UNIVERSITY OF DAIYLA
COLLEGE OF ENGINEERING
CHEMICAL ENGINEERING DEPARTMENT

PHOSPHORIC ACID PRODUCTION

APROJECT SUBMITTED TO THE COLLEGE OF ENGINEERING OF THE UNIVERSITY OF DIYALA IN PARTIAL FULFILLMENT OF THE REQUIREMENETS FOR THE DEGREE OF BACHLOR OF SCIENCE IN CHEMICAL ENGINEERING

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بـــــسم الله الرحمن الرحيم

قَالُواْ سُبْحَانَكَ لاَ عِلْمَ لَنَا إِلاَّ مَا عَلَّمْتَنَا
إِنَّكَ أَنتَ الْعَلِيمُ الْحَكِيمُ

سورة البقرة الآية (٣٢)

DEDICATION

The beginning, "Thank Allah for the completion of this project and ask Allah Almighty to benefit him, and then dedicate this search modest to our families and loved ones and to all who support us and helped us to complete this project of professors and singled them. ***Professor Dr. Anees A. Khadom*** and the rest of esteemed professors who provided us with information We appreciate their efforts so and them sincerely with the thanks and appreciation of us.

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I wish to thank ***Professor Dr.*** **Abdul Mun`em A. Karim** Dean of College of Engineering – Diyala University.

I wish to thank ***Assist. Professor Dr. Anees A. Khadom*** head of the Chemical Engineering Department for providing research facilities.

I repeat thanks to our teacher Mr. ***Assist. Professor Dr. Anees A. Khadom*** guidance, and his continued support for the completion of this research.

Above all, thanks to Allah for the mercy and blesses he showed.

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Supervisor Certification

I certify this project entitled (PHOSPHORIC ACID PRODUCTION) was prepared under my supervision at the Chemical department of Engineering College University of Diyala, by (***JAMAL KH. Ali, OMER MUHSIN, BARQ BAHMAAN***) as partial fulfillment of the requirements for the Degree of Bachelor in Chemical Engineering.

Certification of Committee

We certify that we have read this project and as supervisor and examining committee examined the students in its content and that in our opinion; it meets the standard of a project for the degree of Bachelor of Science in Chemical Engineering.

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SUMMARY

Through the study of the project, we found that phosphoric acid of high acid important commercial value, is of great significance in the field of industry, such as phosphate salts used in industry with fertilizer, which occupies an important place

The phosphoric acid produced in two ways:

1 .Wet method

2. Way electric furnace

It was selected as the wet method based on the comparison between the two methods:

-Wet method uses phosphate rocks with a high-ranking higher than in the way the electric furnace

-The cost of electric furnace method be higher than the wet method

-Produces Al jpson wet method byproduct has commercial significance as it is used directly or indirectly

-Wet method using sulfuric acid, which can be obtained from sulfur and is available in abundance in our country

-Acid resulting from the wet method can be used in the fertilizer industry

Through this comparison shows us that the wet method is most effective in the production of phosphoric acid was used in this way in a number of industrial units such as the reactor Height of reactor (1.5 m), filters and tower absorption length (12.8 m) and heat exchanger length (3.2 m) and a production capacity of 300 tons per year and this amount is only for the purposes of the accounts in this research so she adjustable as needed required.

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CHAPTER ONE

1.1 Introduction (1)

Phosphoric acid was discovered in 1770 by K. W. Scheele and J. G. Gann in bone ash. Scheele later isolated phosphorus from bone ash (1774) and produced (1777) phosphoric acid by the action of nitric acid on phosphorus

Some 9 years later, Albright and Wilson, Ltd was founded at Old bury. In the early days, white phosphorus was obtained from bone ash by treating them with hydrochloric acid to produce precipitated phosphates. Then heating the Meta phosphate for several days in a sealed crucible, in a retort, and distilling off phosphorus vapor, under water. Huge quantities of coal were needed for heating these retort.

The production of white phosphorus was improved by using phosphate rock and sulfuric acid instead of bone ash and hydrochloric acid; and by the use of reverberator furnaces instead of the direct-heated furnace.

White and amorphous phosphorus remained the main product of Albright and Wilson until World War 1.

Phosphoric acid or tri hydroxide phosphorus and other names (ortho phosphoric acid, tri hydroxyl phosphine oxide).

Phosphoric acid is used as an additive and flavoring agent in both human and animal feed It is commonly used in sodas to provide a Sharp or sour flavor. In fact almost all the acidic flavor in soda Comes from phosphoric acid as the carbonic acid contained in the Bubbles has little effect on the overall ph. Phosphoric acid also helps to keep bacteria and fungi from forming in these sugary drinks

Phosphorous is one of the most essential plant nutrients in order to add extra phosphorous to soil , phosphoric acid is converted into Phosphates that are then mixed in with other ingredients to form Fertilizer more than 80 percent of the phosphoric acid produced in the world is used in the manufacture of fertilizer.

1.2 Physical Properties (1)

Table (1) shows the physical properties of phosphoric acid

MOLECULAR FORMULA	H_3PO_4
CHEMICAL NAME	Ortho phosphoric acid
COMMON NAME	phosphoric acid
SOLUBILITY	miscible in water
Molecular Weight	98.00
Boiling Point	213 ^oC
Melting Point	42.35 ^oC
Density/Specific Gravity	tribasic acid (at 25 ^oC)
Vapor Pressure	0.03 mm Hg at 20 ^oC
Vapor Density	3.4
Conversion Factor	1 ppm = 4.01 mg/m^3

1.3 Chemical properties (1)

Mineral acid is the chemical formula H_3PO_4 Phosphoric acid is made up of a dense crystalline solid colorless and odorless and is often used as a solution of water, where it dissolves in the water and reach the boiling point of phosphoric acid 230.5℃. Phosphoric acid is the main source for the first phosphor used in the phosphate fertilizer industry. And phosphoric acid canker cause irritation of the skin and eyes touching and the occurrence of ulcers membranes and tissues, as it leads to poisoning if swallowed or inhaled. As the phosphoric acid is a source of phosphorus necessary for the growth of aquatic plants it is possible to contribute phosphoric acid Lagoon in stagnant water bodies or slow flow, especially those surfaces with low content of phosphorus. And so far it has not been proven scientifically that phosphoric acid from cancer-causing substances. In nature, the minerals that cause water hardness reduce the degree of acid and phosphate salts still remain in the soil until plants use natural fertilizer.

$$3\ H_2SO_4\ (l) + \underset{\text{(ROCK PHOSPHATE)}}{Ca_3\ (PO_4)_2(s)} + 6\ H_2O\ (l) \rightarrow 2\ H_3PO_4(s) + \underset{\text{(GYPSUM)}}{3\ CaSO_4.2\ H_2O(s)}$$

SIDE REACTIONS:

$$CaF_2 + H_2SO_4 + 2H_2O \rightarrow 2HF + CaSO_4.2H_2O$$

$$6HF + SiO_2 \rightarrow H_2SiF_4 + 2H_2O$$

1.4 Method of production (2)

1.4.1 Wet Process Acid Production

1.4.2 Thermal Process Acid Production

Phosphoric acid (H_3PO_4) is produced by 2 commercial methods: wet process and thermal process. Wet process phosphoric acid is used in fertilizer production. Thermal process phosphoric acid is of a much higher purity and is used in the manufacture of high grade chemicals, pharmaceuticals, detergents, food products, beverages, and other no fertilizer products. In 1987, over 9 million mega grams (Mg) (9.9 million tons) of wet process phosphoric acid was produced in the form of phosphorus pent oxide (P_2O_5). Only about 363,000 Mg (400,000 tons) of P_2O_5 was produced from the thermal process. Demand for phosphoric acid has increased approximately 2.3 to 2.5 percent per year.

The production of wet process phosphoric acid generates a considerable quantity of acidic cooling water with high concentrations of phosphorus and fluoride. This excess water is collected in cooling ponds that are used to temporarily store excess precipitation for subsequent evaporation and to allow recirculation of the process water to the plant for re-use. Leachate seeping is therefore a potential source of groundwater contamination. Excess rainfall also results in water overflows from settling ponds. However, cooling water can be treated to an acceptable level of phosphorus and Fluoride if discharge is necessary.

1.4.1 Wet Process Acid Production

In a wet process facility, phosphoric acid is produced by reacting sulfuric acid (H_2SO_4) with naturally occurring phosphate rock. The phosphate rock is dried, crushed, and then continuously fed into the reactor along with sulfuric acid. The reaction combines calcium from the phosphate rock with sulfate, forming calcium sulfate ($CaSO_4$), commonly referred to as gypsum. Gypsum is separated from the reaction solution by filtration. Facilities in the U. S. generally use a dehydrate process that produces gypsum in the form of calcium sulfate with 2 molecules of water (H_2O) ($CaSO_4$ 2 H_2O or calcium sulfate dehydrate). Japanese facilities use a hemihydrate process that produces calcium sulfate with a half molecule of water ($CaSO_4$ ½ H_2O).

This one-step hemihydrate process has the advantage of producing wet process phosphoric acid with a higher P2O5 concentration and less impurities than the dehydrate process. Due to these advantages, some U. S. companies have recently converted to the hemihydrate process. However, since most wet process phosphoric acid is still produced by the dehydrate process, the hemihydrate process will not be discussed in detail here. A simplified reaction for the dehydrate process is as follow:

$$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \rightarrow 2H_3PO_4 + 3(CaSO_4).\,2H_2O$$

In order to make the strongest phosphoric acid possible and to decrease evaporation costs, 94 percent sulfuric acid is normally used. Because the proper ratio of acid to rock in the reactor is critical, precise automatic process control equipment is employed in the regulation of these 2 feed streams.

During the reaction, gypsum crystals are precipitated and separated from the acid by filtration. The separated crystals must be washed thoroughly to yield at least a 99 percent recovery of the filtered phosphoric acid. After washing, the slurred gypsum is pumped into a gypsum pond for storage. Water is syphoned off and recycled through a surge cooling pond to the phosphoric acid Flow diagram of a wet process phosphoric acid plant. Approximately 0.3 hectares of cooling and settling pond area is required for every mega gram of daily P_2O_5 capacity (0.7 acres of cooling and settling pond area for every ton of daily P2O5 capacity).

Considerable heat is generated in the reactor. In older plants, this heat was removed by blowing air over the hot slurry surface. Modern plants vacuum flash cool a portion of the slurry, and then recycle it back into the reactor.

Wet process phosphoric acid normally contains 26 to 30 percent P_2O_5. In most cases, the acid must be further concentrated to meet phosphate feed material specifications for fertilizer production. Depending on the types of fertilizer to be produced, phosphoric acid is usually concentrated 40 to 55 percent P_2O_5($75\%H_3PO_4$) by using 2 or 3 vacuum evaporators.

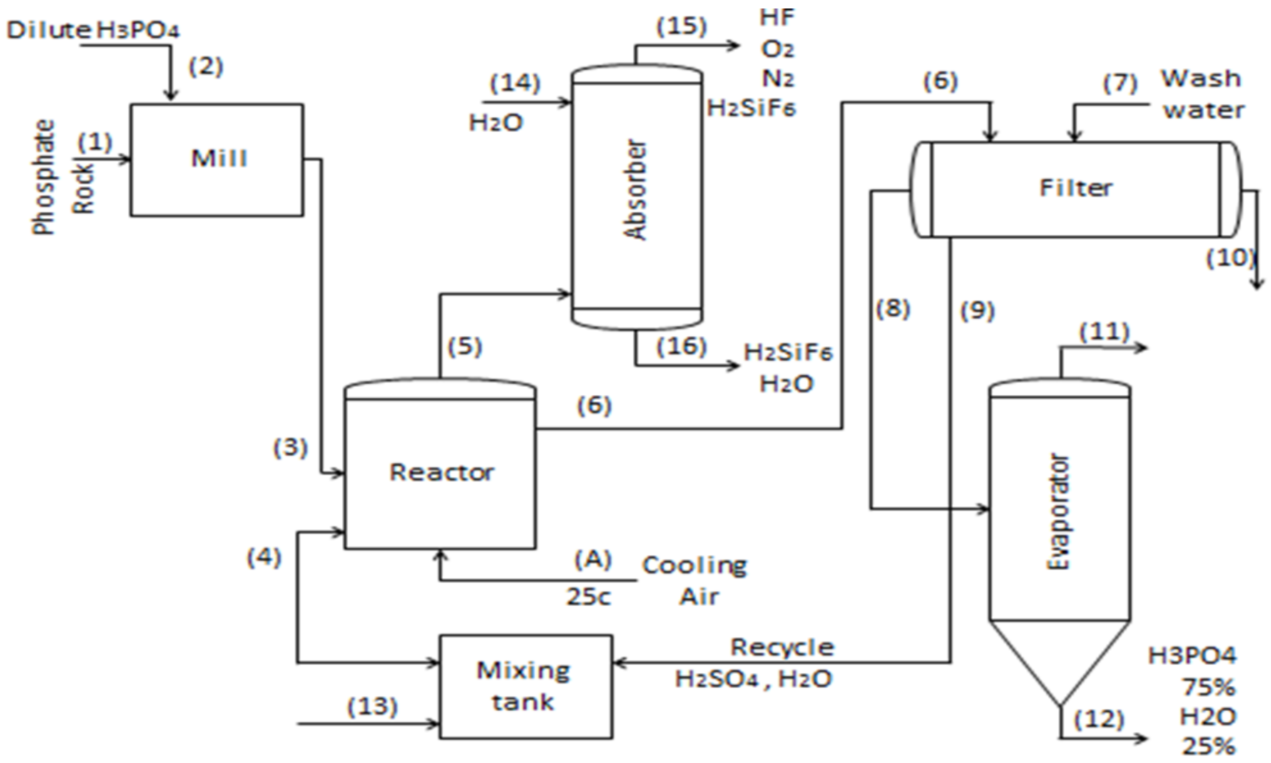


Figure (1.1) Flow diagram of a wet process phosphoric acid plant

1.4.2 Thermal Process Acid Production

Raw materials for the production of phosphoric acid by the thermal process are elemental (yellow) phosphorus, air, and water. Thermal process phosphoric acid manufacture, as shown schematically in Figure 1.2, involves three major steps: (1) combustion, (2) hydration, and (3) demisting

