
FINAL YEAR DESIGN PROJECT

FIRST DRAFT REPORT

(MID TERM-I)

PRODUCTION OF ACRYLIC ACID

&

SCOPE OF IMPLEMENTATION IN INDIA

REPORT BY:

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1 ABSTRACT

A preliminary process design and economic analysis of acrylic acid production by partial oxidation of propylene is presented. The design for the production plant consists of an oxidation reactor, a two stage absorber unit (consisting of packed tower and tray tower), a centrifugal extractor (with extract and raffinate strippers), acrylic acid recovery system which is designed using waste gas incinerator and distillation equipment. The reactor operates at 300°C with an incoming propylene feed of 120.71 kgmole/hr and an incoming oxygen flowrate to maintain the ratio of 2.25:1 in the reactor feed. From the design, the maximum amount of acrylic acid produced is 6504.5 kg/hr. The separation section of the plant allows for 97% extraction.

Two solvents namely (i). 1-Butanol (referred to as Solvent 1/A) and (ii). Toluene (referred to as Solvent 2/B) were compared extensively and it was concluded that 1-Butanol would be the best solvent for extraction. The design also included an incinerator to extract heat from waste gases to provide the over-all utility of the plant. It is estimated that the plant will generate a net revenue of \$27268368/year (before tax deduction; assumed to be 50%) from the sales of both the product and the by-products (assuming the acrylic acid is sold at a minimum of \$0.952/kg). The above profit is inclusive of the sub-charges such as (i). Man-power \$10,356,32/year, (ii). Maintenance charge of \$2818047/year, (iii). Support charges of \$702686/year and (iv). Utilities of \$2990632/year (all the charges quoted in the report are foundationally built upon the national labour norms).

Though the data presented in this report has been extracted from credible sources, it is recommended that:

- 1) The process study be compared to the existing operational plant.
- 2) The impact of reactor conditions be understood by doing the process design for each of the reactor conditions (i.e. different temperature conditions).
- 3) 1-Butanol is chosen; more rigorous physical data be considered instead of just equilibrium data (further research in the field required) for a better comparison.

2 LETTER OF DECLARATION AND GRATITUDE

Dr. Harsh Pandey,

Assistant Professor,

Department of Chemical Engineering,

Manipal University Jaipur,

Respected Sir,

Enclosed is our final year design project titled: "Production of Acrylic Acid via partial oxidation of Propylene". In this project report, we explored the potential of implementation of production of Acrylic Acid domestically. 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°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.

Though other reactions/methods of production were also explored, this process is selected as the desired method of production since (i). Other methods of AA production have not been studied as extensively as this one, (ii). The feed required for partial oxidation i.e. Propylene is more commonly available and (iii). Basic nature of the process allowed for extensive study hence better understanding consequently better results. The report includes an extensive study into the topic regarding the chemical engineering and economic analysis and feasibility.

To the best of our knowledge the work presented in this study has been done by considering literature, data and financial reports from credible sources. Moreover, we would like to acknowledge your imperative support and input through the entire project.

Thanking You,

Regards,

AKSHAY SRIVASTAVA

SHUBHAM PARUTHI

KRITI SINHA

JAI GUPTA

TABLE OF CONTENT

1	ABSTRACT	2
2	LETTER OF DECLARATION AND GRATITUDE	3
3	INTRODUCTION.....	8
4	MARKET RESEARCH.....	9
4.1	DATA ANALYSIS.....	9
5	LITERATURE SURVEY.....	10
5.1	(I). Production via Ethylene:	10
5.2	(II). Propylene to AA:	14
5.3	(III). Acrylic Acid by Sequential Dehydration and Oxidation of Glycerol: ...	16
5.4	(IV). Acrylic acid from propane oxidation:.....	17
5.5	(V). Acrylic acid from Lactic Acid:.....	17
6	PROCESS DESCRIPTION	18
6.1	DETERMINATION OF OXIDATION REACTOR OPERATING CONDITIONS	23
6.2	INTRODUCTION:	24
6.3	DESIGN BASIS:	24
6.4	RAW MATERIAL COSTS:.....	25
6.5	UTILITY COSTS:	26
6.6	MISCELLANEOUS COSTS:	26
6.7	CALCULATION BASIS:	26
6.8	PROCESS ANALYSIS:.....	27
7	CONCLUSION:	36
8	RECOMMENDATIONS:.....	37
9	REFERENCES.....	38
10	APPENDIX: DESIGN CALCULATIONS	39
10.1	APPENDIX A: REACTOR.....	39
10.1.1	REACTOR COST	39
10.1.2	MATERIAL BALANCE AROUND REACTOR:.....	41
10.1.3	HEAT OF REACTION.....	42
10.1.4	PROCESS FLOW AND MATERIAL GRID:	44
10.2	APPENDIX B: ABSORBER QUENCHING SECTION-I	45
10.2.1	ABSORBER QUENCHING SECTION ,T-101.....	45
10.3	APPENDIX C: ABSORBER SECTION-II.....	50

10.3.1	MATERIAL BALANCE:	50
10.4	APPENDIX D: INCINERATOR:	52
10.5	APPENDIX E: EXTRACTION.....	56
10.5.1	Case 1: Solvent A	57
10.5.2	Case 2: Solvent B	61
10.6	ECONOMIC COMPARISON OF 2 CASES	67
10.6.1	Case 1: Solvent A	67
10.6.2	Case 2: Solvent B.....	73
10.7	APPENDIX F: ACRYLIC ACID RECOVERY	78
10.7.1	Acrylic Acid Recovery:	78
10.7.2	Material Balance:.....	81
10.7.3	T106 –Sizing	87
10.7.4	T107 Optimization:.....	96
10.8	APPENDIX G: MISCELLANEOUS EQUIPMENT SIZING & COSTING	104
10.8.1	Air Compressor.....	104
10.8.2	Reactor waste heat boiler E101	105
10.8.3	Cost of boiler water.....	105
10.8.4	E-102 X-chgr for T-101	105
10.8.5	E-105 Waste heat boiler for incinerator.....	106
10.8.6	T102 – Absorber.....	109
10.8.7	Incinerator	110
10.8.8	T-101 Packed Absorber	111
10.9	APPENDIX H: Final Process Flow Diagram and Material Balance for Base Case	113
10.10	APPENDIX I: ECONOMICS	115
10.10.1	Equipment List	115
10.10.2	Utility List	116
10.10.3	Raw Material Cost:	116

List of Tables:

Table 1.....	25
Table 2.....	26
Table 3.....	26
Table 4.....	33
Table 5.....	34
Table 6.....	35
Table 7.....	40
Table 8.....	42
Table 9.....	45
Table 10.....	45
Table 11.....	46
Table 12.....	46
Table 13.....	47
Table 14.....	47
Table 15.....	47
Table 16.....	50
Table 17.....	52
Table 18.....	52
Table 19.....	52
Table 20.....	53
Table 21.....	56
Table 22.....	56
Table 23.....	57
Table 24.....	70
Table 25.....	70
Table 26.....	72
Table 27.....	73
Table 28.....	77
Table 29.....	77
Table 30.....	83
Table 31.....	85
Table 32.....	86
Table 33.....	86
Table 34.....	87
Table 35.....	89
Table 36.....	90
Table 37.....	90
Table 38.....	92
Table 39.....	93
Table 40.....	93
Table 41.....	95
Table 42.....	96
Table 43.....	96
Table 44.....	97
Table 45.....	Error! Bookmark not defined.

Table 46.....	100
Table 47.....	100
Table 48.....	100
Table 49.....	103
Table 50.....	103
Table 51.....	106
Table 52.....	109
Table 53.....	109
Table 54.....	110
Table 55:.....	111

List of Figures:

Figure 1.....	8
Figure 2.....	15
Figure 3.....	25
Figure 4.....	Error! Bookmark not defined.
Figure 5: Process schematic	44
Figure 6.....	61
Figure 7.....	67
Figure 8.....	67
Figure 9.....	80
Figure 10.....	82
Figure 11.....	114

3 INTRODUCTION

Acrylic acid ($\text{CH}_2=\text{CHCOOH}$) is the simplest unsaturated carboxylic acid and exists as a colourless liquid with pungent smell at standard temperature and pressure ($T = 298\text{K}$, $P = 1\text{atm}$). 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:



Figure 1(A): Supply Demand Gap in the national market

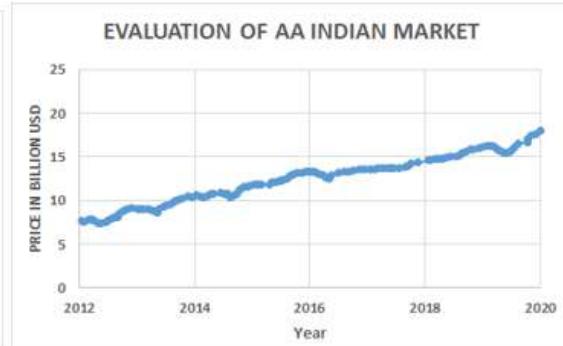


Figure 1(B): Evaluation of AA Indian market.

Figure 1

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. 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.

4 MARKET RESEARCH

The market research was conducted by performing the analysis of the Indian market of Acrylic Acid for the past 10 years. The data was collected from the commodity market performance of acrylic acid published by *ECONOMIC TIMES* until 15th January, 2020. The data was then plotted in excel and the future trends were predicted by data extrapolation. The data that was plotted is as shown in Figure 1(a) and (b). The following deductions can be made about the data set:

4.1 DATA ANALYSIS

- As India is a developing country, the demand for super-absorbent polymers has been increasing. The market for SAPs in India has been increasing at an estimated CAGR of 9.4% for the past decade (a relatively high value). This is attributed to increasing shift in the import v/s domestic production mentality. Dropping labour charges yet rising STEM population in India (Science, technology, engineering and mathematics) has further enhanced the national production of products such as Diapers and sanitary pads. This rise in the demand was constant until 2016.
- There was a rise in the supply v/s demand gap in the market in the late 2016 and then a sudden decrease in the same due to demonetization of the Rs.1000 and Rs.500 notes in the nation which contributed significantly to the sink of the national GDP by ~1.6%.
- The increase in the supply v/s demand gap in the market has been constant up until the late 2019.
- In 2020, the demand for acrylic acid has increased significantly due to increased face mask production (made from polymers made from Acrylic Acid). This is attributed to the pandemic of ncov-2019 virus.
- The evaluation of Acrylic Acid Market size in India has been constantly increasing at a due to the fluctuations in the price of the end product.

It may be concluded that there shall be a sudden surge in the demand of the product until the fluctuations in the market due to ncov-19 pandemic stabilize over time, but even then the demand is increasing as evident from the previous data.

5 LITERATURE SURVEY

The earliest synthesis of acrylic acid occurred in 1843 via the oxidation of acrolein. Since 1927, the commodity chemical has been commercially available through many different production means, such as the following:

- Ethylene cyanohydrins process
- Reppe Process
- β - Propiolactone Process
- Acrylonitrile hydrolysis.

There is also ongoing research on microbe development to produce 3 – hydroxypropionic acid by fermentation, followed by a dehydration reaction to form acrylic acid. There are two main reaction path ways for the production of acrylic acid. These are:

5.1 (I). Production via Ethylene:

Ethylene will be oxidized to its epoxide, which will then be carbonylated to β – propiolactone. The propiolactone intermediate will then be treated with a concentrated phosphoric acid to allow for rearrangement to acrylic acid. The project is envisioned as three main reaction steps including ethylene oxidation, carbonylation of ethylene oxide, and acid catalyzed rearrangement of β -propiolactone (Allen et al, 2012) [2].

Step 1: Oxidation of Ethylene



The most common method for making EO entails a vapor-phase, direct oxidation of ethylene with high purity oxygen in a tubular reactor. The process utilizes silver catalyst on α -alumina support in the form of rings or lobed shapes to increase the surface area. Trace amounts of other elements such as Cs and Re are added to the catalyst to increase selectivity to EO. The reaction occurs in the presence of recycled CO_2 and inert gases including methane, nitrogen, and argon, inerts deliberately included to bring ethylene and oxygen out of the flammability region and terminate free-radicals, thus preventing a runaway reaction. Inerts also help the reaction conditions by increasing heat capacity of the stream and improving heat removal from the exothermic reaction, helping to maintain a smaller temperature profile. Trace amounts of chlorinated compounds are added continuously to the reactor feed to inhibit the undesired combustion of ethylene. The epoxidation of ethylene generally operates at a selectivity of 80-90% based on the industrial grade silver α -alumina catalyst used. Increasing the temperature increases the reaction rate, but decreases the selectivity, Thus, in an effort to maximize selectivity, the per pass ethylene conversion is kept between 10-

20% and the reactor temperature is generally maintained around 240-290°C and 15-25 bar. Heat removal is important to maintain selectivity and prevent hotspots from occurring. Reactors are generally oil or boiling water cooled to maintain a temperature within the acceptable operating range. EO is recovered from the reactor product by absorption in process water to prevent the retention of other gases and light components. It is then stripped from the water, which includes removing water and trace amount of formaldehyde and acetaldehyde. EO is then sent to a purification section where the water and EO mix are separated, resulting in an EO stream around 99.5% purity. The rest of the components need to be recycled due to the low per pass conversion of ethylene (around 10-

20%) to prevent the waste of ethylene. CO₂ needs to be removed to prevent buildup in the system and minimize the purge loss. The presence of other gases with similar molecular weight and polarity such as methane, formaldehyde, and argon makes this separation difficult. Thus, the unreacted ethylene and oxygen, by-product CO₂, and inert gases not absorbed are then sent to a CO₂ scrubbing unit that typically uses MEA to scrub the gas or employs the Benfield Process, owned by UOP, which uses K₂CO₃ as a reagent to react with the CO₂ to produce HKCO₃. The HKCO₃ is then decomposed to CO₂ which is then disposed of. Older versions of EO plants use air, but most air-

based plants have been converted to cryogenic oxygen. This generally increases the selectivity and decreases the amount of nitrogen in the system, thus reducing the size of equipment and permitting a more concentrated reactor inlet feed which increases the selectivity to EO. These advantages generally outweigh the costs of providing and handling pure oxygen. This project proposes the use of a microchannel reactor for the oxidation reaction. Microchannel reactors offer many advantages over a tubular reactor. In a tubular reactor, the massive amount of heat can cause hot spots and heat removal problems which can lead to bad selectivity or runaway, considering that while epoxidation is -105 kJ/mol, total combustion is around -

1250 kJ/mol. As described by Kestenbaum, et al [3], the microchannels are around 500μm by 50μm. This leads to an enhanced surface-to-volume ratio and enables effective heat management. Steif et al [4] showed that microchannels could be controlled to a maximum temperature gradient of 1°C, which would be very beneficial for selectivity. Also, relatively large ratio of surface area acts as a natural free-

radical terminator, so inerts such as nitrogen and argon would not have to be added for the reaction. If an explosion were to occur, microchannel reactors are strong enough to withstand it and continue operation. In theory, the use of a microchannel reactor system would allow for the process to occur without inerts or to allow for operation in the flammable region. Although microchannel reactors are not presently used in commercial EO production, we chose this method in order to obtain the most favorable economics for the ethylene-

based process. In order to minimize reactor volumes, pure oxygen is utilized throughout the process.

Step 2: Carbonylation of Ethylene Oxide:



The ethylene oxide product from the first step is combined with carbon monoxide obtained from the gulf coast in the second step to produce β -propiolactone. Due to its toxicity, the propiolactone will not be isolated. This is a liquid phase, catalyzed reaction with pressure maintained through the use of CO. The catalyst used is $[Co(CO)_4][TPP]$ and the amount of catalyst was determined from the literature to be 15 mM. Church et al. (2006)^[5] have shown that the rate does not vary with the amount of propylene oxide or CO but does vary linearly with the amount of catalyst. Since the structure of propylene oxide is very similar to that of ethylene oxide, a technical assumption was made that the order of kinetics would be similar. However, it should be noted that with a more complex form of epoxide, such as 4,5-

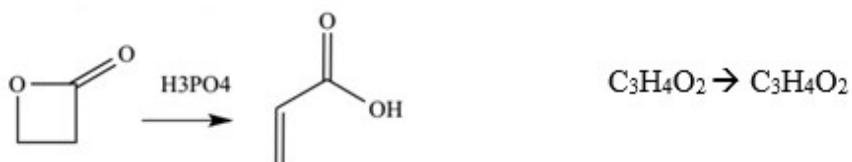
octane oxide, have reported first order carbonylation kinetics (Mulzer et al, 2014, Rowley, 2007)^{[6][7]}. Although Allen et al. (2012)^[8] have presented some data for absorbance of ethylene oxide, acetaldehyde, and β -

propiolactone, the lack of clarification of how the absorbance varies with the concentration did not allow us to make conclusions about the kinetics. Another technical assumption relates to the stability of the catalyst at higher temperatures. As the patent results have shown, the catalyst maintains its stability and does not decompose at 80 °C. The promoter TPP has a boiling point of over 300 °C, and it bonds to metal carbonyl complexes. Therefore, it was assumed to be stable under high temperatures such as 240°C. Production of β -propiolactone from ethylene oxide is a relatively new process, and to our knowledge there are no plants whose operations involve using this reaction. As propylene is the traditional precursor to acrylic acid production, and production from propylene involves only oxidation steps, few similarities are found in this process and the traditional process as propylene as cursor. However, as this reaction is a carbonylation process, it bears many similarities with other processes such as the production of acetic acid. The Monsanto process is a metal carbonyl catalyst-

catalyzed reaction that produces acetic acid from methanol, was later improved upon by BP. A typical acetic acid plant that employs the Monsanto process has a single reactor followed by a series of distillation columns to purify the acetic acid, and the solvent, catalyst, and unused reactants are recycled back into the reactor (Asaoka et al, 1994, Souma, 1991)^{[9][10]}. This process usually takes place under 150-200 °C and 30 bar and has high acetic acid selectivity (Van Leeuwen and Claver, 2001)^[11]. There are key differences between the acetic acid process and the production of β -propiolactone. The former is first order, while the latter is assumed to be zeroth order. The operating temperature of β -propiolactone production is much lower, while the pressure is similar. Because the reaction i

s zeroth order, a large amount of production per year may necessitate parallel CSTRs. A total of 677.57 lb/hr of sulfolane and 2.173 lb/hr of catalyst are recharged into the system to replace the purged sulfolane in the stoichiometric amount. This purged stream will contain, in addition to sulfolane, undistilled β -propiolactone and other side products such as succinic anhydride. Many of these side products, especially high-boiling compounds like succinic anhydride, are difficult to remove fully from sulfolane (Fumagalli, 2001) [12]. Because purchasing the replacement sulfolane and catalyst will be expensive, the purged sulfolane can undergo batch distillation in regular intervals to remove impurities and only leave sulfolane and catalyst. As the sulfolane will need to be heated to high temperatures, the assumption that the catalyst will be stable at high temperatures above 350 C must be tested in the laboratory.

Step 3: Acid-Catalyzed Rearrangement of β -Propiolactone:



Phosphoric acid and propiolactone are mixed in the third step to induce a homogeneously acid catalyzed rearrangement yielding the acrylic acid product. The reaction is carried out in the liquid phase under vacuum at high temperature. The reactor is initially charged with the requisite amount of phosphoric acid catalyst. The catalyst is maintained in the liquid phase by adding trace amounts of water when the viscosity of the acid increases significantly.

As the reaction proceeds, acrylic acid vapors are produced at relatively the same rate as the β -propiolactone feed rate. At this point, mono methyl ether hydroquinone (MEHQ) is added as an inhibitor to prevent the polymerization of the acrylic acid product. In a similar process, which is no longer used, ketene is first reacted with formaldehyde in the presence of a Friedel Crafts catalyst, such as aluminum trichloride, to form β -propiolactone. The lactone is then treated in a process similarly to that described above, with the exception of the MEHQ inhibitor. In this case a copper inhibitor was used.

This process was popularized by Celanese during the years 1957 to 1974 and produced 35,000 tonnes of acrylic acid per plant (Arpe 2010) [13]. Assumptions were made regarding the amount of phosphoric acid catalyst required in the final rearrangement step. As discussed by Schnizer and Wheeler, Phosphoric acid was supplied at a rate so that at least ten percent of the β propiolactone feed rate was present at all times inside of the reactor. This process uses a per pass conversion based on β -propiolactone of 0.97 with an overall conversion of 0.995. Again as discussed by Schnizer and Wheeler, the production rate of any side products were assumed to be negligible.

Reaction kinetics were considered to be first order in lactone, as is typical with most other acid catalyzed reactions. The feed rate of β -propiolactone listed in patent 3,176,042 was 0.116 grams per gram H₃PO₄ per hour. This amount of phosphoric acid requires a reactor volume of 54,000 gallons. Because the example listed in the patent was performed at the bench scale, it was suggested a feed rate of 0.9 gram per gram H₃PO₄ per hour for our scaled up process, provided adequate mixing (personal correspondence).

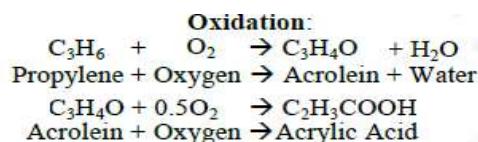
5.2 (II). Propylene to AA:

There are many different alternatives to producing acrylic acid industrially, and as a result many patents exist for these variations. However, the most common method is through the partial oxidation of propylene. The reaction is a two-step oxidation mechanism in which to propylene is

- (i). First oxidised to acrolein and
- (ii). Conversion to acrylic acid

Each of the above is usually under separate catalyst and operating conditions. 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 310-350C 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 [14].



In order to operate a partial oxidation of hydrocarbon safely, the composition of the feed stream must be carefully considered. With high composition, the oxidation can produce large amount of heat making the reaction very dangerous. To overcome this, current design in the industry use fluidized bed reactor which provide isothermal conditions, allowing stable operations [16]. Large amount of steam is also supplied to the reactor for safety reasons. The PFD of the entire process can be found in Figure 1.

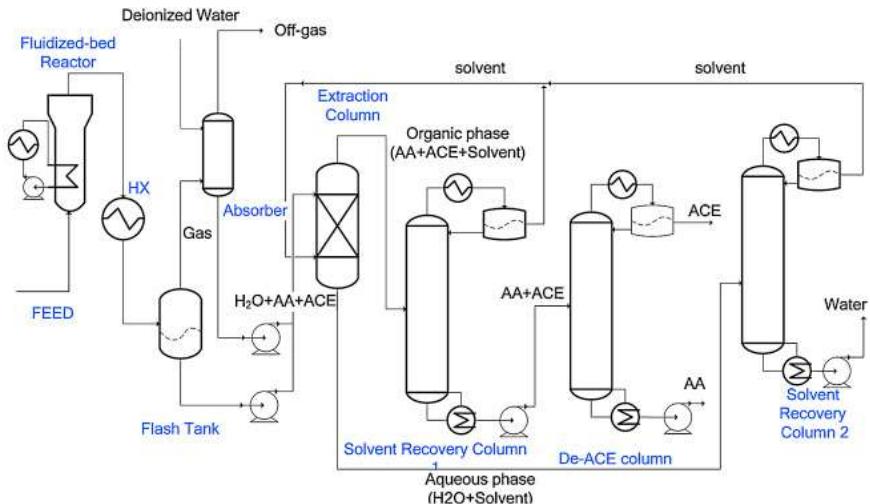


Figure 2

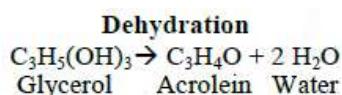
Propylene steam and compressed air are combined and fed to the reactor initially. The reason that air is combined with propylene and steam is to provide the oxygen necessary for oxidation but also minimise side reactions producing carbon dioxide as feeding carbon dioxide will drive equilibrium side reactions in a favourable direction. This stream leaving the reactor must be cooled promptly in order to avoid further oxidation reactions. To accomplish this, the assembly has a quenching tower where it is cooled by a recycle stream of dilute aqueous acrylic acid. The cooled products are then sent to an off-gas absorber where additional acrylic acid is recovered and the major by product is removed. Often DMW is sent to this tower. Vapor leaving this tower contains all unreacted nitrogen, oxygen, carbon dioxide, water and propylene from the reactor as well as trace AA and acetic acid. The off-gas is discharged from the process as waste and sent to incinerator. Product of this column is a dilute acrylic acid solution of is combined with cooling recycled stream described previously and the remainder being sent further into the process. Next the dilute aqueous acidic solution is sent to an acid extractor column to remove the acid fraction from the water. Many solvents can be used to accomplish this so long as they have higher solubility for acrylic acid and lower solubility for water. The most common used in this process is diisopropyl ether [15]. The reason for a liquid-liquid extractor is used prior to purifying the acrylic acid is to achieve greater difference in boiling point between the solvent and acrylic acid. The boiling point of waste is 100°C, while boiling point of acrylic acid is 141°C. To achieve greater separation during purification diisopropyl ether is used, having a boiling point of 69°C. To avoid large solvent cost, diisopropyl Ether is recycled from Waste water tower later in the process. The bottom product from the column consisting mostly of water, is sent to waste water tower where perfect separation between water diisopropyl ether and acetic acid is achieved. The water from Waste tower contains trace amount of Acetic Acid and is sent to waste treatment while the diisopropyl ether and acetic acid in the overhead recycled back into the process. The top product from the Extractor containing tie isopropyl Ether and acrylic acid solution is sent to a recovery Tower. The overhead product containing mostly isopropyl Ether and some water is returned to the extractor, and the bottom

product containing mostly acrylic and acetic acid is sent to an acid purification column. The top product of this column is 95% by mol and the bottom product is 99.9% by Mol acrylic acid. The high concentrated acrylic acid is then cooled and sent to storage. There are several alternatives in this process for the regard changing the raw material composition substituted raw materials include the following:

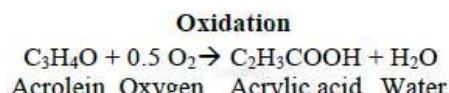
1. Propane is catalytically oxydehydrogenated to form propylene.
2. Acrylic acid is produced via gas derived acetylene complex
3. Acrylic Acid produced via dehydration of glycerol to Acrolein, followed by oxidation
4. Acrylic acid via ethylene oxide carbonylation to B-propiolactone followed by catalytic conversion.

5.3 (III). 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 by-product of the biodiesel production process. Biodiesel is a fast growing industry and so is the production of glycerol that 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 is dehydration of glycerol in Acrolein.



Next step is oxidation of Acrolein in acrylic acid & major by-products are acetaldehyde & acetic acid.



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}PW₁₂O₄₀ supported on Nb₂O₅ (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 by-product 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°C. The reaction is carried out at an optimum temperature of 300°C and ambient pressure because although increasing the temperature increases the conversion but trade-off is decrease in selectivity.

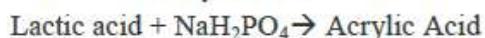
5.4 (IV). Acrylic acid from propane oxidation:

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]

5.5 (V). 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 bio-derivative, which is a renewable source of energy. Various catalysts like sodium dihydrogen phosphate(NaH_2PO_4), silica/alumina, $\text{Ca}_3(\text{PO}_4)_2$ supported on silica are 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 acid to 60-70%. The reaction is carried out at 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]

Dehydration:

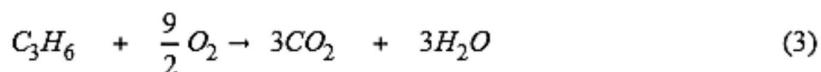
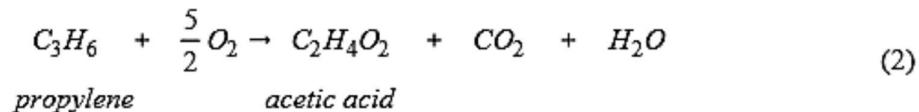
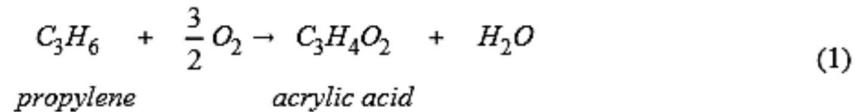


Reaction Phase	Vapor Phase
Reaction Type	Dehydration
Catalyst	NaH_2PO_4 Silica/Alumina $\text{Ca}_3(\text{PO}_4)_2$ on silica
Promoter	CaSO_4 with $\text{CuSO}_4/\text{Cu}_3(\text{PO}_4)_2$
Carrier Gas	CO_2
Selectivity	60-70%
Reaction Temperature	300-400°C
Conversion	~100% conversion

Though more processes have been explored, by a quantitative analysis, production of AA via partial oxidation of propylene was selected.

6 PROCESS DESCRIPTION

The reactions for acrylic acid production from propylene as follows:



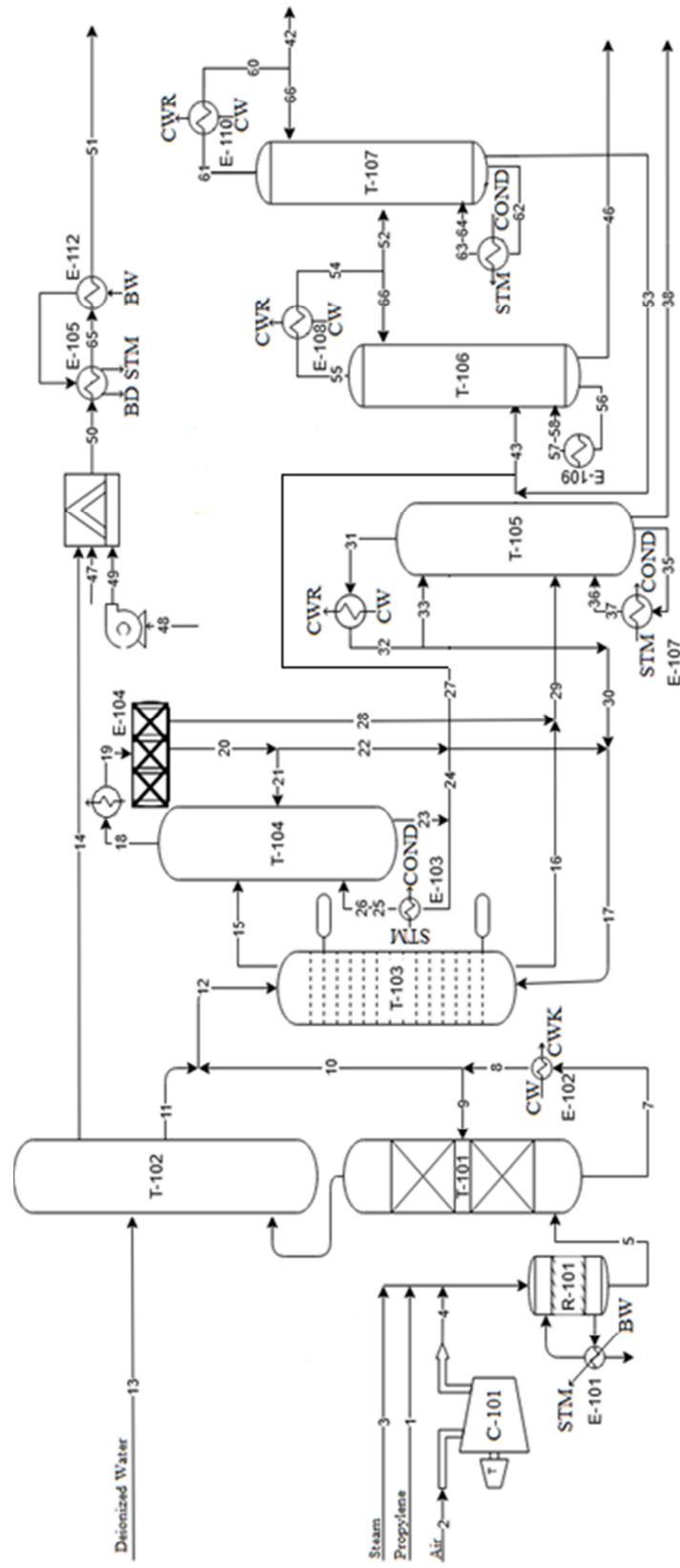
The reaction kinetics are of the form:

$$-r_i = A_i \exp\left[-\frac{E_i}{RT}\right] p_{propylene} p_{oxygen}$$

where i is the reaction number above, and

i	E_i kcal/kmol	A_i kmol/m ³ reactor h (kPa) ²
1	15,000	1.59×10^5
2	25,000	1.81×10^8
3	20,000	8.83×10^5

These reactions are industrially carried out in a fluidized bed reactor at a temperature of 310°C and pressure of 4.3 bars [7]. Molybdenum-Vanadium oxide catalyst (calcined at 1100°C) is used. The operating temperature range is very narrow as above 320°C the catalyst coking begins and below 300°C the kinetics does not support the production of acrylic acid. Therefore, the temperature has to be maintained in the reactor. In our case, the reactor has been assumed to be an ideal mixed flow reactor. The process flow sheet of the whole process designed is as shown below.



