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Production of Acrylic Acid

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Abstract: ***This report provides a complete analysis of an alternate method of production of acrylic acid and highlights the advantages it has over the conventional methods. It covers all the aspects that one needs to consider while setting up a new production plant and thus can act as a suitable guide. Having established the need and profitability prospects of the setup of acrylic acid using ethylene over other possible mechanisms of production in the initial sections, the original hypothesis was proved correct. Following this, the various process flow chart and the process parameters were stated which emphasized the various equipment and raw materials required for the project and gave a rough estimate of the production capacity of the plant. An in-depth analysis was done by using Mass and Energy Balances which can be used to understand the material and power requirements for the process in terms of flow rates and utilities. The ideal break-up of this power requirement through heat exchangers and hot and cold utilities was done and an overall Heat Exchanger Network was drawn. ASPEN Simulations were run which computed and verified the more intrinsic values like all flow rates, composition, entropy and enthalpy values for the different process streams. The Process Flow Diagram gives a crisp representation of the overall design of the plant. Our HEN has 3 heat exchangers, 4 cold utilities and 1 hot utility. Note that this is the MER design which uses minimum utilities and maximum heat recovery. The economic analysis shows that the project has very heavy initial investment of over Rs. 22 crores and thus makes it feasible for only those who have access to such heavy investment at a reasonable interest rate. However, the high rate of return (7.8%) and low payback period (6.3 years) makes our investment worthwhile. Also, the annual profits are very high as compared to the operating cost which implies that in the long run, when the high initial investment is already recovered, our project will be highly profitable.***

Keywords: Acrylic Acid Production, PTA analysis, Pinch analysis

1. Introduction

Acrylic acid (CH_2=CHCOOH) is the simplest unsaturated carboxylic acid and exists as a colourless liquid with pungent smell at standard temperature and pressure (T = 298K, P = 1atm). Acrylic acid finds applications in varied industries due to its varying degree of durability, tackiness & hardness. Crude acrylic acid is(1):

- Used as a monomer in Superabsorbent polymers (SAPs) that have applications in baby diapers, sanitary napkins, soaker pads for food packaging, etc.
- Used to produce acrylates that have applications in masonry and industrial coatings, adhesives, paper and leather coatings, floor polishes, wood paints, tablet coatings etc.
- Used to make acrylic glass which is an alternative to glass in greenhouses
- Used as monomer in production of detergents

The demand for acrylic acid has been constantly increasing due to growth in demand for SAPs, and growth in water-based adhesive market and real estate industry but our country still relies on foreign suppliers. The demand supply gap is increasing very rapidly and is expected to reach 260 kilo tonnes per annum (KTPA) by 2020. (Fig.1A) The Indian market for acrylic acid seems very lucrative as evident from data shown below (Fig.1B)(2).

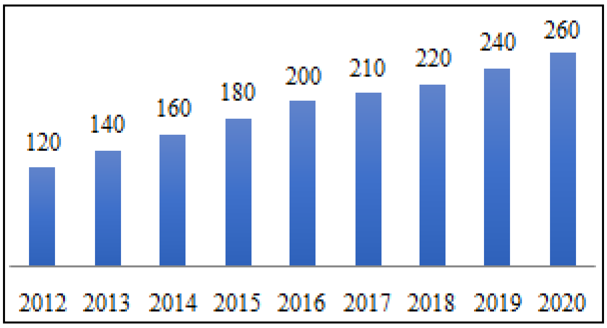


Figure 1 (A): Demand - Supply Gap in India (in KTPA) (2)

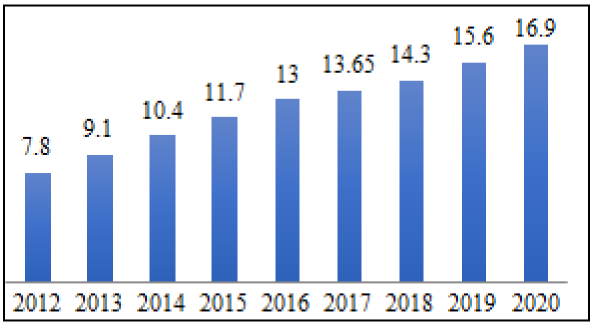


Figure 1 (B): Market Size of Indian Acrylic Acid (in INR billion) (2)

We conclude that setting up an acrylic acid plant will be profitable because of low existing domestic competition. We can sell it at a lower price as compared to foreign manufacturers due to the reduction in transport and custom cost. As demand-supply gap is 210 KTPA, we are planning for plant with production of 100 KTPA to meet the forecasted demand.Mumbai and Gujarat are the best possible locations because of factors like government support, flourishing economy, ease of doing business, good infrastructure and availability of raw material, power and labour etc.

Our major clients will be Delta Chemsol, Meru Chem Private Limited, Madhu Chemicals, Viraj Industries etc. (3)who are importers and suppliersof acrylic acid in India. We'll face competition from companies like A.B Enterprises (Mumbai), Triveni Interchem Pvt. Ltd. (Vapi), A.S. Chemicals (Hyderabad), Alpha Chemika (Mumbai)(4). Amongst foreign players, USA and China are leading exporters.

2. Existing Processes for Acrylic Acid Production

2.1 Acrylic Acid by Sequential Dehydration and Oxidation of Glycerol

The production of acrylic acid using the technique of dehydration–oxidation of glycerol is a very attractive approach. This is mainly because the raw material, glycerol, is available in sufficient amount as the byproduct of the biodiesel production process. Biodiesel is a fast growing industry and so is the production of glycerolthat has increased by 10 times (~1 million to 10 million tonnes) in the last decade. This has led to a drop in the price of glycerol by nearly 60% in the last ten years. (5)

The reaction process involves two steps – first isdehydration of glycerol in acrolein

Dehydration:$C_3H_5(OH)_3 \rightarrow C_3H_4O + 2\ H_2O$
Glycerol → Acrolein + 2 Water(5)

Next step is oxidation of acrolein in acrylic acid&major byproducts are acetaldehyde &acetic acid.
Oxidation:$C_3H_4O + 0.5\ O_2 \rightarrow C_2H_3COOH + H_2O$
Acrolein + 0.5 Oxygen → Acrylic acid + Water(5)

The reaction is carried out using two different types of packed bed catalyst in a single fixed bed reactor because of higher yield (75%) as compared to single packed bed catalyst where the yield is just 25%. For the dehydration step, $Cs_{2.5}H_{0.5}PW_{12}O_{40}$ supported on Nb_2O_5 (CsPW-Nb) is used as a catalyst and for oxidation step, vanadium-molybdenum mixed oxides supported on silicon carbide (VMo-SiC) is used.Acetol and water which are the byproduct of the dehydration step do not have any negative effects on the subsequent oxidation step. Also, there is no deactivation of any of the catalysts and both of them can be completely regenerated by coke burning at 500^{o}C. The reaction is carried out at an optimum temperature of 300^{o}C and ambient pressure because although increasing the temperature increases the conversion but trade off is decrease in selectivity.

Figure 2.A: Input-Output Diagram for production of Acrylic acid by Glycerol

2.2 Acrylic Acid from ethylene

Acrylic Acid can be manufactured from ethylene in a three step process namely oxidation of ethylene, carbonylation of ethylene oxide, and acid catalyzed rearrangement of β-propiolactone. Cost of raw material, ethylene, is the biggest contributor in the total cost of production.

In the first step, ethylene is oxidized to ethylene oxide with highly pure oxygen so as to minimize the reaction volume. This reaction is carried out in vapour phase in a tubular reactor. Silver is used as a catalyst on α-aluminum support. Impurities of Cs and Re are added to the catalyst in order to increase the selectivity of ethylene oxide to 80-90%. The reaction is carried out at a temperature of 240-290^{o}C and a pressure of 15-25 bar. Since the reaction is highly exothermic, heat removal is a major requirement for maintaining high selectivity. After purification, it is possible to obtain a product stream of 99.5% purity.

Oxidation: $C_2H_4 + 0.5O_2 \rightarrow C_2H_4O$
Ethylene + 0.5 Oxygen → Ethylene oxide(6)

In the second step, ethylene oxide is carbonylated to produce β-propiolactone in the presence of [Co(CO)4] as the catalyst and thiamine pyrophosphate (TPP) as the promoter. The product propiolactone is toxic and thus cannot be isolated. The reaction is carried out in liquid phase under isobaric conditions (pressure of CO is kept constant) and at a temperature of 240^{o}C.

Carbonylation: $C_2H_4O + CO \rightarrow C_3H_4O_2$
Ethylene oxide + CO →*β*-Propiolactone(6)

In the last step, β-propiolactone undergoes rearrangement to form acrylic acid in the presence of phosphoric acid as catalyst and mono methyl ether hydroquinone (MEHQ) as inhibitor to prevent the product from polymerizing. Small amount of water is added to catalyst to increase its viscosity & maintain phase as liquid. Reaction is carried in vacuum at high temperature to achieve per pass conversion of 0.97 &the overall conversion of 0.995.(6)

Rearrangement: $C_3H_4O_2 \rightarrow C_2H_3COOH$
β-Propiolactone → Acrylic Acid(6)

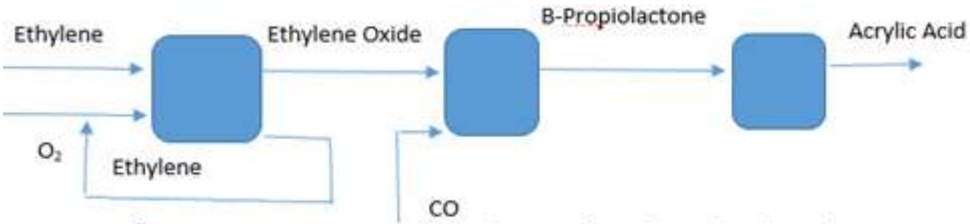


Figure 2.C: Input-Output Diagram for production of Acrylic acid by Glycerol

2.3 Acrylic acid from propylene

Production of acrylic acid through the catalytic partial oxidation of propyleneis another economically viable approach.Propylene is easily available as a byproduct of cracking of naptha.Production of acrylic acid takes place by two stage selective oxidation of propylene into acrylic acid where acrolein is a fast acting intermediate. The oxidation reaction is carried out in a temperature range of 310-350^{0}C and is highly exothermic. It takes place in fixed bed reactor using mixed oxide of arsenic, tantalum and molybdenum as catalyst. 200 MM lb/year glacial-grade acrylic acid can be produced industrially using this technique. This process design is technologically advanced in terms of high production yield, catalysis of unsaturated acid from saturated hydrocarbon and optimization of process component. The process values customer as it is low cost and environmentally responsible and has high purity. Major side reactions are formation of acetic acid through activated propylene or acrolein and oxidation of hydrocarbons to give carbon dioxide. Side reactions can be minimized by maintaining the temperature at desired level by removing the reaction heat(7)

