DESIGN & SIMULATTION OF SOUR GAS TREATMENT PLANT USING SULFINOL & DIPA

(A Report on Dissertation II)

Submitted as a part of course work in
M. Tech (Gas Engineering)

By
Prakash Mukhopadhyay,
B. E (*Chemical Engg.*)

Under the Guidance of

Mr. J. S. Prasad,
Associate Professor
UPES - Rajahmundry

COLLEGE OF ENGINEERING STUDIES
UNIVERSITY OF PETROLEUM AND ENERGY STUDIES
Regional Centre: Rajahmundry
Andhra Pradesh-533 294
2008-2009

DESIGN & SIMULATION OF SOUR GAS TREATMENT PLANT USING SULFINOL & DIPA

A Project report submitted to the University of Petroleum and Energy Studies-Rajahmundry campus in partial fulfillment of the requirement for the degree of

MASTER OF TECHNOLOGY (GAS ENGINEERING)

Approved by

Chair of Committee:

Committee Members:

1. External Examiner: (RUPOM MUKHERJEE).
2. Internal Examiner:
3. Internal Examiner:

Head of Department:

Centre Director:

MAY - 2009

DECLARATION

This is to declare that the project report entitled "**Design & Simulation of Sour Gas Treatment Plant using Sulfinol & DIPA**" has been prepared and submitted by me in all aspects, in partial fulfillment of the requirement for the award of the degree of **Master of Technology [Gas Engineering]** in University of Petroleum and Energy Studies-Rajahmundry.

The content of this report has not been submitted to any university or institution by me for the award of any degree or diploma.

Mukhopadhyay
12/05/09

Rajahmundry
Date:

Mr. Prakash Mukhopadhyay
R030307009
M. Tech (GE)
UPES - Rajahmundry

CERTIFICATE

This is to certify that the project work entitled "**Design & Simulation of Sour Gas Treatment Plant using Sulfinol & DIPA**" being submitted by **Mr. Prakash Mukhopadhyay** (R030307009), in partial fulfillment of the requirement for the award of the degree of **Master of Technology [Gas Engineering]** in University of Petroleum and Energy Studies-Rajahmundry, is a bonafide project work carried out by him under my guidance.

Rajahmundry
Date: 12-04-08

Mr. J. Sarada Prasad
Associate Professor
UPES - Rajahmundry

ACKNOWLEDGEMENTS

First of all I acknowledge, the grace of **God the Almighty**, in this arduous journey.

I would like to thank my **parents** for their love, care and support without which I wouldn't have been able to complete the project.

I avail this opportunity to thank **Prof. K. V. Rao, Academic Head, College of Engineering, UPES, Rajahmundry** and **Dr. K V Mohana Rao, Centre Director, UPES, Rajahmundry** for successful completion of this project.

I express my deepest sense of gratitude to **Mr. J. S. Prasad, Associate Professor, UPES, Rajahmundry** for his guidance and inspiration throughout this project. It has been an experience of great pleasure and satisfaction.

Finally I express sincere thanks to all my colleagues, who helped to complete this arduous task directly or indirectly.

Mukhopadhyay

Prakash Mukhopadhyay

EXTENDED ABSTRACT

The objective of this project is to design a sour gas treatment plant using appropriate solvent, thereby it is divided into two parts.

Part – 1 consists of natural gas with a flow rate of 60MMSCFD (2988.27 Kgmol/hr) with 9 mole% H_2S. Assuming the outlet H_2S in the Natural gas required is 4ppm then using the process selection figures, the suitable solvents for treating this gas stream are Sulfinol-D, Sulfinol-M, DIPA, MDEA, & ADIP.

Sulfinol, as known in basic form consisting of DIPA & Sulfolane is selected as the solvent for the design. The process chemistry of Sulfinol is explained and the equipments involved in the flow sheet are described.

Gas properties like molecular weight, specific gravity, compressibility and density are calculated. For initial separation of liquid (i.e., 2000 BOPD) a vertical separator is designed. Absorption tower is assumed to operate at 973 psia and 109^0F. Sulfinol circulation rate, tower diameter, weir height, Clear liquid Residence Time (CLRT), No of real trays required and shell thickness required for absorption are calculated. After absorption tower, a Flash-drum to vent off the absorbed gases by reducing the pressure to 80.85psia is sized (diameter and length of flash drum is calculated). Two heat exchangers, one for heat transfer between rich/lean solvent and the other for reducing the temperature of lean solvent (after exchanging heat with rich solvent) such that the temperature is approximately 5.5°C (10°F) above inlet gas temperature using cooling water are sized (heat transfer area required for both exchangers is calculated). Stripping tower is assumed to operate at 5 kg_f/cm^2 and 120^0F. Steam circulation rate, tower diameter, weir height, No of real trays required, and shell thickness required for absorption are calculated

Part – 2 consists of simulation of acid gas treatment plant using DIPA as a solvent on HYSYS. Natural gas 20.0786 MMSCFD (1000 kg-mole/hr) natural gas flowrate with is pressure of 1000 psia and at a temperature of 109.4°F is considered for sweetening. During the sweetening process H_2S is to be removed from that present in natural gas to tolerable specifications. In the present case H_2S is reduced from an initial composition of 0.09 to $2*10^{(-6)}$ (mole fraction basis); which is 9 % by mole in sour gas to $2*10^{(-4)}$ % by mole in the sweet gas.

CONTENTS

LIST OF FIGURES

LIST OF TABLES

1. INTRODUCTION

The process which prepares raw natural gas for the market by removing the acidic components, is called "gas sweetening". The legal requirements of natural gas before it reaches the fuel market specifies a maximum H_2S content in the range of 0.25 to 1 grain per 100 SCF (16 ppm) and a total sulphur content no more than 30 grains per 100 SCF (7000 grains = 1 lb, 635 grains per 100 SCF = 1 percent by volume). Natural gas that is transported to the fuel market usually carries a total sulphur content below one grain per 100 SCF. Specifications, such as the above, are fully justified, since hydrogen sulphide is a toxic gas and produces sulphur dioxide as a combustion byproduct.

For acid gas removal Sulfinol Process first appeared in 1965 and offered some very promising features as a gas sweetening method. Patented by Shell Oil (In Nov.2, 1964 by Michael.N.Papadopoulos et al.) and employed in many of their plants, it has shown many advantages surpassing even the conventional MEA system.

Scope - This project consists of two parts.

In part 1 natural gas of a predefined hypothetical composition with a flow rate 2988 Kgmole/hr (60 MMSCFD) is treated using Sulfinol as a solvent. Initially it is subjected to a counter current absorption with Sulfinol in a contactor operating at 973 psia and 43 °C, where solvent enters from the top and sour gas from the bottom. After absorption process sweet gas leaves the contactor from the top and Rich Sulfinol leaving from the bottom, this rich Sulfinol is passed through a valve to cause a pressure drop of 6119 kPa in the rich Sulfinol stream for subjecting it into the flash drum to remove the acid gases dissolved in Sulfinol. This process consists of using a liquid absorbent comprising 15- 45% by weight of diisopropanolamine, 1- 25% by weight of water and essentially all of the balance being at least one sulfolane (tetrahydrothiophene oxide). In another aspect a liquid absorbent containing 45- 65% by weight of diisopropanolamine, 15- 25% by weight of water and essentially remaining balance being at least one sulfolane. Still in accordance with the process , the water content of the effluent sweet gas from the absorption step is substantially reduced by coabsorbtion wih the acid gases and it is further minimized by scrubbing with a slipstream of the same mixture of solvent minus substantially all of the specified water content originally present in the mixed absorption solvent.

In part 2 HYSYS simulation is done for sweetening natural gas of same composition but with flow rate 1000 Kgmole/hr (20.0786 MMSCFD) using DIPA (Di-Iso-Propanol-Anime) which is initially subjected to a counter current absorption with DIPA in a contactor operating at 1000 psia and 43 °C, where solvent enters from the top and sour gas from the bottom. After absorption process sweet gas leaves the contactor from the top and Rich DIPA leaving from the bottom, this rich DIPA is passed through a valve to cause a pressure drop of 6088.07 kPa in the rich DIPA stream for subjecting it into the flash drum followed by a heat exchanger.

The regeneration column is getting feed at 80 psia and 200°F to strip the acid gases present in the solvent. Heat of regeneration is provided by a reboiler providing $1.431*10^{(7)}$ KJ/hr. A condenser is provided at the top of the stripper to recover the solvent vapors regenerated in the process, heat removed is $4.518*10^{(6)}$ KJ/hr. Acid gases leave stripper from the top and lean DIPA from the bottom.