In combustion, the liquid elemental phosphorus is burned (oxidized) in ambient air in a combustion chamber at temperatures of 1650 to 2760°C (3000 to 5000°F) to form phosphorus pent oxide (Reaction 2). The phosphorus pent oxide is then hydrated with dilute H3PO4 or water to produce strong phosphoric acid liquid (Reaction 3). Demisting, the final step, removes the phosphoric acid mist from the combustion gas stream before release to the atmosphere. This is usually done with high-pressure drop demisters.

$$P_4 + 5O_2 \rightarrow 2P_2O_5 \qquad (2)$$

$$2P_2O_5 + 6H_2O \rightarrow 4H_3PO_4 \qquad (3)$$

Concentration of H_3PO_4 produced
from thermal process normally ranges from 75 to 85 percent. This high concentration is required for high grade chemical production and other non-fertilizer product manufacturing. Efficient plants recover about 99.9 percent of the elemental Phosphorus burned as phosphoric acid

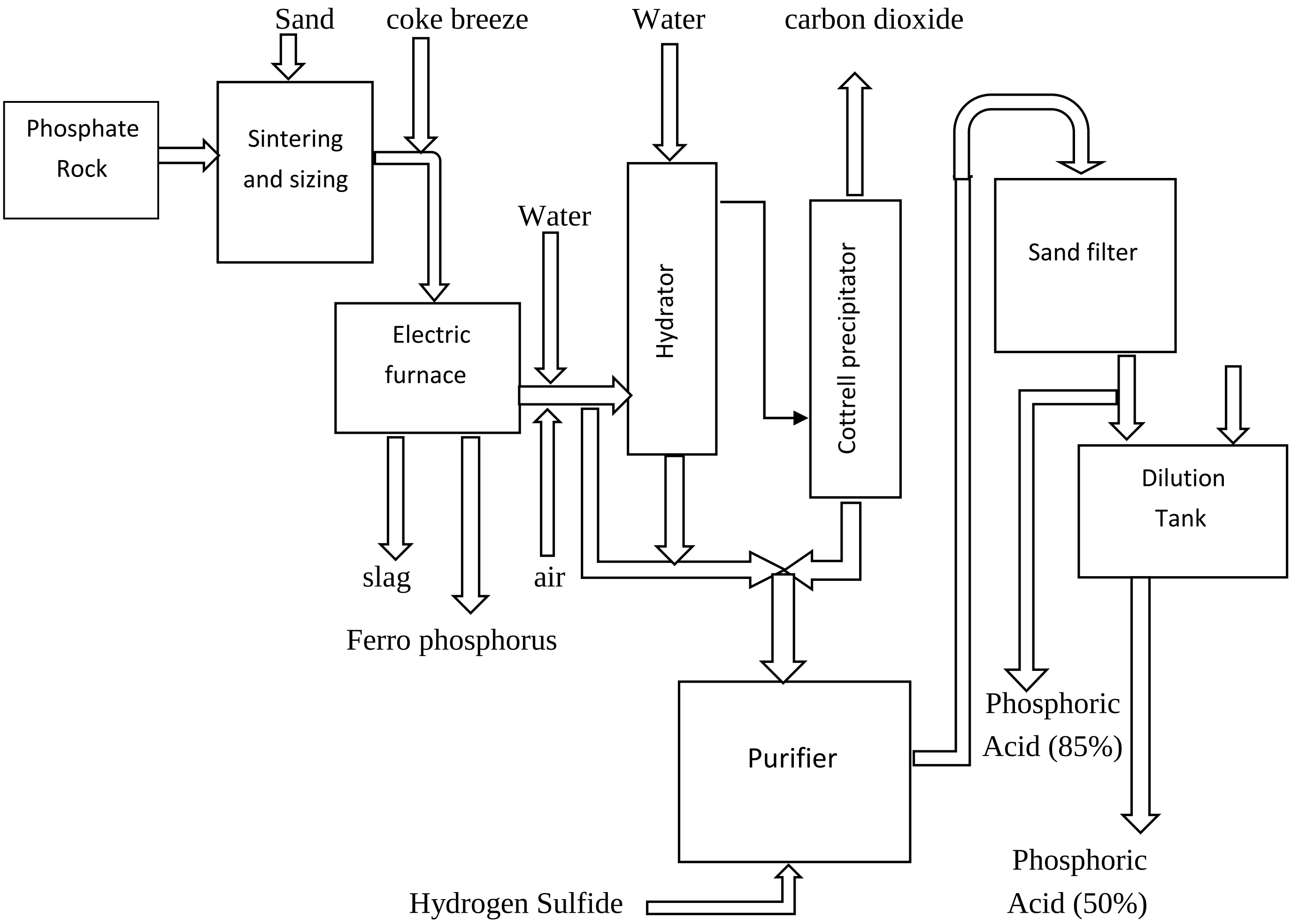


Figure (1.2) Flow diagram of Thermal process phosphoric acid plant

1.5 Choice of production method

1. Wet method uses high-ranking of phosphate rock, while electric furnace method used lower grades
2. Cost method of electric furnace to be conscious as compared with the wet method
3. Aljpson can produce byproduct of the wet method and the task of this article can be used directly
4. Wet method using sulfuric acid, which can be accessed through the availability of sulfur this is an important factor determining the production method for example, "In India there is a problem in the use of the wet method of difficulty provide sulfuric acid while in Iraq can be obtained easily sulfur
5. Wet way despite some disadvantages, but it is the most common way "in the world is approximately 80% of global output using this method

Accordingly, as shown in the above points has been chosen the wet method for the production of phosphoric acid

CHAPTER TWO

2.1 Material Balance

Main Reaction: -

$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \rightarrow 2H_3PO_4 + 3(CaSO_4 . 2H_2O)$ …….. (1)

Side Reactions:

$CaF_2 + H_2SO_4 \rightarrow 2HF + CaSO_4$ …….. (2)

$6HF + SiO_2 \rightarrow H_2SiF_6 + 2H_2O$ ……….. (3)

Raw Materials:

- Phosphate Rock Analysis % by wt.

$Ca_3(PO_4)_2$	75
CaF_2	20
SiO_2	5

- H_2SO_4 concentration 94%

Assumptions: -

- Production of H_3PO_4 =300 Ton/year =1000 Kg/day
- Year = 300day
- Some sulfuric acid (1-1.5) percent is allowed to go with gypsum to make it easily filterable.
- Excess of H_2SO_4 =15%
- Yield = 95%

Component	**M.wt**
H_3PO_4	98
H_2SO_4	98
SiO_2	60
CaF_2	78
H_2O	18
$Ca(PO_4)_2$	310
$CaSO_4$	136
$CaSO_4.2H_2O$	172
HF	20
H_2SiF_6	144
P_2O_5	142

Production of H_3PO_4 =1000 Kg/day = 42Kg/hr. = 0.428Kgmole/hr.

Stream of product = 42/0.75= 56Kg/hr.

H_2O= 56*0.25 = 14 Kg/hr.

From reaction (1)

Reacted of $Ca_3(PO_4)_2$ = 1/2*0.428 = 0.214Kgmole/hr. = 66.34Kg/hr.

$$\text{Yield} = \frac{\text{product H3PO4}}{\text{feed Ca3(PO4)2}}$$

$Ca_3(PO_4)_2$ feed = 0.428/0.95 = 0.45Kgmole/hr. = 139.5Kg/hr.

Feed steam = 139.5/0.75 = 186 Kg/hr.

CaF_2 = 0.2 * 186 = 37.2 Kg/hr.

SiO_2= 0.05 *186 = 9.3 Kg/hr.

Phosphate Rock, Kg/hr.

$Ca_3(PO_4)_2$	139.5
CaF_2	37.2
SiO_2	9.3

2.1.1 Martial Balance on Mill: -

Dilute phosphoric acid stream

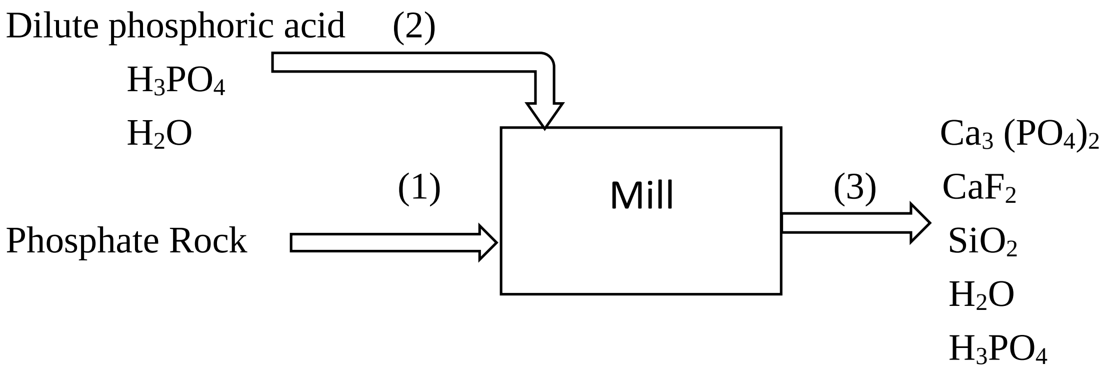
Phosphoric acid	15% wt.
H_2O	85% wt.

Assume 50 Kg of dilute H_3PO_4 /100 Kg Phosphate Rock

186* 50/100 = 93 Kg /hr. (H_2O+H_3PO_4 dilute)

H_3PO_4 = 14 Kg/hr.

H_2O = 79 Kg /hr.



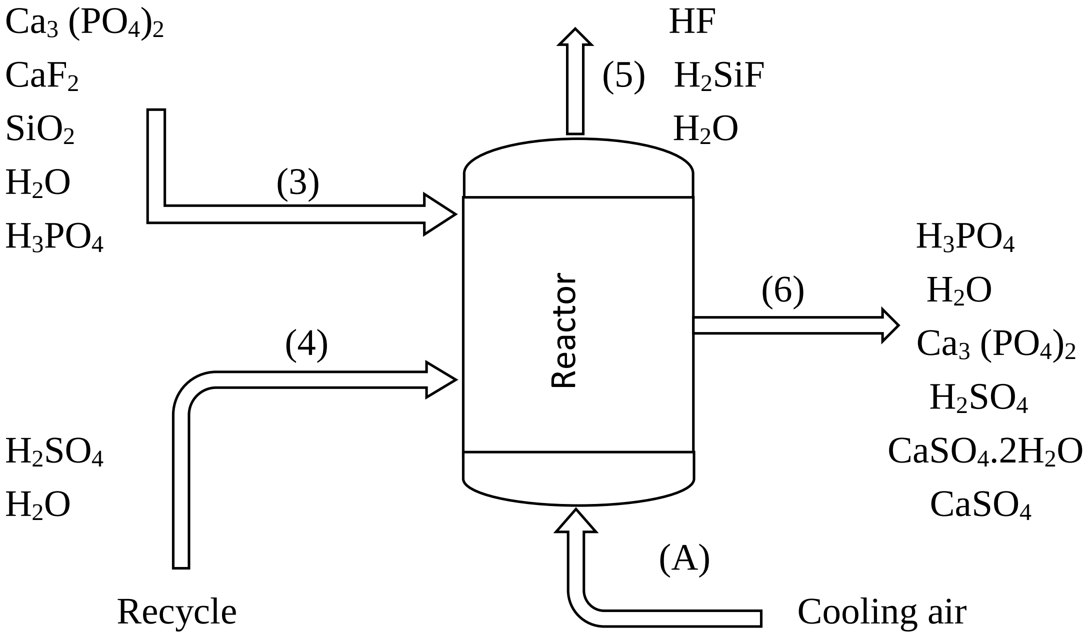
Composition	Stream 1	Stream 2	Steam 3
$Ca_3(PO_4)_2$	139.5		139.5
CaF_2	37.2		37.2
SiO_2	9.3		9.3
H_2O		79	79
H_3PO_4		14	14

Mass in = 139.5 + 37.2 +9.3 +79 +14 = 279 Kg /hr.

Mass out= 139.5 +37.2 +9.3+79+14 = 279 Kg/hr.

Mass in = Mass out = 279 Kg/hr.

2.1.2 Material Balance on Reactor:



Reactions:

$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \rightarrow 2H_3PO_4 + 3(CaSO_4.2H_2O)$ ….. (1)

$CaF_2 + H_2SO_4 \rightarrow 2HF + CaSO_4$ …… (2)

$6HF + SiO_2 \rightarrow H_2SiF_6 + 2H_2O$ ….. (3)

H_3PO_4 product = 0.428 Kg mole/hr.