Oxidation: $C_3H_6 + O_2 \rightarrow C_3H_4O + H_2O$
Propylene + Oxygen → Acrolein + Water
Oxidation: $C_3H_4O + 0.5O_2 \rightarrow C_2H_3COOH$
Acrolein + Oxygen →Acrylic Acid

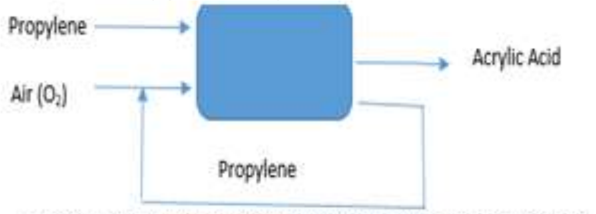


Figure 2.B: Input-Output Diagram for production of Acrylic acid by Glycerol

2.4 Acrylic acid from propane oxidation

Acrylic acid can be manufactured from propane in an attractive way. This is mainly because propane is very less expensive than propylene. In this method, propane is catalytically oxy-dehydrogenated to form a mixture of propylene and propane. Then the mixture of propylene and propane are used to produce crude acrylic acid by using the two stage propylene oxidation process as discussed above. (9)

2.5 Acrylic acid from Lactic Acid

Acrylic acid can be produced by the vapour phase dehydration of lactic acid. This method of production of acrylic acid is highly desirable as it is produced from a bioderivative, which is a renewable source of energy. Various catalysts like sodium dihydrogen phosphate (NaH_2PO_4), silica/alumina, $Ca_3(PO4)_2$ supported onsilicaare used for dehydration of lactic acid. Calcium sulphate with cupric sulphate and phosphate are added as promoter and CO_2 as the carrier gas in order to increase the selectivity of acrylic aid to 60-70%. The reaction is carried out a temperature of 325-400°C. When the reaction temperature increases from 325°C to 400°C, there is a drastic change in the conversion of lactic acid i.e. from 60% to 100%. (8)

Lactic acid + $NaH_2PO_4 \rightarrow$ Acrylic Acid

3. Analyzing the Economic Viability of the Processes

In this section, we'll discuss the economic viability of the processesthat were discussed in the previous section. Out of the three major processes discussed, we need to shortlist one by eliminating the rest. Process 5 is eliminated owing to the very high reaction temperatures (~400^{o}C), which leads to high operation cost. Also, in this reaction, non-volatile oligomers of lactic acid are formed which poison the catalyst (11). Suppressing their formation requires high dilution rates thus this reaction is difficult to scale to industrial levels.Process 3 & 4 are essentially the same except for the way propylene is obtained. Hence we use a common analysis for both. Here, we are assuming air to be used for the supply of oxygen. Thus, air being a cheap raw material; we are ignoring its cost in the calculation of our EP values(6). In process 1, selectivity (S) of acrylic acid is 95% and rest 5% is mainly acetol, which is a byproduct in this process, and for process 3, selectivity of acrylic acid is 60% and 40% is for the byproduct i.e. acetic acid (7).In process 2, the first step i.e. oxidation of ethylene is considered to be reversible process.The same has been included in EP calculation (5). We assume 100% overall conversion in each of these reactions. Cost of streams involved in each process is shown in Figure 3.1 (*www.indiamart.com*).

	Process 1	**Process 2**	**Process 3 & 4**
Raw Material (R1)	Glycerol	Ethylene	Propylene
Price/Kg	Rs. 70	Rs. 52.75	Rs. 52.833
Price/mol	Rs. 6.477	Rs.1.884	Rs. 2.222
Raw Material (R2)		Carbon Monoxide	
Price/Kg		Rs. 14.601	
Price/mol		Rs. 0.41	
Product (P)	Acrylic Acid	Acrylic Acid	Acrylic Acid
Price/Kg	Rs. 150	Rs. 150	Rs. 150
Price/mol	Rs. 10.815	Rs. 10.815	Rs. 10.815
Byproduct (BP)	Acetol		Acetic Acid
Price/Kg	Rs. 150		Rs. 34
Price/mol	Rs. 11.1		Rs. 2.04
General Equation	S*P + (1-S)*BP–R1	P - R1 - R2	S*P + (1-S)*BP – R1
Economic Potential	0.95*10.815 + 0.05*11.1 – 1*6.477 = 4.352	1*10.815 – 1*1.884- 1*0.41 = 8.521	0.6*10.815 + 0.4*2.04 – 1*2.222 = 5.083
Viability	Low	High	Medium

Figure 3.1: Economic Analysis of all different processes

From the Economic Potential (EP) values, it is clear that the process 1, that is, production of acrylic acid by sequential dehydration and oxidation of glycerol, is less viable as compared to the other options. So, it is less preferred due to its low viability. However, we need to conduct a more detailed analysis for choosing between Process 2 (acrylic acid production from ethylene)and Process 3 (acrylic acid production from propylene) since the EP values are almost same. For this, we need to focus on the ease of availability of the raw materials, which can be analyzed on the basis of the production rates of these materials.

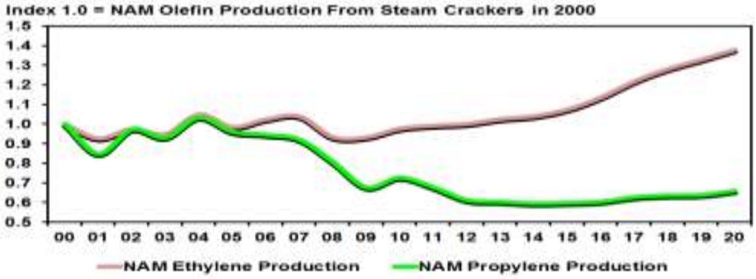


Figure 3.2: Comparison of Ethylene and Propylene Production (10)

As evident from Fig 3.1, the ration of production rates of ethylene and propylene is almost same till 2006 but post that, the production rates of propylene began to fall down while that of ethylene began to shoot up. This can be attributed to the increasing availability of inexpensive shale gas-based ethane. Moreover, as new crackers for pure ethane came into existence, ethylene production is projected to increase further while propylene production is projected to stay low.(10)

4. Process Flow Chart for production of acrylic acid from ethylene

The detailed process flowchart is given in Figure 3.2.Stream 1 (ethylene) and stream 2 (air for oxygen)are mixed and heated to reaction temperature and pressure.The mixed stream is fed to microchannel reactor because of its advantages over tubular reactor. The advantages include the fact that it has a larger surface to volume ratio (S/V)which leads to an effective heat management and decreases downstream downstream volume and vessel sizing.Also, if an explosion occurs (due to high pressure) microchannels can withstand it easily. Since ethylene (Boiling Point -103°C) and ethylene oxide (Boiling Point 10.7°C) can easily be separated, we place a distillation column after first reactor only and send the recycle stream from top to the mixer. This is in accordance with the general heuristics for separation as addition of CO (Boiling Point -191.5°C) to Reactor 2 (R-2) would have increased the separationcost for the two recycle streams (ethylene and CO). Distillate ethylene epoxide is then fed to a CSTR reactor (R-2) along with pure stream of Carbon Monoxide for effective mixing of CO. We use CSTR over tubular reactor because the liquid phase catalyzed reaction has higher yield in CSTR for first order reactions. The product stream from the reactor R-2 is fed todistillation column D-2that is kept at low pressure to maintain the reboiler temperature. β-propiolactane being the heavy key leaves asbottom product and unreacted CO and ethylene epoxide, which come out astop products are recycled back to the reactor R-2.β-propiolactanebeing toxic is not isolated and for that Reactor 2 is closely coupled with the Reactor 3, which is apacked bed reactor (R-3) where rearrangement reaction takes place in presence ofacid catalyst (phosphoric acid). Product stream is entirely in vapor phase which is then flashed to remove the liquid. Flashed liquid products are then recycled back to the reactor and the vapor of acrylic acid are then fed tothe series of distillation column for the further purification.

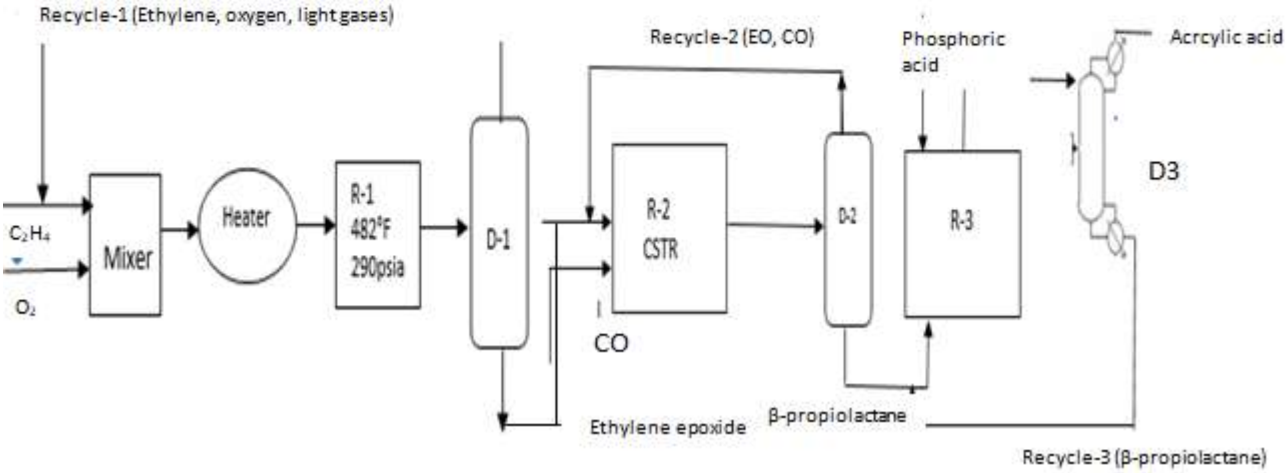


Figure 3.3: Process Flow Chart for acrylic acid production

5. Process Parameters

As concluded in the previous discussion, we'll go ahead with production of acrylic acid from ethylene. This reaction takes place in three steps – oxidation, carbonylation and rearrangement. The first step is oxidation of ethylene, first order reaction with reaction constant (k) at 250°C is $k_{250^{o}C} = 8.91 * 10^{-4}$ /sec and Ea = 52.129 kJ/mol (12). This can be used to calculate k value at any other temperature. Using Vant Hoff's equation [$\ln (k_2/k_1) = Ea/R * (1/T_1-1/T_2)$], the value of k at 285°C is $k_{285^{o}C} = 1.89 * 10^{-3}$ /sec. Here, we assume the reaction to be reversible. Thetemperature is maintained at 250°C and pressure at 20 bar and per pass conversion at 12%, which lie towards the lower end of the permissible limit even though the rate constant increases with temperature. Apart from the fact that it is easier and more viable to maintain lower temperature and pressure, this choice is also based on the observation that increasing temperature and per pass conversion leads to decreased selectivity. Thus, ethylene is recycled and ethylene oxide is separated to 99.5% purity to push the reaction in the forward direction.