	SREAM	4	3	5	SREAM	30	38	42	43	46
TEMP	30	20	155.5	105.1	310	85	100	77.2	90	90
PRESSURE	600	101.3	500	108.8	101	PRESSURE	100.0	105.6	20.08	27.9
PHASE	V	V	V	V	PHASE	L	L	L	L	L
SREAM	6	7	8	9	SREAM	47	48	49	50	51
TEMP	50	50	38	38	TEMP	20	20	20	700	72
PRESSURE	101	176	101	101	PRESSURE	101.3	101.3	108.8	251.3	101.3
PHASE	V	L	L	L	PHASE	V	V	V	V	V
SREAM	11	12	13	14	SREAM	52	53	54	55	59
TEMP	50	50	50	50	TEMP	58.3	90	58.3	58.3	77.2
PRESSURE	101	101	101.3	99.4	PRESSURE	4.66	27.9	4.66	5.96	20.08
PHASE	L	L	V	L	PHASE	L	L	L	L	L
SREAM	16	17	18	19	SREAM	60	61	65	66	
TEMP				38	38	TEMP	77.2	77.2	223.8	58.3
PRESSURE				6.23	4.93	PRESSURE	20.08	21.38	176.3	4.66
PHASE	L	L	V	L	PHASE	L	V	V	V	L
SREAM	21	22	27	28	SREAM					
TEMP	38	38	90	38	TEMP					
PRESSURE	4.93				PRESSURE					
PHASE	L	L	L	L	PHASE					

In the process, 120.77kmol/hr of propylene is fed to the oxidation reactor R-101 where it reacts with oxygen to form acrylic acid. The by-product is acetic acid. The oxygen to propylene feed ratio is 2.25:1. The reactor is at 300 °C and the conversion of propylene is 100%. The catalyst used in the reactor is Molybdenum-Vanadium oxide calcined at 1100 C. The gaseous reactor stream is fed to T-101, packed quenching tower where the stream is quenched with 60,504 kmol/hr of liquid consisting of 3521076 kmol/hr of water, 4947kmol/hr acrylic acid and 359,920 kmol/hr of acetic acid. The gaseous effluent is then sent to T-102, absorber, where the gas is contacted with 227 kmol/her deionized water. Over 99.9% of the acrylic acid is absorbed in the two towers. The gaseous effluent from T-102 which contains acrylic acid and acetic acid in trace amounts and also oxygen, water, nitrogen and carbon dioxide is fed to the incinerator. Organics are combusted at 700 °C with oxygen and methane. The incinerator effluent is cooled to 72 °C which is 10 °C above its dew point by E-105 and E-102. The heat from this stream is used to generate steam at 3500kPa and 260 °C. The cooled steam is vented to the atmosphere. The heat of reaction from the reactor is absorbed by the boiler water and steam is generated at 550kPa at saturated conditions. The liquid effluent from T-102 along with the liquid quenched at T-101 is sent to the extractor where it is extracted with equal amount of solvent A. The solvent rich extract is sent to the extract stripper T-104. 100% of the acids are recovered in the bottom. Energy in the reboiler is supplied by steam. Temperature of the bottoms is 90°C and the pressure is 19.3kPa. All of the solvent and the water go overhead. The overhead temperature is 38 °C and the pressure is 6.23kPa. Energy in the condenser is absorbed by a stream of refrigerated water. Overhead is fed to a reflux decanter where the organics and the water phases separate. The reflux ratio is 0.8. The organic phase is recycled back to be used for extraction in the centrifugal extractor. The water phase along with the raffinate from the extractor is fed to the raffinate stripper where the solvent is removed from water and the trace acids. The overhead is pure solvent at 85°C and 101kPa. The bottom contains no solvent and is 100°C and 105.6kPa. The energy in the reboiler is supplied by 550kPa steam and the condenser utilizes cooling water. The pure solvent overhead is recycled back to the reactor and the bottoms are sent to the waste stream. Acids from the extract stripper bottoms are sent to the acrylic acid distillation train to be separated. Because of low pressures, two towers are involved. The acids from the extract stripper are mixed with recycle acids so that the concentration of the feed is boosted to 30mol%acetic acid. The feed 225,44 kmol/hr is separated, into overhead fractions (50mol% acetic and) and bottom fraction (99.9 mol% acrylic acid).

Temperature and pressure for bottom: 90 °C and 18.09 kPa

Temperature and pressure for overhead: 58.3 °C and 5.96 kPa

Energy for reboiler is supplied by 550kPa saturated steam. Energy in condenser is absorbed by cooling water. Reflux ratio for tower is 0.7795. The distillate is fed to T-107 where the separation yields an overhead containing 95% acetic acid and bottom containing 47.6% acetic acid. Overhead is the acetic acid product and bottom is recycled back to T-106. The bottoms temperature and pressure is 90 °C and 27.9kPa.

Accordingly, the reactor sizing according to the calculation is as shown below:

EQUIPMENT	REACTOR SIZING	DUTY (Kcal/Hr)	COST (\$/yr)
T-101(Incinerator)	G=43,743m ³ /hr		340177
R-101(reactor)	Tubes=7794		3178191
T-101(packed absorber)	5.6mD*29Dm T-T		3634309
T-102(Absorber)	2.13mD*8.83mT-T,4 Trays		77333
T-103(extractor)	11 theoretical stages		1320000
T-104(Extract stripper)	3.8mD*36.27mT-T,28 Trays		788458
T-105(Raffinate stripper)	0.6mD*7.9mT-T,10 trays		80315
T-106(AA recovery)	2.44mD*23.2mT-T,26 trays		462824
T-107(AA recovery)	0.7662mD*10.82mT-T,14 trays		121068
E-101(waste heat boiler)	S=37,642Kg/hr	23084626	350813
E-102(HEX)	A=4900m ²	14284166	1510529
E-103(Reboiler T-104)	A=100m ²	6641111	39068
E-104(condenser T-104)	A=325m ²	6641111	97546
E-105(economizer)	A=785m ²	6178817	12084
E-106(condenser T-105)	A=25m ²	756000	16484
E-107(Reboiler T-105)	A=30m ²	1638000	18223
E-108(condenser T-106)	A=110m ²	1721859	41808
E-109(Reboiler T-106)	A=25m ²	1572121	16484
E-110(condenser T-107)	A=9m ²	254971	10116
E-111(Reboiler T-107)	A=6m ²	401832	8628
E-112(Economizer)	A=210m ²	2360509	6471
Misc. equipment	0.1(total equipment)		1280930

The utility costs are as mentioned below:

Utility	Equipment	Cost/yr
Steam,550kPa	R101,E103,E107,E109,E111	5734620
Cooling water	E102,E106,,E108,E-110	763674
Chilled water	E104	1783984
Boiling water	E101,E112	1043411
Deionized water	T-102	37987
Electricity	C101,C102	944665
Natural gas	I-101	1083764

The incoming stream properties are as mentioned below:

Credits	Equipment	Cost/yr
Steam(550kPa)	E101	5803477

Steam(3500kPa)	E105	2086996
	TOTAL	2990632

The net revenue of the process is as mentioned below:

Description	Cost/year
Propylene	18768794
Miscellaneous chemicals	2007395
Labour, benefits, taxes, etc.	1035632
Maintenance	2818047
Support charges	702686
Utilities	2990632
Revenue	
Acetic acid	1054818

Revenue before taxes: Sales AA (dollars/yr)- 27268368. The present worth calculation yields a minimum selling price of AA as 0.952 dollar/kg.

CALCULATION

6.1 DETERMINATION OF OXIDATION REACTOR OPERATING CONDITIONS

This section deals with determining the reactor operating conditions, a material balance is done around the reactor for all three reactor temperatures. The cost of reactor for each temperature is compared with the corresponding revenue. NPV calculation is done for two temperatures for the comparison. We choose the case with the greatest present worth value. We performed the following calculations so that the material balance can be calculated downstream of the reactor. Energy balance is performed for the waste heat boiler so that the heat duty is known.

CASE 1: (Salt temperature 200°C)

C₃ conversion 90%(assume mole %)

AA selectivity	0.856
HAC	0.064
CO ₂	0.080

The AA recovery following extraction is 97%

$$(6500 \text{ Kg/hr AA}) * (\text{kgmol}/72.06\text{kg}) = (0.856) * (0.9) * (0.97)C_3$$

$$C_3 = 120.71\text{Kgmol/hr}$$

Production rate:

$$\text{HAC} = 120.71 * (0.064) * (0.97) * (0.9) = 6.744\text{Kgmol/hr}$$

$$\text{AA} = 120.71 * (0.856) * (0.97) * (0.9) = 90.205\text{Kgmol/hr}$$

$$\text{O}_2 = 2.256 * (120.71) = 271.60\text{Kgmol/hr}$$

$$\text{N}_2 = (271.6/0.21) - 271.6 = 1021.73\text{Kgmol/hr}$$

$$6.1.1.1 \text{ STM} = \frac{120.71+1021.73+271.60}{0.6} - (120.71 + 1021.73 + 271.6) = 942.7\text{Kgmol/hr}$$

6.1.1.2

6.1.1.3 CATALYST

$$\begin{aligned} & (\text{Kg CAT}/1.44\text{gmolC3}/\text{hr}) * (120.71\text{Kgmol/hr} * 1000\text{gmol/kgmol}) \\ & = 83,826.4\text{Kg catalyst} \end{aligned}$$

CASE 2 (Salt temperature 250°C)

C₃, conversion 95%

AA selectivity 0.826

HAC 0.060

CO₂ 0.114

$$6500/72.06 = (0.826) * (0.95) * (0.97) C_3, C_3 = 118.51 \text{Kgmol/hr}$$

Production rate:

$$\text{HAC} = 118.51 * (0.06) * (0.97) * (0.95) = 6.552 \text{Kgmol/hr}$$

$$\text{AA} = 118.51 * (0.826) * (0.97) * (0.95) = 90.205 \text{Kgmol/hr}$$

$$O_2 = 2.256 * (118.51) = 266.65 \text{Kgmol/hr}$$

$$N_2 = (266.65/0.21) - 266.65 = 1003.11 \text{Kgmol/hr}$$

6.2 INTRODUCTION:

The objective of this project was to design a plant for the production of acrylic acid. Major objectives were:

- Choosing between the two solvents explored in the literature review.
- Deciding on the reactor conditions from rate equations.
- Determine the optimum distillation system to separate acrylic acid and acetic acid.

Design parameters, Reactor conditions, Material energy balances each of these aspects are included in the report. The report is divided into sections including design basis, calculation basis, process analysis, Results and discussions, Conclusions and the Appendix. Design basis summarizes the opportunities, Calculation basis details the assumptions and specifications, Process analysis provides process description, material balance, flow diagram, utility summary and the economics. Every detailed calculation would be presented in the appendix of the report.

6.3 DESIGN BASIS:

Our design for the production plant consists of an oxidation reactor, a two stage absorber unit, consisting of packed tower and tray tower, a centrifugal extractor with extract and raffinate strippers, acrylic acid recovery system which is designed using waste gas incinerator and distillation equipment. Report also includes various heat exchangers and compressors. The project progression path is as described below:

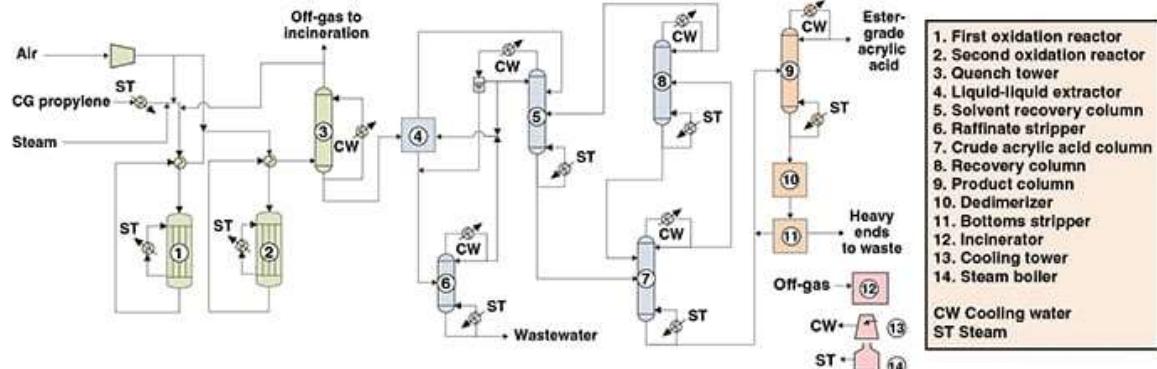


Figure 3

- Propylene is getting oxidized in the reactor to form acrylic acid.
 - The principal by products are carbon dioxide and acetic acid (which can be sold).
 - Furthermore, evaluating and comparing two solvents for extraction allows for economic feasibility study.
 - The economic evaluation will be based upon the capital investment, economic cost and the manufacturing cost as well.
 - We shall also take into account the raw material cost, utility cost and other miscellaneous costs.

6.4 RAW MATERIAL COSTS:

Table 1

Propylene	0.48dollars/kg
By-product HAC credit	0.33dollars/kg
Miscellaneous chemicals	0.04dollars/kg of CAA

*CAA is the Crude acrylic acid produced

6.5 UTILITY COSTS:

Table 2

Cooling water	0.020/K-liter
Deionized water	1.200/K-liter
Electricity	0.074KW/hr
Natural gas	3.700/million-kjoules
Refrigerated water	8.300/million-kjoules
Steam	20.000/K-kg
Boiler feed water	2.500/K-liter
Misc	0.033/kg of AA produced

6.6 MISCELLANEOUS COSTS:

Table 3

Labour, benefits, taxes etc.	0.0147/dollar investment/year
Maintenance	0.04/dollar investment/year
Other support charges(supplies, admin)	0.0140/kg of AA produced/year

Equipment pricing is based upon the purchased equipment cost equations mentioned further into the report. Installed costs of the equipment are found by multiplying the purchased equipment cost by a lang factor of 5. The estimates of the equipment are within $\pm 5\%$. The economic evaluation is based on the fact that the capital investment will be done considering current economic factors. Straight line depreciation is considered. A tax rate of 50% and a discount rate of 15% will be used in the evaluation. Since the price of acrylic acid is not supplied, a present worth calculation was performed to determine the minimum acceptable selling price of acrylic acid in order to ensure a present worth of zero at a 15% discount. Present worth calculation is done to provide the selling cost of acrylic acid. The detailed calculations are shown in the appendix

6.7 CALCULATION BASIS:

We made several assumptions during the course of the solution. Assumptions were made to facilitate the solution and resolve the unclarities. The assumptions along with their impact is discussed. These are mentioned below:

- The reaction is assumed to be gas phase for starters, and products are at 20 °C although at this temperature the products would be liquid. This assumption would result in greater heat duty for the waste heat reboiler than if the reaction was taken as liquid phase.
- It is assumed that methane is available at 20 °C in the incinerator. The heat of combustion for methane was taken at this temperature. The incoming temperature of the methane would have a slight effect on the material energy balances and around the incinerator and would also effect the heat duty of the waste gas/heat boiler.
- It is assumed that the air used in the oxidation reactor is at the same temperature as the air used in the incinerator that is 20 °C and 80% humidity and atmospheric pressure.
- Area of the down comers relative to the total tray area is important for sizing, it is assumed that the area of the down comers is 10% of the total area. This assumption would affect the diameter of the tray towers and ultimately the equipment costs.
- It is assumed that the trays required for the separation are real and not theoretical.
- Another assumption regarding composition of feed in the design of this column is that all of the solvent went overhead and all of the water and acids went to the bottom.
- It is assumed that all the liquids formed ideal solutions and all of the vapours formed ideal gases. This would have an impact on both the energy and the material balance.
- Net 0 heat of mixing or volume change of mixing for both liquids and gases is assumed.
- It is also assumed that the temperature profile for all the reactors are linear.

6.8 PROCESS ANALYSIS:

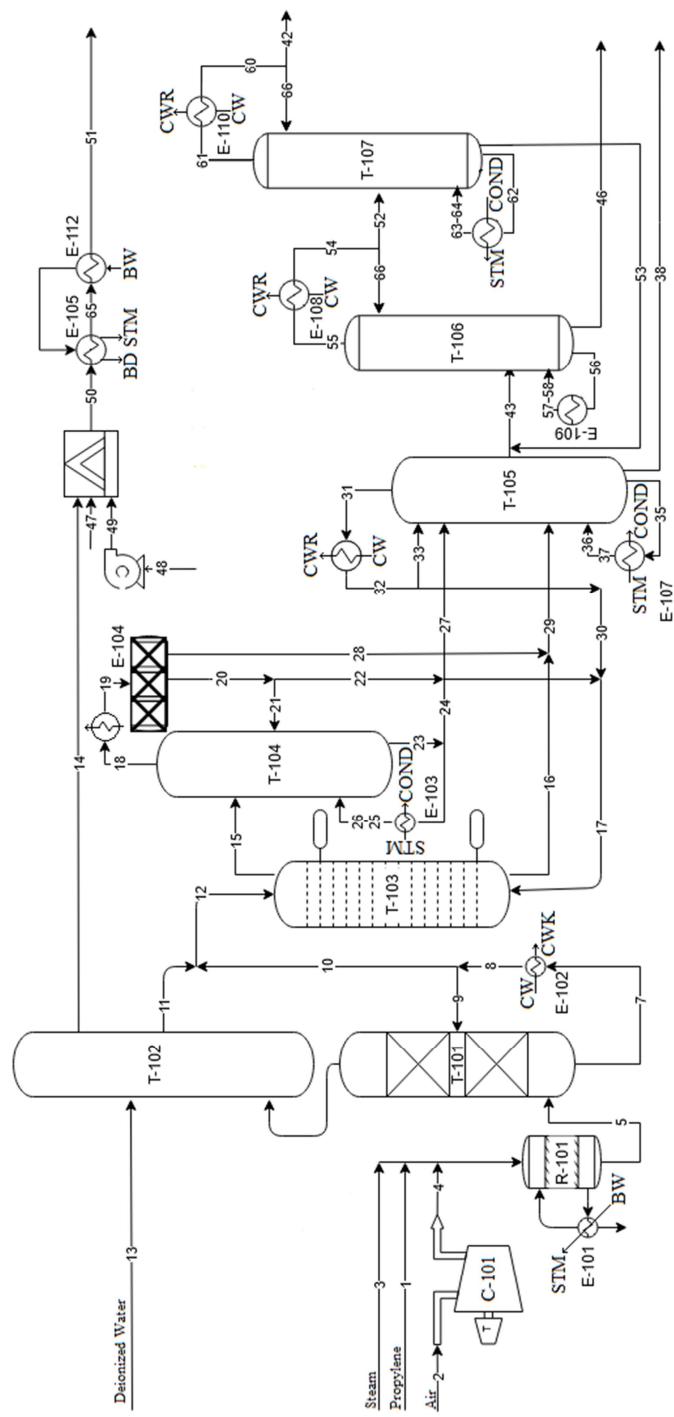
This section provides results of the process studies. The process flow diagram, utility cost, material energy balances are all shown. Economics is study is performed after designing. The material balance contains the temperatures, pressures and the case presented is the optimum case. The maximum production of acrylic acid from our design is 6504.5kg/hr. We estimated 97% recovery of acrylic acid from the recovery stream. A brief description:

SREAM	1	2	3	4	5
C_3H_6	120.77				
CO_2					69.56
AA					92.99
HAC					6.88
O_2		271.70		271.70	20.98
N_2		1022.30		1022.3	1022.3
H_2O		2.46	959.50	24.46	1146.51
CH_4					
SOLVENT A					
TOTAL (Kmol/hr)	120.77	1381.46	959.50	1318.46	2359.22
TEMP	30	20	155.5	105.1	310
PRESSURE	600	101.3	500	251.3	101
PHASE	V	V	V	V	V
SREAM	6	7	8	9	10
C_3H_6					
CO_2	69.56				
AA	2.78	5037.21	5037.21	4947	90.21
HAC	0.56	352.92	359.92	346.6	6.32
O_2	20.98				
N_2	1022.3				
H_2O	139.74	56216.76	56216.76	55210	1006.76
CH_4					
SOLVENT A					
TOTAL Kmol/hr	1255.92	61606.89	61606.89	60503.6	110329
TEMP	50	50	38	38	38
PRESSURE	101	176	101	101	101
PHASE	V	L	L	L	L
SREAM	11	12	13	14	15
C_3H_6					
CO_2				69.56	
AA	2.7283	92.938		69.56	
HAC	0.5016	6.822		0.0584	6.65
O_2				20.98	
N_2				1022.3	
H_2O	227	1233.76	227	139.74	39.99
CH_4					
SOLVENT A					277.86
TOTAL Kmol/hr	230.23	1333.52	227	1252.69	415.11
TEMP	50		50	50	
PRESSURE	101	101	101.3	99.4	
PHASE	L	L	L	V	L
SREAM	16	17	18	19	20
C_3H_6					
CO_2					

AA	2.33				
HAC	0.172				
O_2					
N_2					
H_2O	1196.92	3.15	92.51	42.51	5.676
CH_4					
SOLVENT A	9.53	287.37	500.07	500.07	499.96
TOTAL Kmol/hr	1208.96	290.55	542.58	542.58	505.64
TEMP			38	38	38
PRESSURE			6.23	4.93	4.93
PHASE	L	L	V	L	L
SREAM	21	22	27	28	29
C_3H_6					
CO_2					
AA			90.61		2.33
HAC			6.65		0.172
O_2					
N_2					
H_2O	2.52	3.15		36.83	1233.75
CH_4					
SOLVENT A	222.21	277.76		0.107	9.637
TOTAL Kmol/hr	224.731	280.91	97.26	36.93	1245.89
TEMP	38	38	90	38	
PRESSURE	4.93		19.3		
PHASE	L	L	L	L	L
SREAM	30	38	42	43	46
C_3H_6					
CO_2					
AA		2.33	0.345	67.63	90.265
HAC		0.172	6.56	157.81	0.090
O_2					
N_2					
H_2O		1233.75			
CH_4					
SOLVENT A	9.367				
TOTAL Kmol/hr	9.367	1236.25	6.905	22544	90.355
TEMP	85	100	77.2	90	90
PRESSURE	100.0	105.6	20.08	27.9	18.09
PHASE	L	L	L	L	L
SREAM	47	48	49	50	51
C_3H_6					
CO_2				112.83	112.83
AA					
HAC					
O_2		86.86	86.86	21.57	21.57

N_2		326.75	326.75	134.89	134.89
H_2O		7.82	7.82	233.78	233.78
CH_4	43				
SOLVENT A					
TOTAL Kmol/hr	43	421.43	421.43	1717.08	1717.08
TEMP	20	20	20	700	72
PRESSURE	101.3	101.3	108.8	251.3	101.3
PHASE	V	V	V	V	V

SREAM	52	53	54	55	59
C_3H_6					
CO_2					
AA	67.54	67.195	120.19	120.19	1.79
HAC	67.54	60.98	120.19	120.19	34.05
O_2					
N_2					
H_2O					
CH_4					
SOLVENT A					
TOTAL Kmol/hr	135.08	128.175	240.38	240.38	35.84
TEMP	58.3	90	58.3	58.3	77.2
PRESSURE	4.66	27.9	4.66	5.96	20.08
PHASE	L	L	L	L	L
SREAM	60	61	65	66	
C_3H_6					
CO_2			112.83		
AA	2.135	2.135		52.65	
HAC	40.61	40.61		52.65	
O_2			21.57		
N_2			1348.9		
H_2O			233.78		
CH_4					
SOLVENT A					
TOTAL Kmol/hr	42.745	42.745	1717.08	105.30	
TEMP	77.2	77.2	223.8	58.3	
PRESSURE	20.08	21.38	176.3	4.66	
PHASE	L	V	V	L	



SREAM	1	2	3	4	5	STREAM	30	38	42	43	46
TEMP	30	20	155.5	105.1	310	TEMP	85	100	77.2	90	90
PRESSURE	600	101.3	500	108.8	101	PRESSURE	100.0	105.6	20.08	27.9	18.09
PHASE	V	V	V	V	V	PHASE	L	L	L	L	L
SREAM	6	7	8	9	10	SREAM	47	48	49	50	51
TEMP	50	50	38	38	38	TEMP	20	20	700	72	
PRESSURE	101	101	101	101	101	PRESSURE	101.3	101.3	108.8	251.3	101.3
PHASE	V	L	L	L	L	PHASE	V	V	V	V	V
SREAM	11	12	13	14	15	SREAM	52	53	54	55	59
TEMP	50	50	50	50	50	TEMP	58.3	90	58.3	58.3	77.2
PRESSURE	101	101	101.3	99.4	99.4	PRESSURE	4.66	27.9	4.66	5.96	20.08
PHASE	L	L	V	V	V	PHASE	L	L	L	L	L
SREAM	16	17	18	19	20	SREAM	60	61	65	66	
TEMP						TEMP	77.2	77.2			
PRESSURE						PRESSURE	20.08	21.38			
PHASE						PHASE	L	V			
SREAM	21	22	27	28	29	SREAM					
TEMP	38	38	90	38	38	TEMP					
PRESSURE	4.93	6.23	4.93	4.93	4.93	PRESSURE					
PHASE	L	L	L	L	L	PHASE					

Figure 4

120,77kmol/hr of propylene is fed to the oxidation reactor where it reacts with oxygen to form acrylic acid. The by-product is acetic acid. The oxygen to propylene feed ratio is 2.25:1. The reactor is at 300 °C and the conversion of propylene is 100%. The gaseous reactor stream is fed to T-101, packed quenching tower where the stream is quenched with 60,504 kmol/hr of liquid consisting of 3521076 kmol/hr of water, 4947kmol/hr acrylic acid and 359,920 kmol/hr of acetic acid. The gaseous effluent is then sent to T-102, absorber, where the gas is contacted with 227 kmol/her deionized water. Over 99.9% of the acrylic acid is absorbed in the two towers. The gaseous effluent from T-102 which contains acrylic acid and acetic acid in trace amounts and also oxygen, water, nitrogen and carbon dioxide is fed to the incinerator. Organics are combusted at 700 °C with oxygen and methane. The incinerator effluent is cooled to 72 °C which is 10 °C above its dew point by E-105 and E-102. The heat from this stream is used to generate steam at 3500kPa and 260 °C. The cooled steam is vented to the atmosphere. The heat of reaction from the reactor is absorbed by the boiler water and steam is generated at 550kPa at saturated conditions.

The liquid effluent from T-102 along with the liquid quenched at T-101 is sent to the extractor where it is extracted with equal amount of solvent A. The solvent rich extract is sent to the extract stripper T-104. 100% of the acids are recovered in the bottom. Energy in the reboiler is supplied by steam. Temperature of the bottoms is 90°C and the pressure is 19.3kPa. All of the solvent and the water go overhead. The overhead temperature is 38 °C and the pressure is 6.23kPa. Energy in the condenser is absorbed by a stream ^{251,3} of refrigerated water. Overhead is fed to a reflux decanter where the organics and ^{251,3} the water phases separate. The reflux ratio is 0.8. The organic phase is recycled back to be used for extraction in the centrifugal extractor. The water phase along with the raffinate from the extractor is fed to the raffinate stripper where the solvent is removed from water and the trace acids. The overhead is pure solvent at 85°C and 101kPa. The bottom contains no solvent and is 100°C and 105.6kPa. The energy in the reboiler is supplied by 550kPa steam and the condenser utilizes cooling water. The pure solvent overhead is recycled back to the reactor and the bottoms are sent to the waste stream. Acids from the extract stripper bottoms are sent to the acrylic acid distillation train to be separated. Because of low pressures, two towers are involved. The acids from the extract stripper are mixed with recycle acids so that the concentration of the feed is boosted to 30mol%acetic acid. The feed 225,44 kmol/hr is separated, into overhead fractions (50mol% acetic and) and bottom fraction (99.9 mol% acrylic acid).

Temperature and pressure for bottom: 90 °C and 18.09 kPa

Temperature and pressure for overhead: 58.3 °C and 5.96 kPa

Energy for reboiler is supplied by 550kPa saturated steam. Energy in condenser is absorbed by cooling water. Reflux ratio for tower is 0.7795. The distillate is fed to T-107 where the separation yields an overhead containing 95% acetic acid and bottom containing 47.6% acetic acid. Overhead is the acetic acid product and bottom is recycled back to T-106. The bottoms temperature and pressure is 90 °C and 27.9kPa.

The table below provides the equipment list of the flowsheet.

Table 4

Equipment	Size	Duty (kCal/hr)	Cost (\$/yr)
T-101(Incinerator)	G=43,743m3/hr		340177
C-101(Compressor)	P=4.39*10^4		655179
C-102(Fan)	V=158.2m3/min		18079
R-101(reactor)	Tubes=7794		3178191
T-101(packed absorber)	5.6mD*29Dm T-T		3634309
T-102(Absorber)	2.13mD*8.83mT-T,4 Trays		77333
T-103(extractor)	11 theo stages		1320000
T-104(Extract stripper)	3.8mD*36.27mT-T,28 Trays		788458
T-105(Raffinate stripper)	0.6mD*7.9mT-T,10 trays		80315
T-106(AA recovery)	2.44mD*23.2mT-T,26 trays		462824
T-107(AA recovery)	0.7662mD*10.82mT- T,14 trays		121068
E-101(waste heat boiler)	S=37,642Kg/hr	23084626	350813
E-102(HEX)	A=4900m2	14284166	1510529
E-103(Reboiler T-104)	A=100m2	6641111	39068
E-104(condenserT-104)	A=325m2	6641111	97546

E-105(economizer)	A=785m2	6178817	12084
E-106(condenser T-105)	A=25m2	756000	16484
E-107(Reboiler T-105)	A=30m2	1638000	18223
E-108(condenser T-106)	A=110m2	1721859	41808
E-109(Reboiler T-106)	A=25m2	1572121	16484
E-110(condenser T-107)	A=9m2	254971	10116
E-111(Reboiler T-107)	A=6m2	401832	8628
E-112(Economizer)	A=210m2	2360509	6471
Misc equipment	0.1(total equipment)		1280930

The table below is a summary of utilities:

Table 5

Utility	Equipment	Cost/yr
Steam,550kPa	R101,E103,E107,E109,E111	5734620
Cooling water	E102,E106,,E108,E-110	763674
Chilled water	E104	1783984
Boiling water	E101,E112	1043411
Deionized water	T-102	37987
Electricity	C101,C102	944665
Natural gas	I-101	1083764

Credits	Equipment	Cost/yr
Steam(550kPa)	E101	5803477
Steam(3500kPa)	E105	2086996
		2990632

Table below is a summary of operating costs and revenue. The AA revenue is not known. A present worth calculation was done to determine the minimum selling price.

Table 6

Description	Cost/year
Propylene	18768794
Miscellaneous chemicals	2007395
Labour, benefits, taxes, etc.	1035632
Maintenance	2818047
Support charges	702686
Utilities	2990632
Revenue	
Acetic acid	1054818

Revenue before taxes: Sales AA (dollars/yr)- 27268368. The present worth calculation yields a minimum selling price of AA as 0.952 dollar/kg

7 CONCLUSION:

- 1) The objectives were met. A basic design of acrylic acid production plant was developed. The capital investment and operating costs were determined. The manufacturing costs were found out.
- 2) The optimum reactor conditions were chosen to be set at 300 °C. These conditions give us the greatest present worth. Effect of choice of reactor conditions is not known downstream of the reactor.
- 3) Solvent A is chosen to be optimum solvent for extraction after comparisons and present worth calculations.

8 RECOMMENDATIONS:

- 1) The process study may be compared to the existing plant.
- 2) The impact of reactor conditions can be understood by doing the process design for each of 4the reactor conditions.
- 3) Solvent A is chosen; more rigorous physical data is needed (further research in the field required) for a better comparison.

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10 APPENDIX: DESIGN CALCULATIONS

10.1 APPENDIX A: REACTOR

10.1.1 REACTOR COST

$$36000 * ((8569)^{0.5}) = 3332480 \text{ dollars}$$

Total cost (lang factor)

$$5 * (3332480) = 16662401 \text{ dollars}$$

CASE 1:

$$49904.95 \text{ kg cat} * (\text{m}^3 / 880 \text{ kg cat}) = 56.710 \text{ m}^3$$

$$\# \text{ of tubes} = 7794$$

$$\text{Reactor cost: } 3600 * ((7794)^{0.5}) = 3178191$$

Total cost (lang factor)

$$5 * (3178191) = 15890954 \text{ dollars}$$

CASE 2:

$$\begin{aligned} \text{C3: } 118.51 \text{ kmol/hr} * 24 \text{ hr/day} * 365 \text{ days/yr} * 0.88 * 0.48 \text{ dollars/kg} * 42 \text{ kg/kgmol} \\ = 18417570 \text{ dollars/yr} \end{aligned}$$

$$\text{HAC: } 6.552 \text{ kgmol/hr} * 24 * 365 * 0.88 * 0.33 * 60.05 = 1000060 \text{ dollars/yr}$$

$$\text{AA: } 90.21 \text{ kgmol/hr} * 24 * 365 * 0.88 * 1.32 * 72.06 = 66297257 \text{ dollars/yr}$$

REVENUE: 48879747 dollars/yr

CASE 3:

$$\text{C3: } 120.77 * 24 * 365 * 0.88 * 0.48 * 42 = 18768794.2 \text{ dollars/yr}$$

$$\text{HAC: } 6.8 * 24 * 365 * 0.88 * 0.33 * 60.05 = 1019597 \text{ dollars/yr}$$

$$\text{AA: } 66297257 \text{ dollars/yr}$$

REVENUE: 48548059 dollars/yr

To compare the two cases a present worth analysis is done. A discount rate of 15% and a straight line depreciation of 50% for 11 years is utilized.

CASE 2:

Table 7

EOY	CAPITAL	DEP	REV	C.F.A.T
0	16,662,401	-	-	
1		1,514,732	48,879,474	23,682,508
2		1,514,732	48,879,474	23,682,508
3		1,514,732	48,879,474	23,682,508
4		1,514,732	48,879,474	23,682,508
5		1,514,732	48,879,474	23,682,508
6		1,514,732	48,879,474	23,682,508
7		1,514,732	48,879,474	23,682,508
8		1,514,732	48,879,474	23,682,508
9		1,514,732	48,879,474	23,682,508
10		1,514,732	48,879,474	23,682,508
11		1,514,732	48,879,474	23,682,508

$$PW = -16662401 + 23682508 * (5.2337) = 107284741 \text{ dollars}$$

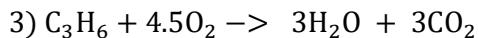
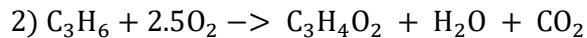
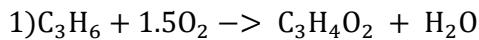
CASE 3:

$$PW = -15890954 + 23551713.4 * (5.2337) = 107371648 \text{ dollars}$$

Therefore, Case 3 will be used for subsequent calculations. (PRESENT WORTH factors were found in PRINCIPLES OF ENGINEERING ECONOMIC ANALYSIS, John A. White, Marnn H. Agee, Kenneth case, 2nd Edition, John Wiley and sons, New York, 1997, Appendix B)

10.1.2 MATERIAL BALANCE AROUND REACTOR:

Reactions:



Feed : 120.77Kgmol/hr

Reaction 1:

AA produced : $(120.77) * (0.77) = 92.99\text{Kgmol/hr}$

H_2O produced : $(120.77) * (0.77) = 92.99\text{Kgmol/hr}$

O_2 consumed: $92.99 * 15 = 139.49\text{Kgmol/hr}$

Reaction 2:

HAC produced: $120.77 * (0.057) = 6.88\text{Kgmol/hr}$

CO_2 produced: 6.88Kgmol/hr

H_2O produced: 6.88Kgmol/hr

O_2 consumed: 17.21Kgmol/hr

Reaction 3:

CO_2 produced : $0.173 * 120.77 * 3 = 62.68\text{Kgmol/hr}$

H_2O produced: 62.68Kgmol/hr

O_2 consumed: 94.02Kgmol/hr

Humidity of air which was neglected would now be included.

