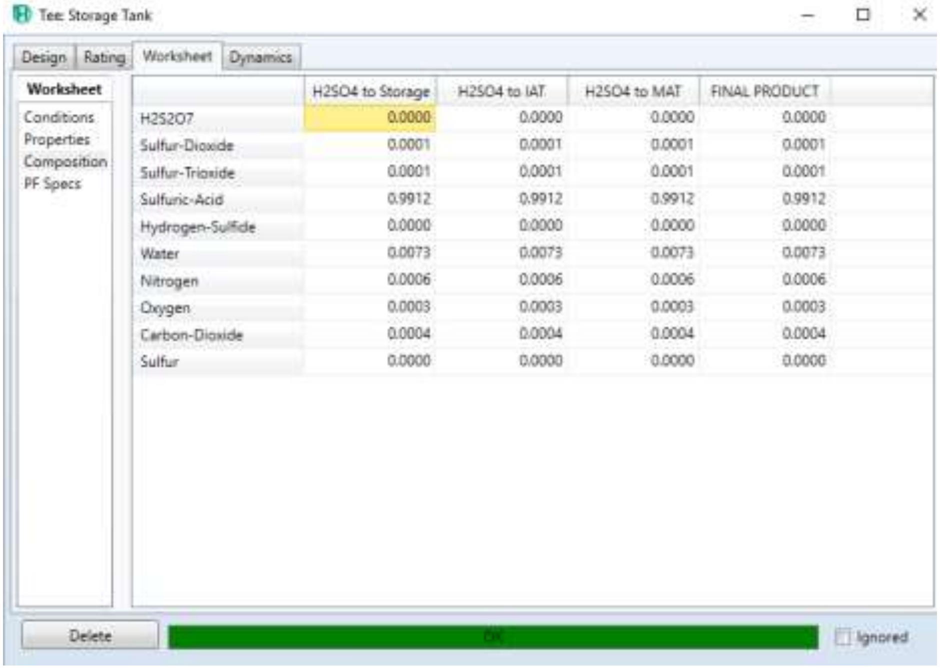
Name	Processed water	To Dilution Tank	H2SO4 to Storage	Novap1	Qdt
Vapour	0.0000	0.0004	0.0000	1.0000	<empty>
Temperature [C]	70.00	92.44	45.00	45.00	<empty>
Pressure [kPa]	101.3	101.3	101.3	101.3	<empty>
Molar Flow [kgmole/h]	24.98	67.05	92.03	0.0000	<empty>
Mass Flow [kg/h]	450.0	8514	8964	0.0000	<empty>
Std Ideal Liq Vol Flow [m3/h]	0.4509	7.133	4.912	0.0000	<empty>
Molar Enthalpy [kJ/kgmole]	-2.824e+005	-5.029e+005	-8.125e+005	-1.543e+004	<empty>
Molar Entropy [kJ/kgmole-C]	-152.6	93.97	-406.3	6.387	<empty>
Heat Flow [kJ/h]	-7.055e+006	-3.372e+007	-7.477e+007	0.0000	-3.400e+007

Figure 45-Dilution Tank Conditions Worksheet

6.2.4.12 Storage Tank



H_2SO_4 is split here for use in absorbers and for extraction as final product.



	H2SO4 to Storage	H2SO4 to IAT	H2SO4 to MAT	FINAL PRODUCT
H2S2O7	0.0000	0.0000	0.0000	0.0000
Sulfur-Dioxide	0.0001	0.0001	0.0001	0.0001
Sulfur-Trioxide	0.0001	0.0001	0.0001	0.0001
Sulfuric-Acid	0.9912	0.9912	0.9912	0.9912
Hydrogen-Sulfide	0.0000	0.0000	0.0000	0.0000
Water	0.0073	0.0073	0.0073	0.0073
Nitrogen	0.0006	0.0006	0.0006	0.0006
Oxygen	0.0003	0.0003	0.0003	0.0003
Carbon-Dioxide	0.0004	0.0004	0.0004	0.0004
Sulfur	0.0000	0.0000	0.0000	0.0000

Figure 46-Storage Tank Streams Compositions

CHAPTER 7

Costing and Economic Analysis

Cost estimation and economic analysis are the most important yet complicated aspects of setting up a plant. Industrial plants, like the one being discussed in our thesis are designed to produce products on a very large scale. In order to maximize the profits generated, the plant investment needs to be minimized. Hence, costing is of utmost importance especially because it is a crucial part of the feasibility analysis, which helps in the selection of the manufacturing process to be used.

The estimation of design cost is the fiscal description of the different costs related to the plant, based on total investment. Working capital and fixed capital make up the total investment. Fixed capital is the overall cost required for erection of a plant that is functional and ready to be set up. This cost is paid for the installation of equipment and will eventually wind up as the salvage value. Whereas, working capital is the additional funds that would be needed to ensure that the plant is up and running. All these values, along with the detailed economic analysis are shown in the tables given below.

7.1 Cost Estimates for the Design

Costing was done with references from Coulson Richardson's Chemical Engineering Design Volume 6. Following are the graphs and figures used in cost estimation. (Vatavuk, 2002)

7.2 Calculation of PCE (Personal Consumption Expenditure)

Table 7.1 Basic data for costing

Equipment Cost Index	Value
2004	444.2
2019	756.9
3.5% decrease in 2020	0.035

2020	730.4
Currency	**Conversion**
Conversion (pound to dollar)	1£ = $1.26

The equation used to calculate the purchase cost of some equipment is:

$$\boldsymbol{Ce = CS^{n}}$$

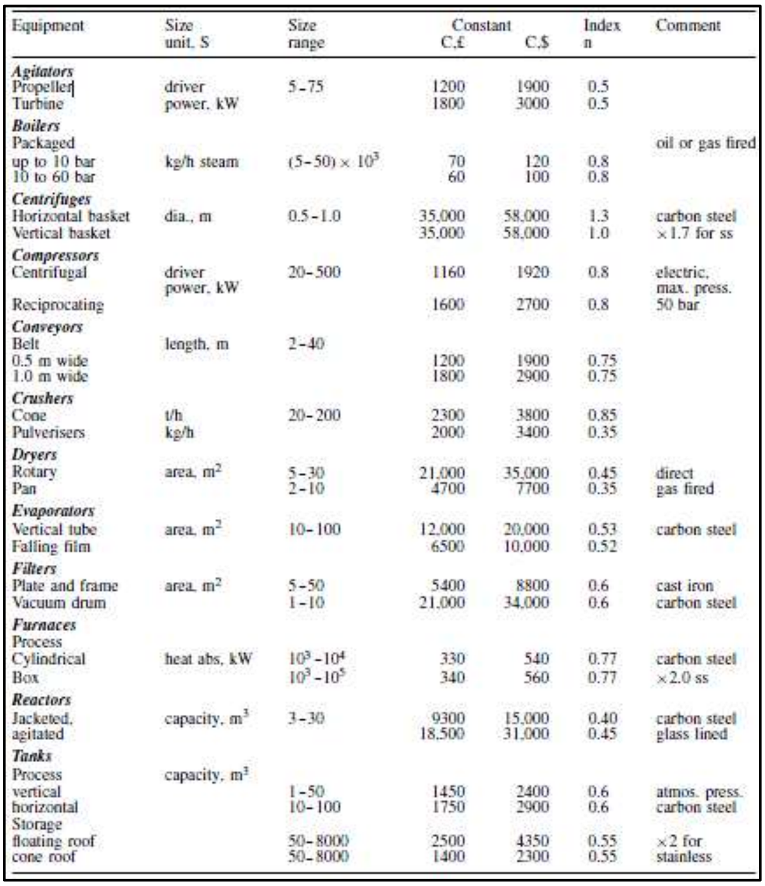
Where Ce is the purchased equipment cost in £,

C is a constant in £,

S is size unit, and

n is the index.

Table 6.2 refers to the equipment for which the purchase cost is calculated using this equation.