A cooler is provided for cooling the lean DIPA to a temperature approximately 5.5°C above the temperature of the sour gas to the absorber, heat removed in the present case is $9.291*10^{(6)}$ KJ/hr. A pump is required to provide sufficient head to lean solvent to enter the contactor for absorption; in the present case 103.1 KW power is required.

The results are given in the datasheets of the equipments. For the present case result outputs are attached in the appendix.

Contrasted to the use of so-called "chemical solvent" (involving the formation of salts or other decomposable reaction products), certain classes of chemicals are regarded as "physical solvents". These includes such chemicals as glycols which appear to act in a purely physical manner, absorbing acid gasses physically without the formation of any apparent reaction product. Mixtures of glycols with amines are known, particularly for treatment of liquid hydrocarbon products or the removal of mercaptans, H_2S and other acidic substance.

Limitations --Aqueous dispertion of glycols are no better in this respect and , moreover, may exhibit extreme corrosiveness toward processing equipment. One of the problems encountered, especilly in the processing of gases initially having high partial pressure of acid components, is efficient removal of entire acid component content both at the time the gas contains relatively high partial pressure of acid gas and after removal of the major proportion, i.e, at the time when the gas contains acid gases only at a relatively low partial pressure. None of the absorbents are efficient in both respects.

Another disadvantage of the use of aqueous solutions is the necessity for a separate dehydration step, since such absorbents donot selectively remove water at the same time they absorb acidic gases such as hydrogen sulfide. Since water cannot be tolerated in almost all instances in the sweet gas stream, its removal constitutes an additional cost factor in the processing of gases. A number of problems have arisen in addition to those outlined above. The thermal stability of organic solvents becomes a problem especially during the stripping step wherein the fat solvent is heated to a sufficient temperature to cause a separation of acid gases from the solvent. Under these conditions it is often found that many solvents show undue decomposition constituting a major cost factor in the operation of the process. Another major problem is the differential in temperature between that of the absorption column and in the stripping column. While a large part of this differential can be readily handeled by heat exchanging the various stream involved in the process, nonetheless, it is often necessary to utilise external cooling means to cool the recycle streams. of clean regenerated solvent to the absorber, as well as to employ heating means for heating the fat solvent coming from the bottom of the absorber column to the stripping column.

2. **SULFINOL** (Material Property Sheet) [1]

Diisopropanolamine (DIPA) is secondary amines, which like the other amines, is used to treat acid gas streams. The major advantage, which offers over other amine processes, is its selectivity for CO_2 in the presence of H_2S. If the gas is contacted at pressures ranging from 5500 to 6900 kPa (800 to 1000 psig), H_2S levels can be reduced to the very low concentrations required by pipelines, while at the same time 40 to 60 percent of the CO_2 present flows through the contactor, un-reacted.

		Diisopropanol Amine	**Sulfolane**
	Mol. Formulae	$(HOC_3H_4)_2NH$	$C_4H_8SO_2$
	Mol.wt	133.19	120.7
	Boiling pt. °C	248.7	285
	Freezing pt. °C	42	27.6
Critical constants	Pressure, psia	546.3	767.3
	Temperature. °C	399.2	545.4
	Specific gravity at 20 °C	0.989	1.288
	Specific heat at 30 °C, Btu/lb	0.69	0.35
	Latent heat of vaporization, Btu/lb	185	225.7
	Viscosity at, cp	870, 198, 86 At 30,45,54 °C	10.3, 6.1, 2.5, 1.4, .97 At 30,50,100,150,200 °C
	Refractive Index,	1.4542 at 45 °C	1.481 at 30 °C
	Flash pt. °C	123.8	176.6

Table 1 : (DIPA, Sulfolane properties) [1]

3. LITRATURE SURVEY --- PHYSICAL PROPERTIES OF SOLVENTS [1]

Monoethanolamine

Aqueous monoethanolamine solutions, which were used almost exclusively for many years for the removal of H_2S and $C0_2$ from natural and certain synthesis gases. are rapidly being replaced by other more efficient systems. particularly for the treatment of high-pressure natural gases. However, monoethanolamine is still the preferred solvent for gas streams containing relatively low concentrations of H_2S and CO: and essentially no minor contaminants such as COS and CS_2. This is especially true when the gas is to be treated at low pressures, and maximum removal of H_2S and CO_2 is required. The low molecular weight of monoethanolamine, resulting in high solution capacity at moderate concentrations (on a weight basis), its high alkality, and the relative ease with which it can be reclaimed from contaminated solutions are advantages, which in many cases more than counterbalance inherent disadvantages. Among the latter, the most serious one is the formation of irreversible reaction products with COS and CS_2, resulting in excessive chemical losses if the gas contains significant amounts of these compounds.

Diethanolamine

Aqueous solutions of diethanolamine (DEA) have been used for many years for the treatment of refinery gases which n o d y contain appreciable amounts of COS and CS_2, besidesH2S and CO. Secondary amines are much less reactive with COS and CS_2 than primary amines, and the reaction products are not particularly corrosive. Consequently, diethanolamine and other secondary amines are the better choice for treating gas streams containing COS and CS_2. The low vapor pressure of diethanolamine makes it suitable for low-pressure operations as vaporization losses are quite negligible. One disadvantage of diethanolamine solutions is that the reclaiming of contaminated solutions may require vacuum distillation. Another disadvantage of DEA is that DEA undergoes numerous irreversible reactions with CO_2, forming corrosive degradation products, and for that -on, DEA may not be the optimum choice for treating gascs with a high C02 content.

Diglycolamine

The use of aqueous solution of Diglycolamine, 2-(2-aminoethoxy) ethanol, was commercialized jointly by the Fluor Corporation (now Fluor Daniel), the El Paso Natural Gas Company, and the Jefferson Chemical Company Inc. (now the Huntsman Corporation) (Holder, 1966; Dingman and Moore, 1968). The process employing this solvent has been named the Fluor Econamine. process. The solvent is in many respects similar to monoethanolamine, except that its low vapor pressure permits its use in relatively high concentrations, typically 40 to 609, resulting in appreciably lower circulation rates and steam consumption when compared to typical monoethanolamine solutions.

Diisopropanolamine

Diisopropanolamine (DIPA) has been used in the ADIP and Sulfinol processes, both licensed by the Shell International Petroleum Company (SIPM). In the Sulfinol process, diisopropanolamine is used in conjunction with a physical organic solvent. The ADP process, which employs relatively concentrated aqueous solutions of diisopropanolamine, has been described by Bally (1961) and by Klein (1970). It has been widely accepted. primarily in Europe, for the treatment of refinery gases and liquids which, besides H,S and CO_2, also contain COS. It is claimed that substantial amounts of COS are removed without detrimental effects to the solution. A theoretical study of the absorption kinetics hV0hed in the selective absorption of H_2S in DIPA has been presented by Ouwerkerk (1978). Equations for mass transfer with chemical reaction are utilized in the study to develop a computer program which takes into account the competition between H_2S and $C0_2$ when absorbed simultaneously.

Methyldiethanolamine

Selective absorption of hydrogen sulfide in the presence of carbon dioxide, especially in cases where the ratio of carbon dioxide to hydrogen sulfide is very high, has recently become the subject of considerable interest, particularly in the purification of non-hydrocarbon gases such as the products from coal gasification processes and Claus plant tail gas. The early work at the Fluor Corp. (now Fluor Daniel) showed that tertiary amines, especially methyldiethanolamine, can absorb hydrogen sulfide reasonably selectively under proper operating conditions involving short contact times (Frazier and Kohl, 1950; Kohl, 1951; Miller and Kohl, 1953). -4 study by Vidaurri and Kahre j1977), in which selective absorption with several ethanolamines was investigated in a pilot and commercial plant, demonstrated that purified gas containing as little as 5 parts per million of hydrogen sulfide could be obtained with absorption of only about 308 of the carbon dioxide contained in the feed gas.