The reactor type and conditions for all the steps have been specified in the table below (6). Here, the rate expression is written in general form since values of other rate constants could not be found in literature.

Reaction	Reactor Type	Temperature	Pressure	Order of reaction	Rate expression
Oxidation	Microchannel	250°C	20 bar	First	k_f[Ethylene] - k_b[Ethylene Oxide]
Carbonylation	CSTR	175 °C	30 bar	Zero	k'
Rearrangement	Packed bed	170 °C	0.133 bar	First	k" [Propiolactone]

Figure 4.1: Reactor Parameters

The parameters for different streams for the 3 reactors have been specified in Figure 4.1. This data is extracted from literature since conducting the experiment at such a large scale is outside the scope of this course. Figure 4.2. includes most of the parameters, which will be required while conducting mass and energy balance in Report 5. Any additional information that might be required will be included later on with proper justification. (13)

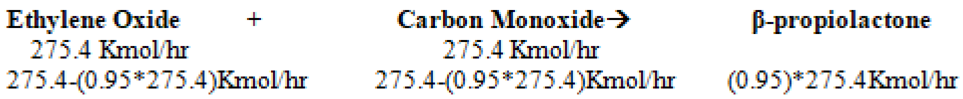
Stream	Flow Rate (kg/hr)	T (°C)	P (bar)	H (kJ/kg)
Ethylene	8126	67	20	1920
Oxygen	6699	67	20	38.5
Ethylene oxide + CO	6614	80	12	-2574
Propiolactone	16918	37.7	0.133	-4556

Figure. 4.2: Flow rates Parameters

6. Mass and Energy Balances

The detailed process flow diagram of acrylic acid production is shown in Figure.5.4with the temperature and flow rates of every stream. Firstly ethylene and oxygen is mixed at 67°C with flow rate of 8126 kg/hr and 6699 kg/hr respectively. The combined flow rate for the mixture of ethylene and oxygen comes out to be 17286.87 kg/hr by assuming ethylene to be limiting reactant in this case.When we apply the energy balance across this stream, temperature of mixture comes out to be 21.8°C. Reactor 1 is operating at 250°C, so to maintain that temperature, a heat exchanger is added in between the mixer and reactor to raise the temperature of mixtureto 250°C. The heat balance equations in the chart are calculated based on the C_P value of Ethylene as 42.9 J/(mol·K)and 29.29 J/(mol·K)for Oxygen. The heat across various heat exchangers is calculated using mC_PT. There is a total of 14592.82kg/hr output flow rate, and contains the product ethylene oxide and a large amount of side products and unreacted ethylene and oxygen. Then the product stream of reactor 1 goes into a distillation column 1 (D1) for separation from which unreacted ethylene and oxygen goes into the recycle stream with the flow rate of 2461.87kg/hrand the product ethylene oxide goes into the reactor 2 with the flow rate of 12130.94 kg/hr.

Sample calculations for Mass Balance
Taking x=0.95
Stream 6= 275.4 Kmol EO = 12130.95 Kg/hr
Stream 7= 275.4 Kmol CO =7711.2 Kg/hr



Ethylene Oxide	+	Carbon Monoxide→	β-propiolactone
275.4 Kmol/hr		275.4 Kmol/hr	
275.4-(0.95*275.4)Kmol/hr		275.4-(0.95*275.4)Kmol/hr	(0.95)*275.4Kmol/hr

Stream8= β-propiolactone+ EO +CO
= 2*(275.4- (0.95*275.4))Kmol/hr + (0.95)*275.4Kmol/hr
= 289.17 Kmol/hr

Stream 10= β-propiolactone
= 0.95*275.4 Kmol/hr
= 261.63 Kmol/hr
= 261.63*72 Kg/hr
= 18853.84 Kg/hr

Stream 9 = Stream 8- Stream 10
= 27.54 Kmol (EO+CO)
= 13.77 Kmol EO + 13.77 Kmol CO
= 606.57 Kg/hr EO+ 385.56 Kg/hr CO
= 992.13 Kg/hr (EO+CO)

In reactor 2, carbon monoxide is added with a flow rate of 7711.2kg/hr at a temperature of 176°C.In presence of CO, ethylene oxide gets converted into β–propiolactane. The total product stream, which is 20837.89kg/hr, enters the distillation column (D2). The heavy components, β-propiolactone leaves as the bottoms product at a rate of 18853.84 kg/hr, while the light key, consisting of unreacted CO and ethylene oxideleave as distillate at a rate of

999.13kg/hr, which is recycled further. To maintain the temperature around the CSTR reactor, a heat exchanger is added to decrease the heat duty of recycle stream. The product stream of reactor 2 goes into a reactor 3, which is a packed bed reactor.The β–propiolactane stream is pure liquid stream with a temperature of 232°C and a total mass flow rate of 18853.84kg/hr. To maintain the temperature 170°C around the packed bed reactor, β–propiolactane stream is cooled down by a heat exchanger of -550KW heat duty as shown in Figure.5.2. In reactor 3, rearrangement reaction takes place in presence of Phosphoric acid which have a flow rate of 4704 kg/hr. The streamsfed to the reactor are in the liquid phase, while the reactor products are in entirely vapor phase, which is then flashed to remove the liquid. So, theproduct stream of reactor 3 enters the distillation column (D3) for the separation from which flashed liquid products (β–propiolactane) are recycled back with a flow rate of 2092.32kg/hr and vapors of acrylic acid are collected as a desired product with a flow rate of 18836.71kg/hr at a temperature of 37.7°C. It is assumed that in reactor 3 only 90% of β–propiolactane gets converted into acrylic acid and remaining goes into the recycle stream.

$$Q=m\ C_p\,(T_f - T_i)$$
$$Q=261.63\ \text{Kmol/hr} * 122.1\ \text{J/mol K} * (443-505)\ \text{K}$$
$$Q=(1980591.426/3600)\ \text{KJ/sec}$$
$$Q=-550\text{KW}$$

Figure 5.1: Heat Balance for stream 10

$$Q= m * (H1 - H2)$$
$$Q = [27.54\ \text{Kmol/hr (EO)} + 27.54\ \text{Kmol/hr (CO)}]\ [-1958.05 + 1644.41]\text{Btu/lb}$$
$$Q= [385.56\ \text{Kg/hr(CO)} + 606.5685\ \text{Kg/hr (EO)}]\ [-313.64\ \text{Btu/lb}]$$
$$Q= 992.1285\ \text{Kg/hr} * (-729.53\text{KJ/kg})$$
$$Q= -201.05\ \text{KW}$$

Figure 5.2: Heat Balance for stream 9

The stream table for the process is given in Figure 5.3.

Stream No.	Components	Total Flow Rates(Kg/hr)	T_s(C)	T_t(C)
1	Ethylene	8126.0	67	21.8
2	Oxygen	6699.0	67	21.8
3	Ethylene + O_2	17286.8	21.8	250
4	C_2H_4 + O2 + Ethylene Oxide	14592.8	250	--
5	C_2H_4 + O_2	2961.8	9.03	67
6	Ethylene Oxide	12130.9	18.4	80
7	CO	7711.2	176	--
8	β–propiolactane + Ethylene oxide + CO	20837.8	80	--
9	CO + Ethylene Oxide	999.1	128.5	37.77
10	β–propiolactane	18853.8	232	170
11	Phosphoric Acid	4707	170	--
12	Acrylic acid + β–propiolactane	20929	170	--
13	Acrylic acid	18836.7	37.7	--
14	β–propiolactane	2092.3	197.0	170