Air has 80% relative humidity and is at 20°C

Vapour pressure at this temperature is 17.535mmHg

$$0.8 * (17.535 / (760 - 17.535)) = 0.189\text{mol H}_2\text{O/mol air}$$

$$H_2O = 0.0189 * (271.7 + 1022.3) = 2446 \text{Kgmol/hr}$$

New steam flowrate:

$$(1022.3 + 271.7 + 120.77 + 24.46)/0.6 - (1022.3 + 271.7 + 120.77 + 24.46) \\ = 959.50 \text{Kgmol/hr}$$

Table 8

COMPONENTS	IN(kgmol/hr)	OUT(kgmol/hr)
C ₃	120.77	0
O ₂	271.70	20.98
N ₂	1022.3	1022.3
H ₂ O	24.46	187.01
STM	959.50	959.50
AA	None	9299
HAC	None	6.88
CO ₂	None	69.56
Total	239872	2359.22

10.1.3 HEAT OF REACTION

$$\Delta H_{rxn} = \sum m_i \int_{100}^{20} C_p \Delta T(\text{reactants}) + \Delta H_{rxn}(20^\circ\text{C}) = \sum m_i \int_{20}^{310} C_p \Delta T(\text{products})$$

ΔH_{rxn} at 20°C (assume gas phase rxn)

$$\text{Rxn1)} 1.61 \times 10^5 \text{ Kcal/Kgmol (92.99) Kgmol/hr} = 14971390 \text{ Kcal/hr}$$

$$\text{Rxn2)} 2.76 \times 10^5 \text{ Kcal/Kgmol (6.88Kgmol/hr)} = 1892000 \text{ Kcal/hr}$$

$$\text{Rxn3)} 4.89 \times 10^5 \text{ Kcal/Kgmol (20.89 Kgmol/hr)} = 10215210 \text{ Kcal/hr}$$

Total : 27978600 Kcal/hr

$$\int_{100}^{20} C_p \Delta T = A(100-20) + (B/2 * (100^2 - 20^2) * 10^{-3} + C/3 * (100^3 - 20^3) * 10^{-7}$$

$$C_3 = 0.339 * (100-20) + 0.504 * (100^2 - 20^2) * 10^{-3} - 1.933 * (100^3 - 20^3) * 10^{-7}$$

$$= 31.76 \text{ cal/g} * 120.77 \text{ Kgmol/hr}^* 42.08 \text{ Kg/Kgmol} * 1000 \text{ g/Kg} * \text{Kcal/1000cal} = \\ 161414 \text{ Kcal/hr}$$

$$O_2 = m_i \int_{100}^{20} Cp \Delta T = m * (18.04 \text{ Cal/g})$$

$$= 18.04 \text{ cal/g} * 271.7 \text{ kgmol/hr} * 32 \text{ kg/kgmol} * 1000 \text{ g/kg} * \text{Kcal/1000cal} = 156847 \text{ Kcal/hr}$$

$$N_2 = m_i \int_{100}^{20} Cp \Delta T = m * (19.99 \text{ Cal/g})$$

$$= 19.99 \text{ cal/g} * 1022.3 \text{ kgmol/hr} * 28 \text{ kg/kgmol} * 1000 \text{ g/kg} * \text{Kcal/1000cal} = 572137 \text{ Kcal/hr}$$

$$H_2O = m_i \int_{100}^{20} Cp \Delta T = m * (35.98 \text{ Cal/g})$$

$$= 35.98 \text{ cal/g} * 24.46 \text{ Kgmol/hr} * 18 \text{ kg/kgmol} * 1000 \text{ g/kg} * \text{Kcal/1000cal} = 15842 \text{ Kcal/hr}$$

STM = at 550KPa , 555 °C to superheated at 101 kPa, 310°C

$$H = 27530 \text{ KJ/kg} \quad H = 3094.69 \text{ KJ/kg}$$

$$\int_{20}^{310} Cp \Delta T = A(310-20) + (B/2 * (310 - 20)^2 * 10^{-3} + C/3 * (310 - 20)^3)$$

$$O_2 = m_i \int_{20}^{310} Cp \Delta T = m * (66.7 \text{ Cal/g})$$

$$= 66.67 * 20.98 * 32 * 1000 / 1000 = 44758 \text{ Kcal/hr}$$

$$N_2 = m_i \int_{20}^{310} Cp \Delta T = m * (73.85 \text{ Cal/g})$$

$$= 73.85 * 1022.3 * 28 * 1000 / 1000 = 2113837 \text{ Kcal.hr}$$

$$H_2O = m_i \int_{20}^{310} Cp \Delta T = m * (138 \text{ Cal/g})$$

$$= 138 * 187.01 * 18 * 1000 / 1000 = 464459 \text{ Kcal/hr}$$

$$AA = m_i \int_{20}^{310} Cp \Delta T = m * (90.2 \text{ Cal/g})$$

$$= 90.2 * 92.99 * 72.06 * 1000 / 1000 = 6044807 \text{ Kcal/hr}$$

$$HAC = m_i \int_{20}^{310} Cp \Delta T = m * (143.6 \text{ Cal/g})$$

$$= 143.6 * 688 * 60.05 * 1000 / 1000 = 59338 \text{ Kcal/hr}$$

$$CO_2 = m_i \int_{20}^{310} Cp \Delta T = m * (66.27 \text{ Cal/g})$$

$$=66.27*69.56*44.01*1000/1000 = 202,8667 \text{ Kcal/hr}$$

$$\Delta H_{\text{HST}} = (3094.69 - 2753) \text{ KJ/kg} * 959.5 \text{ Kgmol/hr} * 18 \text{ kg/kgmol} = 5901328 \text{ KJ/hr}$$

$$5901 * 328 \text{ KJ/hr} * 0.23901 \text{ Kcal/Kg} = 1410476 \text{ Kcal/hr}$$

$$\Delta H_{\text{RXN}} = 161414 + 156847 + 572137 + 15842 + 27078600 - 44758 - 2113837 - 464459 - 604480 - 59338 - 202866 - 1410746 = 23084626 \text{ Kcal/hr}$$

$$(0.50m(655.8-209.33)\text{KJ/kg} + 0.95m(2751.7-209.33)\text{KJ/kg}) * 0.23901 \text{ Kcal/KJ}$$

$$= 23084626 \text{ Kcal/hr}$$

(all property values are taken from **Felder, Richard M., and Ronald W. Rousseau. Elementary Principles of Chemical Processes.** New York: Wiley, 1986., perfect mixing is assumed),

$m = 39623 \text{ Kg/hr}$ water fed to the boiler

$0.95m = 37641.9 \text{ kg/hr}$ steam made

10.1.4 PROCESS FLOW AND MATERIAL GRID:

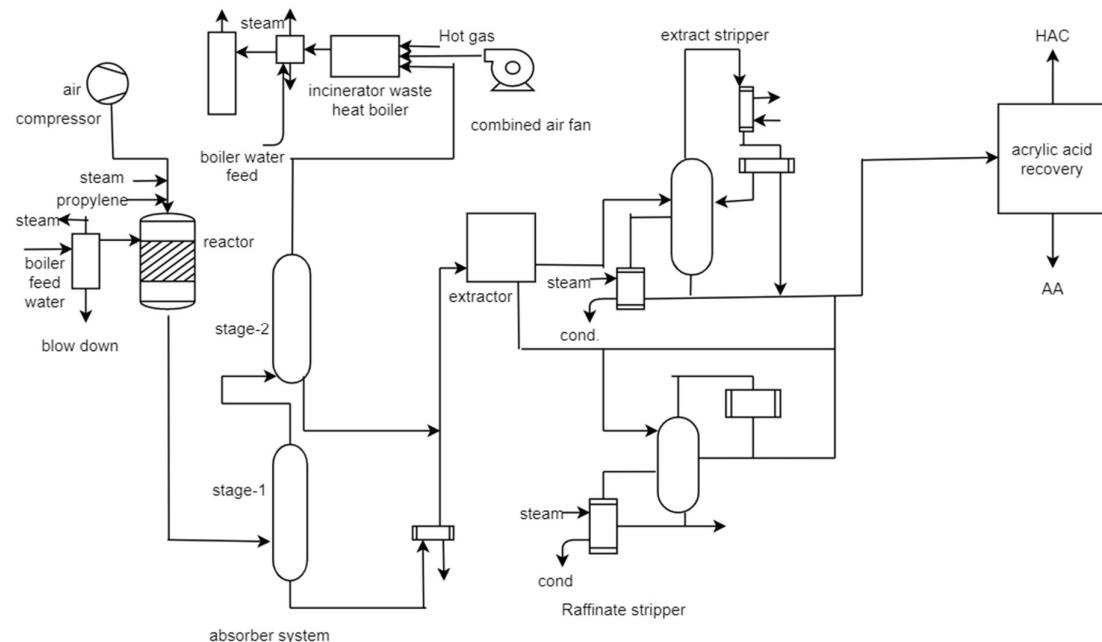


Figure 5: Process schematic

Description:

The purpose of this section is to make a process flow diagram.

10.2 APPENDIX B: ABSORBER QUENCHING SECTION-I

A material balance is done around the packed absorption tower which is used to quench the reactor outlet stream. The composition of gas stream is calculated by assuming that the outlet gas stream is in equilibrium with the liquid leaving the tower at 50 °C. Energy balance is performed on the quencher to determine recycle flowrate of liquid.

10.2.1 ABSORBER QUENCHING SECTION ,T-101

The gas entering the absorber consists of (Kgmol/hr)

Table 9

CO ₂	69.56
AA	92.99
HAC	6.88
O ₂	20.98
N ₂	1022.3
H ₂ O	1146.5
C ₃	0
Total	2359.22

$$\ln(P_{\text{sat}}/P_{\text{critical}}) = A - (B/C + T)$$

P_{sat} : 50 °C

P_{total} : 101.3

Table 10

	P/P _{total}
AA	0.027
HAC	0.074
H ₂ O	0.122

Inerts O₂, N₂, CO₂

$$\text{Total flow rate} = (20.98 + 1022.3 + 69.56) = 1112.84 \text{ Kgmol/hr}$$

Initial guess for the liquid composition X_i for

Table 11

AA	0.20
HAC	0.05
H ₂ O	0.75

$$\sum Y_i = 0.122 * (0.75) + 0.074 * (0.05) + 0.027 * (0.20)$$

$$\text{Total flowrate out of the absorber: } 1112.84 / (1 - 0.1006) = 1237.31 \text{ Kgmol/hr}$$

$$\text{Distributing components in the gas phase} = y_i * (1237.31) = (P_{\text{sat}} / P_{\text{total}}) * (X_i) * (1237.31)$$

$$\text{H}_2\text{O} = 0.122 * (0.75) * (1237.31) = 113.21 \text{ Kgmol/hr}$$

$$\text{AA} = 0.027 * (0.20) * (1237.31) = 6.68 \text{ Kgmol/hr}$$

$$\text{HAC} = 0.074 * (0.05) * (1237.31) = 4.58 \text{ Kgmol/hr}$$

Amount of liquid absorbed in liquid phase:

Table 12

		X _i
AA	92.99 - 6.68 = 86.31	0.077
HAC	6.88 - 4.58 = 2.30	0.002
H ₂ O	1146.5 - 113.21 = 1033.29	0.921
Total	1121.90	

$$\sum Y_i = 0.122 * (0.921) + 0.074 * (0.002) + 0.027 * (0.077) = 0.1146$$

$$\text{New flow rate} = 1112.84 / (1 - 0.1146) = 1256.86 \text{ Kgmol/hr}$$

GAS PHASE:

Table 13

AA	$0.027 * (0.077) * (1256.86) = 2.61 \text{Kgmol/hr}$
HAC	$0.074 * (0.002) * 1256.86 = 0.186 \text{Kgmol/hr}$
H ₂ O	$0.122 * (0.921) * (1256.86) = 141.2 \text{Kgmol/hr}$

Table 14

Gas phase	absorbed	Xi
AA=0.027*0.082*1255.92=2.78	90.21	0.082
HAC=0.074*0.006*1255.92=0.56	6.32	0.006
H ₂ O=0.122*0.912*1255.92=139.74	1006.76	0.912

New flow rate : 1255.92

Exiting gas :

Table 15

CO ₂	69.56
AA	2.78
HAC	0.56
O ₂	20.98
N ₂	1022.3
H ₂ O	139.74
Total	1255.92

10.2.1.1 Determination of Recycle flowrate:

The temperature of the gas went from 310°C to 50°C

The temperature of the recycle went from 38°C to 50°C

$$\Delta H_{\text{gas}} = \left[\int_{\text{BP}}^{310} C_p \Delta T \right]_{\text{gas}} + \Delta H_{\text{vap,BP}} + \left[\int_{38}^{\text{BP}} C_p \Delta T \right]_{\text{liquid}} \quad (\text{if the vapor condensed})$$

$$\Delta H_{\text{gas}} = \left[\int_{50}^{310} C_p \Delta T \right]_{\text{gas}} \quad (\text{for non-condensables})$$

$$\int_{50}^{310} C_p \Delta T = A(310 - 50) + \frac{B \times 10^{-3}}{2} (310^2 - 50^2) + \frac{C \times 10^{-7}}{3} (310^3 - 50^3)$$

$$\text{CO}_2: \int_{50}^{310} C_p \Delta T = 60.15 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{50}^{310} C_p \Delta T: 60.15 \frac{\text{cal}}{\text{g}} \times \frac{44.01\text{g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 69,70 \frac{\text{kgmol}}{\text{hr}} = 184,145 \text{ Kcal/hr}$$

$$\text{O}_2: \int_{50}^{310} C_p \Delta T = 59.93 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{50}^{310} C_p \Delta T: 59.93 \frac{\text{cal}}{\text{g}} \times \frac{32\text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 20.98 \frac{\text{kgmol}}{\text{hr}} = 40,237 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{N}_2: \int_{50}^{310} C_p \Delta T = 66.39 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{50}^{310} C_p \Delta T: 66.39 \frac{\text{cal}}{\text{g}} \times \frac{28\text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 1022.3 \frac{\text{kgmol}}{\text{hr}} = 1,900,296 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{HAC}: \int_{50}^{310} C_p \Delta T = 123.7 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{50}^{310} C_p \Delta T : 123.7 \frac{\text{cal}}{\text{g}} \times \frac{60.05\text{g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 0.56 \frac{\text{kgmol}}{\text{hr}} = 4,160 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{AA}: \int_{50}^{310} C_p \Delta T = 73.1 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{50}^{310} C_p \Delta T = 73.1 \frac{\text{cal}}{\text{g}} \times \frac{72.06\text{g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 2.78 \frac{\text{kgmol}}{\text{hr}} = 14,644 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{H}_2\text{O}: \int_{50}^{310} C_p \Delta T = 124.6 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{50}^{310} C_p \Delta T: 124.6 \frac{\text{cal}}{\text{g}} \times \frac{18\text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 159.74 \frac{\text{kgmol}}{\text{hr}} = 313,335 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{Total} = 2,456,817 \text{ Kcal/hr}$$

Condensed gases:

$$\text{HAC} = \int_{117.9}^{310} C_p \Delta T = 82.6 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{117.9}^{310} C_p \Delta T = 82.6 \frac{\text{Cal}}{\text{g}} \times \frac{60.05\text{g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 6.32 \frac{\text{kgmol}}{\text{hr}} = 31,343 \frac{\text{Kcal}}{\text{hr}}$$

$$\Delta H_r = 5733 \frac{\text{Cal}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 6.32 \frac{\text{kgmol}}{\text{hr}} = 36,233 \frac{\text{Kcal}}{\text{hr}}$$

$$\int_{38}^{117.9} (C_p \Delta T)_{\text{liquid}} = 42.13 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{117.9}^{310} C_p \Delta T = 42.13 \frac{\text{cal}}{\text{g}} \times \frac{60.05\text{g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 6.32 \frac{\text{kgmol}}{\text{hr}} = 15,989 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{Total} = 83,565 \text{ Kcal/hr}$$

$$\text{AA} = \int_{141.6}^{310} (C_p \Delta T)_{\text{vapor}} = 55.23 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{141.6}^{310} C_p \Delta T = 55.23 \frac{\text{cal}}{\text{g}} \times \frac{72.06\text{g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 90.21 \frac{\text{kgmol}}{\text{hr}} = 359,058 \frac{\text{Kcal}}{\text{hr}}$$

$$\Delta H_r = 7872 \frac{\text{Cal}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 90.21 \frac{\text{kgmol}}{\text{hr}} = 710,133 \frac{\text{Kcal}}{\text{hr}}$$

$$\int_{38}^{141.6} (C_p \Delta T)_{\text{liquid}} = 57.2 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{141.6}^{310} C_p \Delta T = 57.2 \frac{\text{Cal}}{\text{g}} \times \frac{72.06 \text{g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 90.21 \frac{\text{kgmol}}{\text{hr}} = 371,677 \frac{\text{Kcal}}{\text{hr}}$$

Total = 1,440,868 Kcal/hr

$$\text{H}_2\text{O} = \int_{100}^{310} (C_p \Delta T)_{\text{vapor}} = 101.99 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{100}^{310} C_p \Delta T = 101.99 \frac{\text{Cal}}{\text{g}} \times \frac{18 \text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 1006.76 \frac{\text{kgmol}}{\text{hr}} = 1,848,364 \frac{\text{Kcal}}{\text{hr}}$$

$$\Delta H_r = 2257 \frac{\text{KJ}}{\text{Kg}} \times 18 \frac{\text{Kg}}{\text{Kg mol}} \times 1006.76 \frac{\text{kgmol}}{\text{hr}} \times 0.239006 \frac{\text{Kcal}}{\text{KJ}} = 9,775,660 \frac{\text{Kcal}}{\text{hr}}$$

$$\int_{38}^{100} (C_p \Delta T)_{\text{liquid}} = 62.7 \frac{\text{Cal}}{\text{g}}$$

$$m \int_{38}^{100} C_p \Delta T = 62.7 \frac{\text{Cal}}{\text{g}} \times \frac{18 \text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 1006.76 \frac{\text{kgmol}}{\text{hr}} = 1,135,709 \frac{\text{Kcal}}{\text{hr}}$$

Total = 12,759,733 Kcal/hr

Total ΔH of gas and liquid absorbing components = 14,284,166 Kcal/hr.

10.2.1.2 Absorbing Liquid ΔH

$$\text{H}_2\text{O} = \int_{38}^{50} C_p \Delta T = 12 \frac{\text{Cal}}{\text{g}}$$

$$= 12 \frac{\text{Cal}}{\text{g}} \times \frac{18 \text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 1006.76 \frac{\text{kgmol}}{\text{hr}} = 218,379 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{AA} = \int_{38}^{50} C_p \Delta T = 6.1 \frac{\text{Cal}}{\text{g}}$$

$$= 6.1 \frac{\text{Cal}}{\text{g}} \times \frac{72.06 \text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 90.21 \frac{\text{kgmol}}{\text{hr}} = 39,662 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{HAC : } \int_{38}^{50} C_p \Delta T = 6.05 \frac{\text{Cal}}{\text{g}}$$

$$: 6.05 \frac{\text{cal}}{\text{g}} \times \frac{60.05 \text{ g}}{\text{gmol}} \times 1000 \frac{\text{mol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} \times 6.694 \frac{\text{kgmol}}{\text{hr}} = 2,432 \frac{\text{Kcal}}{\text{hr}}$$

Total = 260,473 Kcal/hr

$$\frac{14,284,166}{260,473} = 54.84 \quad \text{this implies that recycle flowrate is 54.84 times the amount of vapor which is absorbed}$$

Recycle flowrate :

$$\text{H}_2\text{O} = 54.84(1006.76) = 55210 \text{ kgmol/hr}$$

$$\text{AA} = 54.84(90.21) = 4947 \text{ kgmol/hr}$$

$$\text{H}_2\text{O} = 54.84(6.32) = 346.6 \text{ kgmol/hr}$$

60,503.6 kgmol/hr

The cooler duty is 14,284,166 Kcal/hr

10.3 APPENDIX C: ABSORBER SECTION-II

Description:

The recovery of Acrylic acid in this section is specified as 99.9 %. A material balance around T-102 as well as the number of stages required for absorption is calculated with the kremser equation. The absorber will be sized and priced in the subsequent calculation.

P = 760 mmHg T = 50 °C Isothermal conditions L = 227 Kgmol/hr

10.3.1 MATERIAL BALANCE:

99.9 % recovery of AA in the reactor stream is required

92.99 (0.999) = 92.897 kgmol/hr absorbed in T-101 + T-102

92.99 – 92.897 = 0.093 kgmol/hr of AA in the T-102 effluent

Table 16

Gas Fed to T-102:		K = $\frac{P_{sat}}{P_{total}}$	y_{N_p+1}
Inerts (CO ₂ , N ₂ , O ₂)	1112.84 kgmol/hr	-	-
AA	2.78 kgmol/hr	0.027	2.214×10^{-3}
HAC	0.56 kgmol/hr	0.074	4.459×10^{-4}
H ₂ O	139.74 kgmol/hr	0.122	0.1113
	1255.92 kgmol/hr		

$$\text{Absorption of AA} = \frac{2.78 - 0.093}{2.78} \times 100 = 96.65 \% \text{ absorbed in T-102}$$

The kremser equation will be used to estimate the number of trays and the absorption of the other two components. The absorption is specified at 96.65 %. Kremser equation is taken from Treybal, Robert E., Mass Transfer Operations, 3rd ed., McGraw Hill, N.Y page. 326

$$\frac{Y'_{N_p+1} - Y'_1}{Y'_{N_p+1}} = \left(1 - \frac{L_o X'_o}{A' G_{N_p+1} Y'_{N_p+1}} \right) \frac{A_E^{N_p+1} - A_E}{A_E^{N_p+1} - 1}$$

where, Y' : mole/mole entering gas

N_p : number of equilibrium stages

A : absorption factor, L/mG

G : total gas rate, mole/time

L : total liquid rate, mole/time

X' : mole/mole entering liquid

Subscript o entering top tray

$$A' = \frac{A_{N_p}(A_1 + 1)}{A_{N_p+1}}, \quad A_E = \left[A_{N_p}(A_1 + 1) + 0.25 \right]^{1/2} - 0.5$$

Since the absorbing liquid is pure H₂O the equation reduces to:

$$\frac{Y'_{N_p+1} - Y'_1}{Y'_{N_p+1}} = \frac{A_E^{N_p+1} - A_E}{A_E^{N_p+1} - 1}$$

$$\frac{L}{G} \approx \frac{227 + (2.78 - 0.093) + 0.5}{1255.95} = 0.1833 \text{ at the bottom of the tower}$$

$$\frac{L}{G} \approx \frac{227}{1255.92 (2.78 - 0.093) - 0.5} = 0.1812 \text{ at the top of the tower}$$

$$\text{Average } L/G = 0.1823 , m=0.027$$

$$\text{A rough estimate for } A'_E = \frac{0.1823}{0.027} = 6.75$$

Plugging this in the kremser equation yields:

$$0.9665 = \frac{6.75^{N_p+1} - 6.75}{6.75^{N_p+1} - 1}$$

$$N_p + 1 = 2.7 \Rightarrow N_p = 1.7 \approx 2 \text{ stages}$$

$$\frac{Y'_{N_p+1} - Y'_1}{Y'_{N_p+1}} = \frac{6.75^3 - 6.75}{6.75^3 - 1} = 0.9812 = \frac{2.78 - Y}{2.78} \Rightarrow Y = 0.0523 \text{ kgmol/hr}$$

For HAC

$$m = 0.074$$

$$\text{a rough estimate for } A'_E = \frac{0.1823}{0.074} = 2.46$$

$$\frac{2.46^3 - 2.46}{2.46^3 - 1} = 0.895 = \frac{0.56 - Y}{0.56} \Rightarrow Y = 0.0589 \frac{\text{kgmol}}{\text{hr}} \text{ leaving absorber}$$

For H₂O: (present in both entering liquid and gas)

$$m = 0.122$$

$$\text{a rough estimate for } A'_E = \frac{0.1823}{0.122} = 1.494 , S = \frac{1}{1.494}$$

$$\text{Absorbing: } 1 - \frac{L_o X'_o}{A' G_{N_p+1} Y'_{N_p+1}} = 1 - \frac{227(1)}{1.494(1255.92)(0.1113)} = -0.089$$

$$\text{Stripping: } 1 - \frac{G_{N_p+1} Y'_{N_p+1}}{S' L_o X'_o} = 1 - \frac{(1255.92)(0.1113)}{\frac{1}{1.494}(227)(1)} = 0.065$$

These equations will be checked using the more rigorous Horton-Franklin method (Treybal)

$$G_n \approx G_{N_p+1} \left(\frac{G_1}{G_{N_p+1}} \right)^{(N_p+1-n)/N_p}, \quad G_1 = 1112.84 + 139.74 + 0.0589 + 0.0523 = 1252.69 \frac{\text{kgmol}}{\text{hr}}$$

$$L_n \approx L_o \left(\frac{L_{N_p}}{L_o} \right)^{n/N_p}, \quad L_{N_p} = 227 + 0.5012 + 2.73 = 230.23 \frac{\text{kgmol}}{\text{hr}}$$

$$G_{N_p+1} = 1255.92 \frac{\text{kgmol}}{\text{hr}}, \quad L_o = 227 \frac{\text{kgmol}}{\text{hr}}$$

Table 17

Tray n	G _n	L _n	L _n /G _n
1	1252.7	228.6	0.1825
2	1254.3	230.23	0.1836

Comp	A ₁	A ₂	$\frac{Y'_{N_p+1} - Y'_1}{Y'_{N_p+1}}$
HAC	2.466	2.481	0.8958
AA	6.759	6.8	0.9814

$$\text{AA: } 0.9814 = \frac{2.78 - (\text{AA in extraction})}{2.78} \Rightarrow \text{AA leaving} = 0.0517 \frac{\text{kgmol}}{\text{hr}}$$

$$\text{HAC: } 0.8958 = \frac{0.56 - (\text{HAC in extraction})}{0.56} \Rightarrow \text{HAC leaving} = 0.0584 \frac{\text{kgmol}}{\text{hr}}$$

Table 18

Gas Stream In T-102		Gas Stream Out T-102	
Inerts	1112.84	Inerts	1112.84
H ₂ O	139.74	H ₂ O	139.74
AA	2.78	AA	0.0517
HAC	0.56	HAC	0.0584
	1255.92 kgmol/hr		1252.69 kgmol/hr

Table 19

Water Stream Out	
H ₂ O	227
AA	2.7283
HAC	0.5016
	230.23 kgmol/hr

10.4 APPENDIX D: INCINERATOR:

Description:

The purpose of this section is to do a material and energy balance around the incinerator. A dew-point calculation is also done on the exiting air stream. The exiting airstream will be cooled to 10°C above its dew-point to recover heat in the waste gas heat boiler. The boiler and the incinerator will be sized and priced in later calculations.

Feed gas consists of

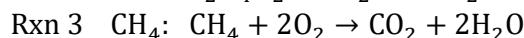
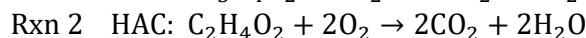
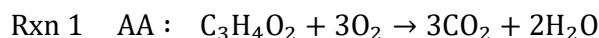
Table 20

CO ₂	69.56	kgmol/hr
AA	0.0517	kgmol/hr
HAC	0.0584	kgmol/hr
O ₂	20.98	kgmol/hr
N ₂	1022.3	kgmol/hr
H ₂ O	139.74	kgmol/hr

Given: Incinerator temp 700°C
25% excess oxygen required
Temp of feed gas 50°C

Assumed: The CH₄ is available at 20°C

Combustion reactions:



ΔH combustion (20 °C):

$$\text{AA: } 317.2 \frac{\text{Kcal}}{\text{gmol AA}} \times 51.7 \frac{\text{gmol}}{\text{hr}} = 16,399.24 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{HAC: } 218.4 \frac{\text{Kcal}}{\text{gmol AA}} \times 58.4 \frac{\text{gmol}}{\text{hr}} = 12,754.56 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{CH}_4: 211.2 \frac{\text{Kcal}}{\text{gmol AA}} \times \text{CH}_4 \frac{\text{Kgmol}}{\text{hr}} \times 1000 \frac{\text{gmol}}{\text{Kgmol}} = 211,200 \text{CH}_4 \frac{\text{Kcal}}{\text{hr}}$$

The ΔH of combustion is given at 20°C but the incinerator is at 700°C so the products of the reaction must be heated to 700°C and the reactions cooled to 20°C:

$$m \int_{T_{\text{feed}}}^{20} [C_p \Delta T_{\text{vapor}}]_{\text{reactants}} + \Delta H_{\text{combustion}} + m \int_{20}^{700} [C_p \Delta T_{\text{liquid}}]_{\text{products}}$$

Rxn 1: Reactants

$$\text{AA} = -17.11 \frac{\text{cal}}{\text{g}} \times 72.06 \frac{\text{g}}{\text{gmol}} \times 51.7 \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = -63.74 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{O}_2 = -6.73 \frac{\text{cal}}{\text{g}} \times 32 \frac{\text{g}}{\text{gmol}} \times 3(51.7) \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = -33.43 \frac{\text{Kcal}}{\text{hr}}$$

Products:

$$\text{H}_2\text{O} = 397.46 \frac{\text{cal}}{\text{g}} \times 18 \frac{\text{g}}{\text{gmol}} \times 2(51.7) \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 739.75 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{CO}_2 = 171.4 \frac{\text{cal}}{\text{g}} \times 44.01 \frac{\text{g}}{\text{gmol}} \times 3(51.7) \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 1170 \frac{\text{Kcal}}{\text{hr}}$$

$$\Delta H_{\text{combustion}} = -16,399.24 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{Total } \Delta H \text{ for reaction: } -14,586.66 \frac{\text{Kcal}}{\text{hr}}$$

Rxn 2: Reactants

$$\text{HAC} = -19.92 \frac{\text{cal}}{\text{g}} \times 60.05 \frac{\text{g}}{\text{gmol}} \times 58.4 \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = -69.85 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{O}_2 = -6.73 \frac{\text{cal}}{\text{g}} \times 32 \frac{\text{g}}{\text{gmol}} \times 2(58.4) \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = -69.85 \frac{\text{Kcal}}{\text{hr}}$$

Products:

$$\text{H}_2\text{O} = 397.46 \frac{\text{cal}}{\text{g}} \times 18 \frac{\text{g}}{\text{gmol}} \times 2(58.4) \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 835.62 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{CO}_2 = 171.4 \frac{\text{cal}}{\text{g}} \times 44.01 \frac{\text{g}}{\text{gmol}} \times 3(58.4) \frac{\text{gmol}}{\text{hr}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 881.06 \frac{\text{Kcal}}{\text{hr}}$$

$$\Delta H \text{ combustion} = -12,754.56 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{Total } \Delta H \text{ for reaction : } -11,32.88 \frac{\text{Kcal}}{\text{hr}}$$

Rxn 3: Reactant

CH_4 assume : fed at 20°C

$$\text{O}_2 = -6.73 \frac{\text{cal}}{\text{g}} \times 32 \frac{\text{g}}{\text{gmol}} \times 2\text{CH}_4 \frac{\text{Kgmol}}{\text{hr}} \times \frac{1000\text{gmol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = -430.85\text{CH}_4 \frac{\text{Kcal}}{\text{hr}}$$

Products:

$$\text{H}_2\text{O} = 397.46 \frac{\text{cal}}{\text{g}} \times 18 \frac{\text{g}}{\text{gmol}} \times 2\text{CH}_4 \frac{\text{Kgmol}}{\text{hr}} \times \frac{1000\text{gmol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 14,308.6\text{CH}_4 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{CO}_2 = 171.4 \frac{\text{cal}}{\text{g}} \times 44.01 \frac{\text{g}}{\text{gmol}} \times \text{CH}_4 \frac{\text{Kgmol}}{\text{hr}} \times \frac{1000\text{gmol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 7543.3\text{CH}_4 \frac{\text{Kcal}}{\text{hr}}$$

$$\Delta H \text{ combustion} = -211,200\text{CH}_4 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{Total } \Delta H \text{ (for oxygen at } 50^\circ\text{C}) : -189,779\text{CH}_4 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{Total } \Delta H \text{ (for oxygen at } 20^\circ\text{C}) : -189,38\text{CH}_4 \frac{\text{Kcal}}{\text{hr}}$$

Inerts: ΔH not affected by rxn

$$\Delta H = m \int_{50}^{700} C_p \Delta T$$

$$\text{CO}_2 : 165.29 \frac{\text{cal}}{\text{g}} \times 44 \frac{\text{g}}{\text{gmol}} \times 69.56 \frac{\text{Kgmol}}{\text{hr}} \times \frac{1000\text{gmol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 506,008 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{N}_2 : 171.34 \frac{\text{cal}}{\text{g}} \times 28.01 \frac{\text{g}}{\text{gmol}} \times 1022.3 \frac{\text{Kgmol}}{\text{hr}} \times \frac{1000\text{gmol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 4,906,256 \frac{\text{Kcal}}{\text{hr}}$$

$$\text{H}_2\text{O} : 384.05 \frac{\text{cal}}{\text{g}} \times 18 \frac{\text{g}}{\text{gmol}} \times 139.74 \frac{\text{Kgmol}}{\text{hr}} \times \frac{1000\text{gmol}}{\text{kgmol}} \times \frac{\text{Kcal}}{1000 \text{ cal}} = 966,009 \frac{\text{Kcal}}{\text{hr}}$$

$$= 6,378,273 \text{ Kcal/hr}$$

Oxygen needed: (25% excess needed)

$$\text{Rxn 1 and 2: } 3(0.0517) + 2(0.0584) = 0.271 \frac{\text{kgmol}}{\text{hr}}$$