Equipment	Size unit, S	Size range	Constant		Index n	Comment
			C,£	C,$		
Agitators						
Propeller	driver power, kW	5–75	1200	1900	0.5	
Turbine			1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5–50) \times 10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	×1.7 for ss
Compressors						
Centrifugal	driver power, kW	20–500	1160	1920	0.8	electric, max. press. 50 bar
Reciprocating			1600	2700	0.8	
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct
Pan		2–10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	$10^3–10^4$	330	540	0.77	carbon steel
Box		$10^3–10^5$	340	560	0.77	×2.0 ss
Reactors						
Jacketed,	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
agitated			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m^3					
vertical		1–50	1450	2400	0.6	atmos. press.
horizontal		10–100	1750	2900	0.6	carbon steel
Storage						
floating roof		50–8000	2500	4350	0.55	×2 for
cone roof		50–8000	1400	2300	0.55	stainless

Figure 47-Data for purchase cost equation

7.2.1 Blower

Table 7.2 Calculation of Blower Purchase Cost

Size Factor (S) / kW	156
Base Cost (C) / £ - 2004	1160
Index Number (n)	0.80
Cost (£)	65742
Cost ($)	82834

7.2.2 Pump

Table 7.3 Estimate of Pump Purchase Cost

Cost ($)	1000

7.2.3 Furnace (Cylindrical)

Table 7.4 Calculation of Furnace Purchase Cost

Size Factor/S (KW)	7600
Base Cost/C ($)	330
Index Number (n)	0.77
Cost (£)	321174
Cost = ($)	404679

7.2.4 Reactor

The cost of catalyst per kg is taken from literature. The total weight of catalyst in the reactor is calculated in our plant design. The product of cost per kg and total weight gives the purchase cost of the catalyst.

$$Total\ reactor\ cost = Cost\ of\ catalyst + Cost\ of\ reactor\ body$$

Table 7.5 Calculation of Reactor Purchase Cost

Size Factor/S (m^3)	13.5
Base Cost/C (£)	9300
Index Number (n)	0.40
Cost (£)	26340
Cost = ($)	33188
Catalyst cost/($/kg)	22.5
Weight of catalyst (kg)	9176
Cost of catalyst	206460
Total reactor cost ($)	238699

7.2.5 Oleum Header (Vertical tank)

Table 7.6 Calculation of Oleum Header Purchase Cost

Size Factor/S (m^3)	7.00
Base Cost/C (£)	1450
Index Number (n)	0.60
Cost (£)	4660
Cost ($)	5872

7.2.6 Dilution tank (Jacketed)

Table 7.7 Calculation of Dilution Tank Purchase Cost

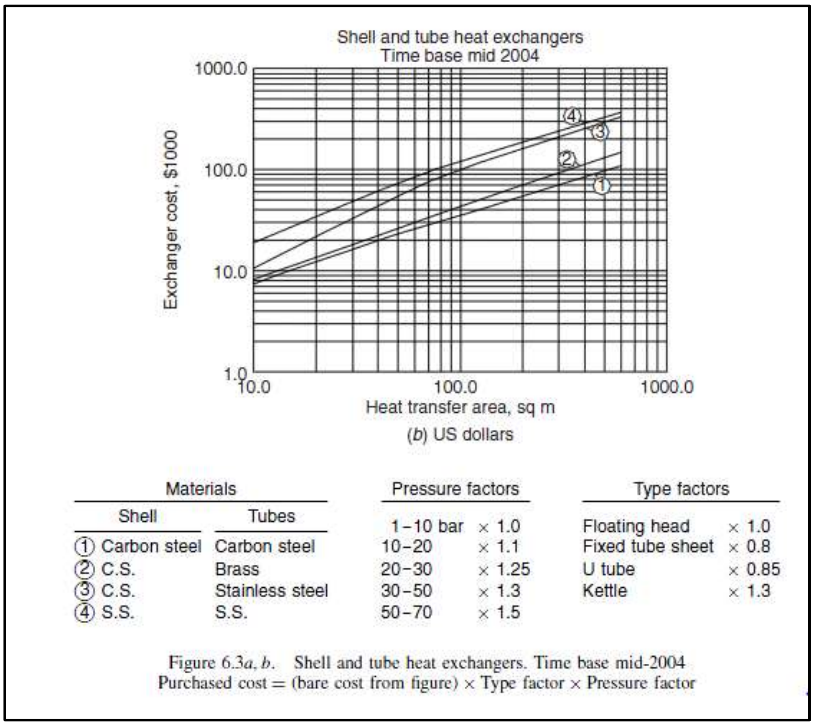
Size Factor/S (m^3)	34.8
Base Cost/C (£)	9300
Index Number (n)	0.40
Cost (£)	38469
Cost ($)	48471

7.2.7 H_2SO_4 tank (Vertical tank)

Table 7.8 Calculation of Storage Tank Purchase Cost

Size Factor/S (m^3)	5.28
Base Cost/C (£)	1450
Index Number (n)	0.60
Cost (£)	3935
Cost ($)	4958

For the calculation of purchase cost of heating and cooling equipment, we use Fig. 6.3 from Richardson Coulson Volume 6.



Materials	
Shell	Tubes
① Carbon steel	Carbon steel
② C.S.	Brass
③ C.S.	Stainless steel
④ S.S.	S.S.

Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure 6.3*a, b*. Shell and tube heat exchangers. Time base mid-2004
Purchased cost = (bare cost from figure) × Type factor × Pressure factor

Figure 48-Purchase Cost for Shell and Tube Heat Exchangers

The heating surface area is taken from our equipment design. The cost of melter is calculated using Fig.6.3 b. The equation used to calculate total cost is:

$$Total\ cost = Bare\ cost * Material\ factor * Pressure\ factor$$

7.2.8 Sulfur Melter

Table 7.9 Calculation of Sulfur Melter Purchase Cost

Heat Transfer Area	90.0
Material Factor (CS)	1.00
Pressure Factor	1.00
Type Factor (Fixed Tube Sheet)	0.80
Bare Cost ($)	32000
Total Cost ($)	25600

7.2.9 Coolers

Table 7.10 Calculation of Cooler Purchase Cost

Heat Transfer Area	75.0
Material Factor (CS)	1.00
Pressure Factor	1.00
Type Factor (Fixed Tube Sheet)	0.80
Bare Cost ($)	30000
Total Cost ($)	24000
Total Cost of both coolers ($)	48000

7.2.10 Reactor pre-heater

Table 7.11 Calculation of Heater Purchase Cost

Heat Transfer Area	60.0
Material Factor (CS)	1.00
Pressure Factor	1.00
Type Factor (Fixed Tube Sheet)	0.80
Bare Cost ($)	27000
Total Cost ($)	21600

7.2.11 Heat Exchanger 2

Table 7.12 Calculation of Heat Exchanger Purchase Cost

Heat Transfer Area	200
Material Factor (CS)	1.00
Pressure Factor	1.00

Type Factor (Fixed Tube Sheet)	0.80
Bare Cost ($)	50000
Total Cost ($)	40000

7.2.12 Heat Exchanger 3

Table 7.13 Calculation of Heat Exchanger Purchase Cost

Heat Transfer Area	150
Material Factor (CS)	1.00
Pressure Factor	1.00
Type Factor (Floating head)	1.00
Bare Cost ($)	42000
Total Cost ($)	42000

7.2.13 Intermediate cooler between 1st and 2nd bed

Table 7.14 Calculation of Cooler Purchase Cost

Heat Transfer Area	47.0
Material Factor (CS)	1.00
Pressure Factor	1.00
Type Factor (Fixed tube)	0.80
Bare Cost ($)	22000
Total Cost ($)	17600

7.2.14 Intermediate cooler between 4th bed and MAT

Table 7.15 Calculation of Cooler Purchase Cost

Heat Transfer Area	60.0
Material Factor (CS)	1.00
Pressure Factor	1.00
Type Factor (Fixed tube)	0.80
Bare Cost ($)	27000
Total Cost ($)	21600

The purchase cost of absorbers is calculated using various figures and design calculations.

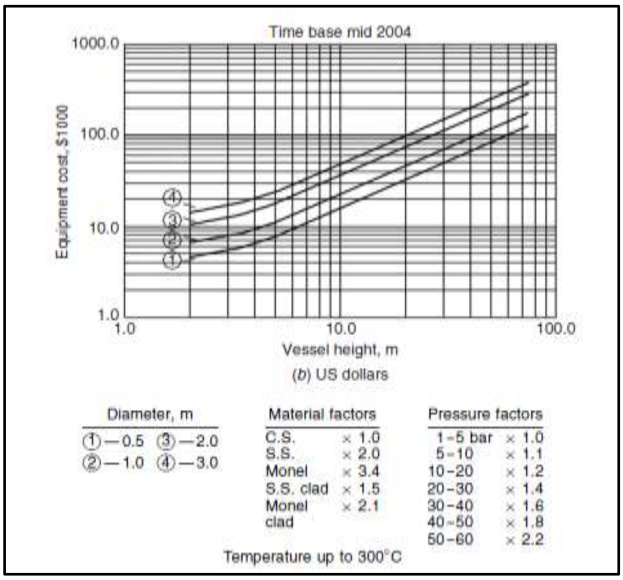


Figure 49-Purchase cost of Vertical Pressure Vessels

The cost of packing is calculated using Table 6.3.

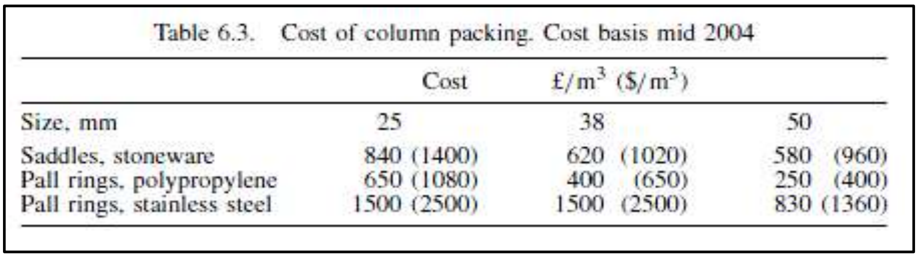


Table 6.3. Cost of column packing. Cost basis mid 2004

	Cost	£/m³ ($/m³)	
Size, mm	25	38	50
Saddles, stoneware	840 (1400)	620 (1020)	580 (960)
Pall rings, polypropylene	650 (1080)	400 (650)	250 (400)
Pall rings, stainless steel	1500 (2500)	1500 (2500)	830 (1360)

Figure 50-Purchase cost of column packing

The dimensions have been calculated in the design of absorbers. Using Figure 6.5b and the dimensions of the column, we calculate the bare vessel cost. Vessel cost is calculated using the formula:

$$Vessel\ cost = Bare\ Vessel\ cost * Material\ factor * Pressure\ factor$$

For the cost of packing, volume of packing and Table 6.3 are used. The volume of packing is taken from our absorption column design.