Figure 5.3: Complete Stream Table

The detailed flow chart for the process is given in Figure.5.4

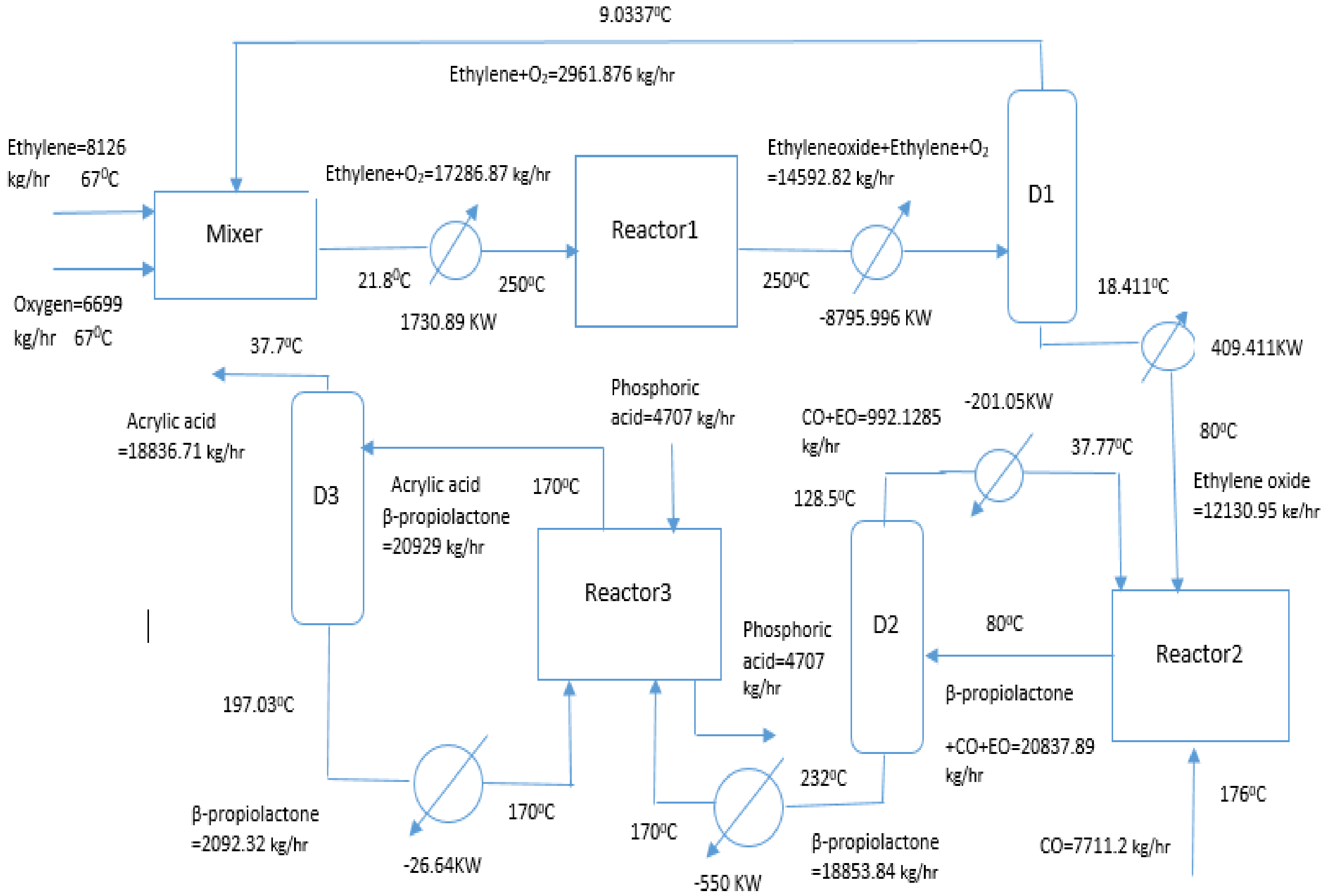


Figure 5.4: Detailed Flow Chart

7. Heat Integration

The detailed flow chart for the process is given in Figure.5.5

We have chosen ΔT_{min}= 20^{o}C. Figure 6.1 contains stream data for all the streams.

Stream Type	$T_{in}(^oC)$	$T^*_{in}(^oC)$	T_{out} (^oC)	T^*_{out} (^oC)	Δ H (KW)	1/mCp	mCp
Cold	21.8	31.8	250	260	1730.89	0.13183	7.5865
Hot	250	240	34.455	24.455	-8795.996	0.0245	40.8163
Cold	18.411	28.411	80	90	409.411	0.1504	6.6489
Hot	128.5	118.5	37.77	27.77	-201.05	0.4512	2.2163
Hot	232	222	170	160	-550	0.11272	8.8715
Hot	197.03	187.03	170	160	-26.64	1.0146	0.9856

Figure 6.1: Stream Data for Flowsheet

The stream population and heat balances for the respective Temperature intervals are given in Figure 6.2.

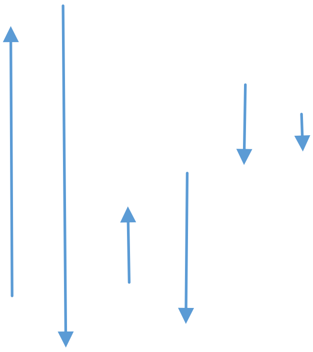
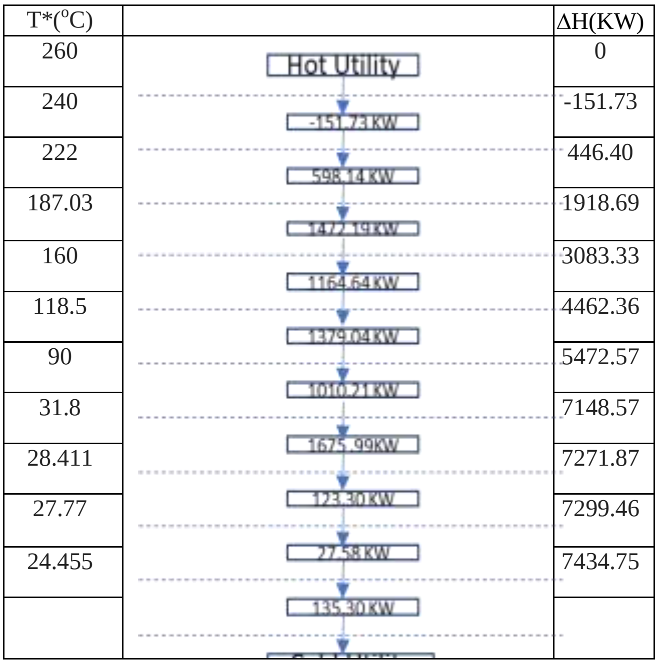
T*(°C)	Stream Population	ΔT interval (°C)	$\Sigma Cp_c - \Sigma Cp_h$ (KW/°C)	Δ H(KW)	Surplus/Deficit
260					
240		20	7.5865	151.73	Deficit
222		18	-33.2298	-598.14	Surplus
187.03		34.97	-42.1013	-1472.19	Surplus
160		27.03	-43.0869	-1164.64	Surplus
118.5		41.5	-33.2298	-1379.04	Surplus
90		28.5	-35.4461	-1010.21	Surplus
31.8		58.2	-28.7972	-1675.99	Surplus
28.411		3.389	-36.3837	-123.30	Surplus
27.77		0.641	-43.0326	-27.58	Surplus
24.455		3.315	-40.8163	-135.30	Surplus

Figure 6.2: Stream population table and temperature interval heat balances

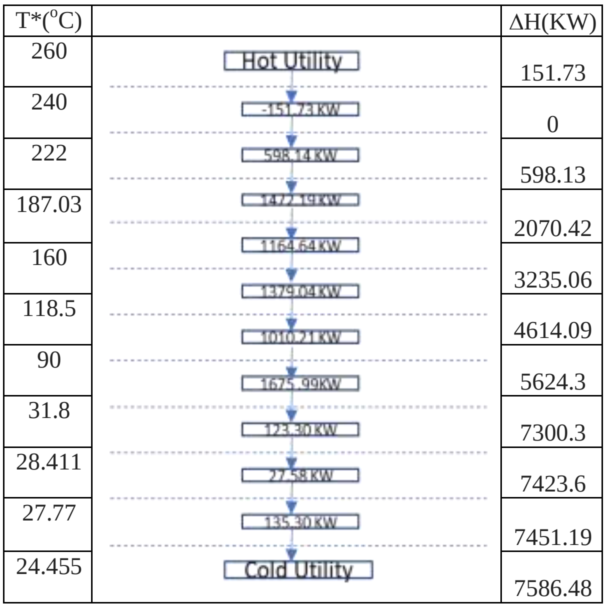
Figure 6.3 & 6.4 depict the cascade heat diagram for the system with and without heat utilities. The Pinch Temperatures comes out to be 250°C. The minimum Hot Utility required is Q_{Hmin}= 151.73KW and the minimum Cold Utility required is Q_{Cmin}= 7586.48KW.



T*(°C)		ΔH(KW)
260	Hot Utility	0
240	-151.73 KW	-151.73
222	598.14 KW	446.40
187.03	1472.19 KW	1918.69
160	1164.64 KW	3083.33
118.5	1379.04 KW	4462.36
90	1010.21 KW	5472.57
31.8	1675.99KW	7148.57
28.411	123.30 KW	7271.87
27.77	27.58 KW	7299.46
24.455	135.30 KW	7434.75

Figure 6.3: Cascade heat diagram for the system without heat utilities

Figure 6.5 depicts the corresponding GCC for the obtained cascade heat diagram.



T*(°C)		ΔH(KW)
260	Hot Utility	151.73
240	-151.73KW	0
222	598.14 KW	598.13
187.03	1472.19 KW	2070.42
160	1164.64 KW	3235.06
118.5	1379.04 KW	4614.09
90	1010.21 KW	5624.3
31.8	1675.99KW	7300.3
28.411	123.30 KW	7423.6
27.77	27.58 KW	7451.19
24.455	135.30 KW Cold Utility	7586.48

Figure 6.4: Cascade heat diagram for the system with heat utilities

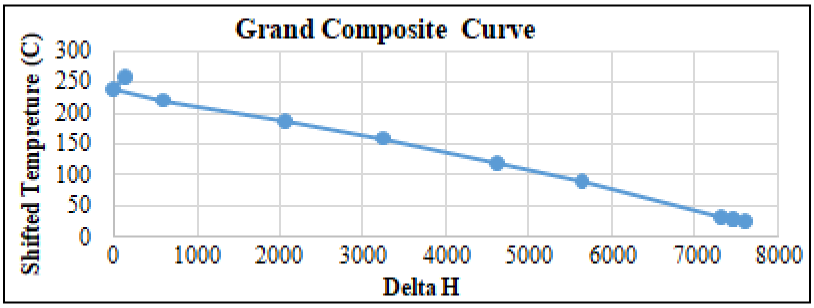


Figure 6.5: Grand Composite Curve

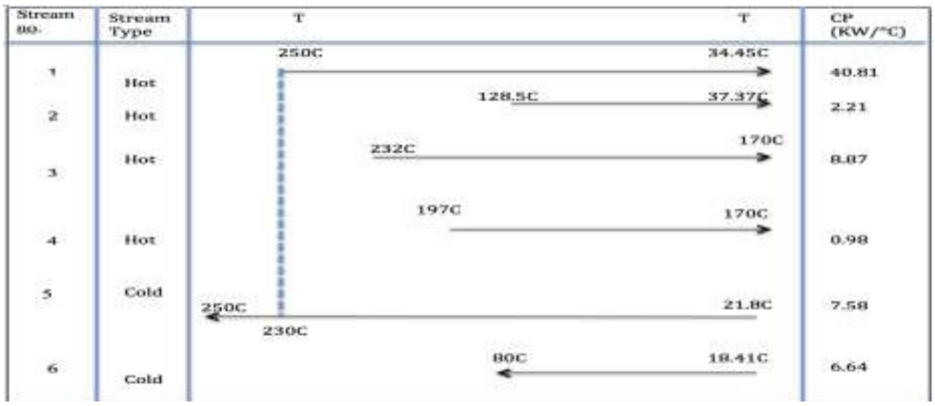


Figure 6.6: Stream Flow Chart

Since no stream is present above the pinch, we only focus on below the pinch analysis. Stream flow chart is shown in Figure 6.6. Figure 6.7 depicts the final Heat Exchange Network after all necessary adjustments. Cp inequality has been kept in mind while designing this network.