Total oxygen used in 3 rxns : $0.271 + 2\text{CH}_4$

$$\text{Total oxygen : } 1.25[0.271 + 2\text{CH}_4] = [0.339 + 2.5\text{CH}_4]$$

Nitrogen: (from additional oxygen from the fan)

$$\left(\frac{0.339 + 2.5\text{CH}_4 - 20.98}{0.21} \right) - (0.339 + 2.5\text{CH}_4 - 20.98) \\ = 3.769(2.5\text{CH}_4 - 20.64)$$

Water: (from additional oxygen from the fan)

$$0.0189 \frac{\text{mole H}_2\text{O}}{\text{mole air}} \text{ (from reactor calculation)}$$

$$= 0.0189 \left(\frac{0.339 + 2.5\text{CH}_4 - 20.98}{0.21} \right) = 0.09(2.5\text{CH}_4 - 20.64)$$

CH_4 : (which can react without the air from the fan)

$$20.98 = [0.339 + 2.5\text{CH}_4]$$

$$\text{CH}_4 = 8.256 \text{ kgmol/hr}$$

Case 1: $\text{CH}_4 < 8.256 \frac{\text{kgmol}}{\text{hr}}$ (no air introduced by fan)

Enthalpy balance:

$$\Delta H_{\text{inerts}} + \Delta H_{\text{rxn1}} + \Delta H_{\text{rxn2}} + \Delta H_{\text{unused O}_2} = \Delta H_{\text{rxn3}}$$

(50 – 700 °C)

$$\begin{aligned}\Delta H_{\text{unused O}_2} &= \int_{20}^{700} C_p \Delta T = 161.48 \frac{\text{Cal}}{\text{g}} \\ &= 161.48 \frac{\text{Cal}}{\text{g}} \times 32 \frac{\text{g}}{\text{gmol}} \times [20.98 - 0.271 - 2\text{CH}_4] \frac{\text{kgmol}}{\text{hr}} \\ &= 5167.36 [20.71 - 2\text{CH}_4] \frac{\text{Kcal}}{\text{hr}}\end{aligned}$$

ΔH for all other terms already calculated

$$6,378,273 - 14586.66 - 11,132.88 + 5167.36[20.71 - 2\text{CH}_4] = 189,779\text{CH}_4$$

$$6459569.5 = 179444.3 \text{ CH}_4$$

$$\text{CH}_4 = 35.99 > 8.256 \quad \therefore \text{ go to case 2}$$

Case 2: $\text{CH}_4 > 8.256 \frac{\text{kgmol}}{\text{hr}}$ (additional O_2 , N_2 and H_2O introduced by fan)

Enthalpy balance:

$$\Delta H_{\text{inerts}} + \Delta H_{\text{rxn1}} + \Delta H_{\text{rxn2}} + \Delta H_{\text{unused O}_2} + \Delta H_{\text{N}_2} + \Delta H_{\text{H}_2\text{O}} = \Delta H_{\text{rxn3}}$$

$$\begin{aligned}\Delta H_{\text{unused O}_2} &= m \int_{20}^{700} C_p \Delta T = 161.48 \frac{\text{Cal}}{\text{g}} \\ &= 161.48 \frac{\text{Cal}}{\text{g}} \times 32 \frac{\text{g}}{\text{gmol}} \times [0.339 + 2.5\text{CH}_4 - (0.271 + 2\text{CH}_4)] \frac{\text{kgmol}}{\text{hr}} \\ &= 5167.36(0.068 + 0.5\text{CH}_4) \frac{\text{Kcal}}{\text{hr}}\end{aligned}$$

$$\begin{aligned}\Delta H_{\text{N}_2} &= m \int_{20}^{700} C_p \Delta T = 178.8 \frac{\text{Cal}}{\text{g}} \\ &= 178.8 \frac{\text{Cal}}{\text{g}} \times 28 \frac{\text{g}}{\text{gmol}} \times 3.76[2.5\text{CH}_4 - 20.64] \frac{\text{kgmol}}{\text{hr}} \\ &= 18824(2.5\text{CH}_4 - 20.64) \frac{\text{KCal}}{\text{hr}}\end{aligned}$$

$$\begin{aligned}\Delta H \text{ of } \text{H}_2\text{O} &= m \int_{20}^{700} C_p \Delta T = 397.5 \frac{\text{Cal}}{\text{g}} \\ &= 397.6 \frac{\text{Cal}}{\text{g}} \times 18 \frac{\text{g}}{\text{gmol}} \times 0.09[2.5\text{CH}_4 - 20.64] \frac{\text{kgmol}}{\text{hr}} \\ &= 643.95(2.5\text{CH}_4 - 20.64) \frac{\text{KCal}}{\text{hr}}\end{aligned}$$

$$\Delta H_{\text{rxn}} = -189779(8.256) - 189348(\text{CH}_4 - 8.256)$$

ΔH for all other terms is already calculated

$$6378273 - 14586.66 - 11132.88 + 5167.36(0.068 + 0.5\text{CH}_4) - 18824(2.5\text{CH}_4 - 20.64) + 643.95(2.5\text{CH}_4 - 20.64) = 189779(8.256) + 189348(\text{CH}_4 - 8.256)$$

$$5951086.3 + 51253.55\text{CH}_4 = 3558.3 + 189348\text{CH}_4$$

$$\text{CH}_4 = 43 \frac{\text{Kgmol}}{\text{hr}}$$

Material balance: (exit stream from incinerator)

$$\text{H}_2\text{O} = (0.0517)(2) + (0.0584)(2) + 43(2) + -0.09[2.5(43) - 20.64] + 139.74 = \\ 233.78 \frac{\text{Kgmol}}{\text{hr}}$$

$$\text{CO}_2 = 3(0.0517) + 2(0.0584) + 43 + 69.56 = 112.83 \text{ Kgmol/hr}$$

$$\text{N}_2 = 1022.3 + 3.76(2.5(43) - 20.64) = 1348.9 \text{ Kgmol/hr}$$

$$\text{O}_2 = [0.339 + 25(43)] - [0.271 + 2(43)] = 21.57 \text{ Kgmol/hr}$$

Air Stream:

$$\text{O}_2 \text{ fed} = [0.339 + 2.5(43) - 20.98] = 86.86 \text{ Kgmol/hr}$$

$$\text{N}_2 \text{ fed} = \frac{86.86}{0.21} - 86.86 = 326.75 \text{ Kgmol/hr}$$

$$\text{H}_2\text{O} = 0.189(326.75 + 86.86) = 7.82 \text{ Kgmol/hr}$$

$$\text{Dew point of this exiting stream: } \frac{y}{x} = \frac{P_{\text{sat}}}{P_{\text{total}}} \quad \text{assume } P_{\text{total}} = 101 \text{ kPa}$$

Table 21

y_1	K (T=100°C)	x_1	K (T = 50°C)	x_1	K (T = 52°C)	x_1
O ₂ :0.0123	-	0	-	0	-	0
N ₂ : 0.786	-	0	-	0	-	0
H ₂ O:0.136	0.9953	0.1366	0.1218	1.117	0.134	1.01
CO ₂ : 0.0657	-	0	-	0	-	0
		$\sum x = 0.1366$		$\sum x = 1.117$		$\sum x = 1.01$

Close enough, T=52°C

Cool to 10°C above dew point, at 62 °C

10.5 APPENDIX E: EXTRACTION

Description:

A material balance and the number of stages required for extraction is done for both solvents. The equipment is seized and the utility requirements are calculated for both the solvents. The present worth is calculated for both solvent cases. The solvent case with the greatest present worth is chosen.

Stream 12 Feed

Table 22

Component	Kgmol/hr	Kg/hr	Mole fraction x (wt%)
AA	92.938	6697	0.2424
HAC	6.822	409.66	
H ₂ O	1233.76	22,207.68	0.7576
	1333.52	29,314.34	

10.5.1 Case 1: Solvent A

The extract is composed of the organic phase from the overhead of the extract stripper and the overhead of the raffinate stripper. A material balance will have to be done to obtain the composition of the extract stream.

97.5 % of the acids are to be recovered

Stream 15 contains:

$$\begin{aligned} \text{AA: } 0.975(92.938) &= 90.61 \text{ kgmol/hr or } 6529.36 \text{ kg/hr} \\ \text{HAC : } 0.975(6.822) &= 6.65 \text{ kgmol/hr or } 399.33 \text{ kg/hr} \\ &\hline & 6928.69 \text{ kg/hr} \end{aligned}$$

Stream 16 contains:

$$\begin{aligned} \text{AA: } 92.938 - 90.61 &= 2.33 \text{ kgmol/hr or } 167.9 \text{ kg/hr} \\ \text{HAC : } 6.822 - 6.65 &= 0.172 \text{ kgmol/hr or } 10.33 \text{ kg/hr} \end{aligned}$$

Stream 27 contains: (100% pure acid bottoms, 100% recovery)

$$\begin{aligned} \text{AA} &= 90.61 \text{ kgmol/hr or } 6529.36 \text{ kg/hr} \\ \text{AA} &= 6.65 \text{ kgmol/hr or } 399.33 \text{ kg/hr} \end{aligned}$$

Composition of Stream 38:

Consists of H₂O, HAC and AA. We are going to assume that there is no solvent in the bottoms because there is no solvent feed to make-up for losses. We will assume that none of the acid goes overhead in T-105.

The composition can be determined because bottoms temperature is known and the overhead pressure is known. The bottoms pressure is calculated by adding the ΔP across the trays to the overhead pressure. Since the bottom pressure and temperature are known then a bubble-point calculation can be done to determine composition. 10 trays (given) assume real trays since the tower is already sized.

$$\Delta P = 10(3.5 \text{ mmHg}) \times \frac{101.3 \text{ kPa}}{760 \text{ mmHg}} = 4.66$$

$$P_{\text{bottoms}} = 4.66 + 101.3 = 105.96 \text{ kPa}$$

$$T_{\text{bottoms}} = 100 \text{ }^{\circ}\text{C}$$

$$K = \frac{P_{\text{sat}}}{P_{\text{total}}} = \frac{Y}{X} \quad P_{\text{sat}} - \text{calculated with coefficients from table 3}$$

Table 23

	Kg mol/hr	K	X	Y = (K * X)
H ₂ O	?	0.9487	H ₂ O / H ₂ O + 2.502	0.9487(H ₂ O / H ₂ O + 2.502)
AA	2.33	0.251	2.33 / H ₂ O + 2.502	0.251(2.33 / H ₂ O + 2.502)
HAC	0.172	0.5237	0.172 / H ₂ O + 2.502	0.5237(0.172 / H ₂ O + 2.502)
	H ₂ O + 2.502			

$$\sum Y = 1 = 0.9487 \left(\frac{H_2O}{H_2O+2.502} \right) + 0.251 \left(\frac{2.33}{H_2O+2.502} \right) + 0.5237 \left(\frac{0.172}{H_2O+2.502} \right)$$

$$H_2O + 2.502 = 0.9487H_2O + 0.5848 + 0.0901$$

⇒ H₂O resulted in a negative value which means there is probably some solvent in the bottoms but there is no vapor pressure data on the solvent in the bottoms. To simplify calculations we will assume that pure solvent is present in the bottoms.

Composition of Stream 38:

$$AA = 2.33 \text{ Kgmol/hr}$$

$$HAC = 0.172 \text{ Kgmol/hr}$$

$$H_2O = 1233.76 \text{ Kgmol/hr}$$

10.5.1.1 T-104 Decanter:

Organic Phase composition : 99.8% solvent ; 0.2% water (wt%)

Liquid Phase composition : 98.4% water ; 1.6% solvent (wt%)

10.5.1.2 Material Balance:

In order to calculate the number of stages for extraction all three of the following things must be known or specified:

1. The flowrate and composition of the feed, F
2. The flowrate and composition of the solvent, S
3. The distribution of the solute in the extract, E₁ and raffinate, R_{np}.

At this point the flowrate and composition of the feed and the distribution of the solute in E₁ and R_{np} is known. The material balance will have to be solved by guessing the solvent composition and flowrate and specifying the composition of the extract and raffinate. We will then do a material balance which yield the composition of the solvent and then compare this to the initial guess. We will keep on doing the material balance until the solvent feed composition does not change.

To begin with we will take pure solvent feed.

The feed is contacted with an equal amount of solvent, so we will assume that mass flowrate of the solvent is equal to the mass flowrate of the feed.

Solvent flowrate = 29,314.34 Kg/hr

The material balance around the extractor will be solved with the use of the ternary diagram:

$$X_M = \frac{Fx_F + Sx_S}{F + S} = \frac{29314.34(0.2424)}{29314.34 + 29314.34} = 0.1212$$

Determining the composition of E₁ and R_{np} is a trial and error process. The E₁ R_{np} line must pass through point X_M on the diagram.

Try X_{E1} = 18%

Composition of Raffinate

Acid 2.9%

Solvent 1.7%

H₂O 95.4%

Composition of Extract

Acid 18%

Solvent 80%

H₂O 2%

$$\text{Flowrate of Extract} : \frac{29314.34(0.2424)(0.975)}{0.18} = 38,492.7 \frac{\text{kg}}{\text{hr}}$$

$$\text{Flowrate of Raffinate} : \frac{29314.34(0.2424)(0.025)}{0.029} = 6125.69 \frac{\text{kg}}{\text{hr}}$$

$$: 38492.7 + 6125.69 = 44618.3 \text{ kg/hr} < 2(29314.34) = 58628.7 \text{ kg/hr}$$

Try $X_{E1} = 19\%$

Composition of Raffinate
Acid 1.4%
Solvent 1.6%
 H_2O 97%

Composition of Extract
Acid 19%
Solvent 79%
 H_2O 2%

$$\text{Extract} = 36463.95 \text{ kg/hr}$$
$$\text{Raffinate} = 10450 \text{ kg/hr}$$

$$46913.95 \text{ kg/hr} < 58,628.7 \text{ kg/hr}$$

Try $X_{E1} = 19.5\%$

Composition of Raffinate
Acid 0.5%
Solvent 1.8%
 H_2O 97.7%

Composition of Extract
Acid 19.5%
Solvent 78.5%
 H_2O 2%

$$\text{Extract} = 35,529 \text{ kg/hr}$$
$$\text{Raffinate} = 35,530 \text{ kg/hr}$$

$$71,059 \text{ kg/hr} > 58,628.7 \text{ kg/hr}$$

Interpolate: Try $X_{E1} = 19.25\%$

Composition of Raffinate
Acid 0.78%
Solvent 1.52%
 H_2O 97.7%

Composition of Extract
Acid 19.25%
Solvent 78.75%
 H_2O 2%

$$\text{Extract} = 35,990 \text{ kg/hr}$$
$$\text{Raffinate} = 22,776 \text{ kg/hr}$$

$$58,956 \text{ kg/hr} \approx 58,628.7 \text{ kg/hr}$$

In Extract:

$$\text{Acid} = 35990(0.1925) = 6928.15 \text{ kg/hr}$$

$$\text{Solvent} = 35990(0.7875) = 28342.12 \text{ kg/hr}$$

$$\text{H}_2\text{O} = 35990(0.02) = 719.8 \text{ kg/hr}$$

In Raffinate:

$$\text{Acid} = 177.65 \text{ kg/hr}$$

$$\text{Solvent} = 346.2 \text{ kg/hr}$$

$$\text{H}_2\text{O} = 22252 \text{ kg/hr}$$

The water and solvent fed do not exactly equal to the water and solvent in E_1 and R_{Np} , this is due to difficulty of accurately reading the ternary diagram. So we will be adjusting the numbers so the material balance will work.

<u>Extract</u>		<u>Raffinate</u>	
Acid	6928.15 kg/hr	Acid	177.65 kg/hr
Solvent	28342.12 kg/hr	Solvent	972.22 kg/hr
H_2O	719.8 kg/hr	H_2O	21487.9 kg/hr

Determination of the Solvent Feed composition (Stream 17)

R_{Np} (stream 16) is fed to T-105 where all of the solvent is recovered. The distillate from the extraction stripping tower is fed to a drum to separate the organic and water phases. The water phase is mixed with stream 16 before it enters T-105. The organic phase is mixed with pre solvent from the Raffinate stripper overhead and is then recycled back to the extractor.

Composition of Phases in Decanter:

Organic Phase composition :	99.8% solvent ; 0.2% water	(stream 22)
Water composition :	98.4% water ; 1.6% solvent	(stream 28)

Flowrate of streams 22+28 = $(28342.12 + 719.8) = 29,062 \text{ kg/hr}$

Trial and error to obtain the actual flowrates

- Organic phase = 28360 kg/hr

Solvent	= 0.998(28360)	= 28303.3 kg/hr
H_2O	= 0.002(28360)	= 56.72 kg/hr

<u>Water phase</u>		(%)
H_2O	= 719.8 - 56.72	= 663.08 kg/hr
Solvent	= 28342.12 - 28303.3	<u>= 38.82 kg/hr</u>
		701.9 kg/hr

- Organic phase = 28380 kg/hr

Solvent	= 0.998(28380)	= 28323.24 kg/hr
H_2O	= 0.002(28380)	= 56.76 kg/hr

<u>Water phase</u>		(%)
H_2O	= 719.8 - 56.76	= 663.04 kg/hr
Solvent	= 28342.12 - 28323.24	<u>= 19.88 kg/hr</u>
		681.92 kg/hr

- Organic phase = 28390 kg/hr

Solvent	= 0.998(28390)	= 28333.22 kg/hr
H_2O	= 0.002(28390)	= 56.78 kg/hr

<u>Water phase</u>		(%)
H_2O	= 719.8 - 56.78	= 663.02 kg/hr
Solvent	= 28342.12 - 28333.22	<u>= 8.9 kg/hr</u>
		681.92 kg/hr

- Organic phase = 28388 kg/hr (interpolate)

Solvent	= 0.998(28388)	= 28331.22 kg/hr
H_2O	= 0.002(28388)	= 56.76 kg/hr

<u>Water phase</u>		(%)
H_2O	= 719.8 - 56.76	= 663.02 kg/hr
Solvent	= 28342.12 - 28331.22	<u>= 10.9 kg/hr</u>
		673.92 kg/hr

The composition of E_1 can now be determined:

Stream 17:

Solvent in stream 28 and stream 16 (R_{Np})

Stream 22

$$\text{Solvent} : 10.9 + 972.22 + 28331.22 = 29,313.34 \text{ kg/hr}$$

$$\text{H}_2\text{O} : 56.76 \text{ kg/hr}$$

Begin 2nd iteration:

$$X_M = \frac{29314.34(0.2424)}{2(29314.34) + 56.76} = \frac{7105.8}{58685.44} = 0.1211$$

The same material balance line will be used because the solvent point and X_M moved insignificantly.

Composition of Raffinate

$$\text{Acid} \quad 0.78\%$$

$$\text{Solvent} \quad 1.52\%$$

$$\text{H}_2\text{O} \quad 97.7\%$$

Composition of Extract

$$\text{Acid} \quad 19.25\%$$

$$\text{Solvent} \quad 78.75\%$$

$$\text{H}_2\text{O} \quad 2\%$$

Extract 35990 kg/hr

$$\text{Raffinate} \quad 22695.4 \text{ kg/hr}$$

$$\text{Acid} \quad 6928.15 \text{ kg/hr}$$

$$\text{Acid} \quad 177.65 \text{ kg/hr}$$

$$\text{Solvent} \quad 28342.12 \text{ kg/hr}$$

$$\text{Solvent} \quad 972.22 \text{ kg/hr}$$

$$\text{H}_2\text{O} \quad 719.8 \text{ kg/hr}$$

$$\text{H}_2\text{O} \quad 21544.66 \text{ kg/hr}$$

The composition of the solvent feed stream will not change because the feed to the extract stripping tower is the same as in the previous iteration. The material balance is now finished and the number of stages can be determined. The stages are stepped off below.

Extraction requires 11 theoretical stages.

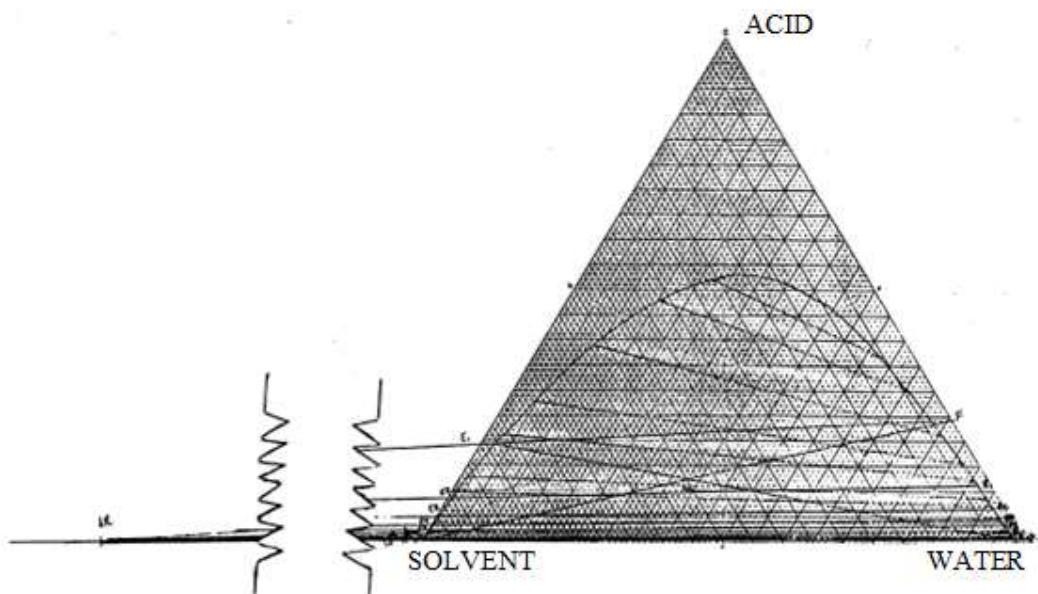


Figure 6

10.5.2 Case 2: Solvent B

The material balance will be solved the same way as case 1.

For 1st iteration- assume Solvent Feed is 98% solvent and 2% water.

$$\text{Solvent Feed (S)} = \frac{29314.34 \text{ kg/hr}}{0.98} = 29912.59 \frac{\text{kg}}{\text{hr}}$$

$$\text{Water} = 0.02(29912.59) = 598.25 \text{ kg/hr}$$

$$\text{Solvent} = 0.98(29912.59) = 29314.34 \text{ kg/hr}$$

$$X_M = \frac{F_{x_F} + S_{x_S}}{F + S} = \frac{29314.34(0.2424)}{29912.59 + 29314.34} = 0.12$$

Determining the composition of E_1 and R_{Np} is a trial and error process. The line $E_1 R_{Np}$ must pass through X_m on the material balance line FS.

Try $E_1 = 19\%$

Composition of Raffinate

Acid	2.5%
Solvent	15.1%
H_2O	82.4%

Composition of Extract

Acid	19%
Solvent	74.6%
H_2O	6.4%

$$\text{Flowrate of Extract : } \frac{97.5\% \text{ of total acid}}{0.19} = \frac{6928.15}{0.19} = 36,463.95 \frac{\text{kg}}{\text{hr}}$$

$$\begin{aligned} \text{Flowrate of Raffinate : } & \frac{2.5\% \text{ of total acid}}{0.025} = \frac{177.65}{0.025} = 7,106 \frac{\text{kg}}{\text{hr}} \\ & = 43,569.95 \text{ kg/hr} \end{aligned}$$

Material balance:

$$F + S = E + R$$

$$F + S = 29912.59 + 29314.34 = 59,226.93 \text{ kg/hr}$$

$$43,569.95 \text{ kg/hr} < 59,226.93 \text{ kg/hr}$$

Try $E_1 = 20\%$

Compostion of Raffinate

Acid	0.5%
Solvent	15.6%
H_2O	83.9%

Composition of Extract

Acid	20%
Solvent	73.3%
H_2O	6.7%

$$\text{Flowrate of } E_1 = 34,640.75 \text{ kg/hr}$$

$$\text{Flowrate of } R_{Np} = 35,530 \text{ kg/hr}$$

$$70,170.75 \text{ kg/hr} > 59,226.93 \text{ kg/hr}$$

Try $E_1 = 19.25\%$

Compostion of Raffinate

Acid	1.3%
Solvent	15.4%
H_2O	83.3%

Composition of Extract

Acid	19.25%
Solvent	74.2%
H_2O	6.55%

$$\text{Flowrate of } E_1 = 35990.39 \text{ kg/hr}$$

$$\text{Flowrate of } R_{Np} = 13,665.38 \text{ kg/hr}$$

$$49,655.77 \text{ kg/hr} < 59,226.93 \text{ kg/hr}$$

Try $X_{E1} = 19.7\%$

Composition of Raffinate	
Acid	0.74%
Solvent	15.95%
H ₂ O	83.31%

Composition of Extract	
Acid	19.7%
Solvent	72.5%
H ₂ O	6.8%

Flowrate of E₁ = kg/hr

$$\text{Flowrate of } R_{Np} = 24,00635,168.27 \text{ kg/hr}$$

$$59,175.03 \text{ kg/hr close enough to } R+S = 59,226.93 \text{ kg/hr}$$

In Extract:

$$\text{Acid} = 35,168.27 (0.1925) = 6928.15 \text{ kg/hr}$$

$$\text{Solvent} = 35,168.27 (0.7875) = 25,848.7 \text{ kg/hr}$$

$$\text{H}_2\text{O} = 35,168.27 (0.02) = 2391.4 \text{ kg/hr}$$

In Raffinate:

$$\text{Acid} = 177.65 \text{ kg/hr}$$

$$\text{Solvent} = 3,465.64 \text{ kg/hr}$$

$$\text{H}_2\text{O} = 20415.4 \text{ kg/hr}$$

E₁ is sent to the extract stripper where the solvent and water all goes overhead. The overhead is separated into two phases in a decanter. The organic phase plus the pure solvent from the Raffinate stripper is recycled back to the solvent feed.

Composition of the phases

Organic Phase composition : 96.5% solvent ; 3.5% water

Water composition : 84.5% water ; 15.5% solvent

$$E_1 = 35168.25 \text{ kg/hr}$$

Trial and error

- Try 1: Organic phase = 25,500 kg/hr

Solvent	= 0.965(25500)	= 24,607.5 kg/hr
H ₂ O	= 0.035(25500)	= 892.5 kg/hr

<u>Water phase</u>		<u>q_o(%)</u>
H ₂ O	= 2391.4 – 892.5	= 1498.9 kg/hr
Solvent	= 25848.7 – 24607.5	<u>= 1241.9 kg/hr</u>
		2740.1 kg/hr

- Try 2: Organic phase = 26,000 kg/hr

Solvent	= 0.965(26000)	= 25,090 kg/hr
H ₂ O	= 0.035(26000)	= 910 kg/hr

<u>Water phase</u>		<u>q_o(%)</u>
H ₂ O	= 2391.4 – 910	= 1481.4 kg/hr
Solvent	= 25848.7 – 25090	<u>= 758.7 kg/hr</u>
		2240.1 kg/hr

- Try 3: Organic phase = 26,500 kg/hr

$$\begin{array}{lll} \text{Solvent} & = 0.965(26500) & = 25,572.5 \text{ kg/hr} \\ \text{H}_2\text{O} & = 0.035(26500) & = 927.5 \text{ kg/hr} \end{array}$$

<u>Water phase</u>		<u>q_o (%)</u>
H ₂ O	= 2391.4 - 927.5	= 1463.9 kg/hr
Solvent	= 25848.7 - 25572.5	<u>= 276.2 kg/hr</u>
		1740.1 kg/hr

- Try 4: Organic phase = 26,800 kg/hr

$$\begin{array}{lll} \text{Solvent} & = 0.965(26800) & = 25,862.8 \text{ kg/hr} \\ \text{H}_2\text{O} & = 0.035(26800) & = 938 \text{ kg/hr} \end{array}$$

<u>Water phase</u>		<u>q_o (%)</u>
H ₂ O	= 2391.4 - 938	= 1453.4 kg/hr
Solvent	= 25848.7 - 25862.8	<u>= 13.3 kg/hr</u>
		1466.7 kg/hr

- Interpolate: Organic phase = 26,508 kg/hr

$$\begin{array}{lll} \text{Solvent} & = 0.965(26508) & = 25,580.22 \text{ kg/hr} \\ \text{H}_2\text{O} & = 0.035(26508) & = 927.78 \text{ kg/hr} \end{array}$$

<u>Water phase</u>		<u>q_o (%)</u>
H ₂ O	= 2391.4 - 927.78	= 1463.62 kg/hr
Solvent	= 25848.7 - 25580.2	<u>= 268.48 kg/hr</u>
		1732.1 kg/hr

New Solvent composition, S

$$\text{Solvent} = 25580 + 268.48 + 3465.64 = 29314.34 \text{ kg/hr}$$

$$\text{H}_2\text{O} = \frac{927.78 \text{ kg/hr}}{30242.12 \text{ kg/hr}}$$

The new solvent composition is plotted and the new FS line is constructed

$$X_M = \frac{0.2424(29314.34)}{30242.12 + 29314.34} = 0.119$$

The location of the line E₁ R_{Np} must be readjusted.

Try X_{E1} = 20.25%

Composition of Raffinate

Acid	0.7%
Solvent	16%
H ₂ O	83.3%

Composition of Extract

Acid	20.25%
Solvent	73%
H ₂ O	6.75%

$$\text{Flowrate of } E_1 = \frac{6928.15}{0.2025} = 34213.09 \text{ kg/hr}$$

$$\text{Flowrate of } R_{Np} = \frac{177.65}{0.007} = 25342.52 \text{ kg/hr}$$

$$\underline{59,591.66 \text{ kg/hr}}$$

$$R+S = 30242.12 + 29314.34 = 59556.46 \text{ kg/hr (close enough)}$$

In Extract:

$$\text{Acid} = 34,213.09 (0.2025) = 6928.15 \text{ kg/hr}$$

$$\text{Solvent} = 34,213.09 (0.73) = 24,975.55 \text{ kg/hr}$$

$$H_2O = 34,213.09 (0.0675) = 2,309.38 \text{ kg/hr}$$

In Raffinate:

$$\text{Acid} = 177.65 \text{ kg/hr}$$

$$\text{Solvent} = 4338.79 \text{ kg/hr}$$

$$H_2O = 20826.08 \text{ kg/hr}$$

The composition of R_{Np} was adjusted so that the material balance work.

E_1 is sent to the extract stripper where the overhead consisting of solvent and water is separated in a decanter.

Composition of organic phase : 96.5 % solvent , 3.5% water

Composition of water phase : 84.52% water , 15.5% solvent

$$E_1 = 354213.08 \text{ kg/hr}$$

Trial and error

- Try 1: Organic phase = 25,500 kg/hr

Solvent	= 0.965(25500)	= 24,607.5 kg/hr
H_2O	= 0.035(25500)	= 892.5 kg/hr

<u>Water phase</u>	(%)	
H_2O	= 2309.38 - 892.5 = 1416.88 kg/hr	79.4
Solvent	= 24975.55 - 24607.5 = 368.05 kg/hr	20.6
	<u>1784.93 kg/hr</u>	

- Try 2: Organic phase = 25,700 kg/hr

Solvent	= 0.965(25700)	= 24,800.5 kg/hr
H_2O	= 0.035(25700)	= 899.5 kg/hr

<u>Water phase</u>	(%)	
H_2O	= 2309.38 - 899.5 = 1409.88 kg/hr	89
Solvent	= 24975.55 - 24800.5 = 175.05 kg/hr	11
	<u>1584.93 kg/hr</u>	

- Interpolate: Organic phase = 26,606.25 kg/hr

Solvent	= 0.965(26606.25)	= 24710.03 kg/hr
H_2O	= 0.035(26606.5)	= 896.22 kg/hr

<u>Water phase</u>			(%)
H ₂ O	=2309.38 – 896.22	= 1413.16 kg/hr	15.9
Solvent	= 24975.55 – 24710.03	<u>= 265.52 kg/hr</u>	84.1
		1678.68 kg/hr	

- Interpolate: Organic phase = 25612.16 kg/hr

Solvent	=0.965(25612.16)	=24715.73 kg/hr
H ₂ O	= 0.035(25612.16)	= 896.43 kg/hr

<u>Water phase</u>		(%)
H ₂ O	=2309.38 – 896.43	= 1412.95 kg/hr
Solvent	=24975.55 – 24715.73	<u>= 259.82 kg/hr</u>
		1466.7 kg/hr

New Solvent composition, S

$$\text{Solvent} = 24715.73 + 4338.79 + 259.82 = 29314.34 \text{ kg/hr}$$

$$\begin{array}{l} \text{H}_2\text{O} \\ \quad = \\ \quad \quad \quad \underline{\quad 896.43 \text{ kg/hr}} \\ \quad \quad \quad 30210.77 \text{ kg/hr} \end{array}$$

The solvent composition for this iteration is sufficiently close to the composition of the last iteration. The difference couldn't be seen on the ternary diagram.

The material balance will be solved with this new solvent composition:

<u>Composition of Raffinate</u>		<u>Composition of Extract</u>	
Acid	0.7%	Acid	20.25%
Solvent	16%	Solvent	73%
H ₂ O	83.3%	H ₂ O	6.75%

$$R+S = 30210.77 + 29314.34 = 59525.11 \text{ kg/hr}$$

<u>Extract</u>	34213.09 kg/hr	<u>Raffinate</u>	25312.02 kg/hr
Acid	6928.15 kg/hr	Acid	177.65 kg/hr
Solvent	24975.55 kg/hr	Solvent	4338.79 kg/hr
H ₂ O	2309.38 kg/hr	H ₂ O	20795.58 kg/hr

The composition of the 2 phases in the decanter will not change because E₁ has not changed.