<u>From Reaction (1)</u>

$Ca_3(PO_4)_2$ reacted = 0.428/2 =0.214 Kg mole /hr.

Out of $Ca_3(PO_4)_2$ =in – req.

= 0.45 – 0.214 = 0.236 Kg mole/hr.

= 73.16Kg/hr.

Reacted of H_2SO_4 = 3/2 * 0.428 = 0.642 Kg mole/hr.

Reacted of H_2O = 6/2 * 0.428 = 1.284 Kg mole/hr.

Product of $CaSO_4.2H_2O$ = 3/2 * 0.428 = 0.642

<u>From Reaction (2)</u>

In CaF_2 = req. = 0.48 Kg mole/hr.

Reacted of H_2SO_4 = 0.48 Kg mole/hr.

Product of HF = 2 * 0.48 = 0.96 Kg mole/hr.

Product of $CaSO_4$ = 0.48 Kg mole/hr.

<u>From Reaction (3)</u>

In SiO_2= req.= 0.155kg.mole/hr.

Reacted of HF= 6*0.155= 0.93kg.mole/hr.

Product of H_2SiF_6= 0.155 kg.mole/hr.

Product of H_2O= 2*0.155= 0.31 kg.mole/hr.

Reacted of H_2SO_4= 0.48 + 0.642= 1.122 kg.mole/hr. = 109.96 kg /hr.

Excess%H_2SO_4= 0.15= $\frac{\text{In}-\text{req.}}{\text{req.}} = \frac{\text{In}-109.96}{109.96}$

In= 126.45 kg /hr.

Out of H_2SO_4= In – req. = 126.45 – 109.96= 16.45 kg/hr.

Stream (4)

$\frac{126.45}{0.94} = 134.53$ kg/hr

H_2SO_4	126.45 kg/hr.
H_2O	8.08 kg/hr.

Stream (6)

In + gen. = req. + out

H_3PO_4= 42 + 14= 56 kg/hr.

Out H_2O= In – req. = 79 + 8.08 + 5.58– 23.1

Out of H_2O= 69.56 kg.

Assume 45% from H_2O is vaporized

H_2O= 38.258kg/hr.

$Ca_3 (PO_4)_2$= 73.16 kg/hr.

$CaSO_2.2H_2O$= 101.4 kg/hr.

H_2SO_4= 16.45 kg/hr.

$CaSO_4$= 65.28 kg/hr.

Stream (5)

H_2SiF_6= 22.32 kg/hr.

H_2O= 31.302 kg/hr.

HF= In–req.= 0.96–0.93= 0.03kg.mole/hr.

HF= 0.6 kg/hr.

Cooling Air

79% N_2

21% O_2

Comp.	Stream(3)	Stream(4)	Stream(5)	Stream(6)	Stream(A)
$Ca_3(PO_4)_2$	139.5			73.16	
CaF_2	37.2				
SiO_2	9.3				
H_2O	79	8.08	31.302	38.258	
H_3PO_4	14			56	
H_2SO_4		126.45		16.45	
$CaSO_4.2H_2O$				110.4	
$CaSO_4$				65.28	
H_2SiF_6			22.32		
HF			0.6		
O_2			21		21
N_2			79		79

Mass in= 139.5+ 37.2+ 9.3 +79 +14+ 8.08+ 126.45+ 21+ 79

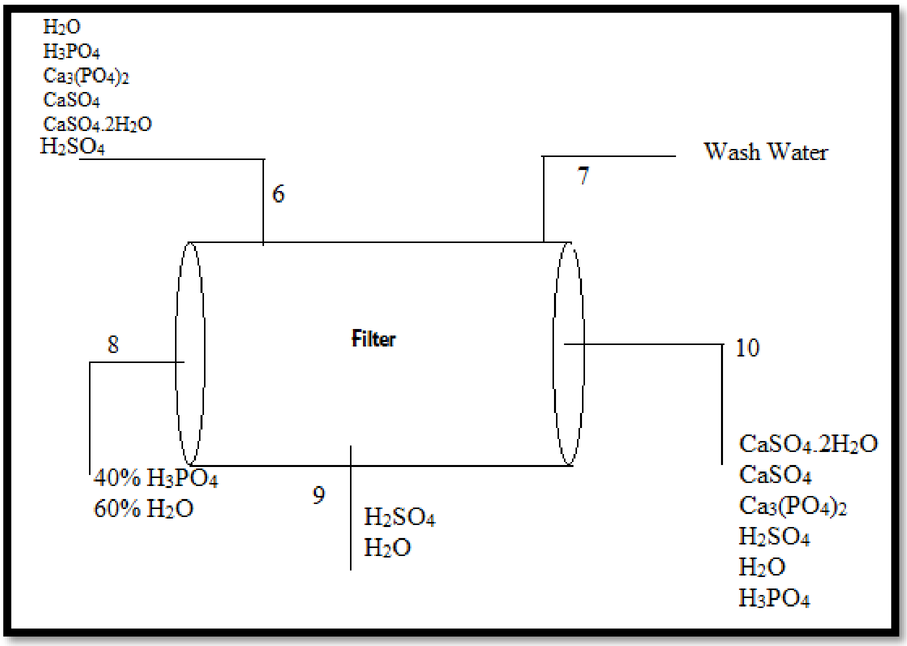
= 513.77 kg/hr.

Mass out= 31.302+ 22.32+ 0.6+ 21+79+73.16+ 38.258+ 56+ 16.45+110.4+65.28

= 513.77 kg/hr.

∴ Mass in= Mass out= 513.77 kg/hr.

2.1.3 Material Balance on Filter



Stream (8)

H_3PO_4= 42 kg/hr.

Stream (8) = $\frac{42}{0.4} = 105\ \frac{\text{kg}}{\text{hr}}$

H_2O= 0.6* 105= 63 kg/hr.

H_2SO_4 in stream (10) = 16.45* $\frac{1.5}{100} = 0.247$ kg/hr.

H_2SO_4 in stream (9) = 16.45–0.247= 16 .2 kg/hr.

H_2O in stream (10) = 38.258* $\frac{1.5}{100} = 0.574$ kg/hr.

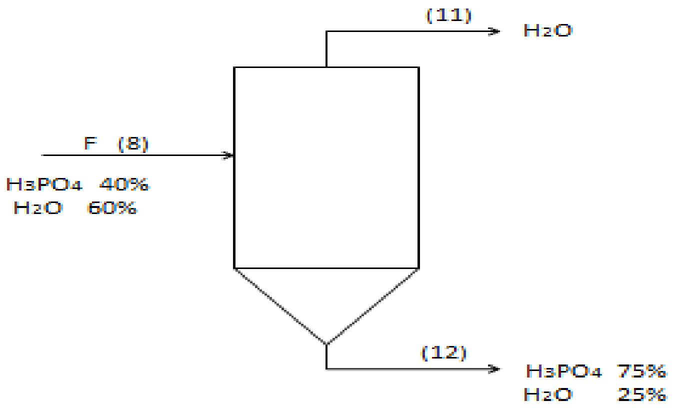
H_2O in stream (9) = 1.035 kg/hr.

H_2O in stream (7) = 0.574+ 1.035+ 63– 38.258 =26.35kg/hr.

H_3PO_4 in stream (10) = 14 kg/hr.

Comp.	**Stream(6)**	**Stream(7)**	**Stream(8)**	**Stream(9)**	**Stream(10)**
$Ca_3(PO_4)_2$	73.16				73.16
H_2O	38.258	26.35	63	1.035	0.574
H_2SO_4	16.45			16.2	0.247
H_3PO_4	56		42		14
$CaSO_4.2H_2O$	110.4				110.4
$CaSO_4$	65.28				65.28

2.1.4 Material Balance on Evaporator: -



M.B on H_3PO_4

IN = Out

0.4 F = 42 $\longrightarrow$ F = 105 Kg/hr.

:. Stream (8)

H_3PO_4 = 42 Kg/hr.

H_2O = 63 Kg/hr.

:. H_2O in stream (11) = 63 – 14 = 49 Kg/hr.

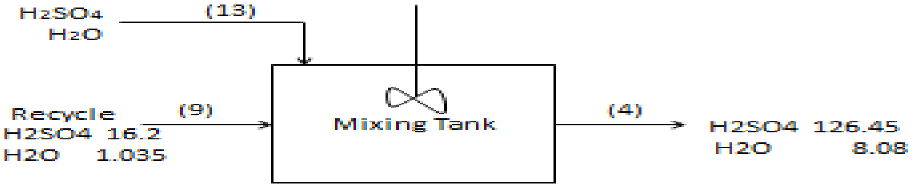
Component	Stream (8)	Stream (11)	Stream (12)
H_3PO_4	42		42
H_2O	63	49	14

Mass In = 42 + 63 = 105 Kg/hr.

Mass Out = 49 + 42 + 14 = 105 Kg/hr.

:. Mass In = Mass Out = 105 Kg/hr.

2.1.5 Material Balance on Mixing Tank.



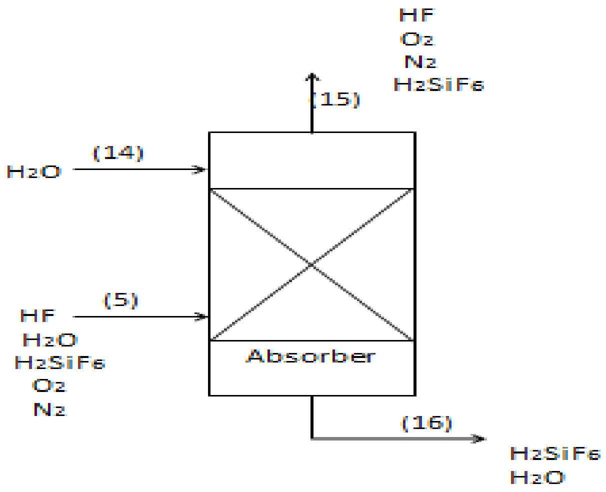
Component	Stream (9)	Stream (13)	Stream (4)
H_2SO_4	16.2	110.25	126.45
H_2O	1.035	7.045	8.08

Mass In = 16.2 + 1.035 + 110.25 + 7.045 = 134.53 Kg/hr.

Mass Out = 126.45 + 8.08 = 134.53 Kg/hr.

:. Mass In = Mass Out = 134.53 Kg/hr.

2.1.6 Material Balance on Absorber.



Recovery = 99%

$Y_2 = Y_1$ (1- recovery) = 0.202 (1- 0.99) = 0.002

$(L_s / G_s)_{min} = (y_1 - y_2)/(y_1/m) = m[1- (y_2/y_1)]$

$m = (P_{H_2SiF_6} / P_T) = (418_{mmHg}/760_{mmHg}) = 0.55$

$(L_s/G_s)_{min}$ = 0.55 [1-(0.002/0.202)] = 0.544

$(L_s/G_s)_{actual} = 1.5(L_s/G_s)_{min}$ = 0.816

Gs = 100.6 Kg/hr.

Ls = 82 Kg/hr.

Comp.	Stream (5)	Stream (14)	Stream (15)	Stream (16)
HF	0.6		0.6	
H_2O	31.302	82		113.3
H_2SiF_6	22.32		0.223	22.097
O_2	21		21	
N_2	79		79	

Mass In = 0.6 + 31.302 + 22.32 + 21 + 79 + 82

= 236.22 Kg/hr.

Mass Out = 0.6 + 0.223 + 21 + 79 + 113.3 + 22.097 = 236.22 Kg/hr.