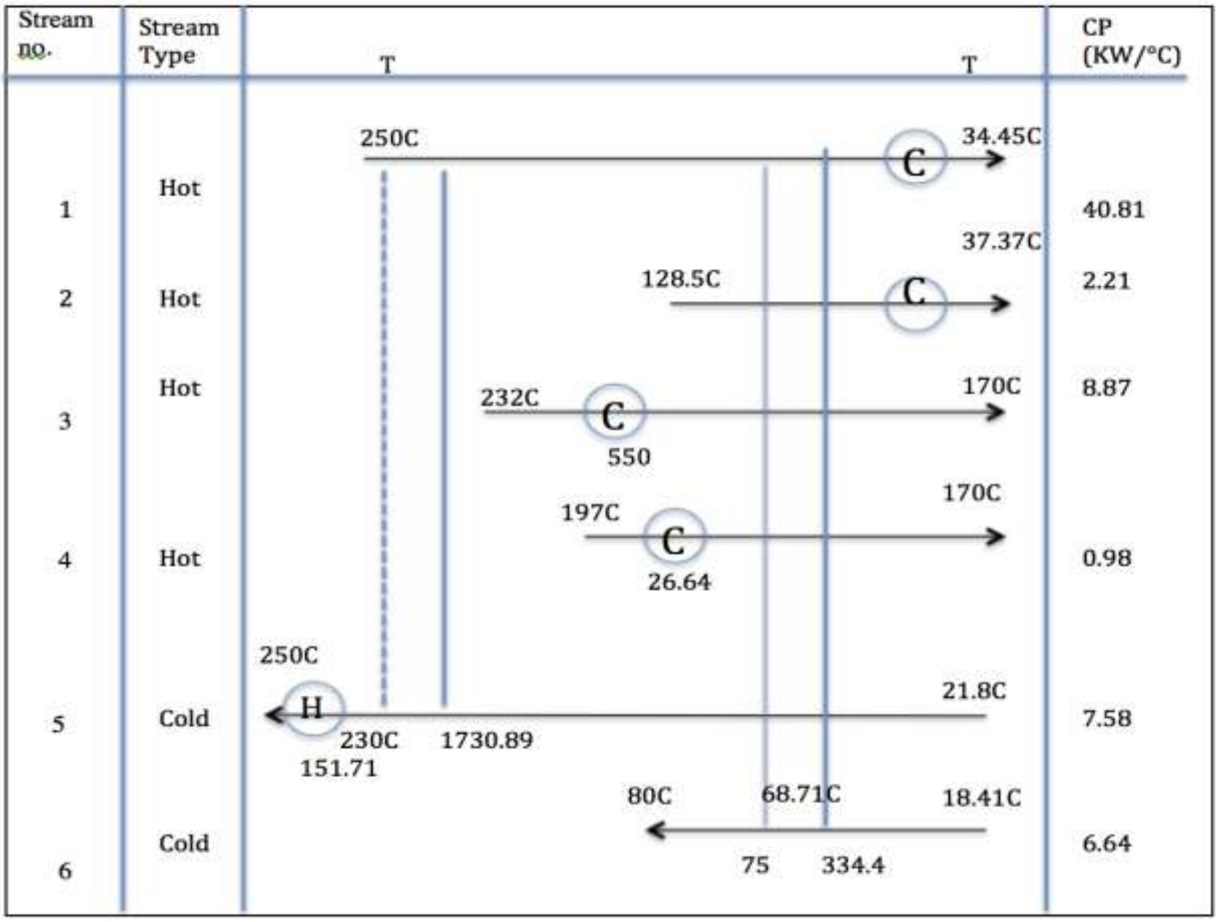


Figure 6.7: Final Heat Exchanger Network

Here, we do not have any loops in our HEN generated so far, so integration by combining HEN's is not possible here. The ΔT_{min} violation and heat balance have been taken care of, and thus Figure 6.7 shows the final Heat Exchange Network which will be used for ASPEN simulation. Details about the type, temperatures and flow rates of utility will be done in the Economic Analysis. Figure 6.8 shows the simplified heat exchanger system showing that there are no loops.

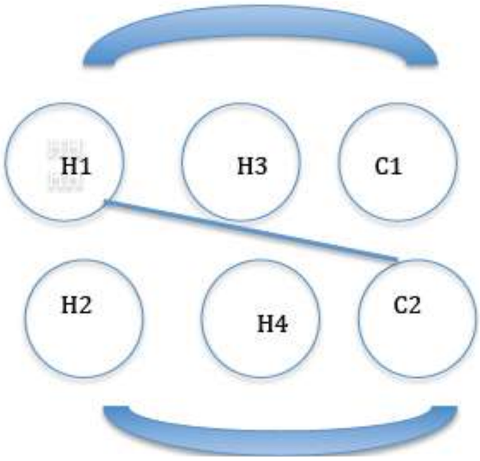


Figure 6.7: Final Heat Exchanger Network

8. Validation by ASPEN Simulation

Figure 7.1 shows the Process Flow Diagram (PFD) generated by the simulator ASPEN. All the streams and equipment (blocks) have been shown and appropriately labelled. The Enthalpy values generated after ASPEN simulation are somewhat deviated from the calculations done in Mass & Energy Balance section. This may be partially because we have taken Cp values to be constant while calculating the respective enthalpies. However, ASPEN, being a scientific simulation program does not make any such assumptions. As the temperature varies across the process, the variation in Cp values leads to variation in Enthalpy values.

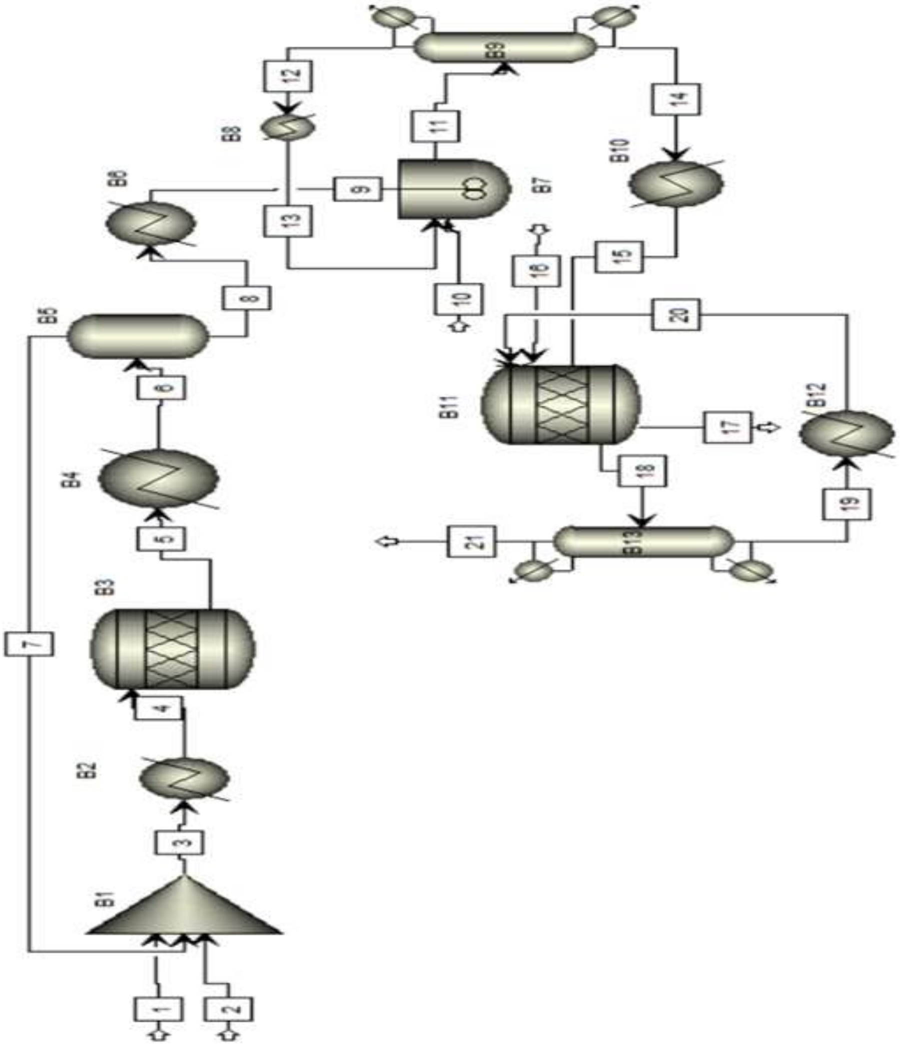
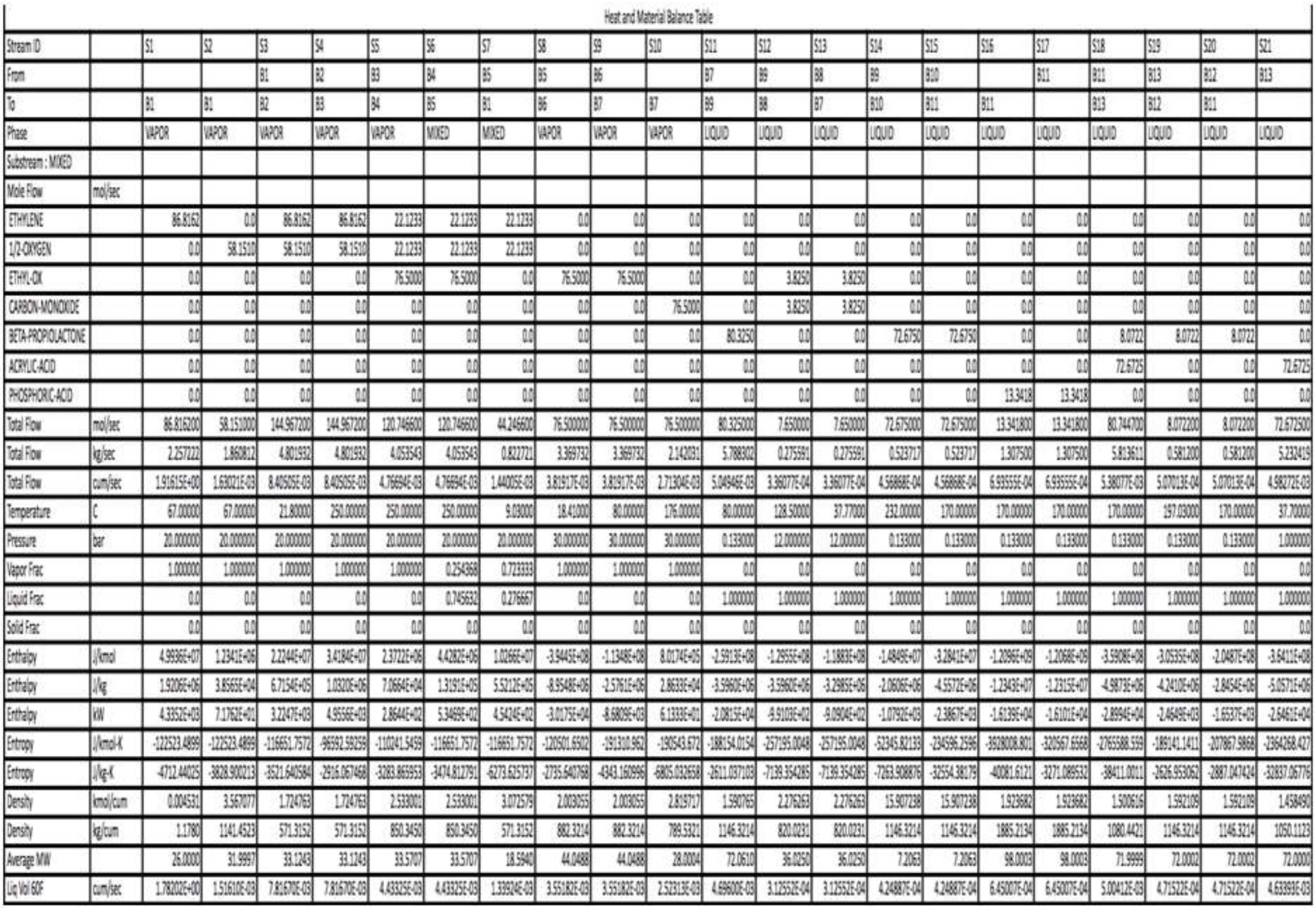