Total cost of column is calculated using the formula:

$$Total\ cost\ of\ column = Vessel\ cost + Cost\ of\ packing$$

7.2.15 Air Dryer

Table 7.16 Calculation of Air Dryer Purchase Cost

Diameter (m)	0.97
Vessel Height (m)	5.18
Bare Vessel Cost (BVC) / $	11000
Material Factor - SS / MF	2.00
Pressure Factor / PF	1.00
Vessel Cost ($)	22000
Packing cost ($/m^3)	960
Volume of packing (m^3)	3.10
Total packing cost ($)	2976
Total column cost ($)	24976

7.2.16 SO_2 Dryer

Table 7.17 Calculation of SO_2 Dryer Purchase Cost

Diameter (m)	1.07

Vessel Height (m)	3.13
Bare Vessel Cost ($) / BVC	9000
Material Factor - SS / MF	2.00
Pressure Factor / PF	1.00
Vessel Cost ($)	18000
Packing cost ($/$m^3$)	650
Volume of packing (m^3)	2.81
Total packing cost ($)	1827
Total column cost($)	19827

7.2.17 IAT

Table 7.18 Calculation of IAT Purchase Cost

Diameter (m)	1.18
Vessel Height (m)	7.84
Bare Vessel Cost ($) / BVC	19000
Material Factor - SS / MF	2.00
Pressure Factor / PF	1.00
Vessel Cost ($)	38000
Packing cost ($/$m^3$)	1020
Volume of packing (m^3)	6.80
Total packing cost ($)	6936
Total column cost ($)	44936

7.2.18 MAT

Table 7.19 Calculation of Oleum Header Purchase Cost

Diameter (m)	0.84

Vessel Height (m)	5.31
Bare Vessel Cost ($) / BVC	6500
Material Factor - SS / MF	2.00
Pressure Factor / PF	1.00
Vessel Cost ($)	13000
Packing cost ($/m³)	1400
Volume of packing (m³)	2.33
Total packing cost ($)	3262
Total column cost ($)	16262

7.3 Calculation of Physical Plant Cost (PPC)

PCE (2004) = $1091822

PCE (2020) = $1795308

The factors needed to calculate PPC are found in Table 6.1. (BUITEMS, 2019) (BUITEMS, 2019)

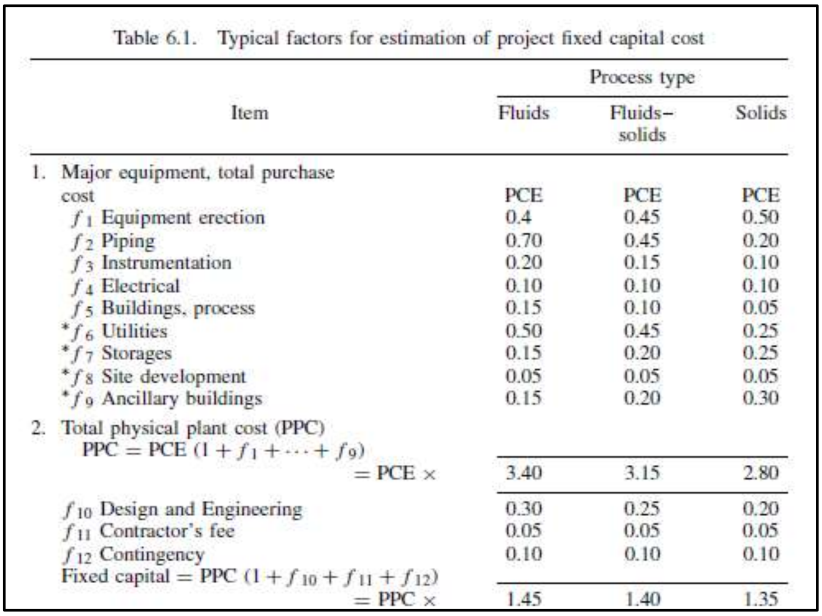


Table 6.1. Typical factors for estimation of project fixed capital cost

Item	Process type		
	Fluids	Fluids–solids	Solids
1. Major equipment, total purchase cost	PCE	PCE	PCE
f_1 Equipment erection	0.4	0.45	0.50
f_2 Piping	0.70	0.45	0.20
f_3 Instrumentation	0.20	0.15	0.10
f_4 Electrical	0.10	0.10	0.10
f_5 Buildings, process	0.15	0.10	0.05
*f_6 Utilities	0.50	0.45	0.25
*f_7 Storages	0.15	0.20	0.25
*f_8 Site development	0.05	0.05	0.05
*f_9 Ancillary buildings	0.15	0.20	0.30
2. Total physical plant cost (PPC) PPC = PCE $(1 + f_1 + \cdots + f_9)$ = PCE ×	3.40	3.15	2.80
f_{10} Design and Engineering	0.30	0.25	0.20
f_{11} Contractor's fee	0.05	0.05	0.05
f_{12} Contingency	0.10	0.10	0.10
Fixed capital = PPC $(1 + f_{10} + f_{11} + f_{12})$ = PPC ×	1.45	1.40	1.35

Figure 51-Data for calculation of PPC

PPC is calculated using the formula:

$$PPC = PCE(1 + f1 + \cdots + f6)$$

Fluids have been selected as the process type because no solids are involved in our plant operation.

Table 7.19 Calculation of PPC

Item	**PCE**
Equipment erection (f1)	0.4
Piping (f2)	0.7
Instrumentation (f3)	0.2
Electrical (f4)	0.1
Buildings/ Process (f5)	0.15
Utilities (f6)	0.5
PPC ($)	5475691

7.4 The calculation of Fixed Capital is done using the equation:

$$FC = PPC(1 + f10 + f11 + f12)$$

Table 7.20 Calculation of Fixed Capital

Item	**PCE**
Design and Engineering (f10)	0.3
Contractors fee (f11)	0.05
Contingency (f12)	0.1
Fixed capital ($)	7939752

7.5 Working Capital is estimated to be 10-20% of the fixed capital. We have assumed an average value of 15% to calculate the working capital.

$$Working\ Capital = 0.15 * Fixed\ Capital$$

Table 7.21 Calculation of Working Capital

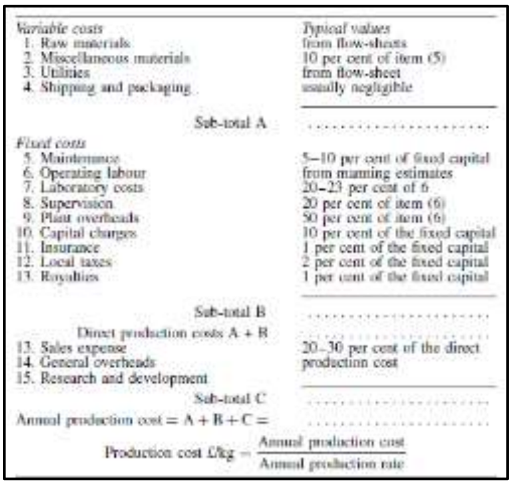
Working Capital ($)	1190962

$$Total\ investment = Fixed\ Capital + Working\ Capital$$

Table 7.22 Calculation of Total Investment

Total Investment ($)	9130715

7.6 During plant operation, the operating costs are calculated using Table 6.6 of Coulson Richardson Chemical Engineering Design Volume 6.

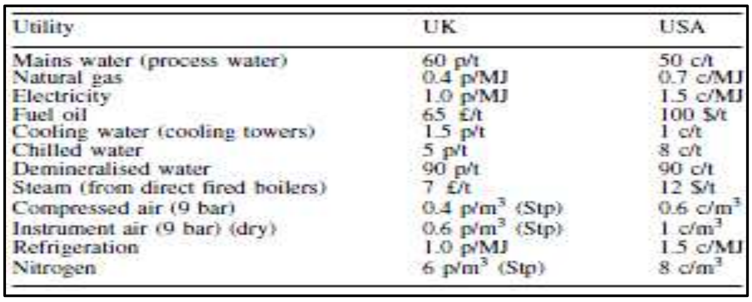


Variable costs	*Typical values*
1. Raw materials	from flow-sheets
2. Miscellaneous materials	10 per cent of item (5)
3. Utilities	from flow-sheet
4. Shipping and packaging	usually negligible
Sub-total A	....
Fixed costs	
5. Maintenance	5–10 per cent of fixed capital
6. Operating labour	from manning estimates
7. Laboratory costs	20–23 per cent of 6
8. Supervision	20 per cent of item (6)
9. Plant overheads	50 per cent of item (6)
10. Capital charges	10 per cent of the fixed capital
11. Insurance	1 per cent of the fixed capital
12. Local taxes	2 per cent of the fixed capital
13. Royalties	1 per cent of the fixed capital
Sub-total B	....
Direct production costs A + B	....
13. Sales expense	20–30 per cent of the direct production cost
14. General overheads	
15. Research and development	
Sub-total C	....
Annual production cost = A + B + C =	....

$$\text{Production cost £/kg} = \frac{\text{Annual production cost}}{\text{Annual production rate}}$$

Figure 52-Data for calculation of Operating Cost

The cost of utilities is calculated using Table 6.5 of Coulson Richardson Volume 6.