Mass In = Mass Out = 236.22

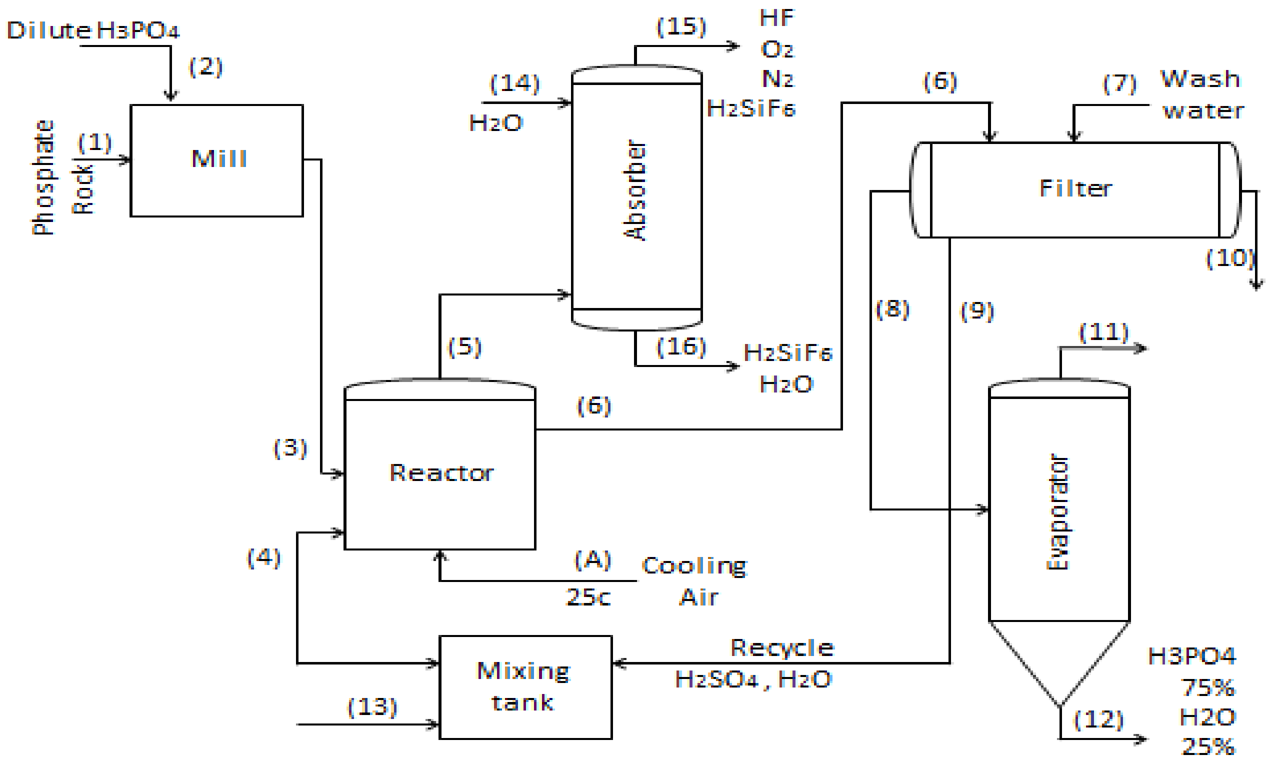
2.1.7 Over all Material Balance:

Stream Input		Stream Output	
Stream (1)	186	Stream (10)	263.66
Stream (2)	93	Stream (11)	49
Stream (A)	100	Stream (12)	56
Stream (13)	117.295	Stream (15)	100.823
Stream (7)	26.35	Stream (16)	135.397
Stream (14)	82		

Mass In = 604.88 & Mass Out = 604.88

:. Mass In = Mass Out = 604.88 Kg/hr.

2.1.8 General Diagram of Process.



2.2 Energy Balance

Table (1) Heat capacity of gases kJ/kg mole

Component	A	B	C	D
H_2O	32.243	$1.923*10^{-3}$	$1.055*10^{-5}$	$-3.596*10^{-9}$
O_2	28.106	$-3.68*10^{-6}$	$1.745*10^{-5}$	$-1.065*10^{-8}$
N_2	31.15	$-1.356*10^{-2}$	$2.679*10^{-5}$	$-1.168*10^{-8}$
HF	29.061	$6.611*10^{-4}$	$-2.032*10^{-6}$	$2.503*10^{-9}$

$\Delta H = m \int Cp \, \Delta T$

$\Delta H = m/M.wt \, [\int (A+BT+CT^2+DT^3) \, dT]$

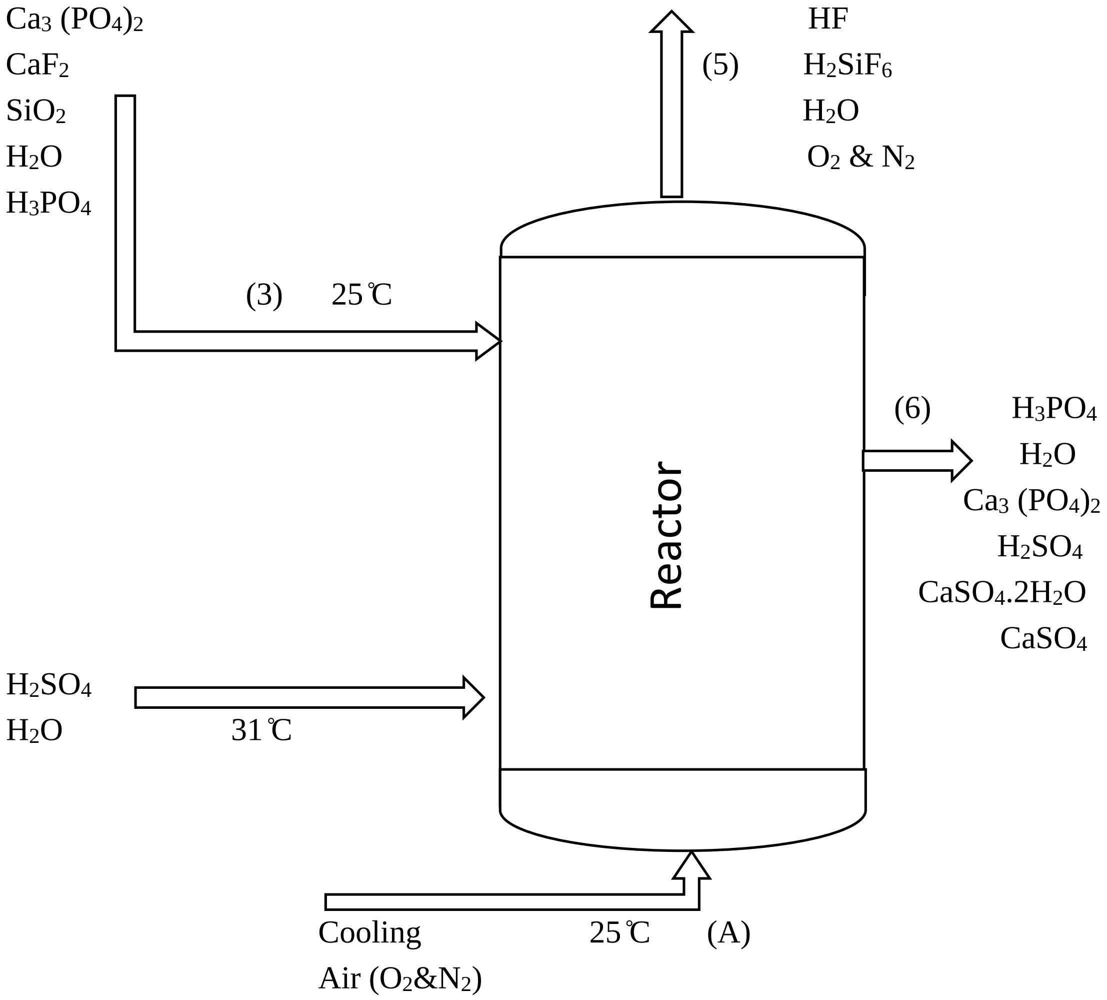
$\Delta H = n \, [A \, (T_2 - T_1) + B/2 \, (T_2^2 - T_1^2) + C/3 \, (T_2^3 - T_1^3) + D/4 \, (T_2^4 - T_1^4)$

$Cp = A + BT + CT^2 + DT^3$ KJ/Kg mole. K

Table (2)

Comp.	TBP ̊C	Cp KJ/kg mole .C	ΔHF Kcal/kg mole. K
HF	20	47.3	-64.2
$Ca_3(PO_4)_2$		262.9	-988.9
H_2SiF_6		236.1	-267.8
$CaSO_4.2H_2O$		315.5	-483.06
$CaSO_4$		164.9	-346.67
CaF_2		92.1	-286.5
SiO_2	223	74.5	-203.4
H_2SO_4		167.4	-193.91
H_3PO_4		185.4	-306.2
H_2O	100	75.3	-68.32

2.2.1 Energy Balance on Reactor



Reaction:

$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \rightarrow 2H_3PO_4 + 3(CaSO_4.2H_2O)$

$CaF_2 + H_2SO_4 \rightarrow 2HF + CaSO_4$

$6HF + SiO_2 \rightarrow H_2SiF_6 + 2H_2O$

Operating Conditions:

T = 75-80℃ , P = 1atm. , Liquid Phase

Heat in = heat out

M Cp $\Delta T_{Ca_3(PO_4)_2}$ + m Cp ΔT_{CaF_2} + m Cp ΔT_{SiO_2}+ m Cp ΔT_{H_2O} + m Cp$\Delta T_{H_3PO_4}$ + m Cp $\Delta T_{H_2SO_4}$

+ m Cp ΔT_{H_2O} + m Cp ΔT_{O_2} + m Cp ΔT_{H_2} + $\Delta Hr°$ = m Cp ΔT_{HF} + m Cp $\Delta T_{H_2SiF_6}$ + mCpΔT_{H_2O}

+ m Cp ΔT_{O_2} + m Cp ΔT_{N_2} + m Cp $\Delta T_{H_3PO_4}$ + m Cp ΔT_{H_2O} + m Cp $\Delta T_{Ca_3(PO_4)_2}$ + m Cp $\Delta T_{H_2SO_4}$ +

m Cp ΔT_{CaSO_4} + m Cp $\Delta T_{CaSO_4.2H_2O}$ + Q + m λ_{H_2O}

$T_{ref..}$ = 25C°

M Cp $_{HF}$ ΔT + m Cp $_{H_2SiF_6}$ ΔT + m Cp $_{H_2O}$ ΔT + m Cp $_{O_2}$ ΔT + m Cp $_{N_2}$$\Delta T$ + m Cp $_{H_3PO_4}$ ΔT + m Cp $_{H_2O}$ ΔT + m Cp $_{Ca3(PO4)_2}$$\Delta T$+ m Cp $_{H_2SO_4}$$\Delta T$+ m Cp $_{CaSO_4}$$\Delta T$+ m Cp $_{CaSO_{4}.2H_2O}$$\Delta T$+ Q + m λ_{H_2O} = $\Delta Hr°$+ m Cp $_{H_2O}$ ΔT + m Cp $_{H_2SO_4}$$\Delta T$

We have three reactions:

$\Delta Hr°$= –693.5 Cal/g.mole = –2901.5 KJ/Kg.mole

$\Delta Hr_2°$= 5.34 Cal/g.mole = 22.343 KJ/Kg.mole

$\Delta Hr_3°$= 184.16 Cal/g.mole = 770.5 KJ/Kg.mole

$\Delta Hr°$= 0.214* (–2901.5) + 0.48* 22.342 + 0.155* 770.5

$\Delta Hr°$= – 490.765 KJ/Kg

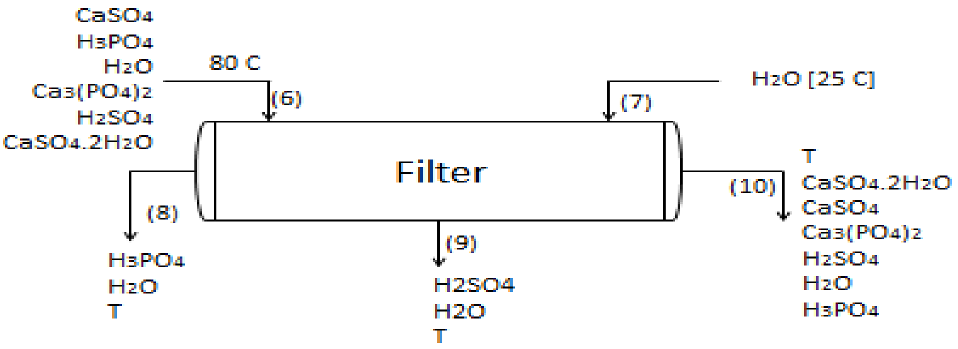
$$\frac{0.6}{20} * [29.061(353-298) + \frac{6.611 * x10^{-4}}{2} * (353^2 - 298^2) - \frac{2.032 * 10^{-6}}{3} * (353^3 - 298^3) + \frac{2.503 * 10^{-9}}{4} * (353^4 - 298^4)\ ] + \frac{31.302}{18} [32.243\ (353 - 298) + \frac{1.923 * 10^{-3}}{2} * (353^2 - 298^2) + \frac{1.055 * 10^{-5}}{3}\ (353^3 - 298^3) - \frac{3.596 * 10^{-9}}{4} * (353^4 - 298^4)] + \frac{22.32}{144} * [\ 0.67 * (353 - 298)]$$

$$+ \frac{21}{32} [28.106\ (353 - 298)\ - \frac{3.68 * 10^{-6}}{2}\ (353^2 - 298^2) + \frac{1.745 * 10^{-5}}{3}\ (353^3 - 298^3) - \frac{1.065 * 10^{-8}}{4}\ (353^4 - 298^4)] + \frac{79}{28} * [31.15\ (353 - 298) - \frac{1.356 * 10^{-2}}{2}\ (353^2 - 298^2) + \frac{2.679 * 10^{-5}}{3}\ (353^3 - 298^3) - \frac{1.108 * 10^{-8}}{4}\ (353^4 -$$

$$298^4) + \frac{56}{98} * 185.4\,(353 - 298) + \frac{38.258}{18} * 75.3\,(353 - 298) + \frac{73.16}{301} * 262.9\,(353 - 298) + \frac{16.45}{98} * 167.4\,(353 - 298) + \frac{16.45}{98} * 167.4\,(353 - 298) + \frac{110.4}{172} * 164.91\,(353 - 298) + \frac{65.28}{136} * 236.1\,(353 - 298)\Big] + Q + \frac{31.302}{18} * 40683 = -490.765 + \frac{8.08}{18} * 75.3\,(304 - 298) + \frac{126.45}{98} * 167.4\,(304 - 298)$$

Q = – 263273.7493 KJ/hr.