Figure 7.1: Final Process Flow Diagram generated by ASPEN

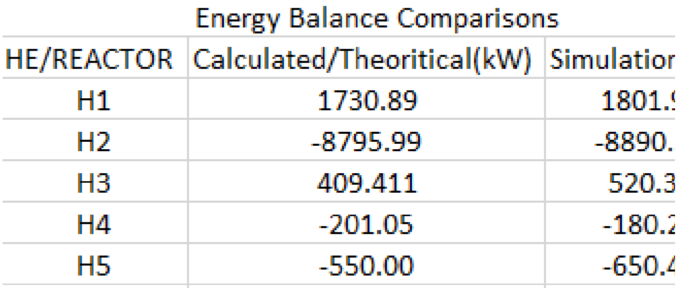


Heat and Material Balance Table

Stream ID		S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11
From				B1	B2	B3	B4	B5	B5	B6		B7
To		B1	B1	B2	B3	B4	B5	B1	B6	B7	B7	B9
Phase		VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	MIXED	MIXED	VAPOR	VAPOR	VAPOR	LIQUID
Substream : MIXED												
Mole Flow	mol/sec											
ETHYLENE		86.8162	0.0	86.8162	86.8162	22.1233	22.1233	22.1233	0.0	0.0	0.0	0.0
1/2-OXYGEN		0.0	58.1510	58.1510	58.1510	22.1233	22.1233	22.1233	0.0	0.0	0.0	0.0
ETHYL-OX		0.0	0.0	0.0	0.0	76.5000	76.5000	0.0	76.5000	76.5000	0.0	0.0
CARBON-MONOXIDE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	76.5000	0.0
BETA-PROPIOLACTONE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	80.3250
ACRYLIC-ACID		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PHOSPHORIC-ACID		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Flow	mol/sec	86.816200	58.151000	144.967200	144.967200	120.746600	120.746600	44.246600	76.500000	76.500000	76.500000	80.325000
Total Flow	kg/sec	2.257222	1.860812	4.801932	4.801932	4.053543	4.053543	0.822721	3.369732	3.369732	2.142031	5.788302
Total Flow	cum/sec	1.91615E+00	1.63021E-03	8.40505E-03	8.40505E-03	4.76694E-03	4.76694E-03	1.44005E-03	3.81917E-03	3.81917E-03	2.71304E-03	5.04946E-03
Temperature	C	67.00000	67.00000	21.80000	250.00000	250.00000	250.00000	9.03000	18.41000	80.00000	176.00000	80.00000
Pressure	bar	20.000000	20.000000	20.000000	20.000000	20.000000	20.000000	20.000000	30.000000	30.000000	30.000000	0.133000
Vapor Frac		1.000000	1.000000	1.000000	1.000000	1.000000	0.254368	0.723333	1.000000	1.000000	1.000000	0.0
Liquid Frac		0.0	0.0	0.0	0.0	0.0	0.745632	0.276667	0.0	0.0	0.0	1.000000
Solid Frac		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Enthalpy	J/kmol	4.9936E+07	1.2341E+06	2.2244E+07	3.4184E+07	2.3722E+06	4.4282E+06	1.0266E+07	-3.9445E+08	-1.1348E+08	8.0174E+05	-2.5913E+08
Enthalpy	J/kg	1.9206E+06	3.8565E+04	6.7154E+05	1.0320E+06	7.0664E+04	1.3191E+05	5.5212E+05	-8.9548E+06	-2.5761E+06	2.8633E+04	-3.5960E+06
Enthalpy	kW	4.3352E+03	7.1762E+01	3.2247E+03	4.9556E+03	2.8644E+02	5.3469E+02	4.5424E+02	-3.0175E+04	-8.6809E+03	6.1333E+01	-2.0815E+04
Entropy	J/kmol-K	-122523.4899	-122523.4899	-116651.7572	-96592.59259	-110241.5459	-116651.7572	-116651.7572	-120501.6502	-191310.962	-190543.672	-188154.0154
Entropy	J/kg-K	-4712.44025	-3828.900213	-3521.640584	-2916.067468	-3283.865953	-3474.812791	-6273.625737	-2735.640768	-4343.160996	-6805.032658	-2611.037103
Density	kmol/cum	0.004531	3.567077	1.724763	1.724763	2.533001	2.533001	3.072579	2.003055	2.003055	2.819717	1.590765
Density	kg/cum	1.1780	1141.4523	571.3152	571.3152	850.3450	850.3450	571.3152	882.3214	882.3214	789.5321	1146.3214
Average MW		26.0000	31.9997	33.1243	33.1243	33.5707	33.5707	18.5940	44.0488	44.0488	28.0004	72.0610
Liq Vol 60F	cum/sec	1.78202E+00	1.51610E-03	7.81670E-03	7.81670E-03	4.43325E-03	4.43325E-03	1.33924E-03	3.55182E-03	3.55182E-03	2.52313E-03	4.69600E-03

Stream ID		S12	S13	S14	S15	S16	S17	S18	S19	S20	S21
From		B9	B8	B9	B10		B11	B11	B13	B12	B13
To		B8	B7	B10	B11	B11		B13	B12	B11	
Phase		LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID
Substream : MIXED											
Mole Flow	mol/sec										
ETHYLENE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
1/2-OXYGEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHYL-OX		3.8250	3.8250	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON-MONOXIDE		3.8250	3.8250	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BETA-PROPIOLACTONE		0.0	0.0	72.6750	72.6750	0.0	0.0	8.0722	8.0722	8.0722	0.0
ACRYLIC-ACID		0.0	0.0	0.0	0.0	0.0	0.0	72.6725	0.0	0.0	72.6725
PHOSPHORIC-ACID		0.0	0.0	0.0	0.0	13.3418	13.3418	0.0	0.0	0.0	0.0
Total Flow	mol/sec	7.650000	7.650000	72.675000	72.675000	13.341800	13.341800	80.744700	8.072200	8.072200	72.672500
Total Flow	kg/sec	0.275591	0.275591	0.523717	0.523717	1.307500	1.307500	5.813611	0.581200	0.581200	5.232419
Total Flow	cum/sec	3.36077E-04	3.36077E-04	4.56868E-04	4.56868E-04	6.93555E-04	6.93555E-04	5.38077E-03	5.07013E-04	5.07013E-04	4.98272E-03
Temperature	C	128.50000	37.77000	232.00000	170.00000	170.00000	170.00000	170.00000	197.03000	170.00000	37.70000
Pressure	bar	12.000000	12.000000	0.133000	0.133000	0.133000	0.133000	0.133000	0.133000	0.133000	1.000000
Vapor Frac		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Liquid Frac		1.000000	1.000000	1.000000	1.000000	1.000000	1.000000	1.000000	1.000000	1.000000	1.000000
Solid Frac		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Enthalpy	J/kmol	-1.2955E+08	-1.1883E+08	-1.4849E+07	-3.2841E+07	-1.2096E+09	-1.2068E+09	-3.5908E+08	-3.0535E+08	-2.0487E+08	-3.6411E+08
Enthalpy	J/kg	-3.5960E+06	-3.2985E+06	-2.0606E+06	-4.5572E+06	-1.2343E+07	-1.2315E+07	-4.9873E+06	-4.2410E+06	-2.8454E+06	-5.0571E+06
Enthalpy	kW	-9.9103E+02	-9.0904E+02	-1.0792E+03	-2.3867E+03	-1.6139E+04	-1.6101E+04	-2.8994E+04	-2.4649E+03	-1.6537E+03	-2.6461E+04
Entropy	J/kmol-K	-257195.0048	-257195.0048	-52345.82133	-234596.2596	-3928008.801	-320567.6568	-2765588.559	-189141.1411	-207867.9868	-2364268.427
Entropy	J/kg-K	-7139.354285	-7139.354285	-7263.908876	-32554.38179	-40081.6121	-3271.089532	-38411.0011	-2626.953062	-2887.047424	-32837.06776
Density	kmol/cum	2.276263	2.276263	15.907238	15.907238	1.923682	1.923682	1.500616	1.592109	1.592109	1.458490
Density	kg/cum	820.0231	820.0231	1146.3214	1146.3214	1885.2134	1885.2134	1080.4421	1146.3214	1146.3214	1050.1123
Average MW		36.0250	36.0250	7.2063	7.2063	98.0003	98.0003	71.9999	72.0002	72.0002	72.0000
Liq Vol 60F	cum/sec	3.12552E-04	3.12552E-04	4.24887E-04	4.24887E-04	6.45007E-04	6.45007E-04	5.00412E-03	4.71522E-04	4.71522E-04	4.63393E-03

Figure 7.3: ASPEN Table



Energy Balance Comparisons

HE/REACTOR	Calculated/Theoritical(kW)	Simulatio
H1	1730.89	1801.
H2	-8795.99	-8890.
H3	409.411	520.3
H4	-201.05	-180.
H5	-550.00	-650.