Mixed Amines (Sulfinol)

The Sulfinol solvent consists of sulfolane (tetrahydrothiophene dioxide) and an alkanolamine, usually diisopropanolamine (DIPA) or methyldiethanolamine (MDEA), and water. The solvent with DIPA is referred to as Sulfinol-D or simply as Sulfinol, and the solvent with MDEA is referred to as Sulfinol-M. Typically, Sulfinol-D is used when essentially complete removal of both hydrogen sulfide and carbon dioxide and deep removal of carbonyl sulfide is desired. Sulfinol-M is used for the selective removal of hydrogen sulfide over carbon dioxide and the partial removal of carbonyl sulfide (Nasir, 1990). Both Sulfinol solvents are reported to be capable of removing mercaptans and alkyl sulfides to very low levels. The Sulfinol process is licensed by the Shell Oil Company in the U.S. and by Shell International Petroleum Maatschappij (SIPM) in the Netherlands. The process has found wide application in the treatment of natural, refinery, and synthesis gases. The Sulfinol process can meet the requirement for deep C02 removal to 50 ppm for LNG plants, as well as the opposite extreme of bulk C02 removal using flash regeneration.

Property	MEA*	DEA*	TEA*	MDEA*	DIPA*	DGA**
Mol wt	61.09	105.14	149.19	119.17	133.19	105.14
Sp-gravity at (20/20^{0}C)	1.0179	1.0919 30/20^{0}C	1.1258	1.0418	0.9890 45/20^{0}C	1.055
Boiling Point(BP),^{0}C						
BP at 760 .mmHg	171	Decomp	360	247.2	248.7	221
BP at 50 .mmHg	100	187	244	164	167	---
BP at 10 .mmHg	69	150	208	128	133	---
Vapor Pressure (.mmHg) at 20^{0}C	0.36	0.01	0.01	0.01	0.01	0.01
Freezing point, ^{0}C	10.5	28.0	21.2	-21.2	42	-9.5
Solubility in water % by wt at 20^{0}C	Complete	96.4	Complete	Complete	87	Complete
Absolute Viscosity Centipoises at 20^{0}C	24.1	380 (30^{0}C)	1,013	101	198 (45^{0}C)	26 (24^{0}C)
Heat of vaporization Btu/lb at 1 atm	355	288 (168.5^{0}C)	230	223	184.5	219.1
Approximate cost $ / lb ***	0.59	0.60	0.61	1.40	---	0.93
MEA: Monoethanolamine; DEA: Diethanolamine TEA: Triethanolamine; MDEA: Methyldiethanolamine DIPA: Diisopropanolamine; DGA: Diglycolamine						

Table 2 : (Physical properties of solvents) [1]

4. METHODOLOGY -- PROCESS SELECTION [8]

Each of the acid gas treating processes has advantages relative to the others for certain applications; therefore, in selection of the appropriate process, the following facts should be considered:

I. Acid contaminant type present in gas stream.

II. The degree of removal required for every concentrations of each contaminant.

III. Amounting treatment required for volume of gas supplied with consideration of temperature and pressure at which the gas is available.

IV. Feasibility amounting for sulfur recovery.

V. The Selectively is desirable for removing one or more of the contaminants without removing the others.

VI. The presence and quantity of heavy lower & higher hydrocarbons and in the gas

VII. The environmental conditions required at the plant site.

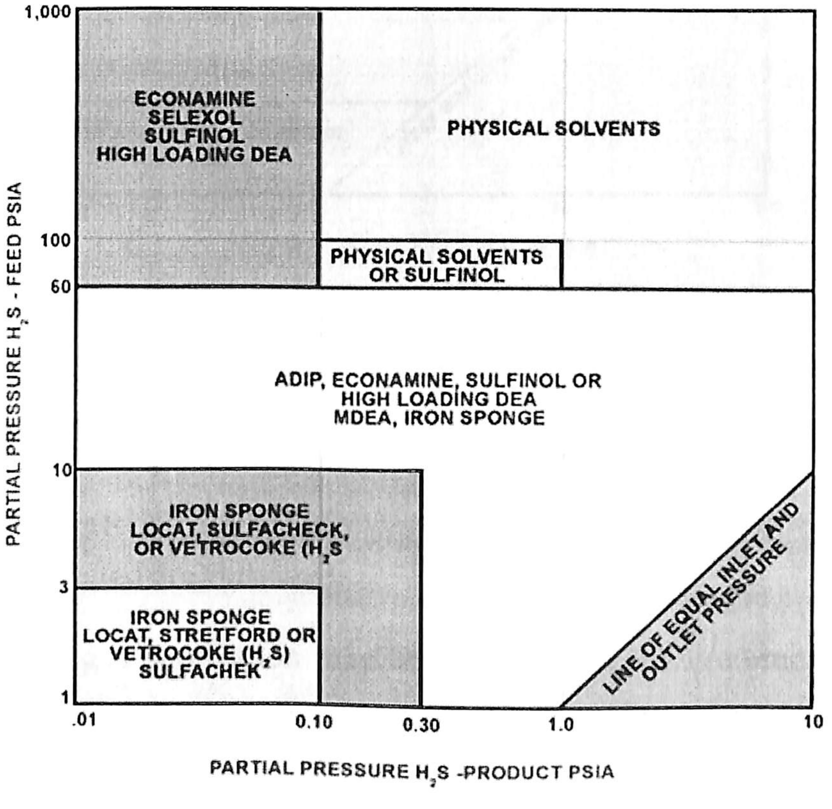
Case 1:- **Removal of H_2S**

For feeds with small acid gas loadings, one of the batch processes should be considered for removal of H_2S. The most common are: iron sponge, Sulfa-Treat, and Sulfa-Check.

As acid gas loadings increase, the disposal and replacement costs become high, and it becomes necessary to choose a process that can be regenerated. The amine systems are most often used for these installation systems. DEA is the most common amine system.

The end product of an amine system is an acid gas stream off the stripper, which must be flared. As acid gas loadings increase, environmental constraints require that this acid gas stream be converted to sulfur. One of the processes that converts acid gas to sulfur, such as LO-CAT®, SulFerox®, Claus, or Stretford, must be added downstream of the amine system. In some cases, it may be feasible to contact the gas stream to be treated directly with LO-CAT® or SulFerox® solution and eliminate the need to separate the acid gas components from the gas stream with an amine unit.

When a Claus unit is used, it may be necessary to add tail gas cleanup downstream of the Claus unit if acid gas loadings are very high.



Figures 1 : Selective removal of H_2S in absence of CO_2 [8]

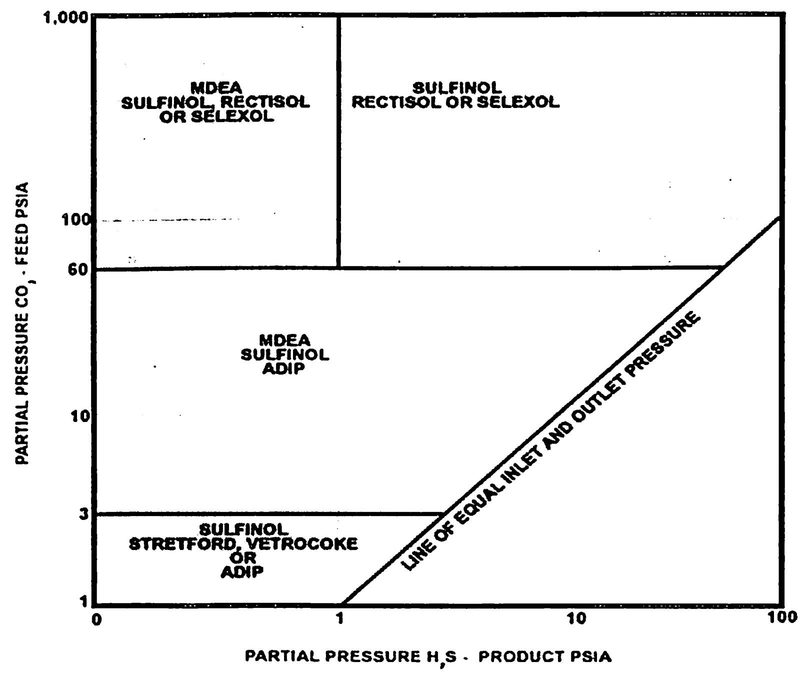


Figure 2 : Selective removal of H2S in presence of CO_2.[8]

Case 2:- **Removal of CO_2**

Removal of CO_2 to meet pipeline quality specifications can be accomplished with an amine-based system since the acid gas from the stripper can be vented (assuming levels of H_2S in the gas being treated are very low).

Removal of CO_2 with gas permeation may be attractive for low volume gas streams in remote areas where the loss of methane is not critical. Permeation systems with a second stage recycle may be competitive with amine systems.