The number of stages is stepped off on the ternary diagram shown below. Required number of stages = 3.

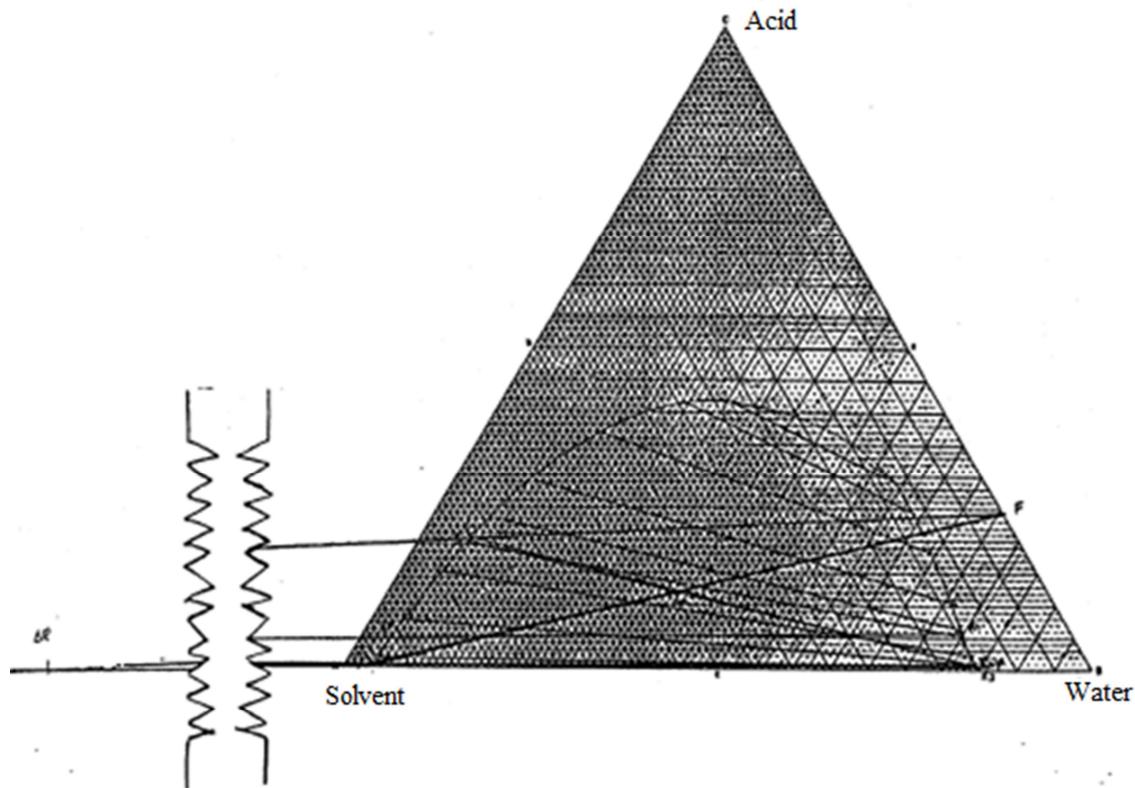
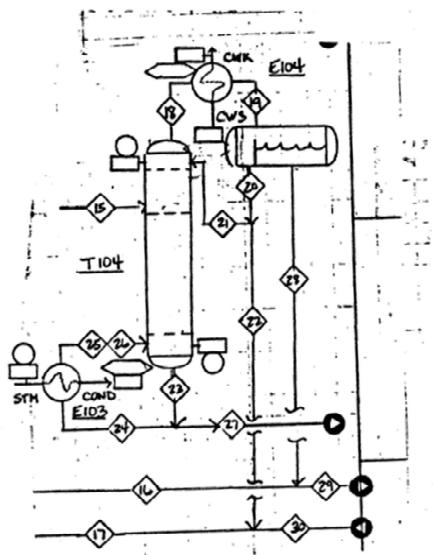


Figure 7

10.6 ECONOMIC COMPARISON OF 2 CASES

10.6.1 Case 1: Solvent A



The reboiler and condenser duties as well as the column diameter must be determined in order to economically compare the 2 cases

Figure 8

10.6.1.1 Extract Stripper T-104:

Overhead condenser duty: We are going to assume the temperature of the condensing fluid is constant in E-04. There would normally be a small temperature difference between the dewpoint and the bubble point but we don't have vapor pressure data on the solvents so that we could determine the temperature range from the dew point to the bubble point.

The reflux ratio (org. ref/org. dist)= 0.8 = stream 21/22

Composition of streams 28 and 22 is calculated without taking the organic reflux into account. This will probably not affect the number of stages but it will definitely affect the material balance. So, we recalculated the material balance.

Stream 15 consists of

Solvent = 28,342.12

Water = 719.8

Acid = 6928.15

Determining the composition of streams leaving Decanter:

We once again assume that all of the water and solvent goes overhead and all the acid to the bottom.

Iteration 1:

Estimate flowrate and composition of stream 22

Stream 22:

Solvent = 28331.22 kg/hr

H₂O = 56.76 kg/hr

= 28388 kg/hr

Stream 20:

Solvent = 1.8(28331.2) = 50996.2 kg/hr

H₂O = 1.8(56.76) = 02.17 kg/hr

Stream 21:

Solvent = 50996.22-28331.22 = 22,664.98 kg/hr

H₂O = 02.17 - 56.76 = 45.41 kg/hr

22,710.39 kg/hr

Stream 18: solvent and water from stream 15+stream 21

Solvent = 22,664.98+28342.12 = 51,007.1 kg/hr

H₂O = 45.41 + 719.8 = 765.2 kg/hr

= 517772.3 kg/hr

Stream 18 = Stream 19

Stream 19 is fed to a decanter where it separates into 2 phases of the following composition.

Organic phase water 0.2% solvent 99.8%

Water phase water 98.4% solvent 1.6%

Equation to find the amounts of each phase

$$\% \text{ solvent in water phase} = \frac{51,007.1 - 0.998(\text{total organic phase})}{51,007.1 - 0.998(\text{TOP}) + 765.2 - 0.002(\text{TOP})} \times 100 = 1.6$$

Total organic phase = 51,098.5 kg/hr

Solvent = 0.998(51098.54) = 50,996.3 kg/hr

H₂O = 0.002(51098.54) = 102.18 kg/hr

Water phase

$$\text{Solvent} = 51007.1 - 50,996.3 = 10.8 \text{ kg/hr}$$

$$\text{H}_2\text{O} = 765.21 - 102.18 = 663 \text{ kg/hr}$$

Stream 19:

$$\text{Solvent} = \frac{0.8}{1.8} (50996.8) + 28342.12 = 51007.1 \text{ kg/hr}$$

$$\text{H}_2\text{O} = \frac{0.8}{1.8} (102.18) + 719.8 = 765.2 \text{ kg/hr}$$

$$51,772.3 \text{ kg/hr}$$

This did not change significantly from the original stream 19 composition.

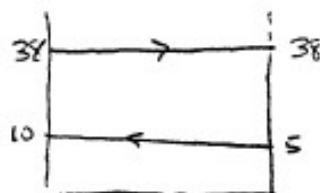
Overhead Condenser Calculations E-104:

$$\text{Condenser duty, } Q_{\text{cond}} = \sum m H_v$$

$$\Delta H_v: \text{H}_2\text{O} @ 38^\circ\text{C} = 2411.7 \frac{\text{KJ}}{\text{Kg}} \times \frac{0.33901 \text{ Kcal}}{\text{KJ}} = 576.64 \frac{\text{Kcal}}{\text{Kg}}$$

$$Q_{\text{cond}} = \frac{51007.1 \text{ Kg}}{\text{hr}} \times 122 \frac{\text{Kcal}}{\text{Kg}} + 765.64 \frac{\text{Kg}}{\text{hr}} \times 576.64 \frac{\text{Kcal}}{\text{Kg}} = 6,664,111.1 \frac{\text{Kcal}}{\text{hr}}$$

$$Q_{\text{cond}} = UA\Delta T_{\text{LMTD}}$$



$$LMTD = \frac{33-28}{\ln(\frac{33}{28})} = 30.43^\circ\text{C}$$

$$A = \frac{Q}{ULMTD} = \frac{6,664,111.1}{(700 \frac{\text{Kcal}}{\text{hr.m}^2.\text{°C}})(30.43)} = 312.85 \text{ m}^2 \approx 325 \text{ m}^2$$

Refrigerated water rate

$$6,664,111.1 \frac{\text{Kcal}}{\text{hr}} \times \frac{\$8.3}{10^6 \text{ KJ}} \times \frac{\text{KJ}}{0.239006 \text{ Kcal}} \times \frac{24 \text{ hr}}{\text{day}} \times \frac{365 \text{ days}}{\text{yr}} \times 0.88 = \frac{\$1,783,984}{\text{yr}}$$

Exchanger cost : $A = 325 \text{ m}^2$

Cost = \$ 97,546.4 $\times 5 = \$ 487,732$

Reboiler Duty, Q_{reb} E103

$$H_{\text{stream15}} + Q_{\text{reboiler}} = Q_{\text{cond}} + H_{\text{stream22}} + H_{\text{stream28}} + H_{\text{stream27}}$$

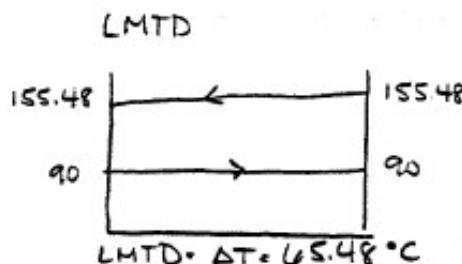
There is no heat capacity data for the solvent so we will have to estimate the reboiler duty.

Assume $Q_{\text{cond}} \cong Q_{\text{reb}}$.

$$\text{So, } Q_{\text{reb}} \cong Q_{\text{cond}} = 6,664,111.1 \frac{\text{Kcal}}{\text{hr}}$$

$$Q = UA\Delta T_{\text{LMTD}}$$

$$U = 1000 \frac{\text{Kcal}}{\text{hr.m}^2.\text{°C}}$$



Right now we are assuming that the acids do not have a temperature range for boiling this will simplify calculations:

$$A = \frac{Q}{U \cdot LMTD} = \frac{6,664,111.1}{1000(65.48)} = 101.7 \text{ m}^2$$

Steam rate:

$$Q = m(\Delta H_v) = m \times 2095.9 \frac{\text{KJ}}{\text{Kg}} \times 0.23901 \frac{\text{Kcal}}{\text{KJ}} = 6,664,111.1 \frac{\text{Kcal}}{\text{hr}}$$

$$m = 13,303.18 \frac{\text{Kg}}{\text{hr}}$$

Steam cost:

$$13,303 \frac{\text{Kg}}{\text{hr}} \times \frac{\$20}{1000\text{Kg}} \times \frac{24\text{hr}}{\text{day}} \times 0.88 = \frac{\$2,051,032}{\text{yr}}$$

Cost of heat exchanger = \$39,068 × 5 = \$195,338

Column Diameter:

$$F_{LV} = \frac{L}{G} \left(\frac{\rho_g}{\rho_L} \right)^{0.5}$$

$$L = 22,710.4 \text{ Kg/hr}$$

$$G = 51,772.3 \text{ Kg/hr}$$

$$\rho g = \frac{P}{RT} \quad \rightarrow \quad P \text{ is unknown}$$

$$\rho_L = \frac{226649}{22710.4} (0.8) \left(995 \frac{\text{Kg}}{\text{m}^3} \right) + \frac{43.41}{22710.4} \left(995 \frac{\text{Kg}}{\text{m}^3} \right) = 796.4 \frac{\text{Kg}}{\text{m}^3}$$

Do a bubble point to find the pressure at the bottom of the tower $K = \frac{P_{sat}}{P} = \frac{y}{x}$

Table 24

		x	P _{sat} at 90 °C	K(P=25 kPa)	y
HAC	6.65 kgmole/hr	6.8 %	38.99 kPa	1.56	0.106
AA	90.61 kgmole/hr	93.2%	17.84 kPa	0.714	0.665
	97.26 Kgmole/hr				$\Sigma y = 0.771$

Table 25

	K(P=20 kPa)	y	K(P=18 kPa)	y
HAC	1.95	0.132	2.1	0.149
AA	0.892	0.831	0.99	0.922
		$\Sigma y = 0.963$		$\Sigma y = 1.07$

Interpolate
 $P=19.3 \text{ kPa}$

Number of trays = $\frac{19.3}{0.7} = 27.5$ or 28 trays

$\Delta P/\text{tray} = 3.5 \text{ mmHg}$

$$19.3 \text{ kPa} \times \frac{760 \text{ mmHg}}{101.3 \text{ kPa}} = 144.8 \text{ mmHg}$$

$144.8 - 28(3.5) = 46.8 \text{ mmHg} = P \text{ at the top of the column}$

The column will be sized using the conditions at the top because the pressure is the lowest

$$\rho_g = \frac{46.8 \text{ mmHg}}{(273.15 + \text{ }) \text{ K}} \times \frac{\text{K.mol}}{62.36 \text{ L.mmHg}} = 0.0024 \frac{\text{mol}}{\text{L}}$$

$$\text{M.W.} = \frac{51007.1}{51772.3} \left(102 \frac{\text{g}}{\text{mol}} \right) + \left[1 - \left(\frac{51007.1}{51772.3} \right) \right] \left[18 \frac{\text{g}}{\text{mol}} \right] = 100.8 \frac{\text{g}}{\text{mol}}$$

$$\rho_g = 0.0024 \frac{\text{mol}}{\text{L}} \times 100.8 \frac{\text{g}}{\text{mol}} \times \frac{\text{kg}}{1000 \text{ g}} \times \frac{1000 \text{ L}}{\text{m}^3} = 0.2418 \frac{\text{Kg}}{\text{m}^3}$$

$$F_{LV} = \frac{22,710.4}{51,772.3} \left(\frac{0.2418}{796.4} \right)^{0.5} = 0.0078$$

$$C_{SB} = 0.39 \frac{\text{ft}}{\text{s}} \times 0.3048 = 0.189 \frac{\text{m}}{\text{s}}$$

From figure 18.10 Perrys , 6th Edition

$$C = UN \left(\frac{\rho_v}{\rho_L - \rho_v} \right)^{0.5} \quad C = C_{se} \cdot Ts \cdot Fs \cdot F_{Hn}$$

$$C = 0.1189 \left(\frac{25}{20} \right)^{0.2} \cdot 1.1 = 0.1243 \text{ m/s}$$

$$C = UN \left(\frac{\rho_v}{\rho_L - \rho_v} \right)^{0.5} = 0.1243 \text{ m/s} = UN \left(\frac{0.2418}{796.4 - 241.8} \right)^{0.5}$$

$$UN = 7.13 \text{ m/s}$$

$$D_T = \left[\frac{4V}{0.8U} \right]$$

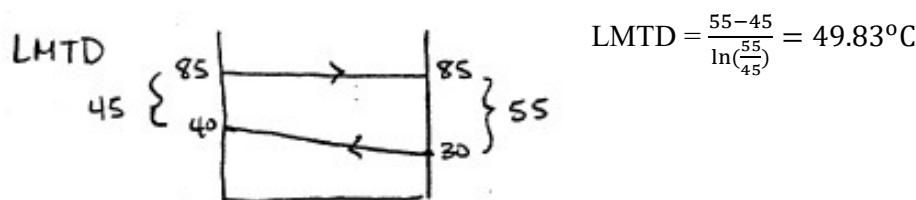
Cost of tower = \$ 788,458 × 5 = \$3,942,290

$$D = 3.8$$

10.6.1.2 Raffinate Stripper

Condenser:

$$Q_{cond} = 0.756 \times 10^6 \text{ Kcal/hr} \quad U = 700 \text{ Kcal/hr.m}^2.\text{°C}$$



$$A = \frac{Q}{U \cdot LMTD} = \frac{0.756 \times 10^6 \text{ Kcal/hr}}{700 \frac{\text{Kcal}}{\text{hr}} (49.84^\circ\text{C})} = 21.67 \text{ m}^2 \text{ around to } 25 \text{ m}^2$$

Cooling water rate

$$Q = mC_p \Delta T = m \left(\frac{\text{Kcal}}{\text{Kg} \cdot ^\circ\text{C}} \right) (10^\circ\text{C}) = 0.756 \times 10^6 \frac{\text{Kcal}}{\text{hr}}$$

$$m = 75,600 \frac{\text{kg}}{\text{hr}}$$

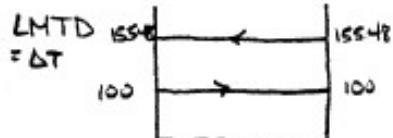
Cost of condenser = \$16484 × 5 = \$82,420

Cost of cooling water:

$$75,600 \frac{\text{kg}}{\text{hr}} \times 24 \frac{\text{hr}}{\text{day}} \times 365 \frac{\text{days}}{\text{yr}} \times 0.88 \times 0.001005 \frac{\text{m}^3}{\text{kg}} \times \frac{\$0.02}{\text{m}^3} = \frac{\$11,714}{\text{yr}}$$

Reboiler:

$$Q_{\text{reb}} = 1.638 \times 10^6 \text{ Kcal/hr} \quad U = 1000 \frac{\text{Kcal}}{\text{hr.m}^2 \text{ }^\circ\text{C}}$$



$F_T = 1$ for isothermal conditions

$$A = \frac{Q}{U\Delta T} = \frac{1.638 \times 10^6 \frac{\text{Kcal}}{\text{hr}}}{\left(\frac{1000 \text{ Kcal}}{\text{hr.m}^2 \text{ }^\circ\text{C}}\right)(55.48 \text{ }^\circ\text{C})} = 29.52 \text{ m}^2 \text{ round } 30 \text{ m}^2$$

Steam rate:

$$Q = m(\Delta H_r) = 1.638 \times 10^6 \frac{\text{Kcal}}{\text{hr}} = m(2095.9 \frac{\text{KJ}}{\text{Kg}} \times \frac{0.23901 \text{ Kcal}}{\text{KJ}})$$

$$m = 3,269.85 \frac{\text{Kg}}{\text{hr}}$$

Seam cost:

$$3269.85 \frac{\text{Kg}}{\text{hr}} \times \frac{\$20}{1000 \text{ Kg}} \times \frac{24 \text{ hr}}{\text{day}} \times 365 \frac{\text{days}}{\text{yr}} \times 0.88 = \frac{\$504,131.7}{\text{hr}}$$

Cost of reboiler = $\$18223 \times 5 = \$91,117.9$

Column Cost:

$$= 2900(D + N) + (3.3D)^{1.9}N.80 + (3.3D)^{0.5}1380 + 5700N$$

D = Diameter, m

N = number of stages

D: 0.6 meters

N: 10 stages, this is unclear whether this is real or theoretical stages. I am going to assume they are real stages.

Column cost = $\$80,315 \times 5 = \$401,575$

Cost of the Extractor:

N = 11 theoretical stages

Cost of extractor = $\$120,000(11) = \$1320000 \times 5 = \$6,600,000$

Cost of Equipment:

Table 26

		Equip Cost	Utility	Utility
T-103	Extractor	6,600,000	-	-
T-104	Extract stripper	3,942,290	-	-
T-105	Raffinate Stripped	401,575	-	-
E-104	T-101 condenser	487,732	-1,783,984 \$/yr	Refrg H ₂ O
E-104	T-104 reboiler	95,338	2,051,032 \$/yr	Steam
E-106	T-105 condenser	82,420	11,714 \$/yr	CW
E-107	T-105 reboiler	91,118	115,530 \$/yr	Steam
		\$ 11,803,473	3,962,260 \$/yr	

10.6.1.3 Economic Evaluation:

Table 27

EOY	Cash flow	Dep	Taxable Income	Tax	After Tax Cash Flow
0	- 11,803,473	-	-	-	- 11,803,473
1	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
2	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
3	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
4	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
5	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
6	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
7	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
8	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
9	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
10	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609
11	- 3,962,260	-1,073,043	-5,035,303	-2,517,651	-1,444,609

$$PW = -11,803,473 - 1,444,609(5.2337) = -\$ 19,364,121$$

(discrete compounding)

10.6.2 Case 2: Solvent B

10.6.2.1 Extract Stripper:

Stream 15:

Acid: 6928.15 kg/hr

Solvent: 24,975.55 kg/hr

Water: 2309.38 kg/hr

34,213.08 kg/hr

Check whether the organic reflux affects the distribution in the decanter. This is an iterative process.

Iteration 1:

Estimate composition of S22:

Stream 22:

Solvent = 24,715.73 kg/hr

H₂O = 896.43 kg/hr

= 25,612.16 kg/hr

Stream 20: (95% reflux)

Solvent = 24,715.73(1.95) = 48,195.7 kg/hr

Water = 896.43(1.95) = 1,748 kg/hr

= 49,943.7 Kg/hr

Stream 21:

Solvent = 98,195.7 - 24,715.73 = 23,479.97 kg/hr

H₂O = 1748 - 896.43 = 851.57 kg/hr

= 24,331.54 kg/hr

Stream 19:

$$\text{Solvent} = 23,479.97 + 24975.55 = 48,455.52 \text{ kg/hr}$$

$$\text{H}_2\text{O} = 851.57 + 2309.38 = 3,160.95 \text{ kg/hr}$$

Organic phase 96.5% solvent , 3.5% H₂O
 Water phase 84.5% H₂O , 15.5% solvent

$$\text{Equation to determine composition of phases: } \frac{48455.52 - 0.965 \text{ TOP}}{51,616.47 - \text{TOP}} \times 100 = 15.5$$

$$\text{TOP} = 49,944.4 \frac{\text{kg}}{\text{hr}}$$

(total organic phase)

Organic phase = 49.9444 kg/hr

$$\text{Solvent} = 0.965(49944.4) = 48196.35 \text{ kg/hr}$$

$$\text{Water} = 0.0356(49944.4) = 1748.05 \text{ kg/hr}$$

Water phase: stream 28

$$\text{Solvent} = 48455.52 - 48196.35 = 259.2 \text{ kg/hr}$$

$$\text{Water} = 3160.95 - 1748.05 = 1412.9 \text{ kg/hr/kg/hr}$$

This did not change from the composition of stream 20 of the 1st iteration. So, I am going to use the previous material balance

Stream 19

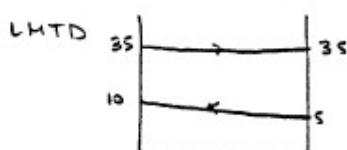
$$\text{Solvent} = 48455.52 \text{ kg/hr}$$

$$\text{H}_2\text{O} = 3160.96 \text{ kg/hr}$$

Condenser Duty

$$Q_{\text{cond}} = mH_{\text{solvent}} + mH_{\text{water}}$$

$$Q_{\text{cond}} = 48,455.52 \frac{\text{kg}}{\text{hr}} * \frac{136 \text{ kcal}}{\text{Kg}} + 3160.95 \frac{\text{kg}}{\text{hr}} * 576.64 \frac{\text{kcal}}{\text{hr}}$$



$$\text{LMTD} = \frac{30 - 25}{\ln(\frac{30}{25})} = 27.4^{\circ}\text{C}$$

$$U = 700 \frac{\text{kcal}}{\text{hr} \cdot \text{m}^2 \cdot {}^{\circ}\text{C}}$$

$$Q_{\text{cond}} = 8412681 \text{ Kcal/hr}$$

$$A = \frac{Q}{U \Delta T} = \frac{8412681 \text{ kcal/hr}}{\frac{27.4 \text{ kmol} \cdot 700}{\text{hr} \cdot \text{m}^2 \cdot {}^{\circ}\text{C}}} = 438.6 \text{ m}^2 \text{ round to } 450 \text{ m}^2$$

$$\text{Cost of heat exchanger} = \exp(8.596 + 0.153 \ln A + 0.06 (\ln(A))^2)$$

$$= 129326 * 5 = 646628$$

Refrigerated water rate Q_{cond}=mc_p ΔT C

$$8412681 \frac{\text{kcal}}{\text{hr}} = m \left(1 \frac{\text{kcal}}{\text{kg}^{\circ}\text{C}} \right) (5^{\circ}\text{C}) \rightarrow m = 1682536 \text{ kg/hr}$$

Refrigerated water cost

$$8412681 \frac{\text{kcal}}{\text{hr}} * \frac{\text{KJ}}{0.2390\text{kcal}} * \frac{\$8.3}{106\text{KJ}} * \frac{24\text{hr}}{\text{day}} * \frac{365\text{day}}{\text{yr}} * 0.88 = \frac{\$2252077}{\text{yr}}$$

Extract Stripper reboiler

Assume $Q_{\text{reb}}=Q_{\text{cond}}=8412681 \text{ kcal/hr}$

LMTD= $\Delta T=15548.90=65.48$

$$U= 700 \frac{\text{kcal}}{\text{hr.m}^2.\text{^oC}}$$

$$A = \frac{Q}{U \Delta T} = \frac{8412681 \text{ kcal/hr}}{700 \frac{\text{kcal}}{\text{hr.m}^2.\text{^oC}} \times 65.48} = 128.5 \text{ m}^2 \text{ round to } 125 \text{ m}^2$$

Cost = \$ 95866*5= \$ 229332

stream rate $Q_{\text{reb}}=m_{\text{steam}} \Delta H_v 155.48$

$$8412681 \frac{\text{Kcal}}{\text{hr}} = m \left(\frac{2095.9 \text{ KJ}}{\text{KG}} * \frac{0.23901 \text{ kCal}}{\text{KJ}} \right) \rightarrow m = 16,794 \frac{\text{Kg}}{\text{hr}}$$

Cost of steam

16794 kg/hr*20.00/1000*24hr/day*365day/yr*0.88=2589,194/yr

Determination of column size

Design for condition at the top of the column

$$F_{\text{Lv}}=L/G(p_g/p_l)^{1/2} L=24331.54 \text{ kg/hr} G=51616.47 \text{ kg/hr}$$

Pressure at the top is 46.8mmHg, T= 35°C

$$\rho_g = \frac{P}{RT} = \frac{46.8 \text{ mm Hg}}{62.36 \frac{\text{mm hg}}{\text{mol K}} \times (273.15+35) \text{ K}} = 0.0025 \frac{\text{mol}}{\text{l}} = 0.938$$

$$\text{Mw: mol fraction solvent is stream 19} = \frac{48455.52*150}{48455.52*150+3161*18} = 0.992$$

$$\text{Mw}=0.992(150\text{g/mol})+(1-0.992)(18\text{g/mol})=149 \text{ g/mol}$$

$$\rho_g=0.0025\text{mol/l}*149\text{g/mol}*\text{kg}/1000\text{g}*1000\text{l/m}^3=0.3724 \text{ kg/m}^3$$

$$\rho_L=\text{wt fraction solvent [spec gravity]} P_{\text{H}_2\text{O}} + \text{wt fraction } P_{\text{H}_2\text{O}}$$

$$\rho_L=0.992[0.85](995)+(1-0.992)(995)=846.9\text{kg/m}^3$$

$$F_{\text{LV}}=\frac{24331.5\text{kg/hr}}{51616.47\text{kg/hr}} * \left(\frac{0.3724\text{kg}}{\frac{\text{m}^3}{846.9\text{kg}}} \right)^{1/2} = 0.0099$$

C_{SB}=0.38 Ft/s or 01158m/s (fig 18.10, perry's)

$$C=C_{\text{SB}} \text{ FST} = 0.1158*(25/20)^{0.2} = 0.1211 \text{ m/s}$$

$$C=UN\left(\frac{\rho_v}{\rho_L-\rho_v}\right)^{1/2}, \text{ Solve for UN}$$

$$UN = \frac{0.1211 \text{ m/s}}{\left(\frac{0.3720 \frac{\text{kg}}{\text{m}^3}}{8469 - 0.3724 \frac{\text{kg}}{\text{m}^3}} \right)^{1/2}} = 5.77 \text{ m/s}$$

Design at 807 = floods

$$DT = \left(\frac{4G}{(0.8UN\left(1-\frac{Ad}{A}\right)\rho_v)} \right)^{1/2} \text{ assume } A_{cl}/A, \text{ area downcomer/total area} = 0.1$$

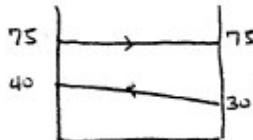
$$DT = \left(\frac{4 \times \frac{(51,616.5) \text{ kg}}{\text{hr}} \times \text{hr}}{(0.8 \times \frac{5.87 \text{ m}}{\text{s}^2} \times 0.9 \times 0.3724 \text{ kg/m}^3)} \right)^{1/2} = 3.4 \text{ meters or } 11.2 \text{ ft round to } 11 \text{ ft or } 3.35 \text{ meters}$$

Cost of tower = $6840360 \times 5 = \$3420183$

10.6.2.2 Raffinate stripper:

Condenser Duty, $Q_{cond} = 1.008 \times 10^6 \text{ Kcal/hr}$

$$LMTD = \frac{45 - 35}{\ln(\frac{45}{35})} = 39.8 \text{ }^\circ\text{C}$$



$U = 700 \text{ Kcal/m}^2 \text{ hr C}$

$$A = \frac{Q}{U \Delta T} = \frac{1.008 \times 10^6 \text{ Kcal/hr}}{700 (39.8)} = 36.2 \text{ m}^2 \text{ round to } 35 \text{ m}^2$$

Cost of heat exchanger = $\$19,898.5 \times 5 = \$99,492.4$

Cost of cooling water $Q_{cond} = mC_p \Delta T$

$1.008 \times 10^6 = m(1.0 \text{ kcal/kg C}) 10 \text{ C}$, $m = 100800 \text{ kg/hr}$

Cost of CW

$100800 \text{ kg/hr} * \$0.02/1000 \text{l} * 0.001005 \text{ m}^2/\text{kg} * 1000 \text{l/m}^3 * 24 \text{ hr/day} * 365 \text{ day/yr} * 0.88 =$

Cost of cooling water = $\$15618.6/\text{yr}$

10.6.2.3 Raffinate stripper reboiler

$Q_{reboiler} = 2.268 \times 10^6 \text{ kcal/hr}$

$$LMTD = \Delta T = 155.48 - 100 = 55.48 \text{ C}$$

$U = 1000 \text{ Kcal/m}^2 \text{ Chr}$

$$A = \frac{2.268 \times 10^6 \text{ Kcal/hr}}{1000 \text{ Kcal/m}^2 \text{ hr C} (55.48 \text{ C})} = 40.9 \text{ m}^2 \text{ round to } 40 \text{ m}^2$$

Cost = $\$21,523 \times 5 = \107615

Steam rate constant

$Q_{reb} * m(\Delta H)$ stream

$$2.268 \times 10^6 \text{ Kcal/hr} = m(2095.9 \text{ KJ/KG} * 0.23901 \text{ Kcal/KJ})$$

M= 4527.5kg/hr

Cost

$$4527.5 \text{ kg/hr} * \$2000/1000 * 24 \text{ hr/day} * 365 \text{ day/yr} * 0.88 = \$698,028/\text{yr}$$

10.6.2.4 Cost of tower

of trays 10, Diameter = 1.2m

$$\text{Cost of tower} = \$109654 * 5 = \$548272$$

10.6.2.5 Cost of Extractor # of stages 3

$$\text{Cost} = \$120000 * 3 = \$360000 * 5 = \$1800000$$

Equipment list

Table 28

Equipment	cost	Utility cost	utility
T103-Extractor	1800000	-	-
T104-Extract Stripper	3420813	-	-
T105-Raffinate stripper	548272	-	-
E104-T104 condenser	646628	22552077/yr	Refrig. water
E-103-T104 reboiler	229332	2589194/yr	stm
E-106-T105 cond	994942	15619/yr	Cooling water
E-107-T105 reboiler	107615	698028/yr	steam
	6851522	5554918/yr	

10.6.2.6 Economic Evaluation

Table 29

EOX	CF	Depreciation	Taxable Income	Tax	ATCF
1	-6851522	-	-	-	-6851522
2	-5554918	-622865	-6177784	-3088892	-2466026
3	-5554918	-622865	-6177784	-3088892	-2466026
4	-5554918	-622865	-6177784	-3088892	-2466026
5	-5554918	-622865	-6177784	-3088892	-2466026
6	-5554918	-622865	-6177784	-3088892	-2466026
7	-5554918	-622865	-6177784	-3088892	-2466026
8	-5554918	-622865	-6177784	-3088892	-2466026
9	-5554918	-622865	-6177784	-3088892	-2466026
10	-5554918	-622865	-6177784	-3088892	-2466026
11	-5554918	-622865	-6177784	-3088892	-2466026

$$PW = -6851522 - 2466026 * (5.2337) = -19757963$$

Therefore, go with solvent A because the present worth of solvent is more negative than the present worth of solvent A.

The material balance and will be filled in for solvent A.

10.7 APPENDIX F: ACRYLIC ACID RECOVERY

Description:

A configuration for the distillation towers was decided upon. The towers were sized using a McCabe Thiele binary distillation program. The optimum case was picked by investigating the reflux ratio versus the number of trays. The optimum design was picked by comparing several cases, recognizing the trade off between capital costs and operating costs.

10.7.1 Acrylic Acid Recovery:

The tower calculations in this section were done with the help of a McCabe-Thiele binary distillation program. This program was used for our tower calculations last semester and has been checked for accuracy. However, I will do a graphical solution. And check the answers that I get with the computers answers. The computer program uses the equilibrium data in the form of α (relative volatility) vs X (mole fraction of HAC in the liquid phase). α is computed by the following equation:

$$\alpha = \frac{y(1-x)}{x(1-y)} = \frac{K_{HAC}}{K_{AA}} \cdot \frac{y_{HAC}/x_{HAC}}{y_{AA}/x_{AA}}$$

The α was computed for every point in Table 4.

POLFIT (MATLAB)

Minimum instruction= 0, All= OTHER #

Fits least square polynomial bivariate data by orthogonal polynomials.