2.2.2 Energy Balance on Filter



Heat IN = Heat Out

$m\,Cp_{H_3PO_4}\,dT + m\,Cp_{H_2O}\,dT + m\,Cp_{Ca_3(PO_4)_2}\,dT + m\,Cp_{H_2SO_4}\,dT + m\,Cp_{CaSO_4.2H_2O}\,dT + m\,Cp_{CaSO_4}\,dT + m\,Cp_{H_2O}\,dT = m\,Cp_{H_3PO_4}\,dT + m\,Cp_{H_2O}\,dT + m\,Cp_{H_2SO_4}\,dT + m\,Cp_{H_2O}\,dT + m\,Cp_{CaSO_4.2H_2O}\,dT + m\,Cp_{CaSO_4}\,dT + m\,Cp_{Ca_3(PO_4)_2}\,dT + m\,Cp_{H_2SO_4}\,dT + m\,Cp_{H_2O}\,dT + m\,Cp_{H_3PO_4}\,dT$

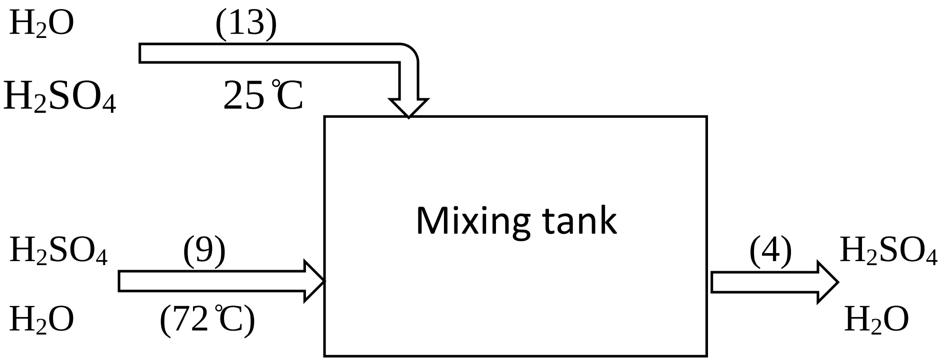
$T_{ref} = T$

$m\,Cp_{H_3PO_4}\,dT + m\,Cp_{H_2O}\,dT + m\,Cp_{Ca_3(PO_4)_2}\,dT + m\,Cp_{H_2SO_4}\,dT + m\,Cp_{CaSO_4.2H_2O}\,dT + m\,Cp_{CaSO_4}\,dT + m\,Cp_{H_2O}\,dT = 0$

>> $[(56/98) * 185.4 + (38.258/18) * 75.3 + (73.16/310) * 262.9 + (16.45/98) * 167.4 + (110.4/172) * 315.5 + (65.28/136) * 164.9] \{353 - T\} + [26.35 * 4.184] \{298 - T\} = 0$

$T = 72$ C°

2.2.3 Energy Balance on mixing tank:



Heat in = Heat out

$m\ Cp_{H2O}\ \Delta T + m\ Cp_{H2SO4}\ \Delta T + m\ Cp_{H2O}\ \Delta T + m\ Cp_{H2SO4}\ \Delta T = m\ Cp_{H2SO4}\ \Delta T + m\ Cp_{H2O}\ \Delta T$

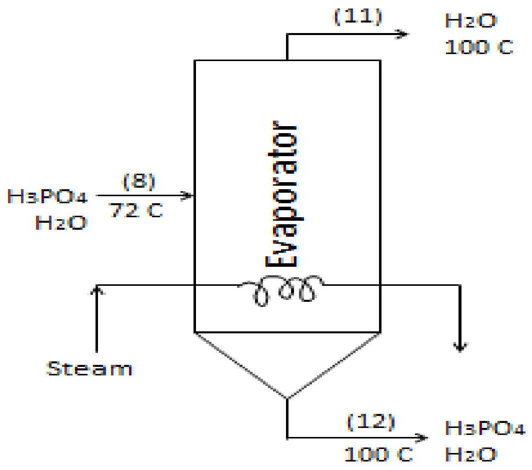
$T_{ref.} = T$

$m\ Cp_{H2O}\ \Delta T + m\ Cp_{H2SO4}\ \Delta T + m\ Cp_{H2O}\ \Delta T + m\ Cp_{H2SO4}\ \Delta T = 0$

$[1.035/18 * 75.3\ (345 - T) + 16.2/98 * 167.4\ (345 - T)] + [7.045/18 * 75.3\ (298 - T) + 110.25/98 * 167.4\ (298 - T)] = 0$

$T = 31$℃ $= 304$K

2.2.4 Energy Balance on Evaporator



Heat In = Heat Out

m $Cp_{H_3PO_4}$ dT + m Cp_{H_2O} dT + Q =

m Cp_{H_2O} dT + m λ_{H_2O} + m $Cp_{H_3PO_4}$ dT + m Cp_{H_2O} dT

T_{ref} = 72 C°

Q = m $Cp_{H_3PO_4}$ dT + m Cp_{H_2O} dT + m λ_{H_2O} + m Cp_{H_2O} dT

Q = (42/98)*185.4*(373-345) + (14/18)*{4.184*(373-345) + (49/18)*32.243*(373-345) + (1.9238*10^-3/2)*(373^2-345^2) + (1.055*10^-5/3)*(373^3-345^3) – (3.596 * 10^-5 /4) *(373^4-345^4)} + (49/18)* 40683

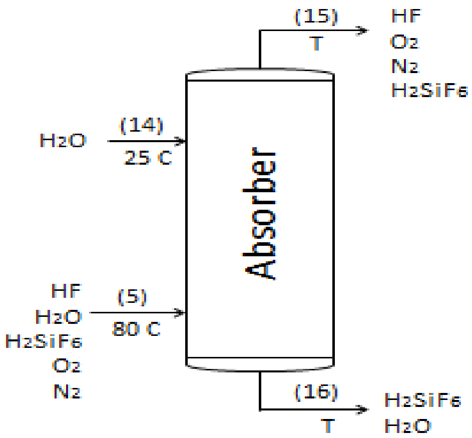
Q = 115665.33 KJ/hr.

This heat is supplied by sat. Steam 150 C°

λ_s = 2113 KJ/Kg.

m_s = Q/λ_s = 115665.33/2113 = 55 Kg/hr.

2.2.5 Energy Balance on Absorber



Heat In = Heat Out

$m\,Cp_{H_2O}\,dT + m\,Cp_{HF}\,dT + m\,Cp_{H_2O}\,dT + m\,\lambda_{H_2O} + m\,Cp_{H_2SiF_6}\,dT + m\,Cp_{O_2}\,dT +$
$m\,Cp_{N_2}\,dT = m\,Cp_{HF}\,dT + m\,Cp_{O_2}\,dT + m\,Cp_{N_2}\,dT + m\,Cp_{H_2SiF_6}\,dT + m\,Cp_{H_2O}\,dT + m\,Cp_{H_2SiF_6}\,dT$

$T_{ref} = T$

$m\,Cp_{H_2O}\,dT + m\,Cp_{HF}\,dT + m\,Cp_{H_2O}\,dT + m\,\lambda_{H_2O} + m\,Cp_{O_2}\,dT + m\,Cp_{N_2}\,dT + m\,Cp_{H_2SiF_6}\,dT = 0$

(82/18)*75.3*(298-T) + (0.6/20)*{29.06(353-T) + (6.611*10^-4/2)*(353^2-T^2) + (2.032*10^-6/3)*(353^3 – T^3) + (2.503*10^-9/4)*(353^4-T^4)} + (31.302/18)*{32.243(353-T) + (1.923*10^-3/2)* (353^2-T^2) + (1.055*10^-5/3)* (353^3 – T^3)

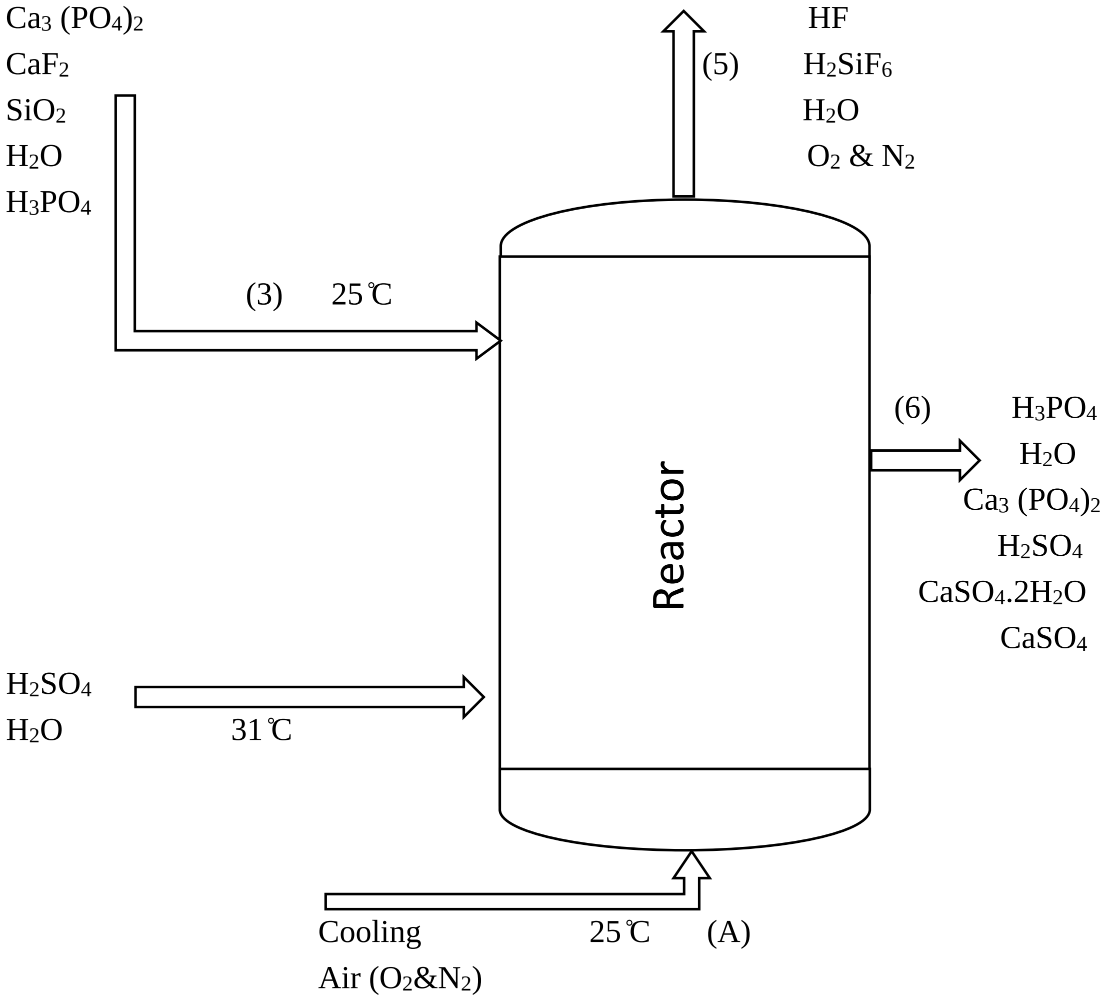
$-3.596 * 10^{-9}/4\,(353^4 – T^4)] + 31.302/18 * 40683 + 21/82\,[28.106(353-T) – 3.68*10^{-6}/2\,(353^2 – T^2) + 1.745*10^{-5}/3\,(353^3 – T^{3)} – 1.065*10^{-8}/4\,(353^4 – T^4)] + 79/28\,[31.15\,(353-T) – 1.35*10^{-2}/2\,(353^2 – T^2) + 2.679*10^{-5}/3\,(353^3 - T^3) – 1.108*10^{-8}/4\,(353^4 – T^4)] + 22.32/144[\,0.67 *(353 – T)] = 0$

By trial and error

T = 60℃

CHAPTER THREE
EQUIPMENT DESIGN

3.1 Equipment Design On Reactor



Reactions:

$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \longrightarrow 2H_3PO_4 +3(CaSO4.2H_2O)$

$CaF_2 + H_2SO_4 \longrightarrow 2HF + CaSO_4$

$6HF + SiO_2 \longrightarrow H_2SiF_6 + 2H_2O$

Operating Conditions

T= 75 – 80C°

P= 1atm

Liquid Phase

$\tau = 4 - 8hr$

We have three chemical reactions.

The reaction is exothermic and temperature is maintained constant by passing air across the reactor.

For continuous stirred tank reactor

$$\frac{V}{F_{A\circ}} = \frac{X_A}{(-r_A)}$$

Where:

V= volume of reactor (m^3)

V° = volumetric flow rate (m^3/time)

FA° = molar rate of A (mole of A/time)

XA= conversion

τ = space time

$$\tau = \frac{V}{\nu_\circ} = \frac{C_{A\circ} V}{F_{A\circ}} = \frac{X_A C_{A\circ}}{(-r_A)} = \frac{C_{A_\circ} - C_A}{(-r_A)}$$

From volume (6)

Comp.	Density Kg/m^3
H_2SO_4	1153
H_2O	988
$Ca_3(PO_4)_2$	1310
CaF_2	1755
SiO_2	1042

$\rho_{mix} = \sum X_i \rho_i$

Or

$V^o = \frac{mass}{\rho}$

Where

V^o = volumetric flow rate (m^3/hr).

ρ = density of component (kg/m^3).

Mass= mass flow rate of component (kg/hr.).