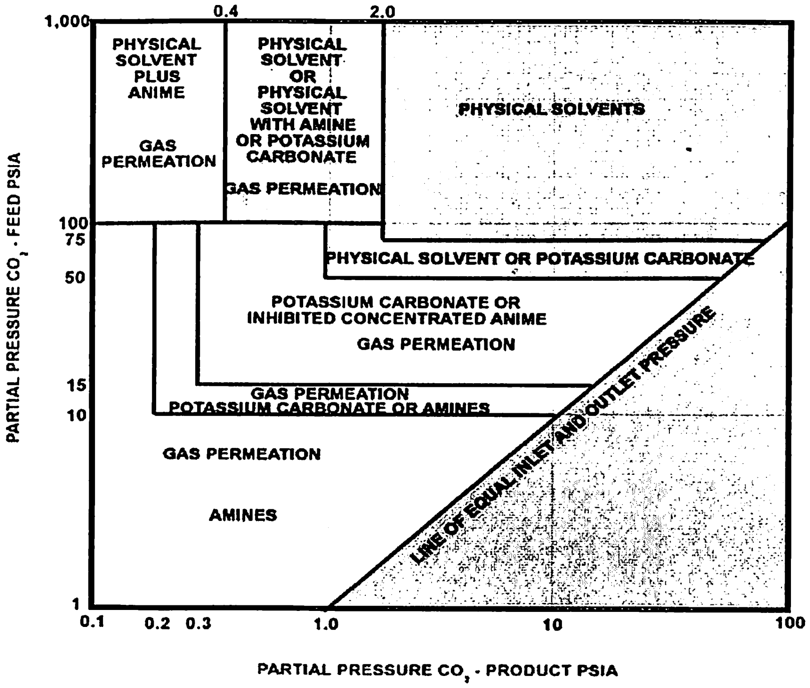


Figure 3: Selective removal of CO_2 in absence of H_2S [8]

Case 3:- **Removal of H_2S and CO_2**

Most commonly, both H_2S and CO_2 are present and must be removed to meet pipeline quality requirements. In most cases, essentially all of the H_2S will have to be removed, while only a fraction of the CO_2 will have to be removed. Use of a non-selective solvent such as MEA or DEA will require that the equipment be sized to essentially remove all the CO_2 so that the H_2S specification can be achieved. This procedure may be the most economical solution for streams with low CO_2 concentrations.

As CO_2 concentrations in the feed increase, it becomes more economical to use a selective process such as MDEA, Sulfinol®, Selexol®, etc., which will remove a higher percentage of H_2S than CO_2 from a stream.

Another alternative is to use gas permeation or a carbonate system for bulk removal of CO_2 upstream of a non-selective amine unit.

Finally, it may be economical to remove both H_2S and CO_2 to a level where the CO_2 content is acceptable with either a selective or non-selective process, and use a sulfur removal process (iron sponge, Sulfa-Treat, Sulfa-Check, LO-CAT®, SulFerox®) for final treating of the residue gas.

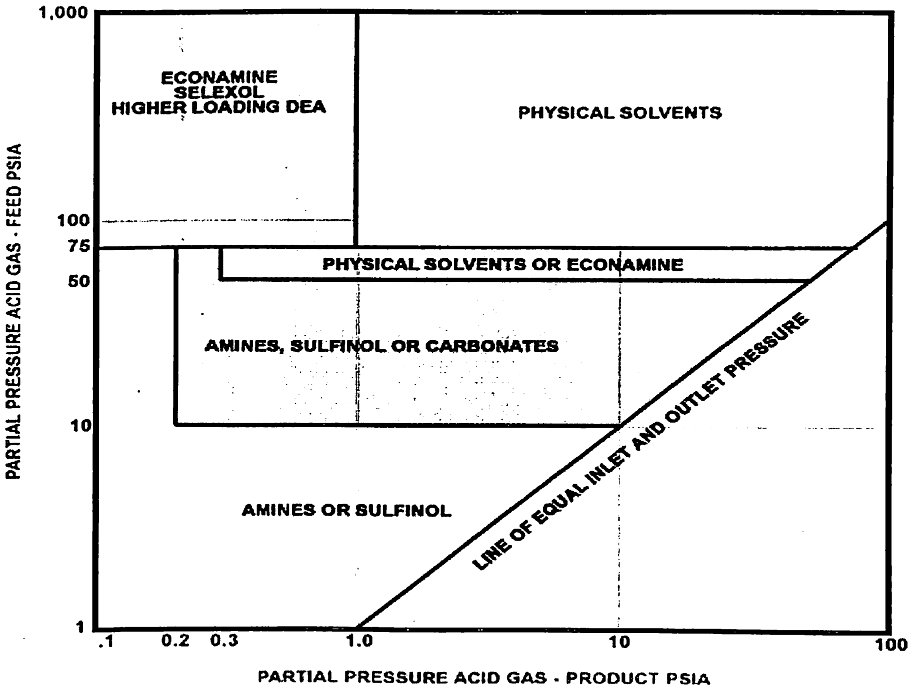


Figure 4 : Simultaneous removal of H_2S and CO_2 [8]

Note :-*These graphs are not meant to supplement engineering judgment nor to cover every possible contingency. New processes are continuously being developed. Modifications to existing proprietary products will change their range of applicability and relative cost. The graphs do enable a first choice of several potential candidates, which could be investigated to determine which is the most economical for a given set of conditions.*

5. METHODOLOGY -- CHOOSING SULFINOL

DESIGN CASE DATA:

Flow rate of gas = 60 MMSCFD

Pressure = 973 psia

Temperature = 109.4°F

H2S content = 9 mole %

PROCESS SELECTION STEPS

Step 1:

Calculate the following:

$$\text{Partial Pressure of } H_2S \text{ in feed} = 973 * 0.09 = 87.569 \text{ psi}$$

$$\text{Partial Pressure of } H_2S \text{ in product} = 973 * 4 * 10^{-6} \text{ psi} = 0.00389 \text{ psi}$$

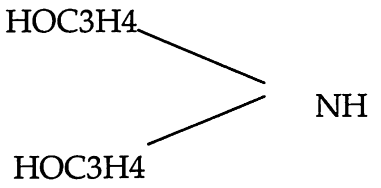
Step 2:

From the appropriate graph i.e fig.1 select the process to be employed

In the present case at pressure near 1000 psig & high H_2S partial pressure, figure shows that Sulfinol, Selexol, Economine or high loading DEA process can be employed.

6. METHODOLOGY -- PROCESS CHEMISTRY OF SULFINOL [21]

Structural formula of DIPA



H_2S & CO_2 chemically reacts with DIPA almost instantaneously by proton transfer as is the case with other commonly used amines.

$RR'\,NH + H_2S \longrightarrow RR'NH_2 + HS$ --------(1)

$2\,RR'\,NH + CO_2 \longrightarrow RR'\,NH_2 + RR'\,NCOO$ --------(2)

$R_3N + H2O + CO_2 \longrightarrow R_3\,NH + HCO_3$ --------(3)

Therefore there is an advantage to be gained by employing 1-25% by weight of water, 15-45% by weight of DIPA remaining sulfolane. This enables DIPA content to increase to 45-65% with the remainder, i.e, 20-40% by weight of sulfolane, this gives a number of desirable physical property changes that occur which are especially effective at enhancing the process economy. For example

1) Positive effect on viscosity. [10]

Solvent Composition		Viscosity, cp	
Percent wt. DIPA	Percent wt. H_2O	Lean Solvent	H_2S loaded solvent
40	10	14.2	45
52	10	26	122
52	25	14.2	45

Table 3 :- (Effect of solvent composition on viscosity)

2) Enhanced sweetening. [10]

Percent H2S in feed gas	S.C.F of feed gas sweetened per gallon of solvent			
	40% wt. DIPA		52% wt. DIPA	
	10% wt. H_2O	25% wt. H_2O	10% wt. H_2O	25% wt. H_2O
5 -----	125	122	158	154
25 ----	53	52	67	65

Table 4 :- (Solvent circulation advantage as change of water percent)

1) Hydrocarbon coabsorbtion reduction [10]

	DIPA / H_2O = 40/10		DIPA / H_2O = 52/25	
	K	Percent H.C Co absorbed	K	Percent H.C Co absorbed
Methane	41	1.2	209	0.24
Ethane	28	1.7	133	0.37
Propane	18	2.7	84	0.59
n-Butane	11	4.3	50	0.99
n-Pentane	6.7	7	33	1.55

Table 5 :- (Hydrocarbon rejection advantage by water percent change)

3) Increase in heat transfer coefficient [10]

Equipment	H_2O percent wt.		
	10	20	25
Lean/ Rich exchanger	70	102	119
Solvent Coolers	37	85	94
Reboilers	131	107	207