Limits: 11th degree fit and max of 100 data points. But fits higher than degree 5 may give poor results. Polfit allows the user to specify the lowest degree polynomial to be fit and then fits polynomial in order of ascending degree. At each stage, the index of determination is printed, and then the user has the choice of:

Going to next higher fit, or getting one of two summaries, or stopping the program.

To use type:

10 data N, D (where N= number of data points to be read and D= initial lowest degree to be fit)

100 data x(1), y(1), x(2), y(2), , x(n), y(n)

Ready

- 10 Data 11, 2
- 100 Data .029, 2.676, 0.62, 2.566, 0.106, 2.434, .208, 2.208, 0.319, 2.059
- 101 Data .415, 1.979, 0.534, 1.898, 0.604, 1.856, 0.682, 1.819, 0.866, 1.756
- 102 Data 0.974, 1.753

- BRN

Polfit (MATLAB)

Least= Squares polynomial

Number of points= 11

Mean value of X= 0.4362727

Mean value of Y= 2.091273

Std error of Y= 0.3325182

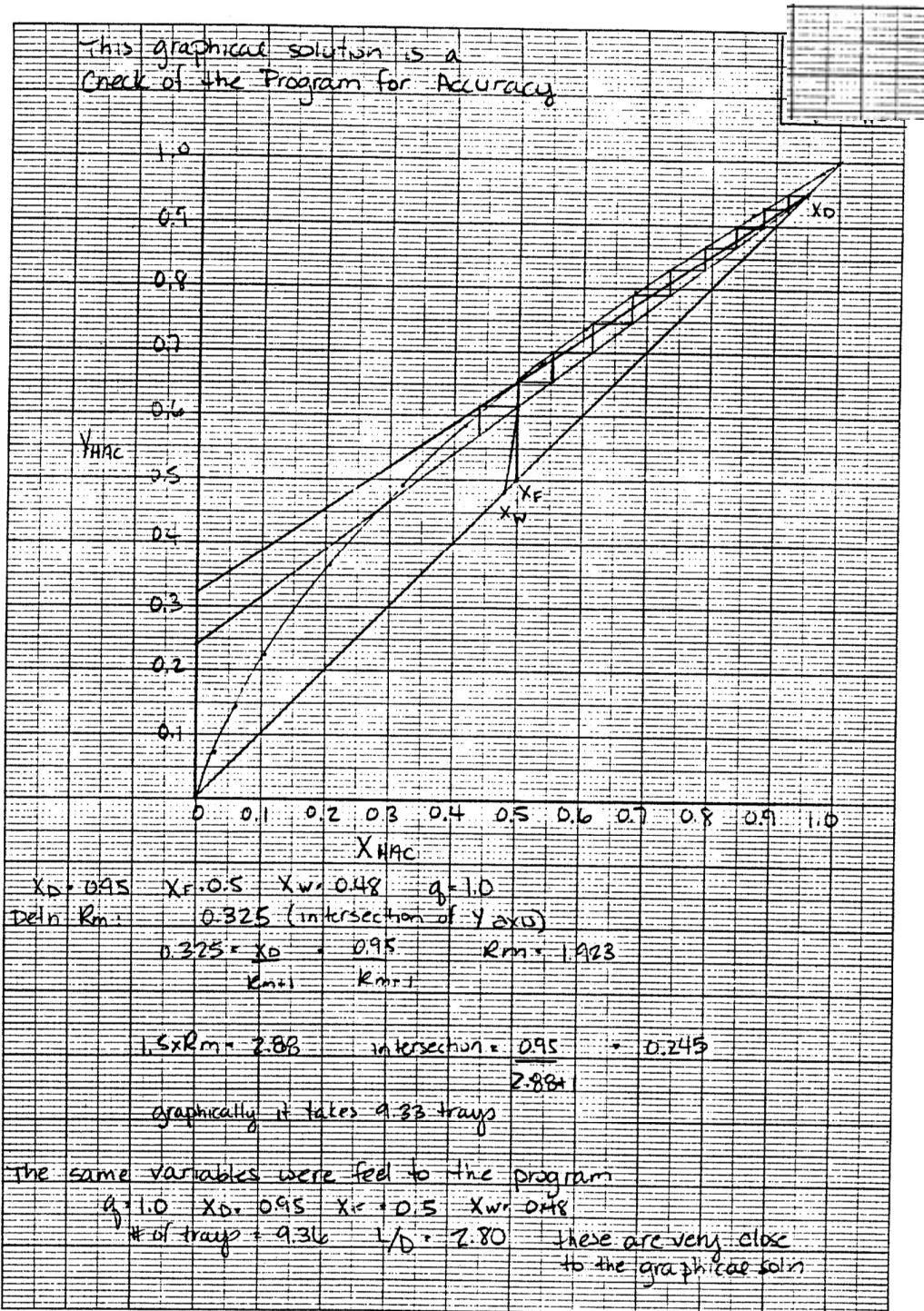


Figure 9

Acrylic acid recovery T106:

Stm 27 feed T= 90°C mol %

HAC 6.65 kmol/hr 6.84

AA 90.61 kmol/hr 93.16
 97.26

Specifications

Bottoms 99.9% AA - 0.170 HAC (assume mol %)

Overhead 95.0% HAC - 5.0% AA (assume mol %)

10.7.2 Material Balance:

Stream 27= stream 42 + stream 46

This is trial and error

Guess stream 46 = 90.67 kmol/hr

$$\text{AA} = 0.999(90.67) = 90.58 \text{ kmol/hr}$$

$$\text{HAC} = 0.001(90.67) = \frac{0.09 \text{ kmol/hr}}{90.67}$$

Guess stream 42 mol %

$$\text{AA} = 90.61 - 90.58 = 0.03 \text{ kmol/hr} \quad 0.45$$

$$\text{HAC} = 6.65 - 0.09 = \frac{6.56 \text{ kmol/hr}}{6.59 \text{ kmol/hr}} \quad 99.55$$

Guess stream 46 = 90.36 kmol/hr

$$\text{AA} = 0.999(90.356) = 90.27 \text{ kmol/hr}$$

$$\text{HAC} = 0.0001(90.35) = 0.09 \text{ kmol/hr}$$

Stream 42 mol %

$$\text{AA} = 90.61 - 90.27 = 0.34 \text{ kmol/hr} \quad 4.93$$

$$\text{HAC} = 6.65 - 0.09 = \frac{6.56 \text{ kmol/hr}}{6.90 \text{ kmol/hr}} \quad 95.07$$

Guess stream 46 = 90.35 kmol/hr

$$\text{AA} = 0.999(90.35) = 90.26 \text{ kmol/hr}$$

$$\text{HAC} = 0.0001(90.35) = 0.09 \text{ kmol/hr}$$

Stream 42 mol %

$$\text{AA} = 90.61 - 90.26 = 0.35 \text{ kmol/hr} \quad 5.07$$

$$HAC = 6.65 - 0.09 = \underline{6.56} \text{ Kmol/hr}$$

$$6.91$$

Interpolate

$$\underline{\text{Stream 46}} = 90.355 \text{ Kmol/hr}$$

$$AA = 0.999(90.355) = 90.265 \text{ Kmol/hr}$$

$$HAC = 0.001(90.355) = \underline{0.090} \text{ kmol/hr}$$

$$90.355 \text{ Kmol/hr}$$

$$\underline{\text{Stream 42}} \quad \text{mol \%}$$

$$AA = 90.61 - 90.625 = 0.3450 \quad 5.0$$

$$HAC = 6.65 - 0.09 = \frac{6.56}{6.905} \quad 95.0$$

Bottoms Pressure

Bubble point of strm 46 $y/x = P_{\text{sat}}/P_{\text{total}}$

X	P _{sat} (90°C) KPa	K(P _{total} = 20 KPa)	Y	K(P _{total} = 18)	Y	k(P _{total} = 17.9 kPa)	Y
HAC	0.001	38.99	1.95	0.00195	2.166	0.0022	2.178
AA	0.999	17.84	0.89	0.8891	0.9911	0.9901	0.997

P bottoms 17.9 KPa or 135.7 mm hg there will have to be more than one tower because the pressure drop across the trays will cause the pressure to be too low at the top of the column. The separator takes 42 real trays and a reflux ratio of 13.37(l/d)

The following configuration will be used

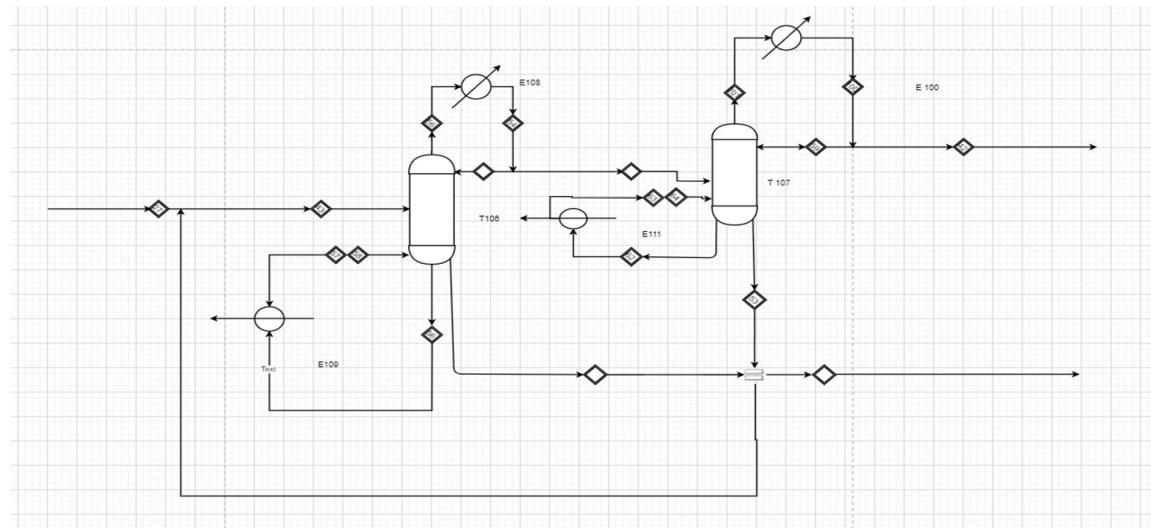


Figure 10

This is a trial and error procedure to solve the material balance. Before the material balance can be solved two specifications must be made concerning the composition of stream 43 and stream 52. To do this I suggest compositions of stream 43 and 52 and used them in m tower program which is listed on the previous pages. I made several guesses and compared these to decide which specification were the best, i.e less trays and smaller L/D ratio. For comparison purposes I used a multiplier of 1.5 times the minimum reflux and a q line of 1.0 for each case.

Guesses:

Table 30

Case	A	B	C	D	E	F	G	H	I	J
X feed HAC	0.3	0.2	0.1	0.07	0.4	0.3	0.3	0.2	0.1	0.2
X bottoms HAC	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001
Xc HAC	0.6	0.6	0.6	0.6	0.6	0.5	0.7	0.4	0.3	0.7
No. of Trays	14	15	16	16	16	16	15	16	16	16
L/D	1.125	2.31	6.8	7.29	7.29	0.25	2	0.405	1.13	3.26

Case E and F are the two best cases; a material balance will be done for both of these cases. In all of these cases the feed was on the top tray so q is probably less than 1.0. I will calculate q after the material balances are calculated. T feed is 90°C the temp at the top of the tower is < 90° C

Material Balance Case E

Trial and error try S43= 180kmol/hr

Stream 43

$$\text{HAC} \quad 0.4(180) = 72 \text{ kmol/hr}$$

$$\text{AA} \quad 0.6(180) = 108 \text{ kmol/hr}$$

$$\text{Stream 52 (Stream 52 =S43-S46)} \quad \% \quad$$

$$\text{HAC} \quad 72 - 0.09 = 71.91 \text{ kmol/hr} \quad 80\%$$

$$\text{AA} \quad 108 - 90.265 = \frac{17.74 \text{ kmol/hr}}{89.645 \text{ kmol/hr}} \quad 20\%$$

$$\text{M bal} \quad \frac{0.4 \text{ S43} - 0.09}{\text{S43} - (90.265 - 0.09)} = 0.6$$

$$\text{S43} = 270.61 \text{ kmol/hr}$$

S43

$$\text{HAC} = 0.4(270.61) = 108.25 \text{ kmol/hr}$$

$$\text{AA} = 0.6(270.61) = 162.36 \text{ kmol/hr}$$

S52 (S52= S43-S46)

$$\text{HAC} = 108.25 - 0.09 = 108.16 \text{ kmol/hr}$$

$$\text{AA} = 162.36 - 90.265 = \frac{72.099}{180.26} \text{ kmol/hr}$$

S53= Strm52- S42 mol%

$$\text{HAC} = 109.16 - 6.56 = 101.6 \text{ kmol/hr} \quad 58.6$$

$$\text{AA} = 72.099 - 0.345 = \frac{71.754}{173.35} \text{ kmol/hr} \quad 41.4$$

S43= strm27+strm53

$$\text{HAC} = 101.6 + 6.65 = 108.25 \text{ kmol/hr}$$

$$\text{AA} = 71.754 + 90.61 = 162.36 \text{ kmol/hr}$$

Material balance case F

$$\frac{0.43 * \text{S43} - 0.09}{\text{S43} - (90.265 + 0.09)} = 0.5$$

$$\text{S43} = 225.44 \text{ kmol/hr}$$

S43

$$\text{HAC} = 0.3(225.44) = 67.63 \text{ kmol/hr}$$

$$\text{AA} = 0.7(225.44) = \frac{157.81 \text{ kmol/hr}}{225.44 \text{ kmol/hr}}$$

S52 = S43 - S46

$$\text{HAC} = 67.63 - 0.09 = 67.54 \text{ kmol/hr}$$

$$\text{AA} = 157.81 - 90.265 = \frac{67.54 \text{ kmol/hr}}{135.08 \text{ kmol/hr}}$$

S53 = S52 - S42 mol %

$$\text{HAC} = 67.54 - 6.56 = 60.98 \text{ kmol/hr} \quad 47.6$$

$$\text{AA} = 67.54 - 0.345 = \frac{67.195 \text{ kmol/hr}}{128.175 \text{ kmol/hr}} \quad 52.4$$

For the rest of the calculations case F will be used because the flowrate of stream 52 in case F is less than the flowrate of stream 52 in case E. This will result in lower utility bills because less overhead will have to be condensed.

Q line

$$Q = \frac{\text{Hg} - \text{Hf}}{\text{Hg} - \text{Hl}} \quad \begin{aligned} \text{P bottoms} &= 135.7 \text{ mm Hg} \\ \text{P top} &= 135.7 \text{ mm Hg} - 3.5 \text{ mm Hg (16trays/.5)} = 23.7 \text{ mm Hg} \end{aligned}$$

Table 31

Y1 (comp. top tower)	K (T=30°C)	X1	K (T=38°C)	X1	K (T=45°C)	X1	K (T=47°C)	X1
HAC = 0.5	0.865	0.578	1.317	0.38	1.86	0.269	2.054	0.243
AA = 0.5	0.274	1.825	0.443	1.13	0.659	0.759	0.735	0.68
		2.4		1.51		1.028		0.923

T at the top of the tower is 46°C

T feed is 90°C

$$q_1 = \Delta H_{vap} (46) + \int_{46}^{90} \Sigma m C_p \Delta T$$

$$\Delta H_{vap} (46)$$

$$\underline{\Delta H_{vap} HAC} \quad (67.54 \text{ kmol/hr}) = H_g - H_L$$

$$M_{48} \int_{117.9}^{141.6} C_p \Delta T + \Delta H_{vap} (117.9) + m_{117.9} \int_{46}^{41.6} C_p \Delta T$$

$$43.7 \frac{kcal}{kg} \times 60.05 \frac{kg}{kgmol} \times 67.73 \frac{kgmol}{hr} + 57.33 \frac{kcal}{kgmol} \times 67.73 \frac{kgmol}{hr} + 38.1 \frac{kcal}{kg} \times \frac{60.06 kg}{kgmol} \times \frac{67.73 kgmol}{hr}$$

$$\Delta H_{vap} HAC @ 46^\circ C = 410,465 \text{ kcal/hr}$$

$$\underline{\Delta H_{vap} AA} = H_g - H_L$$

$$M_{46} \int_{117.9}^{141.6} C_p \Delta T + \Delta H_{vap} (141.6^\circ C) + m_{117.9} \int_{46}^{41.6} C_p \Delta T$$

$$157.81 \frac{kmol}{hr} (61.1 \frac{kcal}{kg} \times 72.06 \frac{kg}{kgmol} + 78.72 \frac{kcal}{kgmol} + 53.12 \frac{kcal}{kg} \times 72.06 \frac{kg}{kgmol})$$

$$\Delta H_{vap} AA @ 90^\circ C = 1,332,942 \text{ kcal/hr}$$

$$H_g - H_f = \Delta H_{vap} - \int_{46}^{90} m C_p \Delta T$$

HAC:

$$410,465 \frac{kcal}{hr} - 23.25 \frac{kcal}{kg} \times 60.05 \frac{kg}{kgmol} \times 67.63 \frac{kgmol}{hr} = 317,539 \frac{kcal}{hr}$$

AA:

$$1,332,942 \frac{kcal}{hr} - 23.25 \frac{kcal}{kg} \times 72.06 \frac{kg}{kgmol} \times 15781 \frac{kgmol}{hr} = 1068595 \frac{kcal}{hr}$$

$$Q = \frac{(317,539 + 1068595) \text{ kcal/hr}}{(410,465 + 1332942) \text{ kcal/hr}} = 0.795$$

T106

Now the number of trays vs reflux ratio will be investigated for case F, this time $q = 0.795$ will be used.

Table 32

Factor (Xemin)	1.2	1.5	1.8	2	2.5	3	4	5	1.35
L/D	0.468	0.585	0.7015	0.7795	0.9743	1.169	1.559	1.9487	0.526
# trays	17.59	14.68	13.91	12.73	11.63	10.94	10.1	9.55	16.05

T107S52

$$X_{\text{feed}} = 0.5 \quad T_{\text{feed}} = 46^{\circ}\text{C} \quad AA = 67.54 \text{ Kmol/hr}$$

$$X_{\text{waste}} = 0.476 \quad T_{\text{waste}} = 90^{\circ}\text{C} \quad HAC = 67.54 \text{ Kmol/hr}$$

$$X_{\text{dist}} = 0.95$$

For T107, $q > 1$ because the feed is on the bottom tray which is 90°C but the feed temperature is 46°C .

$$q = \frac{\Delta H_{\text{vap}}(90) + _{46} \int^{90} m C_p \Delta T}{\Delta H_{\text{vap}}(90)}$$

ΔH_{vap}

$$AA = M_{90} \int^{141.6} C_p \Delta T + \Delta H_{\text{vap}} + m_{141.6} \int^{90} C_p \Delta T$$

$$67.54(29.87 \frac{\text{Kcal}}{\text{Kg}} \times 7206 \frac{\text{Kg}}{\text{kmol}} + 7872 \frac{\text{Kcal}}{\text{kmol}} - 32.88 \frac{\text{Kcal}}{\text{kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}})$$

$$= 517018.25 \text{ Kcal/hr}$$

$$HAC = M_{90} \int^{117.9} C_p \Delta T + \Delta H_{\text{vap}} + m_{117.9} \int^{90} C_p \Delta T$$

$$67.54(15.22 \frac{\text{kcal}}{\text{kg}} \times 60.05 \frac{\text{Kg}}{\text{kmol}} + 5733 \frac{\text{kcal}}{\text{kmol}} - 16.21 \frac{\text{kcal}}{\text{kg}})$$

$$= 383187 \text{ kcal/hr}$$

$$\Delta H_{\text{vap}}(90^{\circ}\text{C}) + _{46} \int^{90} m C_p \Delta T$$

$$AA = 517018.25 + 28.2 \frac{\text{kcal}}{\text{Kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}} \times 67.54 = 654266 \text{ kcal/hr}$$

$$HAC = 383187 + 27.5 \frac{\text{kcal}}{\text{Kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}} \times 67.54 = 494750 \text{ kcal/hr}$$

$$Q = \frac{(654,266 + 494750) \text{ kcal/hr}}{(517018.25 + 383178) \text{ kcal/hr}} = 1.276$$

The following information is generated with the computer program

Table 33

Factor (Xrmin)	1.2	1.5	1.8	2	2.5	3	4	5
L/D	2.076	2.595	3.114	3.46	4.325	5.19	6.92	8.65

# trays	17.59	9.81	8.525	7.94	7.13	6.7	6.22	5.95
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10.7.3 T106 –Sizing

Several cases will be solved to find the optimum case of reflux ratio vs number of trays

Case1: $1.5 \times R_m$ (sized at top of tower conditions)

$$L/D = 0.585, \# \text{ of trays} = 14.67/.5 = 29.3 \text{ or } 30 \text{ trays}$$

$$P_{top} = 135,7 - 3.5(30) = 30.7 \text{ mm hg}$$

T-top (dewpoint calculation) $K_1 = Y_1/X_c$

Table 34

Y1	K1 (T=45°C)	X1	K1 (T=47°C)	X1	K1 (T=50°C)	X1	K1 (T=52°C)	X1
HAC = 0.5	1.44	0.347	1.585	0.316	1.828	0.274	2	0.249
AA = 0.5	0.508	0.984	0.567	0.881	0.667	0.749	0.742	0.674
		1.331		1.197		1.023		0.923

Interpolate T = 50.46°C

10.7.3.1.1 Tower Diameter

$$FLV = L/G(\rho_g/\rho_l)^{0.5}$$

$$L: L = 0.585 D \quad D = 67.54 \text{ kmol/hr} \quad AA = 4866.9 \text{ kg/hr}$$

$$67.54 \text{ kmol/hr} \quad HAC = 4055.8 \text{ kg/hr}$$

$$135.08 \text{ kmol/hr} \quad 8922.7 \text{ kg/hr}$$

$$MW; 135.08 \text{ kmol/hr} \quad mw = 8922.7 \text{ kg/hr}$$

$$Mw = 66.05 \text{ kg/kmol}$$

$$L = 0.585(8922.7 \text{ kg/hr}) = 5219.8 \text{ kg/hr}$$

$$G = L + D = 5219.8 \text{ kg/hr} + 8922.7 \text{ kg/hr} = 14142.48 \text{ kg/hr}$$

$$\rho_L = \Sigma(\text{mol fraction comp})(\text{spec. gravity comp})(\rho_l \text{ H}_2\text{O})$$

$$= 0.59 (1.010)(995\text{kg/m}^3) + 0.5(.973)(995\text{kg/m}^3) = 986.54\text{kg/m}^3$$

$$\rho_g = P/RT = 30.7 \text{ mm hg} / 62.36(273.15+50.46) = 1.52 \cdot 10^{-3}\text{mol/l}$$

$$\rho_g = 1.52 \cdot 10^{-3}\text{mol/l} \times 66 \text{ g/mol} = 0.10 \text{ kg/m}^3$$

$$FLV = L/G (\rho_g/\rho_l)^{0.5}$$

$$FLV = (5219.8/14142.48) (0.10/986.54) = 0.0037$$

$$C_{SB} = 0.39 \text{ ft/sec} \quad (\text{fig 18-10 perry's}) \text{ I had to extrapolate from the graph}$$

$$C_{SB} = 0.1189 \text{ m/s} ; C = C_{SB} F_{ST} = 0.1189 \times (25/20)^{0.2} = 0.1137 \text{ m/s}$$

$$C = UN(\rho_v / \rho_l - \rho_v)^{0.5} \quad \text{solve for } UN$$

$$0.1137 = UN(0.10/986.54 - 0.10)^{0.5}; \quad UN = 11.29 \text{ m/s}$$

$$\text{Design at 50% flood } DT = \left(\frac{4G}{0.8 UN \pi (1 - A_{cl}/A) \rho_v} \right)^{0.5}$$

$$DT = \sqrt{\frac{4(14142.48 \frac{kg}{hr} \times \frac{hr}{3600 sec})}{0.8(11.29 \frac{m}{sec})(3.14)(0.9)(0.1 \frac{kg}{m^3})}} = 2.48 \text{ m or } 8.14 \text{ ft}$$

Cost of tower N=30 D=2.592

$$\text{Cost} = 561286 \times 5 = \$2806428$$

10.7.3.1.2 Condenser:

$$\text{Vapor to be condensed} = L + D$$

$$67.54(1.585) = 107.05 \text{ kmol/hr AA}$$

$$67.534(1.585) = 107.05 \text{ kmol/hr HAC}$$

ΔH_{vap}

$$AA = m_{50.5} \int^{141.6} Cp \Delta T + \Delta H_v + m_{141.6} \int^{50.5} Cp \Delta T$$

$$107.05 \frac{\text{Kmol}}{\text{Hr}} (\frac{58.35 \text{ kcal}}{\text{kg}} \times \frac{72.06 \text{ Kg}}{\text{kmol}} + 7872 \frac{\text{Kcal}}{\text{kmol}} - 50.87 \frac{\text{kcal}}{\text{kg}} \times \frac{72.06 \text{ kg}}{\text{kmol}})$$

$$= 900400 \text{ kcal/hr}$$

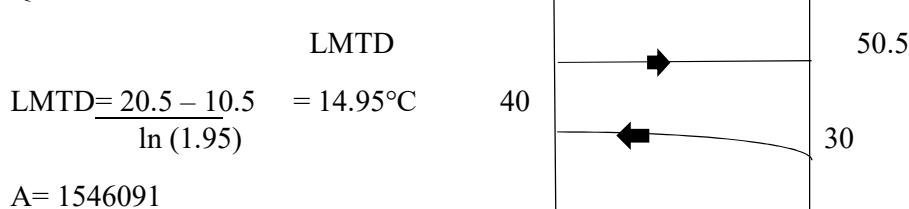
$$HAC = m_{50.5} \int^{117.9} Cp \Delta T + \Delta H_v + m_{117.9} \int^{50.5} Cp$$

$$107.05 \frac{\text{Kmol}}{\text{Hr}} (\frac{40.79 \text{ kcal}}{\text{kg}} \times \frac{60.05 \text{ Kg}}{\text{kmol}} + 5733 \frac{\text{Kcal}}{\text{kmol}} - 35.83 \frac{\text{kcal}}{\text{kg}} \times \frac{60.05 \text{ kg}}{\text{kmol}})$$

$$= 645691 \text{ kcal/hr}$$

$$Q_{\text{cond}} = 900400 \text{ kcal/hr} + 645691 \text{ kcal/hr} = 1546091 \text{ kcal/hr}$$

$$Q = UALMTD$$



$$A = \frac{1546091}{700 (14.95^\circ C)}$$

$$= 147.8 \text{ m}^2 \text{ round to } 150 \text{ m}^2$$

Cost of x-chgr A - 150 m²

$$\text{Cost} - \$52523.5 \times 5 = \$262617$$

Cooling water: $Q_{cond} = m C_p \Delta T$

$$1546091 \frac{\text{kcal}}{\text{hr}} = m(1.0 \frac{\text{Kcal}}{\text{kg}^\circ\text{C}}) 10^\circ\text{C}, m = 154609.1 \frac{\text{kg}}{\text{hr}}$$

$$154609.1 \frac{\text{kg}}{\text{hr}} \times \$0.02 \times 0.001005 \frac{\text{m}^3}{\text{kg}} \times 1000 \frac{\text{L}}{\text{m}^3} \times 24 \frac{\text{hr}}{\text{day}} \times 365 \frac{\text{day}}{\text{yr}} \times 0.88 = \$23956.2/\text{yr}$$

10.7.3.1.3 Reboiler

$$H43 + Q_{reb} = Q_{cond} - H52 + H46$$

$$Q_{reb} = 1546091 + m \int_{90}^{50.5} C_p \Delta t \quad \text{strm 52 (LIQ)}$$

$$\text{AA: } m_{90} \int^{50.5} C_p \Delta t = -20.94 \frac{\text{kcal}}{\text{Kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}} \times 67.54 \frac{\text{kgmol}}{\text{hr}} = -101,936 \text{kcal/hr}$$

$$\text{HAC: } m_{90} \int^{50.5} C_p \Delta t = -20.6 \frac{\text{kcal}}{\text{Kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}} \times 67.54 \frac{\text{kgmol}}{\text{hr}} = -835550 \text{kcal/hr}$$

$$Q_{reb} = 1546091 - 101936 - 835550 = 1356005 \text{kcal/hr}$$

$$Q_{reb} = UALMTD \quad \Delta T = LMTD = 155.48 - 90^\circ\text{C} = 65.48$$

$$A = \frac{Q_{reb}}{U \Delta T} = \frac{1360605 \text{kcal/hr}}{1000 \times 65.48} = 20.8 \text{ m}^2$$

10.7.3.2 Cost of reboiler

$$A = 13 \text{ m}^2$$

$$\text{Cost} = \$14659 \times 5 = \$73295$$

Cost of steam

$$Q_{reb} = m(\Delta H_{st}) , 1360605 \text{kcal/hr} = m (500.94 \text{kcal/kg})$$

$$M = 2716 \text{kg/hr}$$

Cost

$$2716 \text{kg/hr} \times \$20/1000 \text{kg} \times 24 \text{hr/day} \times 365 \text{ day/yr} \times 0.88$$

$$= \$418742/\text{yr}$$

10.7.3.2.1 PW calculation

Table 35

	equipment	cost	utility cost	utility
T106	distl column	2806428		
E108	T106 cond	262617	23956.2	Cw
E109	T106 reboil	73295	418742	Steam
		3142340	442698.2	

Table 36

EOY	Cash flow	Dep	Taxable income	Tax	After Tax Cash Flow
0	-3142340	-	-	-	-3142230
1	-442698.2	-285667	-728365.5	-364183	-78515.5
2	-442698.2	-285667	-728365.5	-364183	-78515.5
3	-442698.2	-285667	-728365.5	-364183	-78515.5
4	-442698.2	-285667	-728365.5	-364183	-78515.5
5	-442698.2	-285667	-728365.5	-364183	-78515.5
6	-442698.2	-285667	-728365.5	-364183	-78515.5
7	-442698.2	-285667	-728365.5	-364183	-78515.5
8	-442698.2	-285667	-728365.5	-364183	-78515.5
9	-442698.2	-285667	-728365.5	-364183	-78515.5
10	-442698.2	-285667	-728365.5	-364183	-78515.5
11	-442698.2	-285667	-728365.5	-364183	-78515.5

$$PW = -3142340 + -78515.5(5.2337) = \$-3553266$$

10.7.3.3 CASE 2 $2.0 \times R_{min}$

$$L/D = 0.7795$$

$$\# \text{ Trays} = 12.93/0.5 = 25.5 \text{ or } 26 \text{ real trays}$$

$$P_{top} \text{ (Dew point calculation)} K_1 = Y_1/X_c$$

Table 37

Y1	K1(T=52)	X1	K1(T=60)	X1
HAC = 0.5	1.37	0.364	1.975	0.253
AA = 0.5	0.509	0.981	0.768	0.651
		1.345		0.904

Interpolate T= 58.3°C

10.7.3.3.1 Tower Diameter

$$L = 0.7795(D); L = 0.7795(135.08 \text{ kmol/hr}) X 66.05 \text{ kg/kmol} = 6954.73 \text{ kg/hr}$$

$$G = L + D = 135.08 \text{ kg/hr} X 66.05 \text{ kg/kmol} + 6954.73 \text{ kg/hr} = 15876.76 \text{ kg/hr}$$

$$\rho_g = P/RT = 44.7 \text{ mm hg} / 62.36(273.15+58.3) = 2.163 \times 10^{-3} \text{ mol/l}$$

$$\rho_g = 2.163 \times 10^{-3} \text{ mol/l} \times 66.05 \text{ Kg/kmol} \times \text{kgmol}/1000\text{mol} \times 1000 \text{ l/m}^3 = 0.143 \text{ kg/m}^3$$

$$\rho_l = 986.54 \text{ kg/m}^3$$

$$FLV = L/G (\rho_g / \rho_l)^{0.5}$$

$$FLV = (6954.73/15876.76) (0.143/986.54) = 0.0053$$

$C_{SB} = 0.39 \text{ ft/sec}$ (fig 18-10 perry's) I had to extrapolate from the graph

$$C_{SB} = 0.1189 \text{ m/s} ; C = C_{SB} F_{ST} = 0.1189 \times (25/20)^{0.2} = 0.1137 \text{ m/s}$$

$$C = UN(\rho_v / \rho_l - \rho_v)^{0.5} \quad \text{solve for } UN$$

$$0.1137 = UN(0.143/986.54 - 0.143)^{0.5} ; UN = 9.44 \text{ m/s}$$

$$\text{Design at 50% f load} \quad DT = \left(\frac{4G}{0.8 UN \pi (1 - A_{cl}/A) \rho_v} \right)^{0.5}$$

$$DT = \sqrt{\frac{4(115876.76 \frac{kg}{hr} \times \frac{hr}{3600 sec})}{0.8(9.44 \frac{m}{sec})(3.14)(0.9)(0.143 \frac{kg}{m^3})}} = 2.4 \text{ m or } 7.88 \text{ ft}$$