To find $\rho_{mix} \xrightarrow{\qquad} V= \frac{mass}{\rho}$

Comp.	Density (kg/m^3)	Mass (kg/hr)	V^o (m^3/hr)
H_2SO_4	1153	126.45	0.11
H_2O	988	8.08	0.008
$Ca_3(PO_4)_3$	1310	139.5	0.1065
CaF_2	1755	37.2	0.0212
SiO_2	1042	9.3	0.009

$V^o = 0.255 m^3/hr.$

Calculation the Volume of Reactor:

Space time = $\frac{volume\ of\ reactor\ (\mathrm{m3})}{volumetric\ flow\ rate\ of\ reactor\left(\frac{\mathrm{m3}}{\mathrm{hr}}\right)}$

$V^o = 0.255 m^3/hr.$

$\tau = 4 - 8hr$

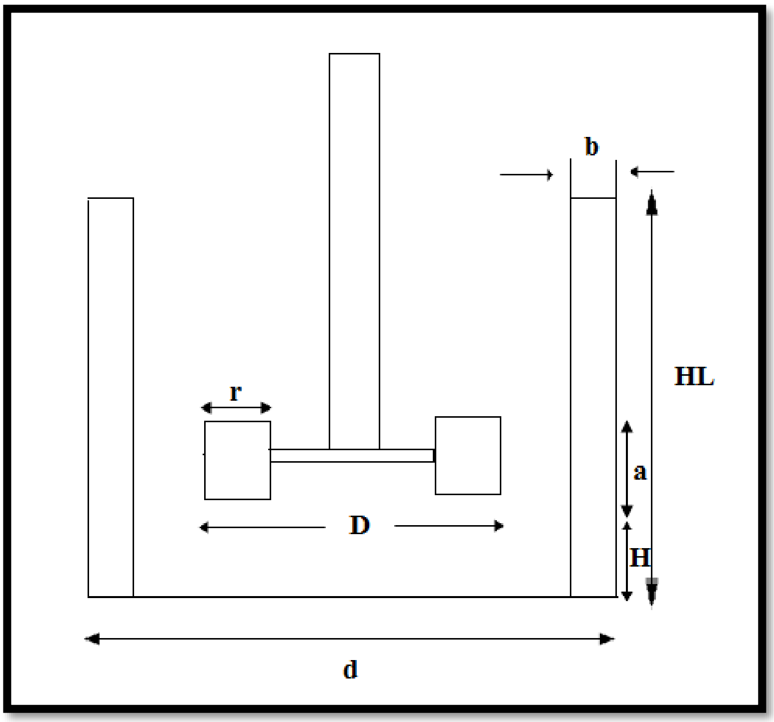
take $\tau = 6$ hr.

$\therefore \tau = \frac{V}{V^o} \xrightarrow{\qquad\qquad} 6 = \frac{V}{0.255}$

$V= 1.53\ m^3$

$\therefore$ Volume of reactor = $1.53\ m^3$

3.1.1 Mechanical design for mixing vessel:



$D = \frac{d}{3}$

$H = \frac{d}{3}$

$a = \frac{d}{5}$

$r = \frac{d}{4}$

$HL = d$

$b = \frac{d}{10}$

Volume of feed = $1.53 m^3$

Volume of feed in the reactor = $V_f + 10\% V_f$

$= 1.53 + 0.1 * 1.53$

$= 1.68 m^3$

Volume of feed = $\frac{\pi}{4} d^2 HL$

HL = d

Volume of feed = $\frac{\pi}{4}$ d^3

1.53 = $\frac{\pi}{4}$ d^3 $\xrightarrow{\qquad\qquad}$ d = 1.249m

D = 0.43m

H = 0.43m

a = 0.258m

r = 0.323m

HL = 1.249m

b = 0.13m

Height of reactor = 1.5m

Volume of reactor = $\frac{\pi}{4}$ d2 H = $\frac{\pi}{4}$ (1.249)^2 * 1.5

$\simeq$ 2m^3

Agitator Type (3)

We can use propellers (three blades pitch diameter)

Agitator tip speed = π DN

For medium agitation = 3.3 – 4.1 m/s

$\therefore$ Agitation tip speed = 3.7m/s

$\therefore$ N= $\frac{3.7}{\pi D}$ = $\frac{3.7}{\pi * 0.43}$ = 2.74rev. /sec

The power required for mixing

$NP = C\ (NRe)m\ (NFr)n$

Where:

N_P = power number = $\frac{P}{N^3\ D^5\ P}$

$NRe = \frac{N\ D^2\ P}{\mu}$

NFr = Froude Number = $\frac{N^2 D}{P}$

$\rho_{\text{mix}} = 1283.8 Kg/m^3$

$\mu_{\text{mix}} = 0.065 mNs/m^2$

$N\ Re = \frac{1283.8 * (0.43)^2 * 2.74}{0.065 * 10^{-3}} = 10 * 10^6$

$\therefore$ Turbulent flow

$P = K_2\ N_3\ D_5 \rho$

For propeller

K2 = 0.32

P = 0.32 * (2.74)2 * (0.43)5 * 1283.8

P = 125W

Mechanical Design:

Thickness of the vessel can be obtained using the following equation [7]

$e = \frac{Pi\ Di}{2fJ - Pi} + C$

Where:

e = thickness of vessel (mm)

Pi = internal pressure (N/mm^2)

Di = internal diameter (mm)

f = design stress

J= joint factor = 0.8 – 085

C = corrosion allowance = 2mm

> The stainless steels are the most frequently used corrosion resistant materials in the chemical industry.
> To impart corrosion resistance, the chromium content must be above 12 percent,
> and the higher the chromium content, the more resistant is the alloy to corrosion in oxidizing conditions. Nickel is added to improve the corrosion resistance in non-oxidizing environments

typical design stress of stainless steel materials (18Cr/ 8Ni)

at $80C^o = 160 N/mm^2$ (4)

Design pressure = 10% above operating pressure

= 1atm + 0.1*1atm = 1.1atm

= $0.11 N/mm^2$

$$e = \frac{0.11 * 1.249 * 10^3}{2 * 160 * 0.8 - 0.11} + 2 = 2.6mm \simeq 3mm$$

Weight of Vessel (5): -

$W_V = C_V \pi \rho_m D_m g (H_V + 0.8D_m) t * 10^{-3}$

Where:

WV = total weight of the shell, excluding internal fittings, … etc.

CV = a factor to account for the weight of nozzle manways, internal supports…etc.

Can be taken as 1.15

HV = height or length between tangent lines (the length of the cylindrical section m).

t = wall thickness

ρ_m = density of vessel material

D_m = mean diameter of vessel = $(Di + t * 10^{-3})$ m

For steel vessel equation reduce to

$WV = 240\ CV\ D_m\ (HV + 0.8D_m)\ t$

$WV = 240 * 1.15 * D_m\ (1.5 + 0.8D_m) * 2.6 * 10^{-3}$

$D_m = (1.29 + 2.6 * 10^{-3}) = 1.293$m

$WV = 240 * 1.15 * 1.293\ (1.5 + 0.8 * 1.293)\ 2.6 * 10^{-3}$

WV = 2.35KN ≃2.5KN

Weight of Insulation:

Thickness = 75mm

Mineral wool density = 130 kg/m^3

Approximate volume of insulation = π d HV t

$= \pi * 1.29*1.5*75 * 10^{-3} = 0.5m^3$

Weight = 0.5 * 130 * 9.81 = 638N = 0.638KN

Double this to allow for fittings, etc. = 1.3KN= 1.5KN

Weight of vessel filled with water = $\frac{\pi}{4} D^2 L \rho_{H2O} * g = 20KN$

Weight of two man = 75 * 2* 9.81= 1.5kN

Total Weight:

Vessel = 2.5 kN

Insulation = 1.5

Vessel with water = 20

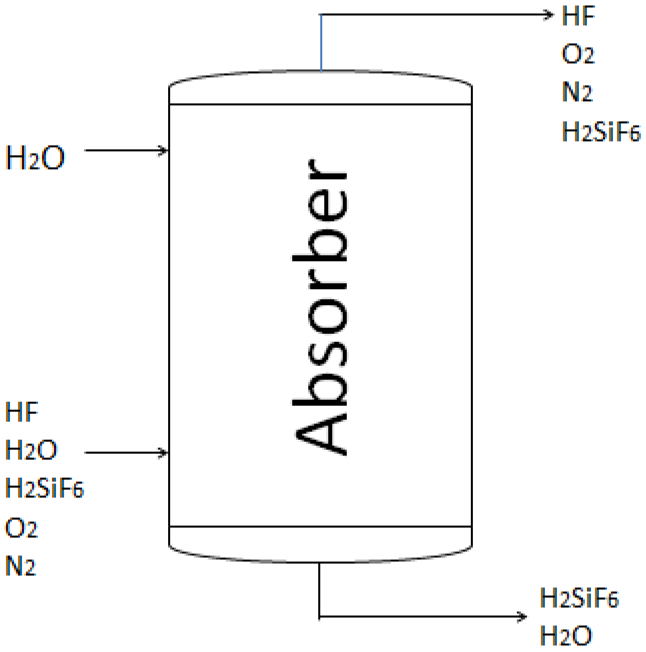
Two men = 1.5

Wt = WV + WW + Wi + W man

= 2.5 + 20 + 1.5 + 1.5

= 25.5kN

3.2 Absorber Design



Recovery = 99%

y_2 = y1 (1- recovery)

y_2 = 0.18 (1- 0.99) = 0.002

(Ls/Gs) min =(Y_1- Y_2) / (Y_1/m) = m (1- Y_2/Y_1)

Ls = (H_2O) Solvent quantity

G_s = Gases

m = PH_2SiF_6 /PT = 0.55

(Ls/Gs) min = 0.55 (1 – 0.002/0.18) = 0.544

(Ls/Gs) actual = 1.5 (Ls/Gs) min

= 1.5 * 0.544 = 0.816

Gs = 100.6 Kg/hr.

Ls = 82 kg/hr

Y = mx

Y = 0.55 x

The equilibrium data is linear

3.2.1 Calculation The Tower Height

$$z = \frac{Gs^{-}}{\mathrm{KoG}} . \mathrm{a} . \mathrm{Pt} \ \int_{y2}^{y1} \frac{dy}{y - y*}$$

Z = HOG. NOG

$$HoG = \frac{GS}{KoG.a.PT}$$

$$NoG = \int_{y2}^{y1} \frac{dy}{y - y*}$$

Where;

Z= The Tower height

$HOG = \ height \ of \ transfer \ unit$

NOG= Number of transfer unit

Y = 0.55 X

$$NoG = \frac{1}{1 - \phi} \ln \left[(1 - \phi) \frac{y1}{y2} + \phi \right]$$

$$\phi = \frac{m \ Gs}{Ls}$$

$$\phi = \frac{0.55 * 100.6}{82} = 0.675$$

$$NoG = \frac{1}{1 - 0.675} \ln \left[(1 - 0.675) \frac{0.18}{0.002} + 0.675 \right]$$

NOG = 10.5 m.

3.2.2 Calculation the Area of column

	Vapor	Liquid
Flow	1.2575	4.56 kg mole /hr.
Density	1.0 (6)	998 kg mole /hr. (7)
M.Wt	29	18 kg mole /hr.
Viscosity	$0.02*10^{-3}$ (8)	$0.89*10^{-3}$ N.S/m^2 (9)

Select 13 mm ceramic INTA lox Saddles

Packing factor (F_p) = 660

$$Flv = \frac{L\text{w}*}{V\text{w}*}\sqrt{\frac{\rho v}{\rho L}}$$

$$= \frac{4.56}{3.47}\sqrt{\frac{1.03}{998}} = 0.043 = Flv$$

Design for pressure drop of 20mm H_2O / m Packing (10)

$K_4 = 0.77$

At flooding $K_4 = 3.25$

Percentage flooding = $\sqrt{\frac{0.9}{5}}$

= 48.6 %

$$K4 = \frac{42.9*Fp*(\frac{\mu l}{\rho l})^{0.1}*(Vw*)^2}{\rho v(\rho l-\rho v)} \quad \ldots.[11]$$

$$0.77 = \frac{42.9*92*(\frac{0.406*10^{-3}}{998})^{0.1}(Vw*)^2}{1.03\,(998-1.03)}$$

$V_w* = 0.597$ kg/m^2. S

Column area required = 0.028 / 0.597 = 0.047 m^2

= 24.5 cm ~ 0.25 m

∴ Area = $\pi/4\ d^2 = \pi/4\ (0.25)^2 = 0.049\ m^2$

Packing Size to column = $0.25 / 13*10^{-3} = 19.2$

Percentage Flooding at Selected diameter

= 48.6 * (0.047 /0.049) = 47 %

3.2.3 Estimate of HOG

Use cornrlls method

$DL = 0.143 * 10^{-6}\ m^2/s$ (10)

$Dv = 1.5 * 10^{-5}\ m^2/s$ (10)

$\mu V = 0.02*10^{-3}\ Ns/m^2$

$(Sc)v = \mu v/\rho v\ .Dv = (0.02*10^{-3} / 1.03 * 1.5 * 10^{-5}) = 1.3$

$(Sc)l = \mu l/\rho l\ .Dl = (0.89 *10^{-3} / 998 * 0.143*10^{-6}) = 6.24$

$Lw^* = 1.368 / 0.049*60 = 0.46\ Kg/m^2.s$

at 47% flooding $K_3 = 0.98$ (11)

at 47% flooding $\Psi n = 52$ (12)

at Lw*, $\varphi h = 0.042$ (13)

$HL = 0.305\ \varphi_h\ (Sc)_L^{0.5}\ K_3\ (Z/3.05)^{0.15}$

Where: -

HL= height of liquid phase transfer unit, m

φ_h = HL factor from fig (11.43)

$(Sc)_L$= liquid Schmidt number

K_3= percentage flooding correction factor from (14)

Z= column height, m

H_L = 0.305 * 0.042 * $(6.24)^{0.5}$ *0.98*$(10.5/3.05)^{0.15}$

H_L = 0.038 m

HG= 0.011 Ψn $(Sc)_v^{0.5}$ $(D_c/0.305)^{1.11}$ $(Z/3.05)^{0.33}$ / $(Lw^* f_1f_2f_3)^{0.5}$ ……[14]

Where: -

HG= height of a gas phase transfer unit, m

Ψn = HG factor from fig (11.42)

Dc= column diameter (m)

F1= liquid viscosity correction = $(\mu L/\mu w)^{0.16}$

F2=liquid density correction = $(\rho_w/\rho_L)^{1.25}$

F3= surface tension correction factor = $(\delta w /\delta L)^{0.8}$

As the liquid temperature (25 °c) and the liquid is water

HG= 0.011 * 52 * $(1.3)^{0.5}(0.25/3.05)^{1.11}(10.5/3.05)^{0.33}$ / $(0.46*1.0)^{0.5}$

HG= 0.09 m

HOG= HG+ m(Gm/Lm) HL

HOG= 0.09+ 0.55* (100.6/82) * 0.037

HOG= 0.115 m

Z = HOG. NOG

Z= 0.115 * 10.5

Z= 1.2 m

Close enough to the estimate value.