Table 6 :- (Heat transfer advantage by water percentage change)

7. SULFINOL AS THE SOLVENT (Solubility comparison) [14]

There was a test done by Combustion Engineering Inc. for the solubility of H_2S in sulfinol , were in steps to the test procedure are;

The flow rate of liquid was established as well as regeneration temperatures. The flow rate was established at 10.7 ml/min., and the temperatures at 250 F for the reboiler, 220 F for outlet of U-tube preheated, and 150 F for the condenser;

Pressure at the absorber vessel was established at either 200, 500 or 800 psig and the test gas mixture containing 10% H_2S, 10% CO_2 & 80% CH_4 was flowed into the absorber vessel at the established pressure;

The test gas flow rate was increased until the gas leaving the absorber vessel reached ¼ gr/ 100 scf. The following tabulation was made of the test results;

	Solution Composition wt %	Absorber Pressure psig	Sorbent Flow Rate mol/hr	Pickup Acid Gas Flow Rate, mol/hr	mol Acid Gas / mol Sorbent
MDEA	50	500	4.04	9.75	0.09
Sulfolane	25	800	4.04	2.55	0.13
H_2O	25				
DIPA	50	200	3.75	0.78	0.21
Sulfolane	25	500	3.75	1.04	0.28
H_2O	25	800	3.75	1.31	0.35

Table 7: (Sulfinol M & Sulfinol D comparison) [14]

8. SCHEMATIC DIAGRAM OF SULFINOL TREATMENT SYSTEM

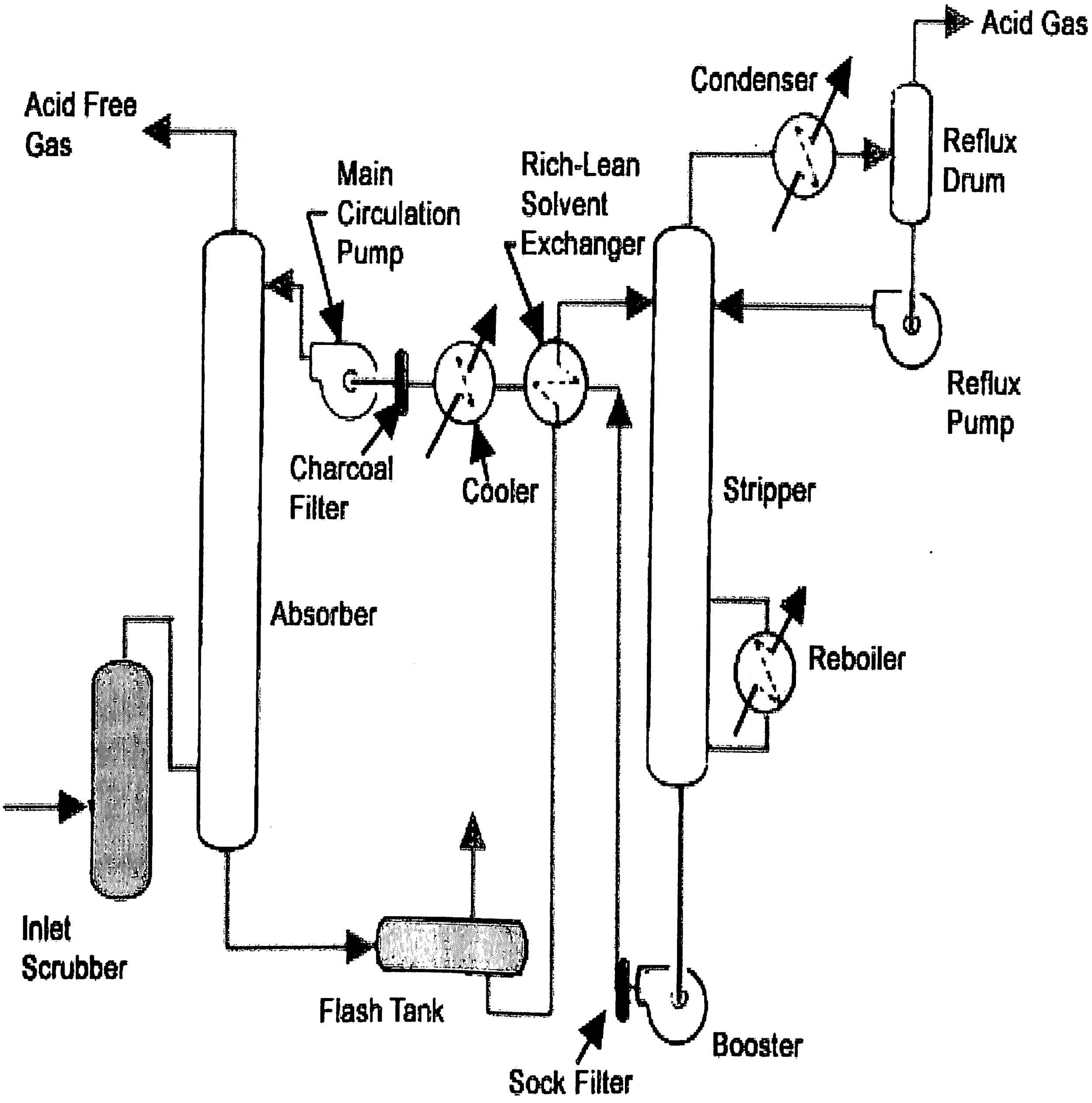


Figure: 5 Schematic Diagram of Typical Natural Gas Treatment System [9]

9. DESCRIPTION OF THE FLOW SHEET

The sour gas enters the system through an inlet scrubber to remove any entrained water or hydrocarbon liquids. Then the gas enters the bottom of the amine absorber and flows counter-currently to the amine solution. The absorber can be either a tray or packed tower with conventional packing usually used for 500.mm (20 in) or smaller diameter towers, and trays or structured packing for larger towers. The sweetened gas then leaves the top of the tower. An optional outlet scrubber may be included to recover entrained amine from the sweet gas. Since the natural gas leaving the top of the tower is saturated with water, the gas will require dehydration before entering a pipeline.

The sulfinol solution after countercurrent contact with the gas leaves at the bottom of the absorber. This solution containing the CO_2 and/or H_2S is referred to as the rich solvent. From the absorber, the rich sulfinol flows to the flash tank to remove most of the dissolved hydrocarbon gases or entrained hydrocarbon condensates and also some composition of acid gases absorbed. From the flash tank the rich sulfinol proceeds to the rich sulfinol /lean sulfinol heat exchanger. This heat exchanger recovers some of the sensible heat from the lean sulfinol stream to decrease the heat duty on the sulfinol reboiler and the solvent cooler. The preheated rich sulfinol then enters the sulfinol stripping tower where heat from the reboiler breaks the bonds between the sulfinol and acid gases. The acid gases are removed overhead and lean sulfinol is removed from the bottom of the stripper.

The hot lean sulfinol flows to the rich sulfinol /lean sulfinol heat exchanger and then to additional coolers, typically aerial coolers, to lower its temperature to approximately 5.5°C (10°F) above the inlet gas temperature. This reduces the amount of hydrocarbons condensed in the sulfinol solution when the sulfinol contacts the sour gas. Typically, a side stream of approximately 3 percent of the sulfinol flow rate is taken off after the rich/lean sulfinol heat exchanger and is flowed through a charcoal filter to clean the solution of contaminants. The cooled lean sulfinol is then pumped up to the absorber pressure and enters the top of the absorber. As the sulfinol solution flows down the absorber it absorbs the acid gases. The rich sulfinol is then removed at the bottom of the tower and the cycle is repeated.

The filters are provided to prevent suspended solids or formed crystals to enter the pump casing. Otherwise the pump casing is prone to damage because of suspended solids.

1) The most common amine processes are monoethanolamine (MEA) and diethanolamine (DEA). Both processes will remove CO_2 and H_2S to pipeline specifications. Among the newer processes, which have been developed are sulfinol, selexol, methyldiethanolamine (MDEA), DIPA. But from the sight of new researches some new solvents having more solubility than sulfinol are found like 4-propanol pyridine, which when used instead of sulfolane gives drastic figures.