Cost of tower N=26 D=2.44

$$\text{Cost} = \$462823.7 \times 5 = \$2314118$$

10.7.3.3.2 Condenser

Vapor to be condensed = L + D

$$67.54(1.7795) = 120.19 \text{ kmol/hr AA}$$

$$67.534(1.7795) = 120.19 \text{ kmol/hr HAC}$$

ΔH_{vap}

$$AA = m_{58.3} \int^{141.6} Cp \Delta T + \Delta H_v + m_{141.6} \int^{58.3} Cp \Delta T$$

$$120.19 \frac{\text{Kmol}}{\text{Hr}} \left(\frac{53.5 \text{ kcal}}{\text{kg}} \times \frac{72.06 \text{ Kg}}{\text{kmol}} + \frac{7872 \text{ Kcal}}{\text{kmol}} - \frac{46.8 \text{ kcal}}{\text{kg}} \times \frac{72.06 \text{ kg}}{\text{kmol}} \right)$$

$$= 1004239.8 \text{ kcal/hr}$$

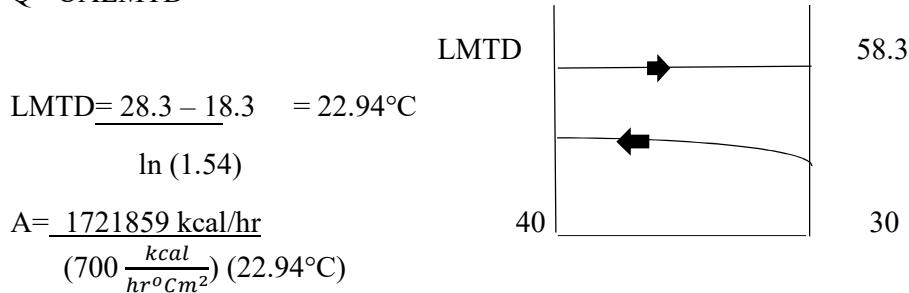
$$HAC = m_{58.3} \int^{117.9} Cp \Delta T + \Delta H_v + m_{117.9} \int^{58.3} Cp$$

$$120.19 \frac{\text{Kmol}}{\text{Hr}} \left(\frac{35.8 \text{ kcal}}{\text{kg}} \times \frac{60.05 \text{ Kg}}{\text{kmol}} + \frac{5733 \text{ Kcal}}{\text{kmol}} - \frac{31.84 \text{ kcal}}{\text{kg}} \times \frac{60.05 \text{ kg}}{\text{kmol}} \right)$$

$$= 717619.4 \text{ kcal/hr}$$

$$Q_{cond} = 1004239.8 \text{ kcal/hr} + 717619.4 \text{ kcal/hr} = 1721859 \text{ kcal/hr}$$

$Q = UALMTD$



= 107.2m² round to 110 m²

Cost of x-chgr A - 110 m²

Cost - \$41808X 5 = \$209040

Cooling water Qcond= m Cp ΔT

$$1721859 \frac{\text{kcal}}{\text{Hr}} = m(1.0 \frac{\text{Kcal}}{\text{kg}^{\circ}\text{C}}) 10^{\circ}\text{C}, m = 172186 \frac{\text{kg}}{\text{hr}}$$

10.7.3.3.3 Reboiler

$$H43 + Q_{reb} = Q_{cond} + H52 + H46$$

$$Q_{reb} = 1721859 + m_{90} \int^{58.3} C_p \Delta t \text{ strm 52}$$

$$\text{AA: } m_{90} \int^{58.3} C_p \Delta t = -16.92 \frac{\text{kcal}}{\text{Kg}} X 72.06 \frac{\text{kg}}{\text{kmol}} X 67.54 \frac{\text{kgmol}}{\text{hr}} = -82348.5 \text{kcal/hr}$$

$$\text{HAC: } m_{90} \int^{58.3} C_p \Delta t = -16.62 \frac{\text{kcal}}{\text{Kg}} X 60.05 \frac{\text{kg}}{\text{kmol}} X 67.54 \frac{\text{kgmol}}{\text{hr}} = -67389 \text{kcal/hr}$$

$$Q_{reb} = 1721859 - 67389 - 82348.5 = 1572121 \text{kcal/hr}$$

$$Q_{reb} = UALMTD \quad \Delta T = LMTD = 155.48 - 90^{\circ}\text{C} = 65.48$$

$$A = \frac{Q_{reb}}{U\Delta T} = \frac{1572121 \text{kcal/hr}}{1000 X 65.48} = 24 \text{ m}^2 \text{ round to } 25 \text{ m}^2$$

Cost of reboiler A= 25 m²

Cost = \$16484.2 X 5 = \$82421

Cost of steam

Qreb= m(ΔHstm), 1572121 kcal/hr= m (500.94kcal/kg)

M= 3138.34kg/hr

Cost

3138.34kg/hr X \$20/1000kg X 24hr/day X 365 day/yr X 0.88 = \$483857/yr

Cooling water

172186kg/hr X \$0.02/1000L X 0.001003 m³/kg X 1000L/ m³ 24hr/day X 365 day/yr X 0.88
=\$26680

10.7.3.3.4 PW Calculation

Table 38

	Equipment	Cost	Utility cost	Utility
T106	distl column	2314118		
E108	T106 cond	209040	26680	Cw
E109	T106 reboil	82421	483857	stm
		2605579	510537	

Table 39

EOY	Cash flow	Dep	Taxable Income	Tax	After Tax Cash Flow
0	-2605579	-	-	-	-2605579
1	-510537	-236870	-747408	-373703	136833
2	-510537	-236870	-747408	-373703	136833
3	-510537	-236870	-747408	-373703	136833
4	-510537	-236870	-747408	-373703	136833
5	-510537	-236870	-747408	-373703	136833
6	-510537	-236870	-747408	-373703	136833
7	-510537	-236870	-747408	-373703	136833
8	-510537	-236870	-747408	-373703	136833
9	-510537	-236870	-747408	-373703	136833
10	-510537	-236870	-747408	-373703	136833
11	-510537	-236870	-747408	-373703	136833

$$PW = -2405579 - 136833(5.2337) = \$-3321722$$

10.7.3.4 Case 3 (3.0 × R_{min})

$$L/D = 1.169$$

$$\# \text{ Trays} = 10.94 \quad \text{real trays} = 22$$

$$P_{top} = 135.7 - 3.5(22) = 58.7 \text{ mm hg}$$

$$T_{top} \text{ (Dew point calculation)} \quad K_1 = Y_1/X_1$$

Table 40

Y1	K1(T=65)	X1	K1(T=60)	X1
HAC = 0.5	1.867	0.268	1.504	0.333
AA = 0.5	0.748	0.669	0.585	0.855
		0.937		1.188

Interpolate T = 63.8°C

10.7.3.4.1 Tower Diameter

$$L = 1.169(D); \quad L = 1.169(135.08 \text{ kmol/hr}) X 66.05 \text{ kg/kmol} = 10429.9 \text{ kg/hr}$$

$$G = L + D = 135.08 \text{ kg/hr} X 66.05 \text{ kg/kmol} + 10429.9 \text{ kg/hr} = 19351.9 \text{ kg/hr}$$

$$\rho g = P/RT = 58.7 \text{ mm hg} / 62.36(273.15+63.8) = 2.794 \times 10^{-3} \text{ mol/l}$$

$$\rho g = 2.794 \times 10^{-3} \text{ mol/l} X 66.05 \text{ Kg/kmol} X \text{kgmol}/1000\text{mol} X 1000 \text{ l/m}^3 = 0.1845 \text{ kg/m}^3$$

$$\rho l = 986.54 \text{ kg/m}^3$$

$$FLV = L/G (\rho g / \rho l)^{0.5}$$

$$FLV = (10429.9/19351.9) (0.1845/986.54) = 0.0074$$

$$C_{SB} = 0.39 \text{ ft/sec} \quad (\text{fig 18-10 perry's}) \text{ I had to extrapolate from the graph}$$

$$C_{SB} = 0.1189 \text{ m/s} ; C = C_{SB} F_{ST} = 0.1189 \times (25/20)^{0.2} = 0.1137 \text{ m/s}$$

$$C = UN(\rho_v / \rho_l - \rho_v)^{0.5} \quad \text{solve for } UN$$

$$0.1137 = UN(0.1845/986.54 - 0.1845)^{0.5} ; UN = 8.31 \text{ ft/s}$$

$$\text{Design at 80% flood } DT = \left(\frac{4G}{0.8 UN \pi (1 - A_{cl}/A) \rho_v} \right)^{0.5}$$

$$DT = \sqrt{\frac{4(19351.9 \frac{kg}{hr} \times \frac{hr}{3600 sec})}{0.8(8.31 \frac{m}{sec})(3.14)(0.9)(0.1845 \frac{kg}{m^3})}} = 2.49 \text{ m or } 8.17 \text{ ft}$$

Cost of tower N=22 D=2.51

$$\text{Cost} = \$405519.25 \times 5 = \$2027596$$

10.7.3.4.2 Condenser

$$\text{Vapor to be condensed} = L + D$$

$$67.54(2.169) = 146.5 \text{ kmol/hr AA and HAC}$$

ΔH_{vap}

$$\text{AA} = m_{63.8} \int_{141.6}^{141.6} Cp \Delta T + \Delta H_v + m_{141.6} \int_{63.8}^{63.8} Cp \Delta T$$

$$146.5 \frac{\text{Kmol}}{\text{Hr}} (49.9 \frac{\text{kcal}}{\text{kg}} \times 72.06 \frac{\text{Kg}}{\text{kmol}} + 7872 \frac{\text{Kcal}}{\text{kmol}} - 43.92 \frac{\text{kcal}}{\text{kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}})$$

$$= 1217328 \text{ kcal/hr}$$

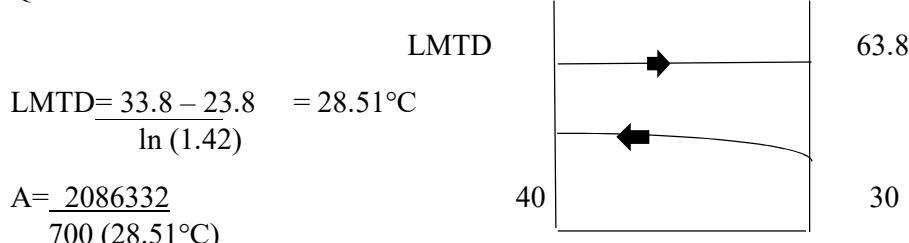
$$\text{HAC} = m_{63.8} \int_{117.9}^{117.9} Cp \Delta T + \Delta H_v + m_{117.9} \int_{63.8}^{63.8} Cp$$

$$146.50 \frac{\text{Kmol}}{\text{Hr}} (32.31 \frac{\text{kcal}}{\text{kg}} \times 60.05 \frac{\text{Kg}}{\text{kmol}} + 5733 \frac{\text{Kcal}}{\text{kmol}} - 29.0 \frac{\text{kcal}}{\text{kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}})$$

$$= 869004 \text{ kcal/hr}$$

$$Q_{\text{cond}} = 1217328 \text{ kcal/hr} + 869004 \text{ kcal/hr} = \text{kcal/hr}$$

$$Q = UALMTD$$



$$\text{LMTD} = \frac{33.8 - 23.8}{\ln(1.42)} = 28.51^\circ\text{C}$$

$$A = \frac{2086332}{700(28.51^\circ\text{C})}$$

$$= 104.5 \text{ m}^2 \text{ round to } 105 \text{ m}^2$$

Cost of x-chgr A - 105 m²

$$\text{Cost} - \$40441.78 \times 5 = \$202209$$

Cooling water $Q_{cond} = m \cdot C_p \cdot \Delta T$

$$2086332 \frac{\text{kcal}}{\text{Hr}} = m(1.0 \frac{\text{Kcal}}{\text{kg}\cdot\text{c}}) 10^\circ\text{C}, m = 208633 \frac{\text{kg}}{\text{hr}}$$

10.7.3.4.3 Reboiler

$$H43 + Q_{reb} = Q_{cond} + H52 + H46$$

$$Q_{reb} = 2086332 + m_{90} \int^{63.8} C_p \Delta t \text{ strm 52}$$

$$\text{AA: } m_{90} \int^{63.8} C_p \Delta t = -14.05 \frac{\text{kcal}}{\text{Kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}} \times 67.54 \frac{\text{kgmol}}{\text{hr}} = -68366 \text{kcal/hr}$$

$$\text{HAC: } m_{90} \int^{63.8} C_p \Delta t = -13.778 \frac{\text{kcal}}{\text{Kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}} \times 67.54 \frac{\text{kgmol}}{\text{hr}} = -55897 \text{kcal/hr}$$

$$Q_{reb} = 2086332 - 68366 - 55897 = 1962069 \text{kcal/hr}$$

$$Q_{reb} = UALMTD \quad \Delta T = LMTD = 155.48 - 90^\circ\text{C} = 65.48$$

$$A = \frac{Q_{reb}}{U \Delta T} = \frac{1962069 \text{kcal/hr}}{1000 \times 65.48} = 30 \text{ m}^2$$

Cost of reboiler A= 30 m²

$$\text{Cost} = \$18223 \times 5 = \$91118$$

Cost of steam

$$Q_{reb} = m(\Delta H_{st}) , 1962069 \text{kcal/hr} = m (500.94 \text{kcal/kg})$$

$$M = 3916.77 \text{kg/hr}$$

Cost

$$3916.77 \text{kg/hr} \times \$20/1000 \text{kg} \times 24 \text{hr/day} \times 365 \text{ day/yr} \times 0.88 \\ = \$603873/\text{yr}$$

10.7.3.4.4 PW Calculation

Table 41

	Equipment	Cost	Utility cost	Utility
T106	distl column	2027596		
E108	T106 cond	202209	32337	Cw
E109	T106 reboil	91118	603873	stm
		2320923	636210	

Table 42

EOY	Cash flow	Dep	Taxable Income	Tax	After Tax Cash Flow
0	-2320923	-	-	-	-2320923
1	-636209	-210993	-847202	-423601	-212608
2	-636209	-210993	-847202	-423601	-212608
3	-636209	-210993	-847202	-423601	-212608
4	-636209	-210993	-847202	-423601	-212608
5	-636209	-210993	-847202	-423601	-212608
6	-636209	-210993	-847202	-423601	-212608
7	-636209	-210993	-847202	-423601	-212608
8	-636209	-210993	-847202	-423601	-212608
9	-636209	-210993	-847202	-423601	-212608
10	-636209	-210993	-847202	-423601	-212608
11	-636209	-210993	-847202	-423601	-212608

$$PW = -2320923 - 212608(5.2337) = \$-3433650$$

Therefore, I will go with the case 2 since it has the least negative worth of all the cases.

10.7.4 T107 Optimization:

CASE1 - 1.5 Rmin

L/D= 2.595

#Trays = 9.81, 9.81/0.5=20 real trays

P bottoms Kc= Y1/X1

Table 43

X1	K1 (160 mmHg)	Y1	K1 (180 mmHg)	Y1	K1 (200 mmHg)	Y1	K1 (210 mmHg)	Y1
AA=0.524	0.836	0.438	0.743	0.389	0.669	0.35	0.637	0.334
HAC =0.476	1.828	0.87	1.625	0.773	1.462	0.696	1.393	0.663
		1.308		1.162		1.046		0.997

Interpolate P= 209.4 mm hg

Ptop= 209.4 – 3.5(20) = 139.4 mm hg

10.7.4.1.1 Column Sizing

L = 2.595(D) = 2.595(6.905 kmol/hr) = 17.92 kmol/hr

D: AA = 0.345 kmol/hr X 72.06kg/mol = 24.86 kg/hr

$$\text{HAC} = 6.56 \text{ kmol/hr} \times 60.05 \text{ kg/mol} = \frac{393.93 \text{ kg/hr}}{418.79 \text{ kg/hr}}$$

$$6.905 \text{ kmol/hr} \times \text{M.W} = 418.79 \text{ kg/hr}, \quad \text{M.W} = 60.65 \text{ kg/kmol}$$

$$L = 17.92 \text{ kmol/hr} \times 60.65 \text{ kg/kmol} = 1086.85 \text{ kg/hr}$$

$$G = L + D = 3.595(6.905 \text{ kmol/hr}) \times 60.65 \text{ kg/kmol} = 1505.5 \text{ kg/hr}$$

T-top (Dew point calculation) $P = 139.4 \text{ mm hg}$ $K_1 = Y_1/X_1$

Table 44

y1	k(T=60°C)	X1	k(T=80°C)	X1	k(T=70°C)	X1
AA= 0.05	0.246	0.999	0.628	0.08	0.399	0.125
HAC= 0.95	0.633	1.499	1.4428	0.659	0.968	0.9811
		2.498		0.739		1.1061

Interpolate T= 72.9°C

$$\rho_g = P/RT = 139.44 \text{ mm hg} / 62.36(273.15+72.9)$$

$$\rho_g = 0.0065 \text{ mol/l} \times 66.65 \text{ kg/kmol} = 0.388 \text{ kg/m}^3$$

$$\rho_l = 969.9 \text{ kg/m}^3$$

$$F_{LV} = L/G (\rho_g/\rho_l)^{0.5}$$

$$F_{LV} = (1028.4.9/1423.8) (0.388/969.9)^{0.5} = 0.0144$$

$C_{SB} = 0.39 \text{ ft/sec}$ (fig 18-10 perry's) I had to extrapolate from the graph

$$C_{SB} = 0.1189 \text{ m/s} ; C = C_{SB} F_{ST} = 0.1189 \times (25/20)^{0.2} = 0.1137 \text{ m/s}$$

$$C = UN(\rho_v/\rho_l - \rho_v)^{0.5} \quad \text{solve for } UN$$

$$0.1137 = UN(0.388/969.9 - 0.388)^{0.5} ; UN = 5.684 \text{ ft/s}$$

Design at 80% f load $DT = (4G/0.8UNT_1(1-Ac_1/a)\rho_v)^{0.5}$

$$DT = \sqrt{\left(\frac{4(1423.83)\left(\frac{1}{3600}\right)}{0.8(5.684)T_1(0.9)0.388}\right)} = 0.56 \text{ m or } 1.85 \text{ ft or round to } 0.61 \text{ m}$$

Cost of tower N=22 D=0.61 m

$$\text{Cost} = \$158462.4 \times 5 = \$792312$$

10.7.4.1.2 Condenser

Vapor to be condensed = L + D

$$67.54(2.169) = 146.5 \text{ kmol/hr AA and HAC}$$

ΔH_{vap}

$$AA = m_{72.9} \int^{141.6} Cp \Delta T + m \Delta H_v + m_{141.6} \int^{72.9} Cp \Delta T$$

$$1.24 \frac{\text{Kmol}}{\text{Hr}} (44.12 \frac{\text{kcal}}{\text{kg}} \times 72.06 \frac{\text{Kg}}{\text{kmol}} + 7872 \frac{\text{Kcal}}{\text{kmol}} - 39.12 \frac{\text{kcal}}{\text{kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}})$$

$$= 10208 \text{kcal/hr}$$

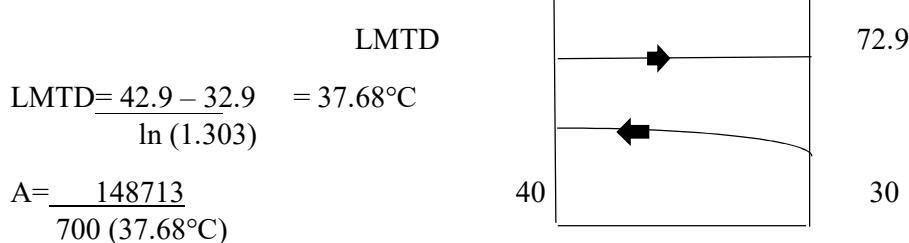
$$HAC = m_{72.9} \int^{117.9} Cp \Delta T + m \Delta H_v + m_{117.9} \int^{72.89} Cp$$

$$23.58 \frac{\text{Kmol}}{\text{Hr}} (26.62 \frac{\text{kcal}}{\text{kg}} \times 60.05 \frac{\text{Kg}}{\text{kmol}} + 5733 \frac{\text{Kcal}}{\text{kmol}} - 24.28 \frac{\text{kcal}}{\text{kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}})$$

$$= 138505 \text{kcal/hr}$$

$$Q_{\text{cond}} = 10208 \text{kcal/hr} + 138505 \text{kcal/hr} = 148713 \text{kcal/hr}$$

$$Q = UALMTD$$



$$\text{LMTD} = \frac{42.9 - 32.9}{\ln(1.303)} = 37.68^\circ\text{C}$$

$$A = \frac{148713}{700 (37.68^\circ\text{C})}$$

$$= 5.63 \text{m}^2 \text{ round to } 5.75 \text{ m}^2$$

Cost of x-chgr

$$\text{Cost} - \$8494.8 \times 5 = \$42474$$

$$\text{Cooling water} \quad Q_{\text{cond}} = m \text{ Cp } \Delta T$$

$$148713 \frac{\text{kcal}}{\text{Hr}} = m (1.0 \frac{\text{Kcal}}{\text{kg}^\circ\text{C}}) 10^\circ\text{C}, m = 14871.3 \frac{\text{kg}}{\text{hr}}$$

Cost of water

$$14871.3 \frac{\text{kg}}{\text{Hr}} \times \$0.02 \times 0.001005 \frac{\text{m}^3}{\text{kg}} \times 24 \frac{\text{hrs}}{\text{day}} \times 365 \frac{\text{days}}{\text{yr}} \times 0.88$$

$$= \$2304.3/\text{yr}$$

10.7.4.1.3 Reboiler

$$H52 + Q_{\text{reb}} = Q_{\text{cond}} + H53 + H42$$

Table 45

S52 (T= 58.3°C)	S42 (T= 72.9°C)	S53 (T= 90°C)
AA=67.54 kmol/hr	0.345 kmol/hr	67.195 kmol/hr
HAC= 67.54 kmol/hr	6.56 kmol/hr	60.98 kmol/hr

$$Q_{reb} = Q_{cond} + m_{58.3} \int^{90} Cp \Delta t \text{ strm 53} + m_{58.3} \int^{72.9} Cp \Delta t \text{ strm 42}$$

$$\text{AA: } m_{58.3} \int^{90} Cp \Delta t \text{ strm 53} = 16.9 \frac{\text{kcal}}{\text{Kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}} \times 67.54 \frac{\text{kgmol}}{\text{hr}} = 81900 \text{kcal/hr}$$

$$m_{58.3} \int^{72.9} Cp \Delta t \text{ strm 42} = 7.67 \frac{\text{kcal}}{\text{Kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}} \times 0.345 \frac{\text{kgmol}}{\text{hr}} = 19.08 \text{kcal/hr}$$

$$\text{HAC: } m_{90} \int^{58.3} Cp \Delta t = 16.6 \frac{\text{kcal}}{\text{Kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}} \times 60.98 \frac{\text{kgmol}}{\text{hr}} = 60844 \text{kcal/hr}$$

$$m_{58.3} \int^{72.9} Cp \Delta t \text{ strm 42} = 7.57 \frac{\text{kcal}}{\text{Kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}} \times 6.56 \frac{\text{kgmol}}{\text{hr}} = 2981 \text{kcal/hr}$$

$$145916 \text{ kcal/hr}$$

$$Q_{reb} = 148713 + 145916 = 294629 \text{ kcal/hr}$$

$$Q_{reb} = UALMTD \quad \Delta T = LMTD = 155.48 - 90^\circ\text{C} = 65.48$$

$$A = \frac{Q_{reb}}{U\Delta T} = \frac{294629 \text{ kcal/hr}}{1000 \times (155.48 - 90)} = 4.5 \text{ m}^2$$

Cost of reboiler

$$\text{Cost} = \$7799.8 \times 5 = \$38999.2$$

Cost of steam

$$Q_{reb} = m(\Delta H_{st}) , 294629 \text{ kcal/hr} = m (500.94 \text{ kcal/kg})$$

$$M = 588.15 \text{ kg/hr}$$

Cost

$$588.15 \text{ kg/hr} \times \$20/1000 \text{ kg} \times 24 \text{ hr/day} \times 365 \text{ day/yr} \times 0.88$$

$$= \$90678/\text{yr}$$

10.7.4.1.4 PW Calculation

Table 46

	Equipment	Cost	Utility cost	Utility
T107	Dist. Column	792312		-
E110	Cond	42474	2304.3	Cw
E111	Reboiler	38999	90679	stm
		873785	92983.3	

Table 47

EOY	CF	Dep	TI	Tax	After Tax Cash Flow
0	-873785	-	-	-	-873785
1	-92983	-79435	-172418	-86209	-6774
3	-92983	-79435	-172418	-86209	-6774
4	-92983	-79435	-172418	-86209	-6774
5	-92983	-79435	-172418	-86209	-6774
6	-92983	-79435	-172418	-86209	-6774
7	-92983	-79435	-172418	-86209	-6774
8	-92983	-79435	-172418	-86209	-6774
9	-92983	-79435	-172418	-86209	-6774
10	-92983	-79435	-172418	-86209	-6774
11	-92983	-79435	-172418	-86209	-6774

$$PW = -873785 - 5.237(6774) = \$-909238$$

10.7.4.2 Case 2: 3.0 X R_{min}

Trays= 6.7 or 14 real

L/D= 2.595

$$P_{top} = 209.4 - 14(3.5) = 160.4 \text{ mm hg}$$

Table 48

y1	k(T=80°C)	Xl	k(T=80°C)	Xl
AA= 0.05	0.546	0.092	0.347	1.265
HAC= 0.95	1.254	0.7576	0.841	0.129
		0.8496		1.394

Interpolate T= 77.2°C

10.7.4.2.1 Column Sizing

$$L = 5.19(D) = 5.19(6.905 \text{ kmol/hr}) \times 60.05 = 2173.5 \text{ kg/hr}$$

$$G = L+D = 6.19(6.905 \text{ kmol/hr}) \times 60.65 \text{ kg/kmol} = 2592.3 \text{ kg/hr}$$

$$\rho_g = P/RT = 160.4 \text{ mm hg} / 62.36(273.15+77.2)$$

$$\rho_g = 0.0073 \text{ mol/l} \times 60.65 \text{ Kg/kmol} = 0.4453 \text{ kg/m}^3$$

$$\rho_{gl} = 969.9 \text{ kg/m}^3$$

$$F_{LV} = L/G (\rho g / \rho l)^{0.5}$$

$$F_{LV} = (2173.5/2592.3) (0.4453/969.9)^{0.5} = 0.018$$

$$C_{SB} = 0.39 \text{ ft/sec} \quad (\text{fig 18-10 perry's}) \text{ I had to extrapolate from the graph}$$

$$C_{SB} = 0.1189 \text{ m/s} ; C = C_{SB} F_{ST} = 0.1189 \times (25/20)^{0.2} = 0.1137 \text{ m/s}$$

$$C = UN(\rho_v / \rho_l - \rho_v)^{0.5} \quad \text{solve for } UN$$

$$0.1137 = UN(0.4453/969.9 - 0.4453)^{0.5} ; UN = 5.31 \text{ ft/s}$$

$$\text{Design at 80% f load } DT = (4G/0.8UNT_1(1-Ac_1/a) \rho v)^{0.5}$$

$$DT = \sqrt{\left(\frac{4(2592.3)\left(\frac{1}{3600}\right)}{0.8(5.31)T_1(0.9)0.4453} \right)} = 0.74 \text{ m or } 2.41 \text{ ft or round to } 0.762 \text{ m}$$

Cost of tower N=14 D=0.762 m

$$\text{Cost} = \$121067.5 \times 5 = \$605338$$

10.7.4.2.2 Condenser

Vapor to be condensed = L + D

$$0.345(6.19) = 2.13 \text{ kmol/hr AA}$$

$$6.56(6.19) = 40.6 \text{ kmol/hr HAC}$$

ΔH_{vap}

$$AA = m_{77.2} \int^{141.6} Cp \Delta T + m \Delta H_v + m_{141.6} \int^{77.2} Cp \Delta T$$

$$2.13 \frac{\text{Kmol}}{\text{Hr}} \left(\frac{41.3 \frac{\text{kcal}}{\text{kg}}}{\text{kmol}} \times 72.06 \frac{\text{Kg}}{\text{kmol}} + 7872 \frac{\text{Kcal}}{\text{kmol}} - \frac{36.82 \frac{\text{kcal}}{\text{kg}}}{\text{kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}} \right)$$

$$= 17455 \text{ kcal/hr}$$

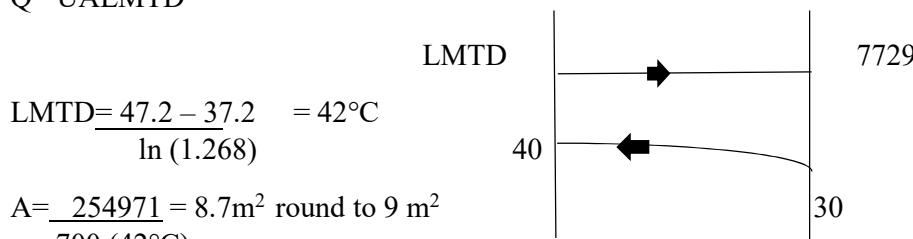
$$HAC = m_{77.2} \int^{117.9} Cp \Delta T + m \Delta H_v + m_{117.9} \int^{77.2} Cp$$

$$40.6 \frac{\text{Kmol}}{\text{Hr}} \left(\frac{23.97 \frac{\text{kcal}}{\text{kg}}}{\text{kmol}} \times 60.05 \frac{\text{Kg}}{\text{kmol}} + 5733 \frac{\text{Kcal}}{\text{kmol}} - \frac{27.01 \frac{\text{kcal}}{\text{kg}}}{\text{kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}} \right)$$

$$= 237516 \text{ kcal/hr}$$

$$Q_{cond} = 237516 \text{ kcal/hr} + 17455 \text{ kcal/hr} = 254971 \text{ kcal/hr}$$

Q = UALMTD



Cost of x-chgr

$$\text{Cost} - \$10115.7 \times 5 = \$42474$$

Cooling water Qcond= m Cp ΔT

$$\frac{254971 \text{ kcal}}{\text{Hr}} = \frac{m(1.0 \text{ Kcal})}{\text{kg}^\circ\text{C}} 10^\circ\text{C}, m = 25497 \frac{\text{kg}}{\text{hr}}$$

Cost of water

$$25497 \frac{\text{kg}}{\text{Hr}} \times \$0.02 \frac{\text{X } 0.001005 \text{ m}^3}{\text{m}^3} \times \frac{\text{X } 24 \text{ hrs}}{\text{kg day}} \times \frac{\text{X } 365 \text{ days}}{\text{yr}} \times 0.88 \\ = \$3951/\text{yr}$$

10.7.4.2.3 Reboiler

$$\text{H52} + \text{Qreb} = \text{Qcond} + \text{H53} + \text{H42}$$

$$\text{A: } m_{68.3} \int^{77.2} \text{Cp} \Delta t \text{ strm 42} = 9.97 \frac{\text{kcal}}{\text{Kg}} \times 72.06 \frac{\text{kg}}{\text{kmol}} \times 0.345 \frac{\text{kgmol}}{\text{hr}} = 247.9 \text{kcal/hr}$$

$$\text{HAC: } m_{68.3} \int^{77.2} \text{Cp} \Delta t \text{ strm 42} = 9.82 \frac{\text{kcal}}{\text{Kg}} \times 60.05 \frac{\text{kg}}{\text{kmol}} \times 6.56 \frac{\text{kgmol}}{\text{hr}} = 2981 \text{kcal/hr}$$

$$\text{AA+ HAC : } m_{58.3} \int^{90} \text{Cp} \Delta t \text{ strm 53} = 81900 + 60844 = \frac{142744 \text{ kcal/hr}}{146861 \text{ kcal/hr}}$$

$$\text{Qreb} = 254971 + 146861 = 401831 \text{ kcal/hr}$$

$$\text{Qreb} = \text{UALMTD} \quad \Delta T = \text{LMTD} = 155.48 - 90^\circ\text{C} = 65.48$$

$$A = \frac{\text{Qreb}}{\text{UAT}} = \frac{401832 \text{ kcal/hr}}{1000 \times (155.48 - 90)} = 6.1 \text{ m}^2$$

Cost of reboiler

$$\text{Cost} = \$8628 \times 5 = \$43140$$

Cost of steam

$$\text{Qreb} = m(\Delta H_{\text{stm}}), 401832 \text{ kcal/hr} = m (500.94 \text{ kcal/kg})$$

$$M = 802.16 \text{ kg/hr}$$

Cost

$$802.16 \text{ kg/hr} \times \$20/1000 \text{ kg} \times 24 \text{ hr/day} \times 365 \text{ day/yr} \times 0.88 \\ = \$123674/\text{yr}$$

10.7.4.2.4 PW Calculation

Table 49

	Equipment	Cost	Utility cost	Utility
T107	Dist. Column	605338		-
E110	Cond	50578	3951	Cw
E111	Reboiler	43140	123674	stm
		699056	127625	

Table 50

EOY	CF	Dep	TI	Tax	After Tax Cash Flow
0	-699056	-	-	-	-699056
1	-127625	-63550	-1911755	-95587.8	-32037
2	-127625	-63550	-1911755	-95587.8	-32037
3	-127625	-63550	-1911755	-95587.8	-32037
4	-127625	-63550	-1911755	-95587.8	-32037
5	-127625	-63550	-1911755	-95587.8	-32037
6	-127625	-63550	-1911755	-95587.8	-32037
7	-127625	-63550	-1911755	-95587.8	-32037
8	-127625	-63550	-1911755	-95587.8	-32037
9	-127625	-63550	-1911755	-95587.8	-32037
10	-127625	-63550	-1911755	-95587.8	-32037
11	-127625	-63550	-1911755	-95587.8	-32037

$$PW = -699056 - 5.2237(32037) = \$-866728$$

Therefore, I will go with Case 2, 3.0 X Rmin, because its PW is less than case 1. For subsequent increase in L/D the # of trays does not decrease enough to overcome the cost of utilities.

10.8 APPENDIX G: MISCELLANEOUS EQUIPMENT SIZING & COSTING

Description: This section deals with the sizing and the costing of the equipment which hasn't been sized in previous calculations for the necessary economic comparisons. The sizing of equipment was done to the extent in which it was necessary to find the price of the equipment. The utility costs associated with the equipment are also calculated in this section.