3.2.4 Pressure drop

$Re_1 = \rho\ u\ dm / \mu$

Where

U1 = U/e

dm = e/s(1-e) µ

to find the pressure drop we can use this equation: -

$$\frac{R_1}{\rho u_1^2} = \frac{dp\ e^3}{s(1-e)L\ \rho u^2} \qquad \frac{R_1}{\rho u_1^2} = \frac{5}{Re} + \frac{0.4}{Re^{0.1}}$$

e = 0.476

$\rho = 1.0\ \mathrm{Kg/m^3}$

Volumetric flow rate = cross sectional area * velocity

100.6/1.0 = 0.049 * u

U = 2053.06 m/s

d particle = 13mm

for spherical particle

$S = 6/d = 6/0.013 = 461.5\ \mathrm{m^2/m^3}$

$\mu = 0.02 * 10^{-3}\ \ \mathrm{Ns/m^2}$

$Re_1 = 1.0 * 2053.06 / 461.5(1-0.476) * 0.02 * 10^{-3} = 424491$

$$\frac{R_1}{\rho u_1^2} = \frac{5}{424491} + \frac{0.4}{424491^{0.1}} = 0.109$$

$$0.109 = \frac{dp\ 0.476^3}{461.5(1-0.476) * 12.8 * 1.0 * 2053.06}$$

$dp = 128.34\ \mathrm{KN/m^2}$

3.2.5 Mechanical Design

Thickened of Column from volume (6)

$e = (Pi\ Di / 2f-Pi) + C$

where

e = thickness of column (mm)

Pi= design pressure (N/mm^2)

Di= internal diameter (mm)

F = design stress (N/mm^2)

C= corrosion allowance = 2 mm

typical design stress of stainless steel material (18 Cr / 8 Ni) at 70 °c = 160 N/mm^2 …… [15]

Design pressure = 10% above operating pressure

$= 1.1 * 1 = 1.1$ atm $= 0.11\ N/mm^2$

$e = (0.11*1200 / 2*160 – 0.11) + 2 = 2.4$ mm ~ 2.5 mm

3.2.6 Design of domed end

From volume 6 use standard ellipsoidal

$$e = \frac{Pi\ Di}{2fJ-0.2\ Pi} + C$$

Where

e = thickness of head (mm)

J = joint factor = 0.8 – 0.85

$$e = \frac{0.11 * 1200}{2 * 0.85 * 160 - 0.2 * 0.11} + 2$$

$e = 2.1$ mm ~ 2 mm

Nozzles

$d = 226\ G0.5\ \rho-0.35$

where

d= pipe diameter (mm)

G= mass flow rate (kg/s)

ρ= density (kg/m^3)

feed gases

G= 0.043 Kg/s

ρ= 997.04 Kg/m^3

$d= 226\ (0.023)^{0.5}\ (997.04)^{-0.35} = 3.05$ mm

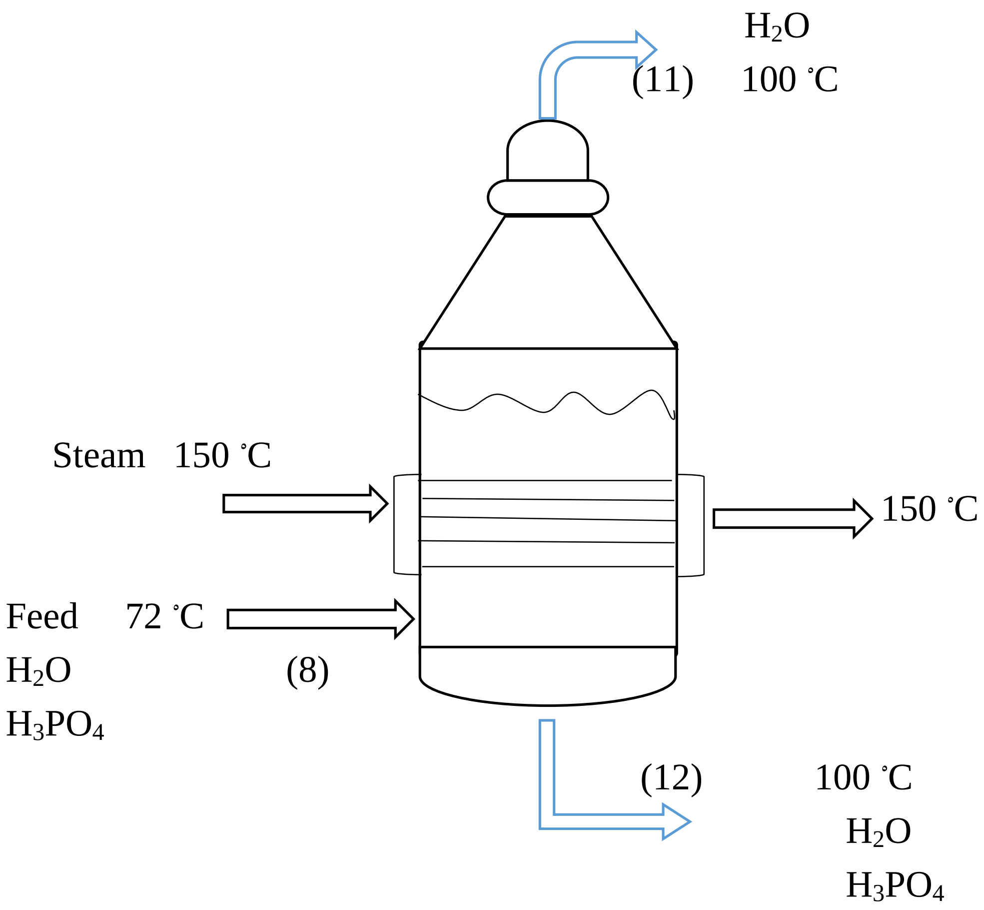
feed water

G= 0.023 kg/s

ρ= 998 kg/m^3

$d= 226\ (0.023)0.5\ (998)-0.35 = 3.1$ mm

3.3 Evaporator Design



Component	Steam (8) kg/hr.	Steam (11) kg/hr.	Steam (12) kg/hr.
H_3PO_4	42	—	42
H_2O	63	49	41

3.3.1 Calculation of Bundle tubes

take tube dimensions as follows (15) :-

$d_o = 16$ mm

di = 14 mm

L = 1.5 m

Heat flux = 2012.5 w/m^2 (16)

From Energy Balance

Heat load = 32.3 Kw

- Heat transfer area = Q/Heat flux

$A= 32.3/2.0125 = 16\ m^2$

Area of one tube $a_\circ = \pi\ d_\circ\ L = \pi *14*10^{-3} *1.5 = 0.066\ m^2$

Number of tube

$N = A/a_\circ = 16/0.066 = 243$

Tubes arranged in triangular pitch one pass

$Pt = 1.25\ d_\circ$

$Pt = 1.25*16 = 20mm$

From volume (6) eq. (12.3b) P. (649)

Bundle diameter $Db = d_\circ\ [Nt/K_1]^{1/n1}$

Where K_1, n_1 constants (17)

$K_1 = 0.319$

$N_1 = 2.142$

$Db = 16\ [243/0.319]^{1/2.142} = 355\ mm$

3.3.2 Tube side pressure drop

Temperature of steam = 150 ˚C

From steam table $v_g = 1.13\ m^3/kg$

$\rho_g = 1/v_g = 1/1.13 = 0.885\ Kg/m^3$

area of flow

$A_t = N_t\ (\pi/4\ d_i^2)$

$A_t = 243 * \pi/4 *(14*10^{-3})^2 = 0.0374\ m^2$

Mass flow rate $G = \rho *u_t *A_t$

$G = 0.016\ kg/s$, $\rho = 0.88\ Kg/m^3$

$U_t = G/ \rho *A_t = 0.016/0.885*0.0374 = 0.53\ m/s$

Steam viscosity ɱ = $1.4*10^{-5}Ns/m^2$

$Re_t = \rho*u_t*d_t/$ɱ $= 0.885*0.5*(14*10^{-3})/ 1.4*10^{-5} = 442.5$

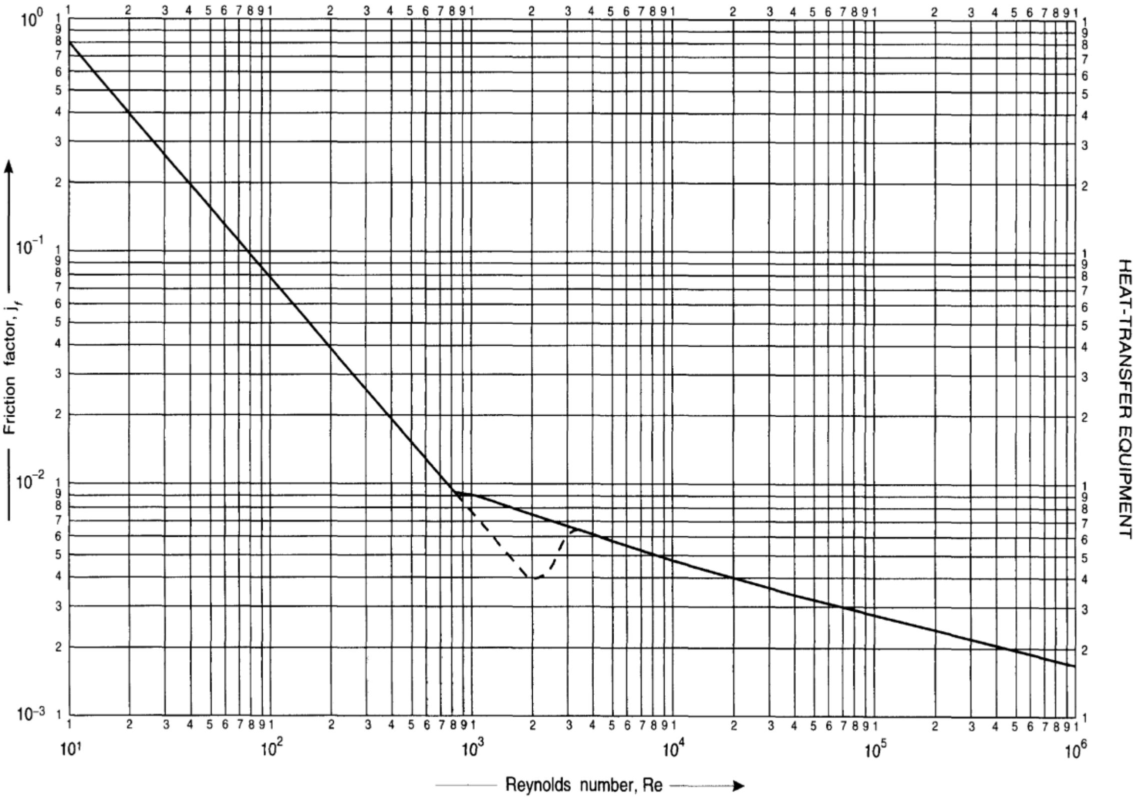


Figure 12.24. Tube-side friction factors
Note: The friction factor j_f is the same as the friction factor for pipes $\phi(= (R/\rho u^2))$, defined in Volume 1 Chapter 3.