	Solution Composition wt %	Absorber Pressure psig	Sorbent Flow Rate mol/hr	Pickup Acid Gas Flow Rate, mol/hr	mol Acid Gas / mol Sorbent
MDEA	50	200	3.86	0.41	0.11
4 PP	25	500	3.86	5.65	0.29
H_2O	25	800	3.86	7.75	0.4
DIPA	50	200	3.57	1.31	0.37
4 PP	25	500	3.57	1.83	0.51
H_2O	25	800	3.57	2.09	0.59

Table 8: (Experimental outcomes for sulfolane replacement) [14]

10. MASS TRANSFER FUNDAMENTALS [21]

The separation is a rate process rather than one dominated by phase equilibrium. An appreciation of the fact that all alkaline solvents are thermodynamically selective for CO_2 but kinetically selective for H_2S is vital to understanding the importance of mass transfer rates to contactor performance.

Sulfinol being a mixed solvent absorbs by both chemical and physical means, therefore because of physical absorption the regeneration of the solvent becomes easy. On the other hand due to chemical absorption less hydrocarbon loading occurs. Due to this duel characteristics the equilibrium solubility of acid gas depends both on the reaction kinetics with DIPA and as well physical absorption by sulfolane. This implies a duel approach for acid gas mass transfer basics also.

For a given lean-solvent acid-gas loadings a high enough tray count or a deep enough packed bed guarantees that the treated gas leaves the contactor in equilibrium with the lean solvent (or for low solvent rates that the rich solvent leaves in equilibrium with the sour gas). However, as the tray count is reduced (or the bed shortened) the treated gas becomes further and further removed from equilibrium. The thermodynamics of acid gas-amine systems is such that CO_2 is the preferred solute because it absorbs by forming a fairly stable chemical bond with the amine. However, the CO_2-amine reaction is of finite rate and, in fact, is quite good in DIPA.

On the other hand, H_2S ionizes instantaneously (to bi-sulfide ion); it does not react with the amine at all, it forms no chemical bonds, and the ionization reaction is immediately reversible. Thus, the chemical reaction kinetics is much faster for H_2S; therefore, CO_2 absorbs more slowly. At short contact times (read low interfacial areas, small tray counts, short packed beds) H_2S absorbs at a higher rate than CO_2, and so H_2S is preferentially absorbed. At long contact times (high interfacial areas, many trays, deep beds), CO_2 absorbs more completely, albeit more slowly, and CO_2 is preferentially absorbed.

Thus, control over selectivity can be achieved by choosing an amine (or a multiple amine mixture) with the right reactivity toward CO2 allowing contact in a column with the right number of trays or the right depth of packing, and choosing the kind of column internals that favor either CO_2 or H_2S absorption. Selectivity depends on rates-not just reaction rates, but mass transfer rates-which implies dependence on all the factors that affect the mass transfer characteristics and mass transfer performance of the actual physical hardware in which the process is carried out. Equilibrium stage models simply cannot capture these effects. The currency of equilibrium stage models is the number of theoretical stages the currency of internals vendors and gas processors is actual tray counts, types, and asses and volumes and depths of packing of specified size, type and material.

With ideal stages, translation between the two is forever an open question. A true mass transfer rate model, on the other hand, always deals in real trays and real packing there is never a question about how many trays are needed or what depth of packing to install. It is equally important to be able to model solvent regeneration accurately if for no other reason than the fact that the loading of the lean solvent produced by the stripper directly and significantly affects contactor performance. Not only does it affect its ability to meet treated gas specifications, but also the actual treated gas composition. Equilibrium stage models don't work very well here either because the reactions and the tower internals type and details affect mass transfer in just as important ways as in absorption. None of the trays in a stripper come even close to an equilibrium stage, and the desorption rate of each acid gas affects the rate of the other. From a technical standpoint, this mass transfer rate based stripper model treats regenerators every bit as rigorously as absorbers. It produces the best possible predictions of regenerator performance without the need for empirical adjustment. And when absorber and stripper models are tied together in a recycle flow sheet, the best possible prediction of treating plant performance is obtained without applying user-supplied or internally generated empirical corrections of any kind. This complete freedom from empiricism allows the engineer to design and predict the

Performance of new facilities for which absolutely no operating data or field experience exists.

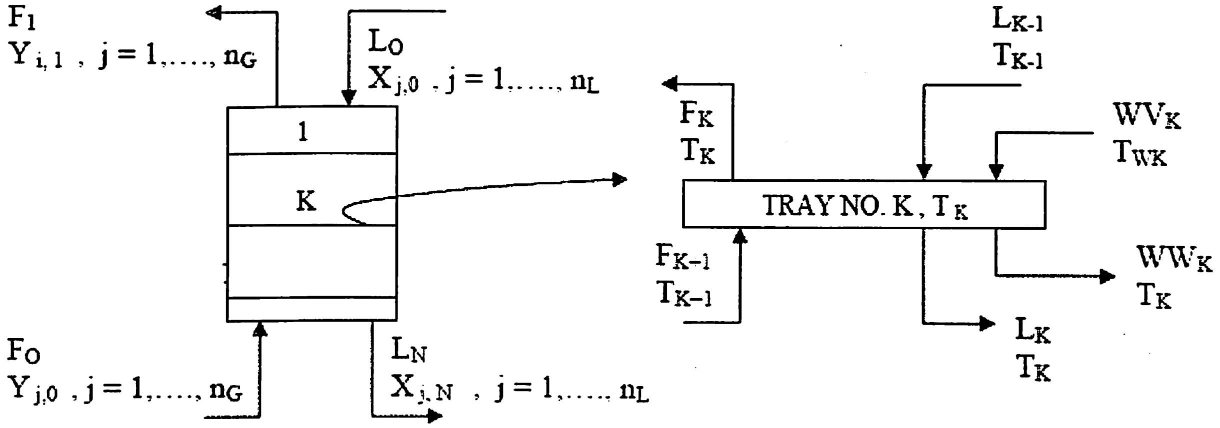


Fig 6 : (Mass transfer in tray)

Material balance:

$$F\frac{dYa}{dZ} = \{-Na : x = 0 + Ya\,[Na : x = 0 + Nb : x = 0]\}APV\;Aa \qquad \text{----------} \qquad 10.1$$

$$F\frac{dYb}{dZ} = \{-Nb : x = 0 + Yb\,[Na : x = 0 + Nb : x = 0]\}APV\;Aa \qquad \text{----------} \qquad 10.2$$

$$\frac{dF}{dZ} = \{-Na : x = 0 + [Nb : x = 0]\}APV\;Aa \qquad \text{----------} \qquad 10.3$$

Non-absorber material will have a constant contrition in tower, therefore will have

DESIGN CALCULATIONS

11. DESIGN CALCULATIONS

General Design Decisions

The designer is required to consider and determine:

1. Entering gas (liquid) flow rate, composition, temperature, and pressure
2. Desired degree of recovery of one or more solutes
3. Choice of absorbent
4. Operating pressure and temperature
5. Minimum absorbent (stripping agent) flow rate and actual absorbent (stripping agent) flow rate as a multiple of the minimum rate needed to make the separation.
6. Number of equilibrium stages
7. Heat effects and need for cooling
8. Type of absorber (stripper)
9. Height and diameter of absorber (stripper)

Entering gas/liquid flow rates and composition

Entering gas composition and flow rates are generally set from the preceding unit operation, the flash drum in our case. The pressure and temperature are also set from this flash calculation.

INPUT DATA

Gas Composition	mol frac
C_1	0.73500
C_2	0.08300
C_3	0.02100
i-C_4	0.00600
n-C_4	0.00200
i-C_5	0.00300

n-C_5	0.00800
C_6	0.00100
C_{7+}	0.00100
N_2	0.05000
H_2S	0.09000
	1.00000

Determination of operating pressure and temperature

a) In general, operating pressure should be high and temperature low for an absorber, to minimize stage requirements and/or absorbent flow rate and to lower the equipment volume required to accommodate the gas flow.

b) Operating pressure should be low and temperature high for a stripper to minimize stage requirements or stripping agent flow rate.