Figure 7.4: Energy Balance Comparison through ASPEN

Figure 7.5 gives a brief description of each of the streams and blocks in the PFD

Stream Name	Stream Description
S1	Input ethylene
S2	Input oxygen
S3	Mixed ethylene & O2 at 21.8°C
S4	Mixed ethylene & O2 at 250°C
S5	Reactor Product (ethylene oxide,ethylene & O2)
S6	Heated reactor product
S7	Distillate (ethylene & O2) as recycle to B1
S8	Bottoms (ethylene oxide) at 18.41°C
S9	Bottoms (ethylene oxide) at 80°C
S10	Input CO
S11	Product beta-propiolactone
S12	Distillate (ethylene oxide & CO) at 128.5°C
S13	Distillate (ethylene oxide & CO) at 37.77°C
S14	Bottoms (beta-propiolactone) at 232°C
S15	Bottoms (beta-propiolactone) at 170°C
S16	Input phosphoric acid
S17	Purge phosphoric acid
S18	Product (acrylic acid &beta-propiolactone)
S19	Bottoms (beta-propiolactone) at 197.03°C
S20	Bottoms (beta-propiolactone) at 170°C as recycle to B11
S21	Distillate acrylic acid

Block Name	Block Description
B1	Mixer
B2	Shell & Tube Heat Exchanger (delta Tmin = 20°C)
B3	Reactor 1
B4	Shell & Tube Heat Exchanger (delta Tmin = 20°C)
B5	Flash Distillation Column
B6	Shell & Tube Heat Exchanger (delta Tmin = 20°C)
B7	Reactor 2
B8	Shell & Tube Heat Exchanger (delta Tmin = 20°C)
B9	Distillation Column
B10	Shell & Tube Heat Exchanger (delta Tmin = 20°C)
B11	Reactor 3
B12	Shell & Tube Heat Exchanger (delta Tmin = 20°C)
B13	Distillation Column

Figure 7.2: Stream & Block Description Table

9. Economic Analysis

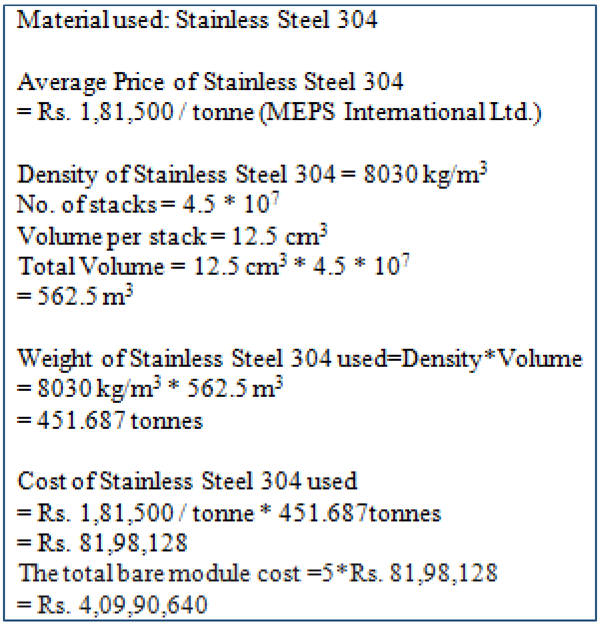
9.1 Total Capital Investment

In this section, we'll calculate the total investments by using the empirical relations involving on-site costs. We assume our on-site cost to include cost of equipment such as the three reactors as well as all the shell and tube heat exchangers. Note that we are ignoring the cost associated with mixers and distillation columns as they are very less as compared to these heavy investments. These calculations are based on the production capacity of our plant i.e., 18,836 kg acrylic acid per hour.

9.1.1 Cost of Reactors

We will calculate the costs of two different types of reactors used –microchannel reactor and CSTR. We assume the material of these reactors to be Stainless Steel 304. For microchannel reactors, we assume that it is composed of 4.5*10^^{7} stacks as per our production capacity. The standard volume per stack is 12.5 cm^3. These, along with other assumptions are stated below. Data has been obtained from MEPS International Ltd(15). Total bare module cost has been taken to account for both direct and indirect costs.

Reactor 1 & Reactor 3:



Material used: Stainless Steel 304

Average Price of Stainless Steel 304
= Rs. 1,81,500 / tonne (MEPS International Ltd.)

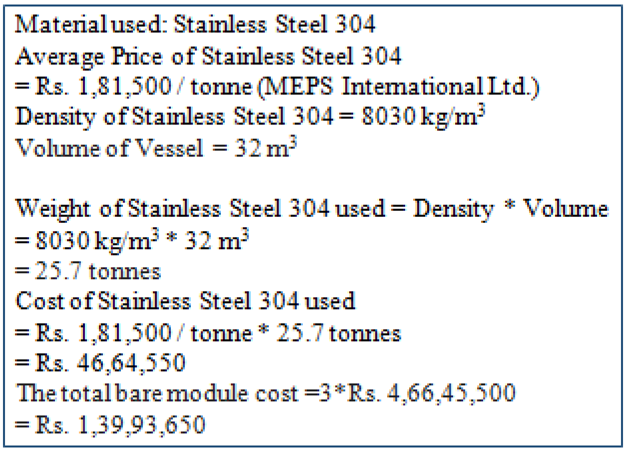
Density of Stainless Steel 304 = 8030 kg/m^3
No. of stacks = 4.5 * 10^7
Volume per stack = 12.5 cm^3
Total Volume = 12.5 cm^3 * 4.5 * 10^7
= 562.5 m^3

Weight of Stainless Steel 304 used=Density*Volume
= 8030 kg/m^3 * 562.5 m^3
= 451.687 tonnes

Cost of Stainless Steel 304 used
= Rs. 1,81,500 / tonne * 451.687tonnes
= Rs. 81,98,128
The total bare module cost =5*Rs. 81,98,128
= Rs. 4,09,90,640

Taking into consideration the large manufacturing costs associated, we choose a bare module factor of 5 (Alaskan Natural Gas to Liquids).

Reactor 2:



Material used: Stainless Steel 304
Average Price of Stainless Steel 304
= Rs. 1,81,500 / tonne (MEPS International Ltd.)
Density of Stainless Steel 304 = 8030 kg/m^3
Volume of Vessel = 32 m^3

Weight of Stainless Steel 304 used = Density * Volume
= 8030 kg/m^3 * 32 m^3
= 25.7 tonnes
Cost of Stainless Steel 304 used
= Rs. 1,81,500 / tonne * 25.7 tonnes
= Rs. 46,64,550
The total bare module cost =3*Rs. 4,66,45,500
= Rs. 1,39,93,650

Considering the medium level of manufacturing costs associated, we choose a bare module factor of 3.

Total Costs of Reactors = 2*4,09,90,640 + 1,39,93,650 = Rs. 9,59,74,930

9.1.2 Cost of Heat Exchanger Costing:

Cost of heat exchangers has been estimated using the area approach. Area calculations have been shown below in Figure 9.1. These calculations are based on our HEN shown above which comprises of 3 heat exchangers and 5 utilities. Calculations for cost of heat exchanger use the following formula:

$A_i = Q/U*LMTD$
$C_b = EXP(8.202 + 0.01506*lnA + 0.06811*(lnA)^2)$
$F_d = EXP(-0.9003 + 0.0906*lnA)$
$F_P = 0.8955 + 0.04981*lnA$
$F_m = 1.4144 + 0.23296*lnA$
Cost of exchanger =Cb*Fd*Fp*Fm

	Stream No	T(in)	T(out)	LMTD	U	Q	A	Cb	Fd	Fp	Fm	Cost of exchanger
H	1	250.0	207.6	74.38	12.68	1730.89	1835.21	191303.1	0.8030	1.2698	3.16507	617365.1
C	5	230.0	21.8									
H	1	207.6	205.7	132.25	12.68	75.00	44.72	10332.1	0.5735	1.0848	2.29977	14783.1
C	6	80.0	68.7									
H	1	205.7	197.5	157.14	12.68	344.40	172.85	24046.0	0.6482	1.1521	2.6147	46957.7
C	6	68.7	18.4									
hot utility	steam	270.0	270.0	28.85	23.78	151.71	221.14	28811.0	0.6629	1.1644	2.6721	59422.4
	5	250.0	230.0									
cold utility	1	197.5	34.5	42.92	31.71	6644.06	4882.31	564213.9	0.8774	1.3186	3.39302	2214737
	water	40.0	30.0									
cold utility	2	128.5	37.4	32.64	31.71	201.39	194.61	26204.9	0.6552	1.1580	2.64233	52541.0
	water	40.0	30.0									
cold utility	3	232.0	170.0	164.63	31.71	550.00	105.37	17147.4	0.6198	1.1275	2.49941	29951.1
	water	40.0	30.0									
cold utility	4	197.0	170.0	148.34	31.71	26.64	5.66	4596.1	0.4756	0.9819	1.8184	3902.8
	water	40.0	30.0									
											Total	30,39,660.3

Figure 9.1: Installation cost of heat exchangers

9.1.3 Total Onsite Cost

Total onsite cost = Cost of reactors + Cost of heat exchangers
= Rs. 9,59,74,930 + Rs. 30,39,660
= Rs. 9,90,14,590
Total capital investment = 2.36 * Total onsite cost
= 2.36 * Rs.9,90,14,590
= Rs.22,37,72,973

Note that since we do not have any additional data to account for other costs such as cost of land, machinery, etc. Thus, total capital investment is obtained by multiplying the onsite cost by a factor of 2.36 (empirical correlation)

9.2 Total Production cost

Production cost or the operating cost is assumed to consist of only the utility cost as it is the only quantity in this category that can be calculated from the data available because there is no such empirical correlation for estimating the other components of operating cost like rent, labour, maintenance etc. Also, we are using water at room temperature as the cold utility here. Thus, we can ignore any cost incurred for cold utility.