Utility	UK	USA
Mains water (process water)	60 p/t	50 c/t
Natural gas	0.4 p/MJ	0.7 c/MJ
Electricity	1.0 p/MJ	1.5 c/MJ
Fuel oil	65 £/t	100 $/t
Cooling water (cooling towers)	1.5 p/t	1 c/t
Chilled water	5 p/t	8 c/t
Demineralised water	90 p/t	90 c/t
Steam (from direct fired boilers)	7 £/t	12 $/t
Compressed air (9 bar)	0.4 p/m^3 (Stp)	0.6 c/m^3
Instrument air (9 bar) (dry)	0.6 p/m^3 (Stp)	1 c/m^3
Refrigeration	1.0 p/MJ	1.5 c/MJ
Nitrogen	6 p/m^3 (Stp)	8 c/m^3

Figure 53-Data for calculation of Utility cost

The fixed operating cost is the sum of all the items ranging from 5 to 12.

$$Total\ base\ salary = Annual\ Base\ Salary * Number\ of\ operators$$

$$Allowances = 0.5 * Total\ Base\ Salary$$

$$Total\ operating\ labor = Total\ base\ salary + Allowances$$

$$Lab\ costs = 0.2 * Total\ Operating\ Labour$$

$$Supervision = 0.2 * Total\ Operating\ Labour$$

$$Plant\ Overheads = 0.5 * Total\ Operating\ Labour$$

$$Capital\ Charges = 0.1 * Fixed\ Capital$$

$$Local\ taxes = 0.02 * Fixed\ Capital$$

Table 7.23 Calculation of Fixed Operating Cost

Item	**%**	**Cost**
Maintenance = 0.05*FC	0.05	396987.6309
Operating Labor		
Annual Base salary ($)	5700	
Shifts	2	
Operators	10	
Total Base Salary ($)	57000	
Allowances ($)	28500	
Total operating labor ($)	85500	
Lab costs ($)	0.2	17100
Supervision ($)	0.2	17100
Plant Overheads ($)	0.5	42750
Capital Charges ($)	0.1	793975.2618
Local Taxes ($)	0.02	158795.0524
Total ($)		**1512208**

The variable operating cost covers the following items:

Table 7.24 Calculation of Variable Operating Cost

Raw materials	
Sulphur Cost ($/tonne) -1998	36.85
Cost index - 1998	436
Cost Index -2020	592
Sulphur Cost ($/tonne) -2020	50.0
Sulphur Weight / (tonne/hr)	1.4576
Sulphur Weight / (tonne/yr)	12768
Total Cost	**638982**

The cost of miscellaneous materials is calculated using the formula:

$$Miscellaneous\ Materials = 0.1 * Maintenace\ cost$$

Table 7.25 Calculation of Miscellaneous Material Cost

Miscellaneous Materials	39699

Table 7.26 Calculation of Utilities Cost

Utilities	
Process Water Weight (tonne/yr)	154402
Variable cost ($/tonne)	0.5
Total cost of PW ($)	77201
Steam Cost ($)	0
Total Utilities Cost ($)	**77201**

$$Total\ variable\ cost = Raw\ material + Utilities + \ Miscellaneous\ materials$$

Table 7.27 Calculation of Variable Operating Cost

Total Variable costs ($)	**755881**

$$Direct\ production\ cost = Fixed\ operating\ cost + Variable\ operating\ cost$$

Table 7.28 Calculation of Direct Production Cost

Direct Production Cost ($)	2268089

7.7 Calculation of Annual production Cost

$$Annual\ Production\ Cost = Direct\ production\ cost + Sales\ expense + General\ overheads + R\&D$$

Table 78-Calculation of Annual Production Cost

Annual Production Cost ($)	
Sales expense	453618
General overheads	
R&D	
Annual Production Cost ($)	**2721707**

Annual Production rate (kg)	38605320

7.8 Calculation of Production Cost per kg

The production cost per kg is calculated using the formula:

$$Production\ cost\ \left(\frac{\$}{kg}\right) = \frac{Annual\ Production\ Cost}{Annual\ Production\ Rate}$$

Production cost ($/kg)	0.0705

7.9 Calculation of payback period

$$\begin{aligned} &Total\ selling\ price\ of\ sulphuric\ acid \\ &= Selling\ price\ per\ unit\ mass * \ mass\ of\ acid\ produced \end{aligned}$$

$$Annual\ Cash\ Flow = Total\ selling\ price - Annual\ production\ cost$$

$$Payback\ period = \frac{Total\ investment}{Annual\ Cash\ Flow}$$

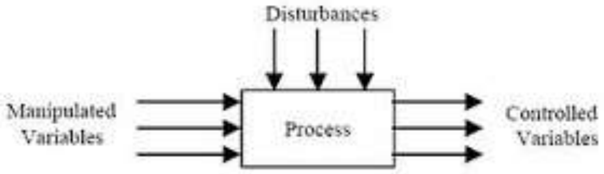
Table 79-Calculation of Payback Period

Payback Period	
Selling price H_2SO_4 / ($/tonne)	150
H_2SO_4 Produced (tonne/ hr)	4.41
H_2SO_4 Produced (tonne/ year)	38605
Total price	5790798
Annual Cash Flow ($)	3069090
Payback period (years)	**2.94**

CHAPTER 8

Instrumentation & Process Control

Process control system is a set of controls that is used to constantly monitor any deviation from the pre-set conditions and troubleshoot the errors to ensure a safe and efficient process. (Holloway, 2012)



8.1 Overview of the Controllers

8.1.1 Temperature Controller

A temperature controller measures the temperature (process variable) of the stream to which it is connected and compares it to the required value (set value). It does so using a closed or open loop control system. In the open loop system, the temperature of the output or the process variable does not influence the heating/cooling system. However, in the closed loop system, the temperature of the output is fixed and the process variable must be adjusted if there is a deviation from the set value.

A temperature controller uses this error (deviation) and sends an output signal to the manipulated variable that can be a control valve or heater or any other Final Control Element. Any variation in the manipulated variable subsequently has an effect on the Controlled Variable (Process Variable).

For our project, closed loop temperature controllers will be used as the temperatures affect the kinetics of the reactions.

8.1.2 Pressure Controller

Pressure Sensor detects any variation in the pressure of the equipment or the stream and Pressure Controller adjusts it to the set value if need be. Like other controllers, Pressure Controllers are crucial for use in production line as any deviations from the pre-set limits can be hazardous. Over pressurization may lead to explosions and low pressures can sometimes affect the kinetics and hence the quality of the product.

8.1.3 Flow Controller

It is an electrical control device that sends signals to troubleshoot deviations in the pre-set value of a controlled variable. It is connected to a flow control valve or an actuator. The actuator is adjusted by the controller in response to the signal detected by the sensor. The valve opening is adjusted as per requirement.

8.1.4 Level Controller

It uses a displacer which is connected to the sensor to detect fluid level. Displacer is immersed into the fluid and if the level increases and reaches the higher set value, the actuator will receive a signal from the level controller. The control valve opens to allow the level to drop to the set point.

8.1.5 Analyzer Controller

The composition control of the product is imperative for economic operation of the plant. To ensure stringent composition controls, process analyzers are installed that measure the composition (purity) of the product.

8.1.6 Differential Pressure Controller

It has a valve and a pressure actuator that are used to keep a constant and low pressure difference (pressure drop) across a system. It can also be used to control flow across a valve that affects the differential pressure

8.2 Process Control on Equipment

8.2.1 Controls in pump

8.2.1.1 Pressure Controller: The flow rate through the valve determines the amount of work the pump will exert on the incoming stream of molten sulfur that will subsequently affect the pressure of the outlet stream.

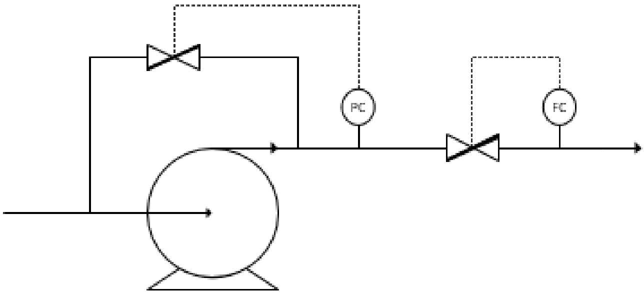


Figure 54-Controls on Pump

8.2.1.2 Flow Controller: The valve opening controls the flow-rate of molten sulphur. Higher the amount of product required, larger the opening.

8.2.2 Controls in Heat Exchanger

8.2.2.1 Flow Controller: The flow rate of SO_2/SO_3 gas mixture is controlled via a control valve. Higher the flow-rate of the SO_2/SO_3 gas mixture from the reactor, larger the opening of the valve to ensure required supply of the gas mixture to IAT and vice versa.