Pressure: (psia) -------	972.99	
TEMPERATURE(^{o}F)---	109.4	
oil API--------------------	40	
Dm = (micron)----------	140	Particle Dia
Qg = (MMSCFD)------	60	Gas Flow
Ql = (BOPD)------------	2000	Liquid flow
t = s-----------------------	180	Retention time

y2, Outlet N.G purity required, H_2S = 4 ppm

Sulfinol composition used (by .wt)

DIPA = 40 %

Sulfolane = 40 %

Water = 20 %

H_2S = 0 % Assumption

Molecular wt.	DIPA =	133.19	
	Sulfolane =	120.7	
	Water =	18	
	H_2S =	34	
Density	DIPA =	989	kg/m3
	Sulfolane =	1288	kg/m3
	Water =	1000	kg/m3
	H_2S =	102.52913	kg/m3
Viscosity	DIPA =	198	cp
	Sulfolane =	6.1	cp
	Water =	1	cp
	H_2S =	0.3	cp
Specific heat	DIPA =	0.69	Btu/lb °F
	Sulfolane =	0.35	Btu/lb °F
	Water =	1.8013127	Btu/lb °F
	H_2S =	0.23839	Btu/lb °F

CALCULATION :

Solvent Properties

		mol. Fraction
DIPA	=	0.172
Sulfolane	=	0.190
Water	=	0.637
H_2S	=	0

Avg	Viscosity of sulfinol	=	36.916	cp
Avg	Specific heat	=	0.430	kcal/kg °C
Avg	Molecular wt. of sulfinol	=	57.377	
Avg	Density of sulfinol	=	1110.8	kg/m3

SPESIFIC GRAVITY - Gas-specific gravity (yg) is defined as the ratio of the apparent molecular weight of a natural gas to that of air, itself a mixture of gases. The molecular weight of air is usually taken as equal to 28.97 (approximately 79% nitrogen and 21% oxygen).

VISCOSITY - Gas viscosity is a measure of the resistance to flow exerted by the gas. Dynamic viscosity (jug) in centipoises (cp) is usually used in the natural Engineering. Gas viscosity is very often estimated with charts or correlations developed based on the charts. The gas viscosity correlation of Carr, Kobayashi, and Burrows (1954) involves a two-step procedure which is executed bellow. The gas viscosity at temperature and atmospheric pressure is estimated first from

gas-specific gravity and inorganic compound content. The atmospheric value is then adjusted to pressure conditions by means of a correction factor on the basis of reduced temperature and pressure state of the gas.

Gas Properties							
Compound	y_i	MW_i	y_iMW_i	p_{ci} (psia)	y_ip_{ci} (psia)	T_{ci}, (°R)	y_iT_{ci} (°R)
C_1	0.735	16.04	11.79	673	494.66	344	252.84
C_2	0.083	30.07	2.50	709	58.85	550	45.65
C_3	0.021	44.10	0.93	618	12.98	666	13.99
i-C_4	0.006	58.12	0.35	530	3.18	733	4.40
n-C_4	0.002	58.12	0.12	551	1.10	766	1.53
i-C_5	0.003	72.15	0.22	482	1.45	830	2.49
n-C_5	0.008	72.15	0.58	485	3.88	847	6.78
C_6	0.001	86.18	0.09	434	0.43	915	0.92
C_{7+}	0.001	114.23	0.11	361	0.36	1024	1.02
N_2	0.050	28.02	1.40	227	11.35	492	24.60
CO_2	0.000	44.01	0.00	1073	0.00	548	0.00
H_2S	0.090	34.08	3.07	672	60.52	1306	117.54
	1.000	MW_a =	21.14	p_{pc} =	649	T_{pc} =	472
		g_g =	0.73				

Table: 9 (Sour gas properties)

ρv =	63.74	kg/m^3
Pseudocritical pressure:	710.78	psia
Pseudocritical temperature:	398.50	R
Uncorrected gas viscosity at 14.7 psia:	0.009	cp
N_2 correction for gas viscosity at 14.7 psia:	0.00042	cp
CO_2 correction for gas viscosity at 14.7 psia:	0.000000	cp
H_2S correction for gas viscosity at 14.7 psia:	0.00023	cp
Corrected gas viscosity at 14.7 psia (m_1):	0.009	cp
Pseudo-reduced pressure:	1.37	
Pseudo-reduced temperature:	1.43	
$\ln (m_g/m_1 * T_{pr})$	0.52	

a_0 =	-2.462
a_1 =	2.97
a_2 =	-0.2862
a_3 =	0.008054
a_4 =	2.808
a_5 =	-3.498
a_6 =	0.3603
a_7 =	-0.01044
a_8 =	-0.7933
a_9 =	1.396
a_{10} =	-0.1491
a_{11} =	0.00441
a_{12} =	0.08393
a_{13} =	-0.1864
a_{14} =	0.02033
a_{15} =	-0.0006095

	$\mu 1$
N_2	0.0004215
CO_2	0
H_2S	0.0002311
HC	0.0179041
TOTAL	0.018
SPGR =	0.729

COMPRESSIBILITY FACTOR - Gas compressibility factor is also called deviation factor, or z-factor. Its value reflects how much the real gas deviates from the ideal gas at given pressure and temperature. Very often the z-factor is estimated with the chart developed by Standing and Katz (1942). The chart has been set up for computer solution by a number of individuals. Brill and Beggs (1974) yield z-factor values accurate enough for many engineering calculations, which is followed here too.

Tr = 1.4288449
Pr = 1.3956401
μr = 0.5350266
μg = **0.0221756** Cp

A = 0.3771496
B = 0.5608933
C = 0.0824048
D = 0.9612878
E = 3.8596042
F = -0.017147

Z = **0.846**

11.1 VERTICAL SEPARATOR DESIGN (ref: 3)

FACTORS AFFECTING SEPARATION

Gas and liquid flow rates (minimum, average, and peak)

- Operating and design pressures and temperatures
- Surging or slugging tendencies of the feed streams
- Physical properties of the fluids such as density and compressibility
- Designed degree of separation (e.g., removing 100% of particles greater than 10 microns)
- Presence of impurities (paraffin, sand, scale, etc.)
- Foaming tendencies of the crude oil
- Corrosive tendencies of the liquids or gas

In vertical separators, a minimum diameter must be maintained to allow liquid drops to separate from the vertically moving gas. The liquid retention time requirement specifies a combination of diameter and liquid volume height. Any diameter greater than the minimum required for gas capacity can be chosen. Figure 4-15 shows the model used for a vertical separator.

Drop Diameter, Dm = 140 micron

$$V_t = 0.0119\left[\left(\frac{\rho_l - \rho_g}{\rho_g}\right)\frac{d_m}{C_D}\right]^{1/2} \qquad \text{-----(11.1.1)}$$

Assume Cd = 0.34

Terminal Velosity — Respective Vt = 0.834

Renolds No. — Re = 198.546

2nd

New Cd = 0.673

Respective Vt = 0.592

Re = 141.03

3rd

New Cd = 0.723

Respective Vt = 0.572

Re = 136.148

4th

New Cd = 0.729

Respective Vt = 0.569

Re = 135.57

5th

New Cd = 0.729

Respective Vt = 0.569

Re = 135.507

6th

New Cd = 0.730

Respective Vt = 0.569

Re = 135.498

7th

New Cd = 0.730

Respective Vt = 0.569

Re = 135.497

Final Cd = 0.730

Terminal Velocity $V_t = V_g$ Final Vt = 0.569

Gas Capacity Constaint

$$d^2 = 5{,}040\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{1/2} \qquad \text{---- (11.1.2)}$$

Min. Vessel Internal dia d = 55.939 in

Liquid Capacity Constraint

Liquid Flow	(Q = 6.49* 10^(-5)*Q1)	Q =	0.129	ft³ /s
	(Vol = 4.55 * 10^(-4)*d²h	vol =	23.364	
Min. retention time		d2 h =	50000	

$$t = \frac{Vol}{Q}$$ ---- (11.1.3)

$$d^2h = \frac{t_r\,Q_l}{0.12}$$ ----- (11.1.4)

Diameter vs Height for Liquid capacity constraint

tr	d	H	Vol (1/2) ft3	Seam To Seam Leangth	Slenderness Ratio
3	24	86.80	22.75	13.5	6.783
	30	55.55	22.75	10.96	4.385
	36	38.5	22.75	9.548	3.182
	42	28.34	22.75	8.695	2.484
	48	21.70	22.75	8.141	2.035
	52	18.49	22.75	7.874	1.817
	56	**15.94**	**22.75**	**7.661**	**1.641**

Choose a reasonable size with a diameter greater than that determined by the gas capacity. A 56-in. X 10-ft separator is selected

11.2. ABSORBER DESIGN

This design will not be covered in great detail because it is of little relevance to our process design problem. However, for some applications it may be necessary to use inter-stage cooling and force isothermal condition if the temperature change in the absorption column is great enough to cause large solvent losses.

where

L' -- molar flow rate of pure absorbent

G' -- molar flow rate of pure gas

X -- mole ratio of solute to pure absorbent in liquid

Y -- mole ratio of solute to pure gas in the vapor

$Y_{n+1}=X_n(L'/G')+Y_1-X^o(L'/G')$ ----- (11.2.1)