Total Production Cost = Manufacturing Cost = Utility cost
Cold Utility = 0
For Hot Utility:
Latent Heat of Steam = 2260kJ/kg
Q_H = 151.71 Kw
Steam requirement = 0.0671 kg/s
Cost of steam = Rs. 4/kg
Cost of steam generation = Rs.4/kg*0.0671 kg/s*3600 s/hr * 24 hours/day * 365 days/year
= Rs. 84,64,262 per year

9.3 Annual Profit

Annual Profit has been obtained on the basis of the production capacity of the plant, i.e.,18,836.71kilogram of acrylic acid/hr. Also note that we have assumed that the plant runs for an average of 20 hours per day and 300 days per year (To account for maintenance, cleaning and shut down time of the plant). Annual Profit after tax comes out to be Rs. 4,23,34,667.

Annual Profit = Total Sales – Manufacturing Cost
Economic Potential = Product Cost – Feed Cost
= Rs. 8.521/mol
= Rs. 0.61/kg Acrylic Acid produced

From the stream data,
Acrylic acid produced
= 18836.71 kg/hr
= 18836.71kg/hr * 20hr/day * 300 days/year
= 11,30,20,260 Kg/year

Total Sales = EP*Sales per annum
= 11,30,20,260 Kg/year * Rs. 0.61/kg
= Rs. 6,89,42,358

Annual Profit = Rs. 6,89,42,358 - 84,64,262
= Rs. 6,04,78,097

Assuming 30% tax,
PAT = 0.7*6,04,78,097
= Rs. 4,23,34,667

9.4 Profitability Analysis

A cash flow analysis of 10 years is done in Figure9.2. We have assumedthe first year to be no profit period to account for the time for setting up the plant.

Year	Total Capital Investment	Annual Cash Flows	Cumulative CF (Rs.) = = ∑ {After-Tax Cash Flow / (1+i)^t} - Initial Investment			
			i=0%	i=6%	i=8%	i=10%
1	22,37,72,973		-223772973	-223772973	-223772973	-223772973
2		4,23,34,667	-18,14,38,306	-183834608	-184574207.3	-185286912.1
3		4,23,34,667	-13,91,03,639	-146156905	-148279053.8	-150299584
4		4,23,34,667	-9,67,68,972	-110611902	-114672430.2	-118492922.1
5		4,23,34,667	-5,44,34,305	-77078881	-83555186.16	-89577774.89
6		4,23,34,667	-1,20,99,638	-45443955	-54742923.15	-63291277.45
7		4,23,34,667	3,02,35,029	-15599685	-28064901.85	-39394461.59
8		4,23,34,667	7,25,69,696	12555286	-3363030.264	-17670083.54
9		4,23,34,667	11,49,04,363	39116580	19509073.05	2079351.058
10		4,23,34,667	15,72,39,030	64174404	40686946.49	20033382.51

Figure 9.2: Cumulative Cash Flow for 10 years

9.4.1 Payback Period

Payback period does not consider the Time Value of Money (TVM) and the principles of discounting/ compounding. However, it is agood measure to compare similar projects. Payback period is that time when cumulative cash flow becomes 0 considering no interest rate (i=0). Payback period for this project comes out to be 6.28 years (by interpolation), which can be clearly seen from Figure 9.3.

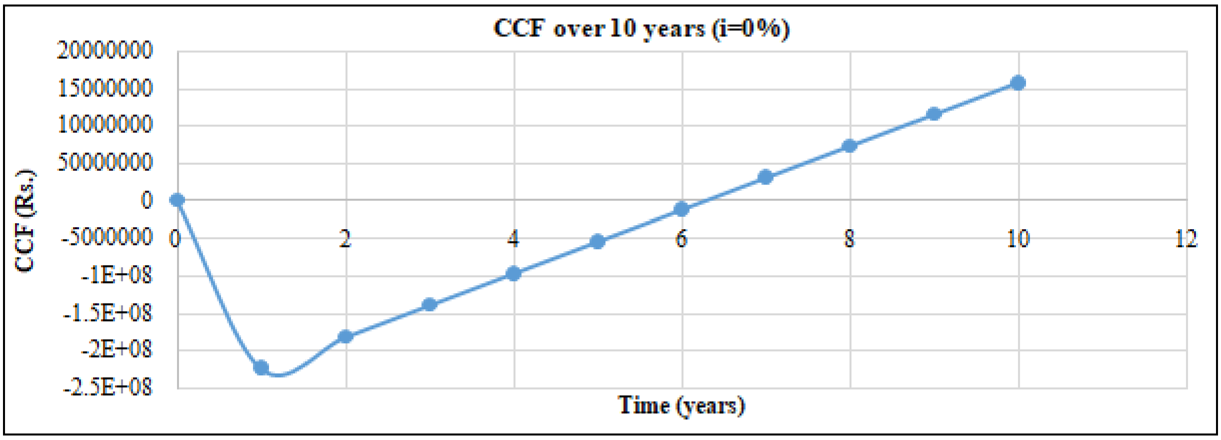


Figure 9.3: Cumulative Cash Flow plot for 10 years for i=0%

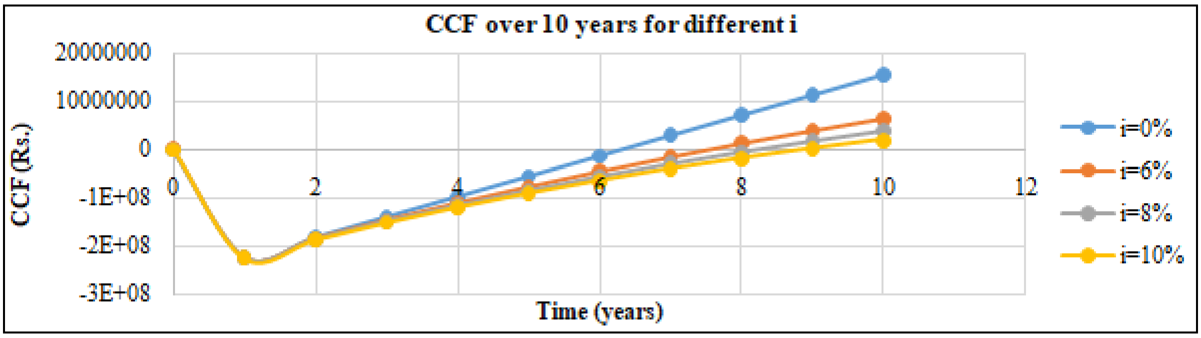
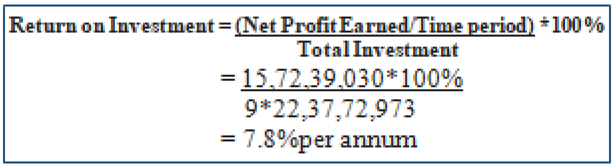


Figure 9.4: Cumulative Cash Flow plot for 10 years for different interest rates

9.4.2 Return on Investment

Return on investment (ROI) is a direct measure of profitability of the investors. For our project, ROI comes out to be 7.8% which is good enough from an investor's point of view. Calculations have been shown below:

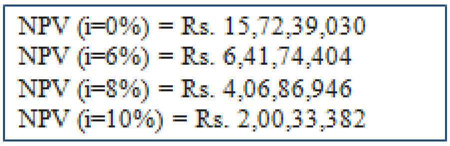


Return on Investment = (Net Profit Earned/Time period) *100%
Total Investment
= 15,72,39,030*100%
9*22,37,72,973
= 7.8%per annum

9.4.3 Internal Rate of Return and Net Present Value

The formula for Net Present Value (NPV) is given below. The corresponding calculations have been shown in Figure 9.2. Note that the appropriate discounting corresponding to different interest values has been done while calculating the Cumulative Cash Flows for interest rates of 6%,8%,10%.

NPV = NPV= ∑ {After-Tax Cash Flow / (1+i)^t} - Initial Investment



NPV (i=0%) = Rs. 15,72,39,030
NPV (i=6%) = Rs. 6,41,74,404
NPV (i=8%) = Rs. 4,06,86,946
NPV (i=10%) = Rs. 2,00,33,382

IRR is that value of interest rate when NPV becomes zero. Considering payback period of approximately 7 years, NPV becomes 0 somewhere between 0% and 6%. We estimate the IRR value using interpolation.

IRR = 6% + (-15599685/(3,02,35,029+15599685)*6 = 3.96%

9.5 Comment on Economics

From above economic analysis, we observe that the net present value of our plant is Rs. 4,06,86,946 for a rate of return of 8% over a period of 10years, which is considerably high. This rate of return gives the true idea of our overall profits since it is very close to the ROIwhich comes out to be 7.8%. The total capital investment for the project is Rs. 22,37,72,973 while the operating cost is Rs. 84,64,262 per year. We can compare only those projects with this project that has similar capital investment. From calculation, it is shown that it is a profitable method because we get a positive NPV at the end of 10 years and also the payback time is 6.2 years, which is quite less. All the above analysis is done assuming salvage value to be zero which is not applicable in real life scenario. If we also include the salvage value, a significant portion of fixed cost will be recovered which will increase our profit over 10 years and our project with become more profitable with time.Another important factor worth noticing is the high annual profit as compared to the low operating cost which makes the project even more profitable in the long run once the initial investment is paid back. Hence, from our analysis, this project comes out to be highly profitable.

10. Conclusion

This report provides a complete analysis of an alternate method of production of acrylic acid and highlights the advantages it has over the conventional methods. It covers all the aspects that one needs to consider while setting up a new production plant and thus can act as a suitable guide. Having established the need and profitability prospects of the setup of acrylic acid using ethylene over other possible mechanisms of production in the initial sections, the original hypothesis was proved correct. Following this, the various process flow chart and the process parameters were stated which emphasized the various equipment and raw materials required for the project and gave a rough estimate of the production capacity of the plant. An in-depth analysis was done by using Mass and Energy Balances which can be used to understand the material and power requirements for the process in terms of flow rates and utilities. The ideal break-up of this power requirement through heat exchangers and hot and cold utilities was done and an overall Heat Exchanger Network was drawn. ASPEN Simulations were run which computed and verified the more intrinsic values like all flow rates, composition, entropy and enthalpy values for the different process streams. The Process Flow Diagram gives a crisp representation of the overall design of the plant. Our HEN has 3 heat exchangers, 4 cold utilities and 1 hot utility. Note that this is the MER design which uses minimum utilities and maximum heat recovery. The economic analysis shows that the project has very heavy initial investment of over Rs. 22 crores and thus makes it feasible for only those who have access to such heavy investment at a reasonable interest rate. However, the high rate of return (7.8%) and low payback period (6.3 years) makes our investment worthwhile. Also, the annual profits are very high as compared to the operating cost which implies that in the long run, when the high initial investment is already recovered, our project will be highly profitable.

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