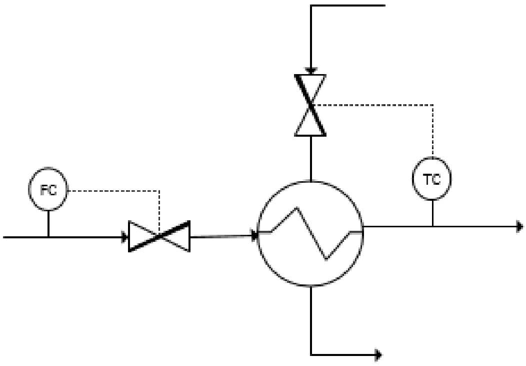


Figure 55-Controls on Heat Exchangers

8.2.2.2 Temperature Controller: The flow rate of the incoming shell-side stream of unconverted SO_2/SO_3 gases from the absorbers will determine the tube side outlet temperature that is coming from the reactor.

8.2.4 Controls in Reactor

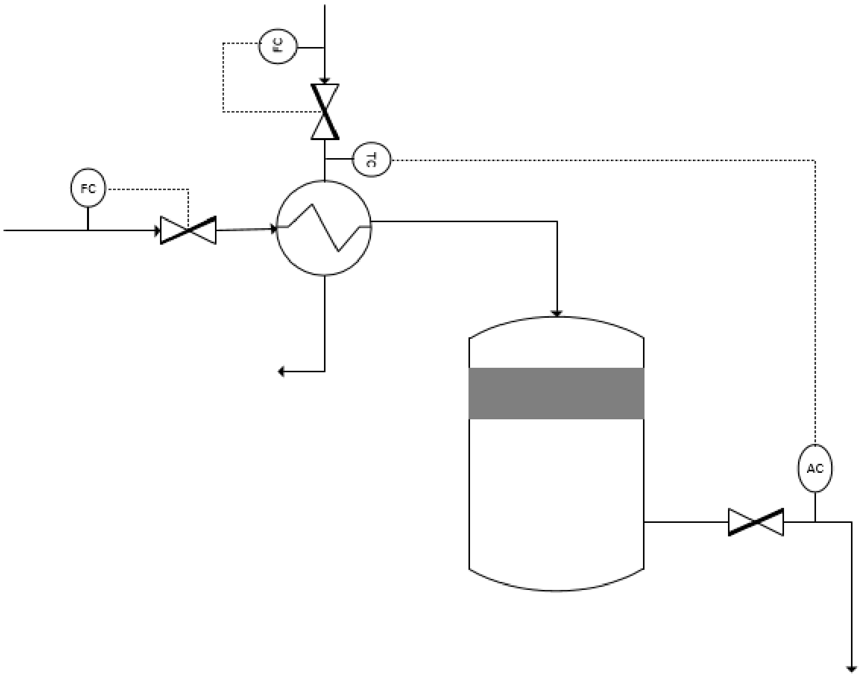


Figure 56-Controls on Reactor

The heat exchanger ensures optimum reactor inlet temperature to allow maximum SO_3 yield.

- Flow Controller: Flow-rate of the SO_2 tube side inlet stream is adjusted by a control valve. The opening of the valve is varied as per the required flow of the SO_2 gas mixture in the reactor.
- Pressure Controller: The optimum pressure of the SO_2 inlet stream is 2 atm. A pressure controller is used to ensure that the pressure remains at this value.
- Temperature Controller: The shell side heating stream temperature will ensure optimum tube side SO_2 outlet temperature entering the reactor that will subsequently affect the composition of the SO_3 product formed. For maximum conversion and purity, the temperature of the SO_2 stream must be optimum i.e. 450°C.
- Analyzer Controller: This is a crucial controller that ensures maximum purity of the SO_3 product which affects the quality of our final products i.e. sulphuric acid and oleum as well as the percentage of flue gases released into the atmosphere.

8.2.5 Controls in Absorption Column

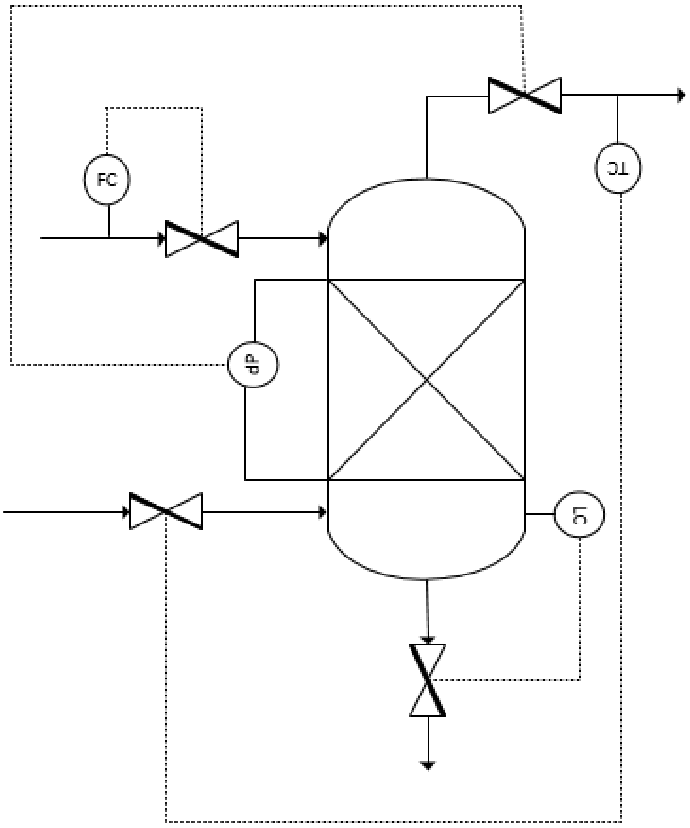


Figure 57-Controls on Absorbers

- Flow Controller: The incoming sulphuric acid flow rate will determine absorption extent of the gas mixture in sulphuric acid. The valve opening controls the flow rate. Higher the flow-rate of sulphuric acid required, larger the opening of the valve.
- Differential pressure controller: Absorbers work best at low temperatures and high pressures. However to ensure that there is not a large pressure drop in the packing as it will affect the quality of product, a differential pressure controller is used to allow low and constant pressure drop across the column.
- Level Controller: The controller is connected to a control valve at the column outlet to allow for only required level of gas mixture and sulphuric acid in the column. It is typically controlling the level of the interface between the two phases.

- Temperature Controller: The temperature of the outgoing SO_2/SO_3 gas mixture is controlled via the flow rate of the SO_2/SO_3 inlet. Absorption is favorable at low temperatures. Since our reaction is exothermic and the outgoing gas mixture stream has to be at a higher temperature before entering the Reactor Bed 4, the flow rate of the incoming stream must not be low so that maximum absorption takes place and higher amount of energy is produced.

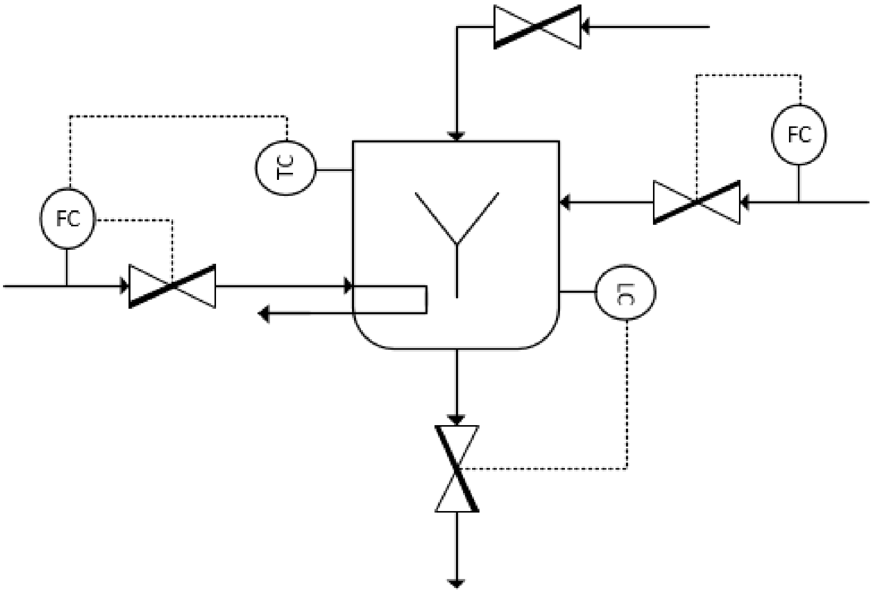


Figure 58-Controls on Dilution Tank

8.2.6 Controls in Dilution Tank

- Flow Controller (Coolant): The flow rate of the coolant ensures proper cooling of the oleum stream. Since the absorption of oleum in water to form sulphuric acid is a highly exothermic reaction, large amount of coolant is required to ensure acceptable level of temperature of the tank.
- Temperature Controller: The coolant flow rate controls the tank temperature. To ensure sufficient cooling of the tank and its products after an exothermic reaction, large amounts of coolant is required.
- Flow Controller: The flow rate of oleum inlet stream must be adjusted according to the need of the final product i.e. sulphuric acid.
- Level Controller: The level controller allows for the output of sulphuric acid as per requirement. If the level gets too high, it will allow the

sulphuric acid to pass through. It will also ensure that only required amount of oleum is entered so that the level does not rise too high or fall too low.