Equipment sizing costing:

10.8.1 Air Compressor

$$\text{Cost} = \exp(0.73\ln(p)-2.9)$$

Turbine efficiency = 75%

P= Rated power in joules/hr

Motor efficiency = 90%

$$Kwad = 2.78 \times 10^{-4} \frac{K}{K-1} Q_1 P_1 \left\{ \frac{(P_2)^{k-1/k}}{P_1} - 1 \right\}$$

Q1= m³/hr Volumetric flow rate

P1= inlet pressure, kPa 101.3 kPa

P2= outlet pressure, kPa 101.3+ 150 = 251.3 kPa

K= Cp/Cv = 1.39 for air

$$Q_1 = \rho g = P/RT = 101.3 \text{ kPa} / 8.314(273.15+20) = 0.0416 \text{ kmol/m}^3$$

Flow rate= 1318.46kmol/hr

$$Q_1 = 1318.46/0.0416 = 31722 \text{ m}^3/\text{hr}$$

$$Kwad = 2.78 \times 10^{-4} (1.39/0.39)(31722)(101.3)\{251.3/101.3\}^{0.28} - 1$$

$$Kw = 924.5 \text{ kw}$$

$$\text{Rated power} = 924.5/(0.9)(0.75) = 1369.6 \text{ kw}$$

Cost of electricity

$$1369.6 \text{ kw} \times \$0.074 \frac{\text{x}}{\text{Kw. Hr}} \times \frac{24\text{hr}}{\text{day}} \times \frac{365 \text{ days}}{\text{year}} \times 0.88 = \$781288/\text{yr}$$

Cost of compressor

$$P = 1369.6 \text{ Kw} \times \frac{\text{KJ/sec}}{\text{Kw}} \times \frac{1000 \text{ J}}{\text{KJ}} \times \frac{3600\text{sec}}{\text{hr}} = 4.9306 \times 10^9 \text{ J/hr}$$

$$\text{Cost} = \exp(0.73\ln(p)-2.9) = \exp(0.73 \times \ln(4.9306 \times 10^9) - 2.9) = \$655179$$

$$\text{Cost} = \$655179 \times 5 = \$3275897$$

$$T_2 = T_1(P_2/P_1)^{K-1/K} = 293.15(251.3/101.3)^{0.28}$$

$$T_2 = 378.3^\circ\text{K} \text{ or } 105.1^\circ\text{C}$$

10.8.2 Reactor waste heat boiler E101

$$Q_{reb} = 23084626 \text{ kcal/hr}$$

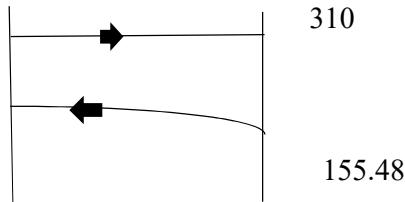
$$A = \frac{Q_{reb}}{U\Delta T}$$

Assume molten salt temperature goes from 300°C to 310°, 300°C is the temperature of reactor and 310°C is the exit temperature of product.

$$\text{LMTD} = \frac{154.52 - 144.52}{\ln(1.06)} = 149.5^\circ\text{C}$$

$$A = \frac{23084626}{700 (149.5^\circ\text{C})}$$

$$= 220.6 \text{ m}^2 \text{ round to } 220 \text{ m}^2$$



$$M_{\text{boiler water}} = 39623 \text{ kg/hr}$$

10.8.3 Cost of boiler water

$$\frac{\$2.5}{\text{m}^3} \times \frac{0.001005 \text{ m}^3}{\text{Kg}} \times 39623 \text{ kg} \times \frac{24 \text{ hr}}{\text{day}} \times \frac{365 \text{ days}}{\text{yr}} \times 0.88 = \$767433/\text{yr}$$

$$\text{Credit for steam} = 37641.9 \text{ kg/hr}$$

$$\frac{\$20}{1000 \text{ Kg}} \times 37641.9 \text{ kg} \times \frac{24 \text{ hr}}{\text{day}} \times \frac{365 \text{ days}}{\text{yr}} \times 0.88 = \$580347.7/\text{yr}$$

$$\text{Cost of boiler} = \exp(0.79 \ln(S) + 4.47)$$

$$S = \text{steam production kg/hr} = 37641.9 \text{ kg/hr}$$

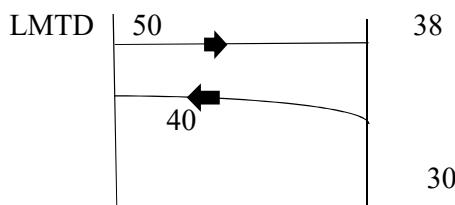
$$\text{Cost} = \$359813 \times 5 = \$1799065$$

10.8.4 E-102 X-chgr for T-101

$$Q = 14284166 \text{ kcal/hr}$$

$$A = \frac{Q}{UF_t(LMTD)}$$

$$\text{LMTD} = \frac{10 - 8}{\ln(1.25)} = 8.96^\circ\text{C}$$



$$U = 500 \text{ Kcal/hr m}^2 \text{ }^\circ\text{C}$$

$$F_t R = 50 - 38 / 40 - 30 = 1.2, \quad S = 40 - 30 / 50 - 30 = 0.5$$

$$F_t = 0.65 \text{ for a 1-shell pass 2 tube pass xChgr}$$

$$A = \frac{14284166 \text{ kcal/hr}}{50 \text{ kcal/hr}^{\circ}\text{C} (8.16)(0.65)} = 4905 \text{ m}^2 \text{ round to } 4900 \text{ m}^2$$

Cost of xchgr = \$1510529 x 5 = \$7552644

Cooling water $Q = mC_p\Delta T$

$$M = \frac{14284166 \text{ kcal/hr}}{1 \text{ kcal/kg}^{\circ}\text{C} (10^{\circ}\text{C})} = 1428416.6 \text{ kg/hr}$$

$$1428416.6 \frac{\text{kg}}{\text{hr}} \times \frac{\$20}{\text{m}^3} \times 0.001005 \frac{\text{m}^3}{\text{kg}} \times \frac{24\text{hr}}{\text{day}} \times \frac{365\text{days}}{\text{yr}} \times 0.88 = \$221329/\text{yr}$$

10.8.5 E-105 Waste heat boiler for incinerator

$T = 700 \text{ }^{\circ}\text{C}$, dewpoint = 62°C

$\text{CO}_2 = 112.83$

Assume stream 50 is at atmospheric pressure.

$\text{O}_2 = 21.57$

$\text{N}_2 = 1384.9$

$$\text{H}_2\text{O} = \frac{233.78}{1717.08 \text{ kgmol/hr}}$$

Stream 50 is to be cooled to 10°C above its dewpoint, which is 72°C .

$$Q \text{ boiler duty} = \Sigma m c p \Delta T \quad \Delta T = 700 - 72$$

$$\text{CO}_2 = 112.83 \text{ kmol/hr} \times 160.7 \text{ kcal/kg} \times 44.01 \text{ kg/kmol} = 7978934 \text{ kcal/hr}$$

$$\text{O}_2 = 21.57 \text{ kmol/hr} \times 149.8 \text{ kcal/kg} \times 32.0 \text{ kg/kmol} = 103385.7 \text{ kcal/hr}$$

$$\text{N}_2 = 1348.9 \text{ kmol/hr} \times 165.8 \text{ kcal/kg} \times 28.01 \text{ kg/kmol} = 6263565.3 \text{ kcal/hr}$$

$$\text{H}_2\text{O} = 233.78 \text{ kmol/hr} \times 374.1 \text{ kcal/kg} \times 18 \text{ kg/kmol} = 1574481.6 \text{ kcal/hr}$$

$$Q = 8739326 \text{ kcal/hr}$$

Steam is to be generated at 3500 kPa $\sim 260^{\circ}\text{C}$.

Table 51

Water 50°C (Sat'd liq)	Sat'd water (3500 kPa, 242.6°C)	Sat'd vapor (3500 kPa, 242.6°C)	Superheated vapor (3500 kPa, 260°C)
209.33 KJ/kg	1049.7 KJ/kg	2803.4 KJ/kg	2866.3 KJ/kg

Trial and error to determine the temperature of stream B and stream 65. There should be a maximum ΔT approach of 10°C in the x-chgr.

Specific temp of S = 270°C sat'd liq

$$\Sigma m_{72}^{270} C_p \Delta T$$

$$N_2 = 1348.9 \text{ kmol/hr} \times 50.47 \text{ kcal/kg} \times 28.01 \text{ kg/kmol} = 1906892 \text{ kcal/hr}$$

$$O_2 = 21.57 \text{ kmol/hr} \times 45.57 \text{ kcal/kg} \times 32.0 \text{ kg/kmol} = 31454 \text{ kcal/hr}$$

$$H_2O = 233.78 \text{ kmol/hr} \times 93.97 \text{ kcal/kg} \times 18 \text{ kg/kmol} = 395869 \text{ kcal/hr}$$

$$CO_2 = 112.83 \text{ kmol/hr} \times 45.57 \text{ kcal/kg} \times 44.01 \text{ kg/kmol} = 226294 \text{ kcal/hr}$$

$$2560509 \text{ kcal/hr}$$

Flow rate of boiler water, m

$$Q = [0.95m (2866.3 - 209.33) + 0.05m (1049.7 - 209.33)] \text{ KJ/kg} \times 0.23901 \text{ kcal/kJ}$$

$$m = 14248.9 \text{ Kg/hr}$$

T of stream 65

$$2560509 \text{ kcal/hr} = 14248.9 \text{ kg/hr} (H_s - 209.33) \text{ KJ/kg} \times 0.23901 \text{ kcal/kJ}$$

$$H_s = 961.18 \text{ KJ/kg}$$

At H = 961.18 KJ/kg , T = 223.79°C sat'd liq

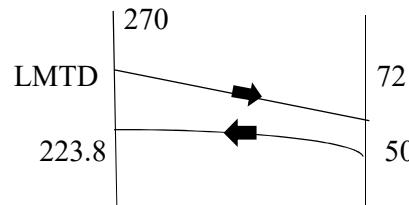
The ΔH is greater than 10°C so the exchangers can now be sized.

Sizing of E12

$$A = Q/U(LMTD)F_t$$

$$LMTD = \frac{46.2 - 22}{\ln(2.1)} = 36.62^\circ\text{C}$$

$$U = 100 \text{ Kcal/hr m}^2 \text{ }^\circ\text{C}$$

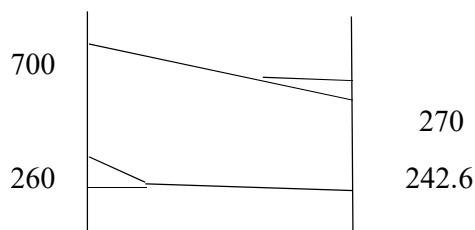


$$F_t \quad R = 270 - 72 / 223.8 - 50 = 1.14, \quad S = 223.8 - 50 / 270 - 50 = 0.79$$

F_t = 1.0 for a 1- shell pass 2 tube pass xChgr

$$A = \frac{25604.509 \text{ kcal/hr}}{100 \text{ kcal/hr m}^2 \text{ }^\circ\text{C} (32.62)(1.0)} = 784.9 \text{ m}^2 \text{ round to } 785 \text{ m}^2$$

Cost of xchgr = \$12084 x 5 = \$60422



$$Q_1 = 14248.9 \text{ kg/hr} (0.95)(2866.3 - 2803.4) \text{ KJ/kg} \times 0.23902 \text{ kcal/KJ}$$

$$Q_1 = 203503 \text{ kcal/hr}$$

$$Q_2 = 8739326 - 203503 - 2560509 = 5975314 \text{ Kcal/hr}$$

That ↓ error to determine the intermediate temp of the waste gas

$$Q = \Sigma m 685 \int_{T_1}^{T_2} C_p \Delta T \quad \text{try } 685^\circ\text{C}$$

$$N_2 = 1348.9 \text{ kmol/hr} \times 4.15 \text{ kcal/kg} \times 28.01 \text{ kg/kmol} = 156798 \text{ kcal/hr}$$

$$O_2 = 21.57 \text{ kmol/hr} \times 3.74 \text{ kcal/kg} \times 32.0 \text{ kg/kmol} = 2582 \text{ kcal/hr}$$

$$H_2O = 233.78 \text{ kmol/hr} \times 12.51 \text{ kcal/kg} \times 18 \text{ kg/kmol} = 52701 \text{ kcal/hr}$$

$$CO_2 = 112.83 \text{ kmol/hr} \times 4.185 \text{ kcal/kg} \times 44.01 \text{ kg/kmol} = \frac{20781 \text{ kcal/hr}}{232865 \text{ Kcal/hr}}$$

$$Q_1 = 203503 \text{ kcal/hr close enough to } 232865 \text{ kcal/hr}$$

$$T = 685$$

$$A_1 = Q/U(LMTD)F_t$$

$$LMTD = \frac{440 - 442.4}{\ln(0.994)} = 441.2^\circ\text{C}$$

$$U = 50 \text{ kcal/hr m}^2\text{C}$$

$$F_t \quad R = 700 - 685 / 740 - 242.6 = 0.86, \quad S = 260 - 242.6 / 700 - 242.6 = 0.04$$

$$F_t = 1$$

$$A = \frac{203503 \text{ kcal/hr}}{50 \text{ kcal/hr m}^2\text{C} (441.2) (1.0)} = 9.23 \text{ m}^2 \quad \text{round to } 10 \text{ m}^2$$

$$Q_2 = 5975314 \text{ kcal/hr}$$

$$T = 685$$

$$A_2 = Q/U(LMTD)F_t$$

$$LMTD = \frac{442.4 - 27.4}{\ln(16.14)} = 149.2^\circ\text{C}$$

$$U = 200 \text{ kcal/hr m}^2\text{C}$$

$$A_1 + A_2 = 210 \text{ m}^2$$

$$\text{Cost of boiler} = \$6471 \times 5 = \$32355$$

Cost of boiler water

$$14248.9 \text{ kg/hr} \times 0.001005 \text{ m}^3/\text{hr} \times 24 \text{ hrs/day} \times 365 \text{ days/year} \times 0.88 = \$275978/\text{yr}$$

Credit for stm

$$0.95(14248.9 \text{ Kg/hr}) \times \$20/1000 \text{ Kg} \times 24 \text{ hrs/day} \times 365 \text{ days/yr} \times 0.88 = \$2086996/\text{yr}$$

10.8.6 T102 – Absorber

of theoretical trays = 2

Real trays = $2/0.5 = 4$ trays

Assume the bottom is at atmospheric pressure

$T = 50^\circ\text{C}$

Table 52

Gas	(stream 6)	Y1	Ym.w
CO ₂ =	69.56	0.0554	2.438
AA =	2.78	0.0022	0.159
HAC =	0.56	0.0004	0.024
O ₂ =	20.98	0.0167	0.534
N ₂ =	1022.3	0.814	22.792
H ₂ O =	139.74	0.1113	2.003
	1255.92kmol/hr		27.95kg/kmol

Table 53

Liquid		X1	X1ρl	X1 m.w
H ₂ O	227	0.986	981.07	17.748
AA	2.7283	0.012	12.06	0.865
HAC	0.5016	0.002	1.94	0.12
	230.2299		995.07	18.733

$$L = (230.23 \text{ kmol/hr}) \times 18.733 \text{ kg/kmol} = 4312.9 \text{ kg/hr}$$

$$G = (1255.92 \text{ kmol/hr}) \times 27.95 \text{ kg/kmol} = 35103 \text{ kg/hr}$$

$$\rho_g = P/RT = 1 \text{ atm} / 0.08206(273.15+50)$$

$$\rho_g = 0.0434 \text{ mol/l} \times 27.95 \text{ Kg/kmol} = 1.186 \text{ kg/m}^3$$

$$\rho_l = 995.07 \text{ kg/m}^3$$

$$F_{LV} = L/G (\rho_g / \rho_l)^{0.5}$$

$$F_{LV} = (4312.9/35103) (1.186/995.07)^{0.5} = 0.0042$$

$$C_{SB} = 0.39 \text{ ft/sec} \quad (\text{fig 18-10 perry's}) \text{ I had to extrapolate from the graph}$$

$$C_{SB} = 0.1189 \text{ m/s} ; C = C_{SB} F_{ST} = 0.1189 \times (25/20)^{0.2} = 0.1137 \text{ m/s}$$

$$C = UN(\rho_v / \rho_l - \rho_v)^{0.5} \quad \text{solve for UN}$$

$$0.1137 = UN(1.186/995.07 - 1.186)^{0.5} ; UN = 3.29 \text{ ft/s}$$

Design at 80% f load DT= (4G/0.8UNT1(1-Ac1/a) ρv)^{0.5}

$$DT = \sqrt{\left(\frac{4(35103)\left(\frac{1}{3600}\right)}{0.8(3.29)3.14(0.9)1.186} \right)} = 2.1 \text{ m or } 6.9 \text{ ft or round to } 7 \text{ ft} = 2.13 \text{ m}$$

Cost of tower N=4 D=2.13 m

Cost = \$77333 X 5 = \$386664

10.8.7 Incinerator

Feed is stream 14, 47 & 49

Table 54

20°C, 101.3KPa stream 47	50°C, 101.3KPa stream 14		20°C, 108.8 pa stream 49	
CH4= 43.0	AA	0.0517	O2	86.86
	HAC	0.0584	N2	326.75
	H2O	139.74	H2O	7.82
	N2	1022.3		
	O2	20.98		
	CO2	69.56		
		1252.69 kmol/hr		421.43 kmol/hr

Stream 47

$$\rho_g = \frac{P}{RT} = \frac{(101.3 \text{ kPa} \times 1000 \frac{\text{Pa}}{\text{kPa}})}{(8.314 \frac{\text{Pa m}^3}{\text{mol.K}})(273.15+50)} = 41.56 \frac{\text{mol}}{\text{m}^3} \times \frac{\text{kgmol}}{1000 \text{ mol}} = 0.0416 \frac{\text{kgmol}}{\text{m}^3}$$

$$G_{47} \left(\frac{\text{m}^3}{\text{hr}} \right) = \frac{43 \frac{\text{kgmol}}{\text{hr}}}{0.0416 \frac{\text{kgmol}}{\text{hr}}} = 1034.6 \frac{\text{m}^3}{\text{hr}}$$

Stream 14

$$\rho_g = P/RT = 101.3 \text{ kPa} \times 1000 \text{ pa/kPa} / 8.314(273.15+50)$$

$$\rho_g = 37.7 \text{ mol/l} \times \text{Kg}/1000 \text{ kmol} = 0.0377 \text{ kgmol/ m}^3$$

$$G_{14}(\text{m}^3/\text{hr}) = 1252.69 \text{ kgmol/hr} / 0.0377 \text{ kgmol/ m}^3$$

$$= 33224 \text{ m}^3/\text{hr}$$

Stream 49

$$\rho_g = P/RT = 108.8 \text{ kPa} / 8.314(273.15+20)$$

$$\rho_g = 0.0444 \text{ kmol/ m}^3$$

$$G_{49}(\text{m}^3/\text{hr}) = \frac{421.43 \frac{\text{kgmol}}{\text{hr}}}{0.0444 \frac{\text{kgmol}}{\text{m}^3}} = 9484 \text{ m}^3/\text{hr}$$

$$\Sigma G = 43743 \text{ m}^3/\text{hr}$$

$$\text{Cost} = \$340177 \times 5 = \$1700878$$

10.8.8 T-101 Packed Absorber

Table 55:

Bottom stream 5			Bottom stream 7		
Gas		X1 (M.W)	Liq		X1 (M.W)
CO ₂	69.56	1.298	AA	5037.21	5.89
AA	92.99	2.838	HAC	352.92	0.343
HAC	6.88	0.175	H ₂ O	56216.76	16.426
O ₂	20.98	0.285		61606.89	22.66
N ₂	1022.3	12.133			
H ₂ O	1146.51	8.748			
	2359.22 kmol/hr	25.477 kg/kmol		61606.89 kmol/hr	22.66 kg/kmol

$$L = (61606.89 \text{ kmol/hr}) \times 22.66 \text{ kg/kmol} = 1396012 \text{ kg/hr}$$

$$G = (2359.22 \text{ kmol/hr}) \times 24.48 \text{ kg/kmol} = 57754 \text{ kg/hr}$$

$$\rho_l = \text{same as H}_2\text{O at same temp} = 988 \text{ kg/m}^3 \text{ at } 50^\circ\text{C}$$

$$\rho_g = P/RT = 760 \text{ mm hg} / 62.36(273.15+50)$$

$$\rho_g = 0.923 \text{ kg/m}^3$$

Properties of loosely packed 1" Raschig rings, 25 mm

Assume they are ceramic

$$A_p = 190 \text{ m}^2/\text{m}^3 \text{ total area of packing}$$

$$\varepsilon = 0.74 \text{ Fractional voids}$$

$$\mu H_2O @ 50^\circ\text{C} = 0.575 \text{ Cp (m Pa. s)} \quad [\text{figure 14, appendix, process heat transfer, kern}]$$

$$FLV = L/G (\rho_g / \rho_l)^{0.5}$$

$$FLV = (1396012/57754) (0.923/988)^{0.5} = 0.7388$$

$$\frac{U_T^2 A_p}{g \cdot \varepsilon^3} (\rho_g / \rho_l) \mu^{0.2} = 0.031$$

$$\frac{U_T^2 \times 190}{9.5067(0.74)^3} (0.923/988)(0.575)^{0.2} = 0.031$$

UT = 0.881 m/s

Design at 80% flood

$$\text{Area} = G/.8U\varrho g = 57754/0.8(0.881)(0.923) = 24.66\text{m}^2$$

$$\Pi r^2 = 24.66\text{m}^2 , r = 2.8\text{m}, d = 5.6\text{m}$$

Cost of tower

$$\text{Cost} = D^2(3000+3400H) + D(9800+1500H)+14000+350(3.3D)^{1.9}+ 1400(3.3D)^{1.2}$$

$$D = 5.6 \text{ m}$$

$$H = 29 \text{ m}$$

$$\text{Cost} = \$3634309 \times 5 = \$18171543$$

Combustion air fan

Strm 49 20°C, 108.8 KPa

$$O_2 = 86.86$$

$$N_2 = 326.75$$

$$H_2O = \frac{7.82}{421.43} \text{ kmol/hr}$$

$$\rho_g = P/RT = 108.8/8.314(273.15+20)$$

$$= 0.0444 \text{ kmol/m}^3$$

$$V(\text{m}^3/\text{min, inlet conditions}) = 42.143/0.0444 \times \text{hr}/60 \text{ min} = 158.2 \text{ m}^3/\text{min}$$

$$\text{Cost} = \$18097 \times 5 = \$90484$$

Cost of electricity KW = $2.72 \times 10^{-5} Q_p$

$$Q = \text{fan volume } \text{m}^3/\text{hr}$$

$$P = \text{operating pressure cm H}_2\text{O}$$

$$Q = 158.2 \text{ m}^3/\text{min} \times 60 \text{ min/hr} = 9492 \text{ m}^3/\text{hr}$$

$$P = 108.8 \text{ kPa} \times 10.333 \text{ m H}_2\text{O}/101.3 \text{kPa} \times 100 \text{cm/m} = 1109.5 \text{ cm H}_2\text{O}$$

$$KW = 2.72 \times 10^{-5}(9492 \text{ m}^3/\text{hr})(1109.5 \text{ cm H}_2\text{O})$$

$$KW = 286.45 \text{ KW} = 1031149.7 \text{ KJ/hr}$$

Cost

$$1031149.7 \text{ kj/hr} \times \$0.074/\text{KW.hr} \times 2.778 \times 10^{-7} \text{ KW.hr/J} \times 1000 \text{ J/KJ} \times 24 \text{ hrs/day} \times 365 \text{ days/yr} \\ \times 0.88$$

Electricity = \$163377/yr

Miscellaneous Cost

CH4 – Assume it is the same price as natural gas

Stream 47 = 43 kgmol/hr

$$43 \text{ kmol/hr} \times 211.2 \text{ kcal/gmol} \times 1000 \text{ gmol/kmol} \times 1\text{Kj}/0.23901 \text{ kcal} \times \$3.7/10^6\text{Kj} \times 24\text{hrs/day} \times 365\text{days/yr} \times 0.88$$

Miscellaneous cost = \$0.04 /kg of AA produced

$$\$0.04/\text{Kg AA} \times 90.355\text{kmol/hr} \times 72.05 \text{ kg/kmol} \times 24\text{hrs/day} \times 365\text{days/yr} \times 0.88$$

$$= \$2007395/\text{yr}$$

Revenue

$$\text{HAC} = 6.905\text{kmol/hr} \times 60.05 \text{ kg/kmol} \times \$0.33/\text{kg} \times 24\text{hrs/day} \times 365\text{days/yr} \times 0.88$$

$$= \$1054818/\text{yr}$$

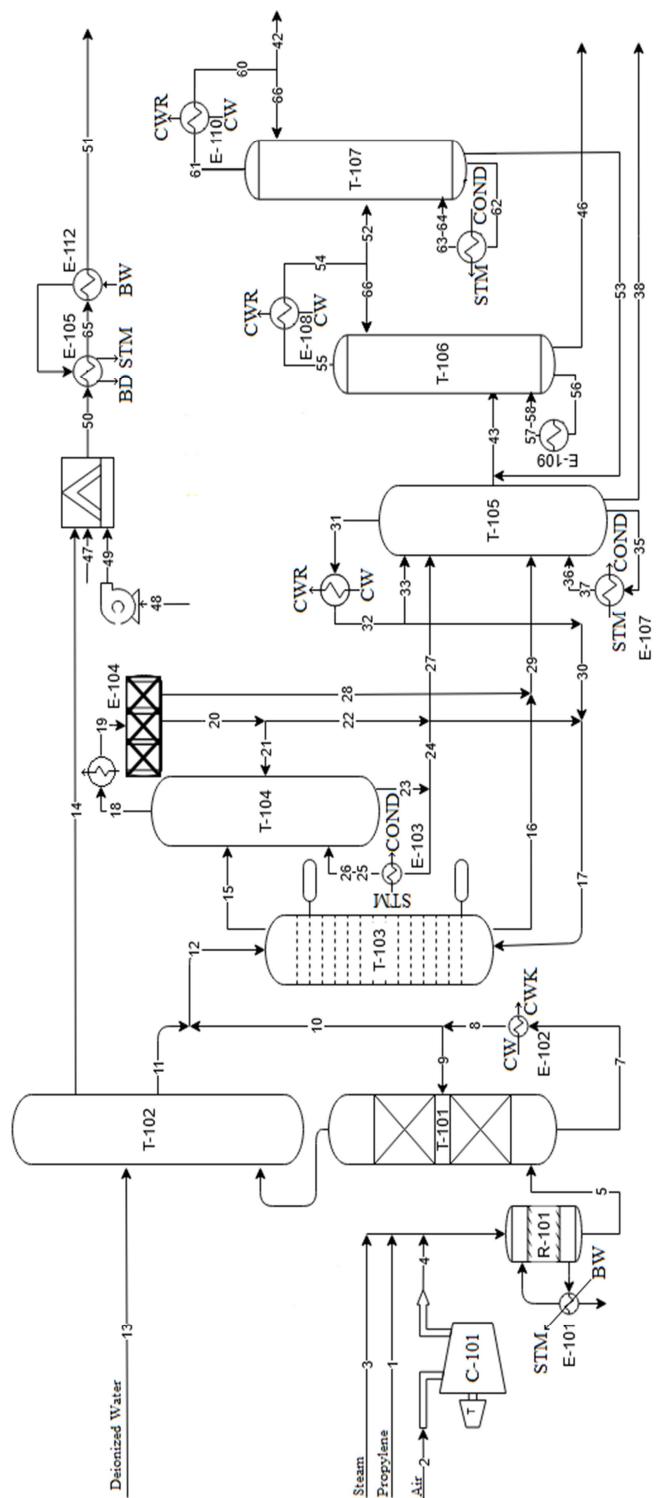
$$\text{AA} = 90.355\text{kmol/hr} \times 72.06 \text{ kg/kmol} \times \$7/\text{kg} \times 24\text{hrs/day} \times 365\text{days/yr} \times 0.88$$

$$= \$ 50191853/\text{yr}$$

10.9 APPENDIX H: Final Process Flow Diagram and Material Balance for Base Case

Description:

This section deals with the final PFD and material balance grid. Equipment which hasn't been sized such as pumps and reflux tanks will not appear on the material balance grid. The flow rates of the streams around the raffinate stripper will not appear on the grid. Since the necessary information for sizing the equipment appears in the problem statement. The material balance around the reboilers has also not been calculated because the material balance was not necessary to size any equipment or wasn't integral to the solution of the overall material balance. The energy balance will not appear on the grid because too many streams contained solvent for which there was not enthalpy or heat capacity information. The energy balance were one on all pieces of equipment to determine such things as heat exchanger duties. The pressure will appear only on streams where is known the pressure is not known for many streams because pumps were not designed or shown on the PFD. The battery condition were also not known so the temperature and the pressure of the product stream were not altered with pumps or heat exchangers.



SREAM	1	2	3	4	5	30	38	42	43	46
TEMP	30	20	155.5	105.1	310	85	100	77.2	90	90
PRESSURE	600	101.3	500	108.8	101	100.0	105.6	20.08	27.9	18.09
PHASE	V	V	V	V	V	L	L	L	L	L
SREAM	6	7	9	10	9	SREAM	47	48	49	51
TEMP	50	50	38	38	38	TEMP	20	20	700	72
PRESSURE	101	176	101	101	101	PRESSURE	101.3	101.3	251.3	101.3
PHASE	V	L	L	L	L	PHASE	V	V	V	V
SREAM	11	12	13	14	15	SREAM	52	53	54	59
TEMP	50	50	50	50	50	TEMP	58.3	90	58.3	77.2
PRESSURE	101	101	101.3	99.4	99.4	PRESSURE	4.66	27.9	4.66	5.96
PHASE	L	L	V	V	V	PHASE	L	L	L	L
SREAM	16	17	18	19	20	SREAM	60	61	65	66
TEMP						TEMP	77.2	77.2	223.8	58.3
PRESSURE						PRESSURE	20.08	21.38	175.3	4.66
PHASE						PHASE	L	V	V	L
SREAM						SREAM				
TEMP						TEMP				
PRESSURE						PRESSURE				
PHASE						PHASE				

Figure 11

10.10 APPENDIX I: ECONOMICS

10.10.1 Equipment List

Table 56

EQUIPMENT	REACTOR SIZING	DUTY (Kcal/Hr)	COST (\$/yr)
T-101(Incinerator)	G=43,743m ³ /hr		340177
R-101(reactor)	Tubes=7794		3178191
T-101(packed absorber)	5.6mD*29Dm T-T		3634309
T-102(Absorber)	2.13mD*8.83mT-T,4 Trays		77333
T-103(extractor)	11 theoretical stages		1320000
T-104(Extract stripper)	3.8mD*36.27mT-T,28 Trays		788458
T-105(Raffinate stripper)	0.6mD*7.9mT-T,10 trays		80315
T-106(AA recovery)	2.44mD*23.2mT-T,26 trays		462824
T-107(AA recovery)	0.7662mD*10.82mT-T,14 trays		121068
E-101(waste heat boiler)	S=37,642Kg/hr	23084626	350813
E-102(HEX)	A=4900m ²	14284166	1510529
E-103(Reboiler T-104)	A=100m ²	6641111	39068
E-104(condenser T-104)	A=325m ²	6641111	97546
E-105(economizer)	A=785m ²	6178817	12084
E-106(condenser T-105)	A=25m ²	756000	16484
E-107(Reboiler T-105)	A=30m ²	1638000	18223
E-108(condenser T-106)	A=110m ²	1721859	41808
E-109(Reboiler T-106)	A=25m ²	1572121	16484
E-110(condenser T-107)	A=9m ²	254971	10116
E-111(Reboiler T-107)	A=6m ²	401832	8628
E-112(Economizer)	A=210m ²	2360509	6471
		TOTAL	12809305
Misc. equipment	0.1(total equipment)		1280930
Total Purchased Equipment			14090235
Total Capital Investment	= (Total Purchased Equipment) x (Lang Factor (5))		70451175

10.10.2 Utility List

Table 57: The utility costs are as mentioned below

Utility	Equipment	Amount Used	Cost/yr
Steam,550kPa	R-101	16610.4 kg/hr	2,560,925
	E-103	13,303 kg/hr	2,051,032
	E-107	3,269.9 kg/hr	504,132
	E-109	3,138.4 kg/hr	483,857
	E-111	802.2 kg/hr	123,674
			5,734,620
Cooling water	E-102	1,428,417 kg/hr	221,329
	E-106	75,600 kg/hr	11,14
	E-108	172,186 kg/hr	26,680
	E-110	25,497 kg/hr	3,951
			763,674
Chilled water	E-104	6,664,111 kg/hr	1,783,984
Boiling water	E-101	14,249 kg/hr	275,978
	E-112	39,623 kg/hr	767,433
			1,043,411
Deionized water	T-102	4,086 kg/hr	37,987
Electricity	C-101	1,369.9 kg/hr	781,288
	C-102	286.5 kg/hr	163,377
			944,665
Natural gas	I-101	688 kg/hr	1,083,764
Credits (-)			
Steam(550kPa)	E101	37,642 kg/hr	-5803477
Steam(3500kPa)	E105	13,536 kg/hr	-2086996
	TOTAL		2990632

10.10.3 Raw Material Cost:

Table 58: The net revenue of the process is as mentioned below

Raw Material	Cost/year
Propylene	18768794
Miscellaneous chemicals	2007395
Other Costs	
Labour, benefits, taxes, etc.	1035632
Maintenance	2818047
Support charges	702686
Utilities	2990632
TOTAL (OP)	28,323,186

Revenue	
Acetic acid	1054818
Acrylic acid	47782644.06
TOTAL	48,837,462.06

Table 59: DCFROR

Year	0	1 to 9	10
Revenue	-	48837462.06	48837462.06
[OP]Operating cost(-)	-	-28323186	-28323186
Salvage(+)	-	-	1409023.5
Depreciation(-)	-	-1268121.15	-1268121.15
Taxable Income	-	19246154.91	20655178.41
Tax (40%)	-	7698461.962	8262071.362
Net Profit	-	11547692.94	12393107.04
Depreciation (+)	-	1268121.15	1268121.15
Capital investment(-)	-71296589.1	-	-
After Tax Cash Flow	-71296589.1	12815814.09	13661228.19

ROR Calculation (PW method):

$$71296589.1 = 12815814.09 \left[\frac{P}{A} \right]_{i,10 \text{ yr}} + 13661228.19 \left[\frac{P}{F} \right]_{i,10 \text{ yr}}$$

$$71296589.1 = 12815814.09 \left[\frac{(1+i)^n - 1}{i(1+i)^n} \right] + 13661228.19 \left[\frac{1}{(1+i)^{10}} \right]$$

By trial and error, we have found

$i = 0.18622$ or 18.62%

So, the project ROR is 18.62%.