$j_f = 2*10^{-2}$ friction factor (18)

$\Delta P = Np*[\, 8*j_f*(L/d)*(ɱ/ɱ_w)^{-m} + 2.5] * \rho*u_t^2/2$ (19)

Where:

Np = number of passes

J_f = fraction factor

L = tube length (m)

di = inside diameter (m)

ρ = density (Kg/m^3)

u_t =velocity (m/s)

neglect $(ɱ/ɱ_w)^{-m} = 1$

$$\Delta P= [\, 8*2*10^{-2} *(1.5/14*10^{-3}) +2.5] *0.885 * (0.5)^2/2 = 2\ N/m^2$$

$$= 2.9*10^{-4}\ Psia$$

If a high-pressure drop must choose the other dimensions of the tube to get to the proper pressure drop is less than (1Psia)

3.3.3 Volume of evaporator

$V = 1.7261[\, e^{-p/RT} + K\, (ɱ_l/ɱ_v)^{0.2}]$ (20)

Where:

V= volume of evaporator (m^3)

P= pressure of evaporator (N/m^2)

R= gas constant

T= temperature (K)

K= constant depend on type of tubes bundle (steam bundle)

K= 2.125 for vertical tube

K= 1.1942 for horizontal tube

$\mu_v = 1.4*10^{-5}\ Ns/m^2$

$\mu_l = 0.28*10^{-3} Ns/m^2$

take horizontal tube

operating pressure = 10 psi = 0.6 atm. = $0.69*10^5\ N/m^2$

$V = 1.7261[\, e^{-0.69*10-5/8.314+373} + 1.1942(0.28/0.014)^{0.2}] = 5.5\ m^3$

For standard tubes diameter select shell diameter = 1.5 m

(shell diameter must be equal to tube length of steam bundle)

3.3.4 Length of evaporator

$V = π/4\ D^2 *L$

$5.5 = π/4\ (1.5)^2 *L$

L = 3.114 m = 3.2 m

3.3.5 Calculation of residence time (Ʈ)

$Ʈ = v/v_{\circ}$

Where:

V = volume of evaporator (m^3)

$v_{\circ}$ = feed volumetric flowrate (m^3/s)

$Ʈ = v/v_{\circ} = \frac{5.5}{mass/\rho}$

$P_{mix.} = 1333.6\ Kg/m^3$

$Ʈ = \frac{5.5}{24*105/1333.6} = 2.9$ hr. = 174 min

3.3.6 Mechanical Design

t= $\frac{pi*Di}{2fJ-pi}+C$ (21)

where:

t= thickness of shell (mm)

Pi = operating Pressure (N/mm^2)

Di= shell diameter (mm)

J = joint factor (0.8)

C= corrosion allowance (2mm)

f= design stress (N/mm^2)

Operating pressure = 10% above design pressure = 1.1*0.069= 0.076 N/mm^2

Stainless steel (18cr/8Ni) (22)

f= 150 N/mm^2

t= $\frac{0.076*1500}{2*0.8*150-0.076}+2=2.5\,mm=3mm$

3.3.7 Thickness of cover

We can use hemispherical cover thickness of cover = 0.6 * thickness of shell

t = 0.6 *3 = 1.8mm

3.3.8 Weight of evaporator

For stainless steel

W_v = 240 C_v D_m (H_v+0.8 D_m)t

Where

Cv= constant (1.08)

D_m= mean shell diameter (m)

= (Di + t * 10^{-3})= 1.503 m

H= length of vessel (m)

t = shell thickness (mm)

Wv= 240 * 1.08 * 1.503 (3.2+ 0.8 * 1.503) * 3 = 5145 N

3.3.9 Weight of vessel filled with water

$Ww = \pi/4\ D^2L * \rho_{H2O} * g$

$= 5.5 * 1000 * 9.81 = 53955\ N$

3.3.10 Weight of tubes

Weight of tube (1.05 Kg/m)

Weight of one tube = 1.5 * 1.05 = 1.6 Kg

Weight of tube (Wt) = 243 *1.6 * 9.81 = 3815 N

4.3.11 Weight of cover

$r_i = D/2 = 1.5/2 = 0.75\ m$

$r_\circ = r_i + t = 0.75 + 0.0018 = 0.7518\ m$

∴ Total height of evaporator = 3.2 + 2 * 0.7518 = 4.7 m

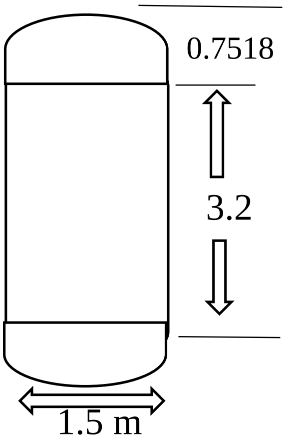
Volume of sphere ($4\pi/3 * r^3$)

∴ volume of two cover = $4\pi/3\ (r_\circ^3 - r_i)^2$

for carbon steel ($\rho = 7.7*10^{-4}\ Kg/m^3$)

$Wc = 4\pi/3[(0.7518)^2 - (0.75)^2] * 7.7*10^4*9.81$

$= 8549*2 = 17098\ N$



4.3.12 Total weight

$W_T = W_v + W_t + W_w + W_c + W_{man}$

$W_T = 5145 + 3815 + 53955 + 8549*2 + 1500$

$W_T = 81513\ N$

$= 81.5\ KN$

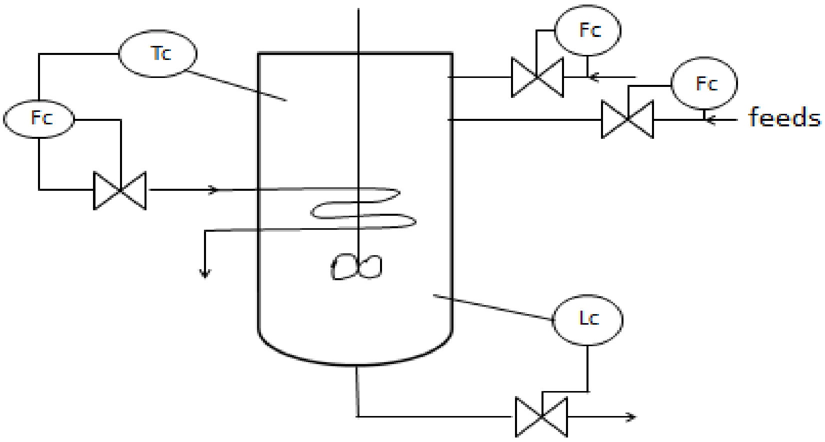
CHAPTER FOUR
PROCESS CONTROL

4.1 Introduction to Process Control

Process control is an engineering discipline that deals with architectures, mechanisms and algorithms for maintaining the output of a specific process within a desired range. For instance, the temperature of a chemical reactor may be controlled to maintain a consistent product output.

Process control is extensively used in industry and enables mass production of consistent products from continuously operated processes such as oil refining, paper manufacturing, chemicals, power plants and many others. Process control enables automation, by which a small staff of operating personnel can operate a complex process from a central control room.

4.2 Control system for Reactor



Where

Tc = temperature controller

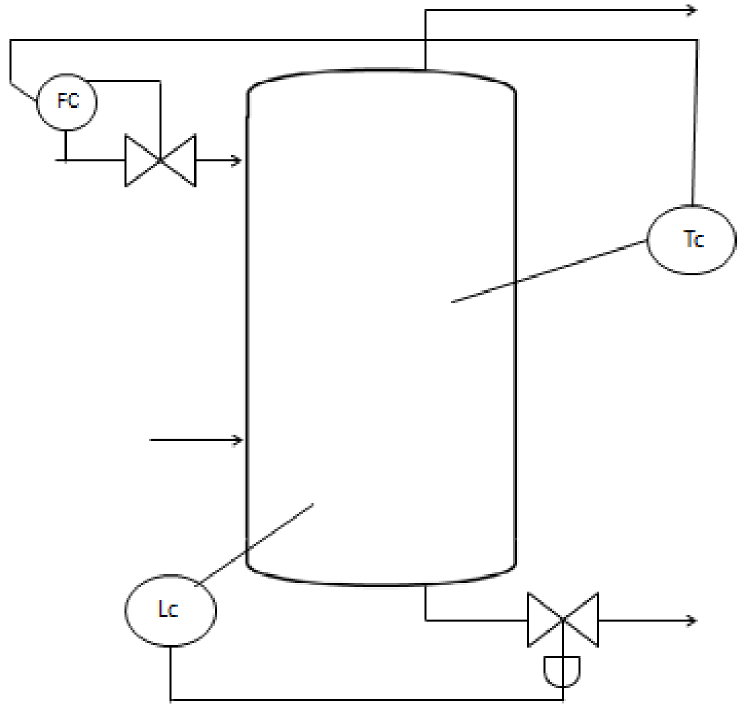
Fc = Flow controller

Lc = Level controller

$$\frac{1}{\mu m} = \frac{W1}{\mu 1} + \frac{W2}{\mu 2}$$

Log (log 10 μm) = ρ_m [x1 I1 +x2 I2 /x_1 μ_1 +y_2 μ_2] * 10-3 – 2.9

4.3 Control system for Absorber



Where

T_c= temperature controller

F_c= Flow controller

L_c= Level controller

CHAPTER FIVE

5.1 PLANT LOCATION AND SITE SELECTION

The location of the plant can have a crucial effect on the profitability of project, and the scope for future expansion. Many factors must be considered when selecting suitable site, and only a brief review of the principal factors will be given in this section.

1. Location, with respect to the marketing area
2. Raw material supply.
3. Transport facilities.
4. Availability of labor.
5. Availability of utilities: water, fuel, power.
6. Availability of suitable land.
7. Environmental impact, and effluent disposal.
8. Local community considerations.
9. Climate.
10. Political and strategic considerations.

5.2 Safety and Environmental

Any organization has a legal and moral obligation to safeguard the health and welfare of its employees and the general public. Safety is also good business; the good management practices needed to ensure safe operation will also ensure efficient operation

The term “loss prevention” is an insurance term, the loss being the financial loss caused by an accident. This loss will not only be the cost of replacing damaged plant and third party claims, but also the loss of earnings from lost production and lost sales opportunity.
All manufacturing processes are to some extent hazardous, but in chemical processes there are additional, special, hazards associated with the chemicals used and the process conditions. The designer must be aware of these hazards, and ensure, through the application
of sound engineering practice, that the risks are reduced to acceptable levels.

Safety and loss prevention in process design can be considered under the following broad headings:

1. Identification and assessment of the hazards.

2. Control of the hazards: for example, by containment of flammable and toxic materials.
3. Control of the process. Prevention of hazardous deviations in process variables
(Pressure, temperature, flow), by provision of automatic control systems, interlocks,
Alarms, trips; together with good operating practices and management.
4. Limitation of the loss. The damage and injury caused if an incident occurs: pressure
relief, plant layout, provision of fire-fighting equipment.

THE HAZARDS

In this section the special hazards of chemicals are reviewed (toxicity, flammability and corrosively); together with the other hazards of chemical plant operation

1.Toxicity Most of the materials used in the manufacture of chemicals are poisonous, to some extent. The potential hazard will depend on the inherent toxicity of the material and the frequency and duration of any exposure.
2. Corrosion and erosion: despite good design and materials selection, some corrosion problems may arise, both internally and externally. The factor to be applied depends on the anticipated corrosion rate.
3. Explosions an explosion is the sudden, catastrophic, release of energy, causing a pressure wave (blast wave). An explosion can occur without fire, such as the failure through over-pressure of
a steam boiler or an air receiver
4. Temperature deviations Excessively high temperature, over and above that for which the equipment was designed, can cause structural failure and initiate a disaster.
5. Fire protection to protect against structural failure, water-deluge systems are usually installed to keep vessels and structural steelwork cool in a fire
6. Noise
Excessive noise is a hazard to health and safety. Long exposure to high noise levels can cause permanent damage to hearing
7.Rotating equipment: this factor accounts for the hazard arising from the use of large pieces of rotating equipment: compressors, centrifuges, and some mixers.
8. Leakage joints and packing: this factor accounts for the possibility of leakage from gaskets, pump and other shaft seal

5.3 Economic

The economic construction and efficient operation of a process unit will depend on how well the plant and equipment specified on the process flow-sheet is laid out

The principal factors to be considered are:

1. Economic considerations: construction and operating costs.
2. The process requirements.
3. Convenience of operation.
4. Convenience of maintenance.
5. Safety.
6. Future expansion.
7. Modular construction.

The cost of the devices used in the factory (23)

The cost of a reactor height of (1.5 m) = 30,000 $

The cost evaporator area (16 m^2) = 20,000 $

The cost of absorbing high tower (12.8 m) = 55,000$

Total = 105,000 $

factory note that "these accounts by plans in place for the year 2004

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- *Sulfuric/Phosphoric Acid Plant Operation*, American Institute of Chemical Engineers, New York, 1982.

3. From volume (6) fig 10.75 we have viscosity and volume we get suitable is propellers (three blades pitch diameter)

4. From volume (6) table 13.2

5. From volume (6)

6. From Ideal Gas Law (p v = n R T)

7.Antoine.Frostburg.edu/chem/sense/javascript/water- density.html

8. www.Viscopedia.com/viscosity-tables/substances/water/

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11. From volume (6) fig (11.41)

12. From volume (6) fig (11.42)

13. From volume (6) eq. (11.111)

14. From volume (6) fig (11.41)

15. From volume (6) table (12.3) p. (645)

16. From "Heat transfer mechanism of vaporization water" by Dangler

17. from volume (6) p. (649), Table (12.4)

18. From volume (6) fig. (12.24) (P. 667)

19. from volume (6) (P. 668)

20. Using brooks and badger equation "from heat transfer mechanism of vaporization water " by dangler

21. Thickness of evaporator shell from volume (6)

22. From volume (6) table (13.2)

23. From volume (6) fig.(6.3a) & fig. (6.5a) page. (256)

24.Graduation Project / University of Technology Faculty of Engineering / Chemical Engineering Department for the academic year 2005