Following is the brief on the function of various controllers in each of our equipment.

Table 80-Proposed Plant Control System

Equipment	Name of Controller	Type of Controller	Manipulated Variable	Controlled Variable
Pump	PC	PI	Flow rate	Pressure
	FC	PI	Valve Opening	Flow rate
Air Dryer	FC	PI	Valve Opening	Flow rate of Air
	LC	PI	Flow rate of H_2SO_4 Outlet	Level
	TC	PI	Flow rate of $H_2S_2O_7$ Inlet	Temperature of Dried Air
	dP	PI	Flow rate of Dried Air Outlet	Pressure Drop
SO_2 Dryer	FC	PI	Valve Opening	Flow rate of Air
	LC	PI	Flow rate of H_2SO_4 Outlet	Level
	TC	PI	Flow rate of $H_2S_2O_7$ Inlet	Temperature of Dried Air
	dP	PI	Flow rate of Dried Air Outlet	Pressure Drop
Reactor	PC	PI	Flow rate	Pressure
	FC	PI	Valve Opening	Flow rate
	TC	PI	Inlet Flow rate	Reactor Temperature
	AC	PI	Reactor Inlet Temperature	Product composition
Heat	TC	PI	Shell side inlet	Tube Side

Exchanger 2			from HEX 3 flow rate.	SO_2/SO_3 Outlet Temperature
	FC	PI	Valve Opening	Flow rate of SO_2/SO_3
Heat Exchanger 3	TC	PI	Shell side Inlet from IAT flow rate	Tube Side SO_2/SO_3 Outlet Temperature
	FC	PI	Valve Opening	Flow rate of SO_2/SO_3
IAT	FC	PI	Valve Opening	Flow rate of Sulphuric Acid
	LC	PI	Flow rate of Oleum Outlet	Level
	TC	PI	Flow rate of SO_2/SO_3 Inlet	Outlet Temperature of SO_2/SO_3 mixture
	dP	PI	Flow rate of SO_2/SO_3 Outlet	Pressure Drop in the column
MAT	FC	PI	Valve Opening	Flow rate of Sulphuric Acid
	LC	PI	Flow rate of Oleum Outlet	Level
	TC	PI	Flow rate of SO_2/SO_3 Inlet	Outlet Temperature of flue gases
	dP	PI	Flow rate of flue gases	Pressure Drop in the column
Dilution Tank	FC	PI	Valve Opening	Flow rate of Oleum Inlet
	TC	PI	Flow rate of coolant	Temperature of the tank
	LC	PI	Flow rate of H_2SO_4 Outlet	Level
	FC	PI	Valve Opening	Flow rate of coolant

CHAPTER 9

Safety Analysis

As we look at different manufacturing plants, including the Oleum and Sulphuric Acid Plant, it is important to adopt a wide range of procedures to ensure the well-being and safety of the staff as well as the smooth and reliable plant operation. (Wong, 2010) It is also imperative to prevent the emergence of any hazards and operability issues. These safety measures are classified into four different categories as mentioned below: (Industries, 2004) (Baldwin, 2002)

9.1 Design strategies and continuous inspection

- Location of administration building away from the plant
- Plant boundaries consolidation with strong walls for safety and security
- Incorporation of noise controlling measures
- Proper illumination
- Proper ventilation for displacement of harmful oxides with air
- Minimization of fire risks by construction of roads
- Proper accessibility throughout the plant
- Easy accessibility of safety lifts, ladders and other instrumentation equipment
- Proper insulation of equipment for conservation of temperatures and safety of operators
- Proper guarding and inspection of rotating and moving parts
- Walkthrough audits to monitor the effectiveness of existing procedures and policies

9.2 Health and safety communication

- Periodic training of all employees regarding prevention of risks and vocational health and safety
- Induction of safety signs in compliance with national and international standards

- Proper tagging and labeling of all equipment
- Demarcation of different areas of the plant and walk ways
- Provision of instruction manuals for different equipment
- Establishment of access point to material safety data sheets

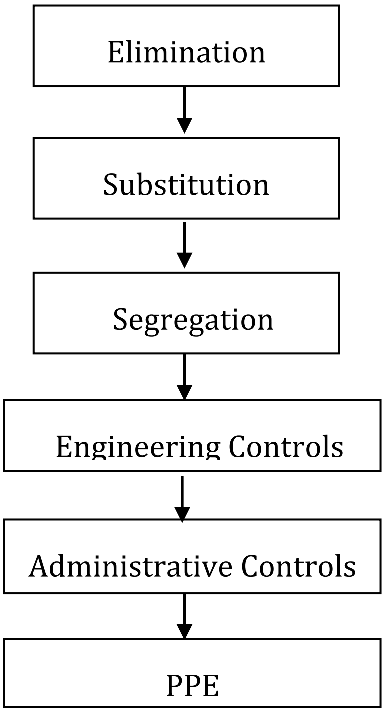
9.3 Safety practices and personal protective equipment (PPE)

- Preparation of suitable work schedules to avoid fatigue
- Proper cleaning facilities for workers to use during and after shifts
- Medical testing for proper monitoring of employee health
- Usage of personal protective equipment

Table 9.1 Types of PPE and its application

Type of PPE	**Application**
Hard Hat	Protects skull from impacts, penetration and electrical injuries especially in compact areas (converter and absorption column) and in multilevel units with restricted head space
Safety Shoes	Protection against rolling and falling objects, electrical hazards, hot surfaces and hot liquid splashes Recommended to be worn throughout the plant
Goggles	Protection against flying fragments, dusts and hot liquid splashes Recommended to be worn near high concentration oleum and Sulphuric acid production units
Respirators	Protection against inhalation of toxic vapors such as oxides of Sulphur, Sulphuric Acid mist etc.

- Implementation of safety protocols other than personal protective equipment depending on likelihood and severity of the hazard



9.4 Emergency Planning

- Allocation of assembly points near admin building in case of emergencies
- Incorporation of a functioning, frequently tested system of alarms
- Induction of fire extinguishers at different locations
- Sufficient stock of first aid material due to excessive exposure to sulphuric acid and oleum
- Directions for all first aid equipment (showers, eye-wash stations, stretchers, kits) throughout the plant
- Installation of health care center near the plant
- Detailed procedures for immediate evacuation in case of emergency
- Back up services such as extra water tanks and fire ponds

9.5 Detailed HAZOP Analysis

A Hazard and operability study (HAZOP) is an overall systematic examination and structured representation of a complex planned operation or process for the identification and evaluation of problems that may pose a risk to either the equipment or personnel.

The first step is a detailed analysis of the three major materials that are entering as reactants, or coming out as products namely Sulphur, Oleum and Sulphuric Acid. The major hazards associated with them and the course of action that can be employed for their mitigation have been mentioned below. (Khan, 2018)

Table 9.2 Hazard Analysis of Raw Materials and Products

Sulfur	Flammable	Isolated area for storage that must be monitored Charging in melter using Automated Conveyer System
Oleum	Corrosive	Provision of level gauge Provision of Personal Protective Equipment to all employees Provision of double drain valve to Sulfuric Acid storage tank
Sulfuric Acid (98.5%)	Corrosive	Display of precautions and first aid Safety training for all employees Provision of eye wash and safety shower Closed process for Sulphuric Acid handling

Following the analysis of the materials, we then moved onto the major equipment. This was done by using certain guidewords to look for any probable causes of incidents in equipment, their consequences, as well as the necessary action required for their mitigation. We will be dealing with the equipment one by one with the help of tables mentioned below. (BUITEMS, 2019)

9.5.1 Sulfur Melter HAZOP Worksheet

Table 9.3 Property data for Sulfur Melter

Title	**Melting of Sulfur**				
Part Considered	Storage tank and Pump			**Activity**	Melting
Design Intent	**Pressure**	**Temperature**	**Flow**	**Material**	Sulfur
	101.3 kPa	150^0C	1457.6 kg/hr	**Source**	Sulfur Storage Tank

Table 81-HAZOP Analysis on Sulfur Melter

Parameter	**Guide Word**	**Possible Causes**	**Possible Consequences**	**Actions required**
Temperature	More	• High temperature of steam • High ambient • temperature • Low sulfur quantity	• Explosion due to ignitable mixture of H_2S vapors, sulfur steam and dust • Change in molecular structure of sulfur causes pump failure	• N_2 blanketing of vapor phase to eliminate source of ignition • Controls on flow, temperature & pressure
	Less	• Low temperature of inlet steam • Low ambient temperature • Increased quantity of sulfur to the burner	• Plant shutdown due to blockage of sulfur in lines • Incomplete melting of sulfur may cause failure of pump	• Install a bypass heater • Periodic maintenance and Calibration of heating and control system

Parameter	Guide Word	Possible Causes	Possible Consequences	Actins Required
Pressure	More	• Temperature of incoming stream is very high • Choking in the outlet valve	• Over pressurization may cause adverse effect on unit operation	• Controls on temperature and pressure • Periodic inspection of the Sulfur melter • Vent valves must be installed
	Less	• Low Temperature • Low flow • Failure in PSV (Pressure Safety Valve)	• Wrong operation procedures due to different process conditions	• Install flow and temperature indicator • PIAL (Pressure Indicator Alarm Low) installed
Level/Flow	More	• Outlet valve choked • Faulty level indicator • High flow rate or pump running over-speed	• Accumulation of material may cause over-pressurization that will cause leakage and thus poor quality products will be formed • Overflow and spillage of molten sulfur in sulfur melter	• Pressure and level indicators must be used • Usage of Vent valves • Installation of high level alarms in melter • Use a diversion tank with proper valve arrangements

	Less	• Leak in feed line or melter • Insufficient supply of raw material • Inlet valve choked	• Probability of running dry • Change in molecular structure of sulfur and an increased pumping power leading to its failure	• Install automatic flow control valve • Use NDT such as liquid penetration inspection or radiography
	Reverse	• Higher pressure and flow rate at the outlet causes a back flow	• Product set off	• Install NRV (Non-Return Valve)
Sulfur and Steam	As well as	• Leakage in the steam coil • Foreign impurities	• The quality of sulfur will deteriorate due to exposure with steam	• Non-invasive techniques to detect leaks • Regular Quality Control
	Power Failure		• Sulfur blockage will cause pump failure	• DG supply

9.5.2 Furnace HAZOP Worksheet

Table 9.5 Property data for Furnace

Title	Burning of Sulfur with Air				
Part Considered	Sulfur Burner			**Activity**	Combustion
Design Intent	**Pressure**	**Temperature**	**Flow**	**Material**	Sulfur
	101.3 kPa	2014^{0}C	1458 kg/hr	**Source**	Sulfur Melter

Table 82-HAZOP Analysis on Furnace

Parameter	Guide Word	Possible Causes	Possible Consequences	Actions required
Temperature	More	• High flow rates of incoming streams	• Rupture of furnace walls	• Closed process • Controls on flow, temperature & pressure
	Less	• Low incoming air temperature • Less amount of fuel incoming	• Incomplete or no combustion of sulfur to produce SO_2	• Install a bypass heater • Periodic Calibration of heating and control system
Pressure	More	• Inlet air flow is higher than set point • High inlet pressure • Choking prevents SO_2 gas discharge	• Rupture of furnace • Heavy leakage may start	• Use flow control valves • Use PSV • Install bypass line for restrictions

Parameter	Guide Word	Possible Causes	Possible Consequences	Actions Required
Level/Flow	More	• Choking at the outlet • High flow rates of incoming streams • Faulty level indicator	• Over pressurization and overheating may cause rupture of walls	• Vent valve should be connected • Regular check of Level indicators • Installation of high level alarms in furnace
	Less	• Less flow of raw materials than set value	• Incomplete combustion of sulfur to produce SO_2	• Automatic flow control valve installation • Periodic inspection and maintenance
	No	• No flow of raw materials due to pump or valve malfunction • Power failure	• No combustion of sulfur • Blockage of sulfur in burner	• Installation of flow controller • Connect to DG supply

9.5.3 Reactor HAZOP Worksheet

Table 9.7 Property data for Reactor

Title	Combustion of SO_2 with air				
Part Considered	Packed Bed Reactor			**Activity**	Combustion
Design Intent	**Pressure**	**Temperature**	**Flow**	**Material**	Sulfur
	101.3 kPa	450^0C	11658.38 kg/hr	**Source**	Sulfur Melter

Table 83-HAZOP Analysis on Reactor

Parameters	**Guide Words**	**Possible causes**	**Possible Consequences**	**Action Required**
Temperature	More	• Incoming streams are at a higher temperature than set value	• Inefficient reaction due to temperature deviation • Over pressurization may occur	• Monitoring of pre coolers to ensure optimum temperature of the inlet stream • Periodic inspection of PSV
	Less	• Raw material inlet is at a lower temperature than set value	• Reaction kinetics will change and reaction rate will slow down	• Higher concentration of SO_2, more formation of SO_3
Pressure	More	• High temperature • Leakage in pipelines	• Over pressurization of the reactor	• Installation of PSV • Use NDT such as robotic crawlers etc

Parameter	Guide Word	Possible Causes	Possible Consequences	Actions Required
	Less	- Failure of PI controller which sends a bad signal	- Decrease in the operation conditions causing inefficiency in reactor	- Monitoring of controllers to avoid undesirable reactions and maintain SO_3 production
Level/Flow	More	- Excessive pump speed - Failure of flow control valves	- Increase in pressure - Decrease in temperature leading to change in reaction kinetics	- Regular inspection of pump motors - Install high level alarms - LCV (level control valve) inspection
	Less	- Less air flow - Valve is partially closed - Leak in the pipeline	- Unconverted SO_2 due to incomplete combustion - Product specifications not met	- Recycling of unconverted SO_2 to temporary storage - Control loop to automatically switch on/off the valve - Use of NDT such as electro-magnetic testing
	No	- No air flow - Closed valve - Line blockage - Pipeline rupture	- No product formed (SO_3) - Excess SO_2 in the reactor.	- Inspection of transfer lines - Efficient communication with operators - Possible Shutdown

Parameters	**Guide Words**	**Possible Causes**	**Possible Consequences**	**Actions Required**
	Part of	• Normal flow of decreased concentration of O_2 • Environmental conditions/ polluted air	• Unreacted SO_2 • Composition of product will alter due to impurity	• Check O_2 supply equipment after charging • Inspection of pump design and associated pipeline
	Other than	• Material flow other than O2 in line • Environmental pollution (main source of oxygen)	• Poisoning of catalyst • Equipment failure	• Equipment shutdown • Inspection of equipment

9.5.4 Absorber HAZOP Worksheet

Table 9.9 Property data on Absorber

Title	Absorption of SO_3 in Oleum				
Part Considered	Absorption Column			**Activity**	Absorption
Design Intent	**Pressure**	**Temperature**	**Flow**	**Material**	SO_3
	101.3 kPa	200^0C	3654.93 kg/hr	**Source**	Reactor

Table 84-HAZOP Analysis of Absorber

Parameter	Guide Word	Possible Causes	Possible Consequences	Actions required
Temperature	More	• High inlet stream temperature • Excess flow of heated raw materials	• Disturbed rate of reaction • Increased temperature leads to excess buildup of pressure	• Installation of high level alarms • Installation and inspection of Pressure Relief Valve
	Less	• Low temperature of inlet streams • Low temperature of column	• Inadequate absorption due to change in reaction kinetics	• Installation of temperature controls at inlets and outlets
Pressure	More	• High pressure of inlet streams • Closed flow control valve causing restriction • Malfunction of valve	• Breakage of packing • Bursting of column	• Installation of Vent Valve or PSV • Use of Radiography and Ultrasonic Testing
	Less	• Low pressure of inlet streams • Malfunction of pressure regulating valve	• Leads to improper reaction • Leads to abnormal temperature and pressure in column	• Using Differential Pressure Control to ensure safe limit • Installation and Inspection of Pressure Relief Valve

Parameter	Guide Word	Possible Causes	Possible Consequences	Actions Required
Flow/Level	More	• FCV (Flow Control Valve) malfunction • Valve left open by the operator	• High pressure drop across the column • Flooding	• Regular inspection of valves
	Less	• Leakage of the pipe • Valve partially opened	• V/L ratio is lower than required. • Inefficient absorption/ reaction	• Use of Non Destructive Techniques • Install PIAL (Pressure Indicator Alarm Low)
	No	• Blockage of the pipeline • Valve failure • Failure of flow in previous units	• High liquid hold up in the column • V/L ratio is lower than required	• Installation of level controller alongside alarm in case of excessive liquid hold-up

9.5.5 Heat Exchanger HAZOP Worksheet

Table 9.11 Property data on Heat Exchanger

Title	Temperature Optimization for reactor				
Part Considered	Heat Exchanger			**Activity**	Heating/ Cooling
Design Intent	**Pressure**	**Temperature**	**Flow**	**Material**	SO_2/SO_3
	101.3 kPa	471^0C	15338.7 kg/hr	**Source**	Reactor

Table 85-HAZOP Analysis of Heat Exchanger

Parameters	Guide Words	Possible Causes	Possible Consequences	Action Required
Temperature	More	• Higher shell side inlet stream temperature • Increase in the shell side inlet stream flow • Higher tube side inlet temperature	• Increase in temperature difference increases power requirement and fluctuation in outlet temperatures • Increase in pressure	• Maintenance of control loops • Installation of a flow indicator • Implement a differential pressure gauge to work with input and output currents for the tube side
	Less	• Leakage in the pipes due to cracks or corrosion • Shell side temperature is lower • Decrease in shell side flow rate	• Decrease in the contact area • Over pressurization • Increased outlet stream temperature • Temperature difference decreases	• Use differential pressure gauge to work with input and output currents for the tube side • Eddy current probes insertion in tubes to check for corrosion • Maintenance of flow control loops

Parameter	Guide Words	Possible Causes	Possible Consequences	Actions required
Pressure	More	• Higher tube side pressure due to valve malfunction	• Rupture of tube	• Installation of high pressure alarm
Flow/level	More	• Increase in shell side inlet flow • Failure of shell side inlet valve to close	• Higher decrease in tube side outlet temperature than set value	• Low temperature alarm • Installation of temperature indicators before and after the heat exchanger
	Less	• Less shell side inlet flow due to pipe leakage or blockage	• Constant tube side temperature	• High temperature alarm installation • Installation of flow meter
	Reverse	• Flow of tube side is reversed due to malfunction of tube side valve	• Product specifications are not met	• Installation of Non Return Valve

	No	• No shell side inlet • Failure of shell side inlet valve	• Insufficient decrease in tube side temperature	• Installation of temperature indicators at the inlets and outlets • High temperature alarm installed

CHAPTER 10

Optimization

One of the major objectives of our project, after design, was to present an optimized and workable solution of the conventional problems in Sulphuric Acid production process and plants.

While keeping an economic investment in mind, taking into consideration all the requirements of the Industry and assuring a compliance with all the issued guidelines, a few optimization steps were thus introduced as a part of the project and well incorporated into it. These plans and propositions are given below along with their ample elaborations and annexing details.

10.1 Dual functionality

The plant is dual functional as it produces both Sulphuric acid and Oleum as per the requirement.

Table 86-Proposed Capacity

Percentage	**Sulfuric Acid (kg/hr)**	**Oleum (kg/hr)**
100% H_2SO_4	4410	0
100% $H_2S_2O_7$	0	4005
50% H_2SO_4, 50% $H_2S_2O_7$	2205	2002.5
25% H_2SO_4, 75% $H_2S_2O_7$	1102.5	3003.75
75% H_2SO_4, 25% $H_2S_2O_7$	3307.5	1001.25

10.2 Reactor efficiency

Inter-stage cooling is introduced between the catalyst beds of the reactor to keep temperature uniform in order to assume an adiabatic reactor which results in efficiency increase, better product quality, no catalyst deactivation and cost optimization.

10.3 Yield maximization

Due to the fact that all recycle streams and equipment outlets, or inlets are interlinked, the final value of plant capacity for oleum or sulfuric acid production or the definite yield becomes a function of a number of variables i.e. the split ratios of all these interlinked streams/recycles. Any change in one/more of these variables brings some change in the product amount and these changes when brought collectively can either increase or decrease our product yield. Thus, by employing the hit and trial methods on multiple drafts of the simulation model, the resulting yield of each constraint application was observed and crosschecked until a well-suited optimum value was reached. This capacity obtained can hence be termed as the optimized maximum and efficient capacity of our subject plant. The numerical value achieved for capacity here was 4407kg/hr.

Table 87-Optimization of yield based on split ratios

Split Ratio Oleum	Split Ratio H2SO4	Yield (H_2SO_4)
0.397	0.4920	3499
0.45	0.4420	3772
0.55	0.4920	4410

10.4 Energy recovery and its cost effectiveness

Pinch Analysis is performed on the entire plant to optimize the energy requirements as depicted in the table. This saves the need to supply any excess energy thus making the process more cost effective and completely green. The energy values are provided in the table below.

Table 88-Energy recovery

Heating equipment	**Energy value (MJ)**	**Cooling equipment**	**Energy value (MJ)**
H-100	285	C-100	13890
H-101	3090	C-101	11320
Sulfur Burner	27412	C-102	1236
		C-103	4258
		Air Dryer	62.3
		SO_2 Dryer	0.10
		IAT	827
		MAT	248
		Dilution Tank	356
Total energy consumed (MJ)	30787	Total energy released(MJ)	32197
Remaining energy (MJ) = 1410			

Table 89-Energy contribution of equipment

Equipment	**Percentage energy contribution**
Sulphur Burner coolers	52.0%
Reactor coolers	17.1%
Dryers	1.30%
Absorption tanks (IAT + MAT)	22.2%
Dilution Tank	7.40%

The energy recovered from the Sulphur burner and reactor is high grade energy and should be used in process heating or power requirements where the energy requirement is large. Whereas, energy coming from dryers, absorption and dilution tanks is low grade energy and should only be used for processes such as water heating or inline heating of process lines such as molten Sulphur transfer lines.

Table 90-Steam energy consumption

Unit	Utility	MJ/hr	kW	Cp	Δ T	m(kg/s)	m(kg/hr)	m(ton/yr)
Melter (H100)	S	285	79.2	1.87	125	0.339	1219	10680
H-101	S	3090	858	1.87	265	1.73	6226	54540
Sulphur Burner	S	27412	7614	1.87	1864	0.115	413	3618
Total steam = 68840 tons								
Cost of steam per ton ($/ton) = 12								
Cost saved ($) = 826082								

Table 91-Cooling Water consumption

Unit	Utility	MJ/hr	kW	Cp	Δ T	m(kg/s)	m(kg/hr)	m(ton/yr)
Burner (C-100)	CW	13890	3858	4.18	1014	0.910	3277	28707
Burner (C-101)	CW	11320	3144	4.18	930	0.809	2912	25509
Reactor (C-102)	CW	1236	343	4.18	75.8	1.08	3903	34186
Reactor (C-103)	CW	426	1182	4.18	323	0.875	3152	27609
Dryers	CW	62.4	17.3	4.18	15	0.276	995	8718
MAT	CW	248	68.9	4.18	113	0.146	525	4599
IAT	CW	827	229	4.18	90	0.611	2198	19257
DT	CW	356	98.9	4.18	47.4	0.499	1797	15739
Total CW = 164326.3								

Cost per ton ($/ton) = 0.5

Cost saved ($) = 48118

$$Total\ cost\ saved = Steam\ cost + Cooling\ Water\ cost$$

Total cost saved ($) = 874200

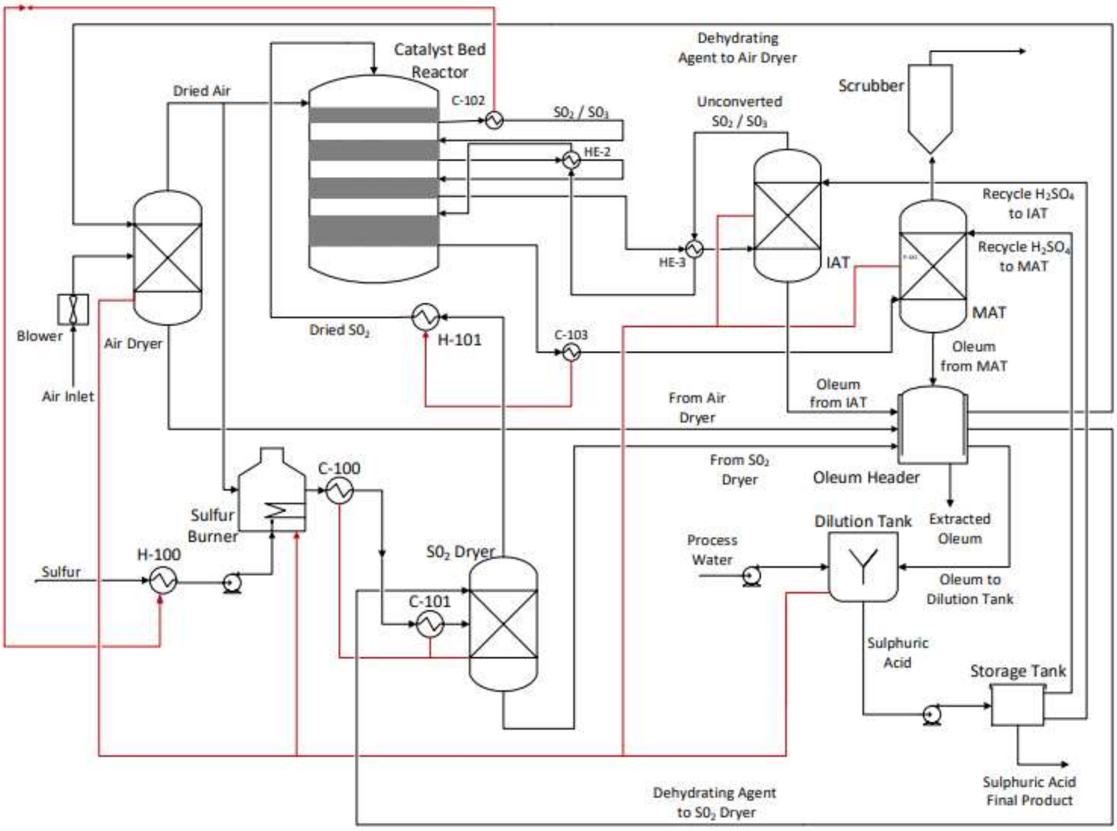


Figure 59-Energy Recovery Process Flow Diagram

CONCLUSION

The plant thus designed has a capacity of 4410 kg/hr of Sulphuric Acid with a dual functional facility provision along with variant optimization techniques employed at every stage. It offers inter-stage cooling between adiabatic catalytic beds of the reactor thus saving catalyst deactivation and excessive running cost. A pinch analysis of the entire plant is offered as an add-on which indicates maximum energy recovery about 70% of which is offered by the plant's own coolers. The conversion efficiency for SO_3 formation is maximized by properly tested optimization of all contributing recycle streams and inlets/outlets. Process is ensured to be designed keeping well in mind the green considerations for the plant and it offers an efficiency of 99.7%. With an overall payback period of 2.94 years, the production plant fulfills all the set objectives and can hence be termed as workable, cost friendly, Eco-system catering, efficient, practical and green.

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