$Y_n=X_{n+1}(L'/G')+Y^o-X_1(L'/G')$ ----- (11.2.2)

These equations are called operating line equations and when plotted represent the conditions at the top and bottom of the towers. The slope of the line yields the liquid to gas ratio (L'/G'). Given the gas flowrate, the minimum solvent flowrate can be calculated from the relation: $L'_{min}=G'm$

Once the minimum solvent flow rate is found, the actual solvent can be found by multiplying the minimum flow rate by a factor of 1.4. To obtain the equilibrium line for the plot, one would use the following derivation. A balance around a section of the tower will give:

$y = Y / 1 + Y$ ----- (11.2.3)

And similarly for the liquid component:

$x = X / 1 - X$ ----- (11.2.4)

Applying the equilibrium constraint on the system:

$y = K(x)$ ----- (11.2.5)

giving: $(Y / 1 + Y) = K(X / 1 - X)$ ----- (11.2.6)

solving for Y: $Y = KX / 1 + X - KX$ ----- (11.2.7)

This is the equation that determines the equilibrium line in the plot of the McCabe-Thiele diagram. But in this report tray calculations are done from The Colburn equation (Perry's 7th edition) for the number of equilibrium stages viz:

$Nt = \log(((y1 - K^*xn) / (yn - K^*xn))^*(1-1/A) + 1/A) / \log A$ ----- (11.2.8)

Where, A --absorption factor = L'/mG'

y1 -- mole fraction of solute in entering gas

yn -- mole fraction of solute in exiting gas

Once the theoretical number of equilibrium stages has been determined, the actual number of equilibrium stages can be calculated from the relation:

$Eo=Nt/Na$ ----- (11.2.9)

Where, Eo -- stage efficiency

Na -- actual number of stages

Nt -- theoretical number stages

The stage efficiency can be found to be a complex function of the flow rates of vapor and liquid streams, the geometry and design of the contacting trays, and the properties of the vapor and liquid streams. The stage efficiency can be found multiple ways such as from performance data, semi-theoretical models, scaled up laboratory data, and the O'Connell correlation for plate efficiency. The O'Connell correlation can be used given that the operating conditions fall within those covered by the correlation. Once the efficiency has been estimated, the number of actual stages can be determined.

Gas flowrate, G1 in Kg.moles /s = 0.830 Kg.moles/s
= 19.664 std. m^3 / s
= 0.275 act.m^3 / s

as, y1 (H_2S) = 0.090
So, Y1 (H_2S) = 0.0981 mole H_2S / mole gry gas
yn (H_2S) = 0.000004
So, Y n (H_2S) = 4E-06
So, H_2S to be absorbed = 0.074 Kg.moles H_2S / s
Partial pressure of H_2S in N.G = 87.569 Psia

So, from the graph loading, xs
= 14 scf H_2S / gal solvent
= 0.016 Kg.moles H_2S / gal solvent
= 0.228 moles H_2S / moles solvent

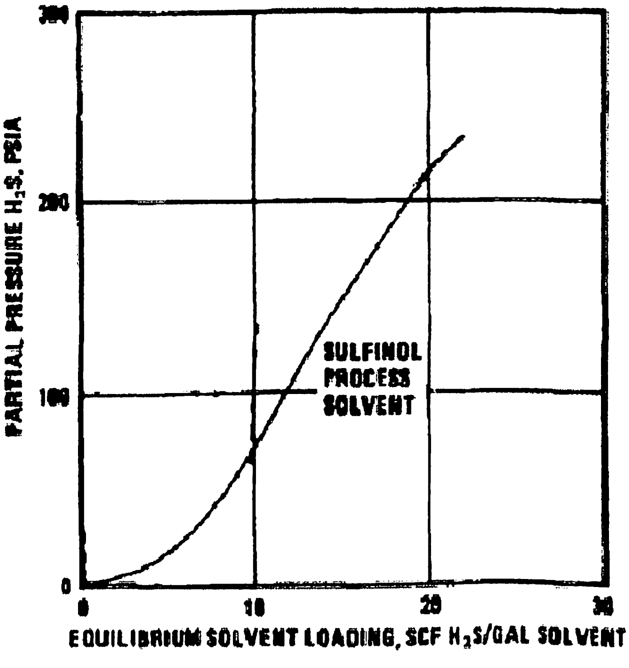


Fig 7 : Equilibrium solubility data

TOP ---So, Solvent (Sulfinol) feed required, Ls = 267.84 GPM

= 0.016 m^3/s
= 18.76 Kg/s
= 0.327 Kg.moles/s

BOTTOM---So,rich Solvent (Sulfinol) out, L1 =
Ls(1+xs) = 0.401 Kg.moles/s
26.427 Kg/s

Density of rich sulfinol --- 818.586 kg/m^3

Natural gas composition :					
	Input (bottom)		Output (Top)		
Compound	mole fraction	Kg.moles/s	Kg.moles/s		y_i
C_1	0.735	0.61	0.610	80.768	**0.808**
C_2	0.083	0.068	0.068	9.120	**0.091**
C_3	0.021	0.017	0.0174	2.307	**0.023**
$i\text{-}C_4$	0.006	0.004	0.0049	0.659	**0.007**
$n\text{-}C_4$	0.002	0.001	0.0016	0.219	**0.002**
$i\text{-}C_5$	0.003	0.002	0.0024	0.329	**0.003**
$n\text{-}C_5$	0.008	0.006	0.006	0.87	**0.009**
C_6	0.001	0.0008	0.0008	0.109	**0.001**
C_{7+}	0.001	0.00083	0.0008	0.109	**0.001**
N_2	0.050	0.041	0.041	5.494	**0.055**
H_2S	0.090	0.074	3.3203E-06	0.0004	**0.000**
Total	1.000	0.830	0.755	100	1.000

Table : 10 Sour & sweet gas composition

Output Natural Gas Properties						
Compound	MW_i	y_iMW_i	p_{ci} (psia)	y_ip_{ci} (psia)	T_{ci}, (°R)	y_iT_{ci} (°R)
C_1	16.04	12.96	673	543.57	344	277.84
C_2	30.07	2.74	709	64.67	550	50.16
C_3	44.10	1.02	618	14.26	666	15.37
i-C_4	58.12	0.38	530	3.49	733	4.83
n-C_4	58.12	0.13	551	1.21	766	1.68
i-C_5	72.15	0.24	482	1.59	830	2.74
n-C_5	72.15	0.63	485	4.26	847	7.45
C_6	86.18	0.09	434	0.48	915	1.01
C_{7+}	114.23	0.13	361	0.40	1024	1.13
N_2	28.02	1.54	227	12.47	492	27.03
H_2S	34.08	0.00	672	0.00	1306	0.01
Total	MW_a =	**19.86**	p_{pc} =	**646**	T_{pc} =	**389**
	g_g =	**0.68**				
	ρv =	**59.82**	kg/m³			

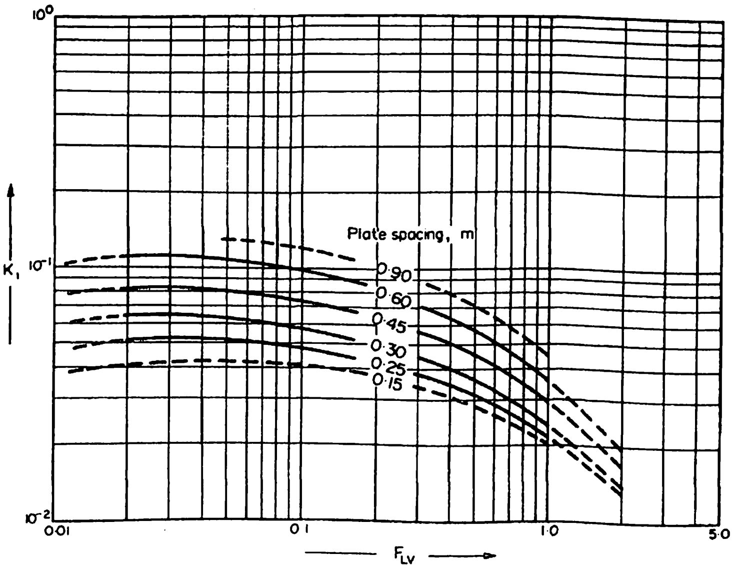
Table 11 : Sweet gas properties

Top, Gs = 0.75 Kg.moles/s

= 0.25 act.m³/s

Flv bottom = (L/G)*(ρv / ρl)^.5 = 0.360

Flv top = (L/G)*(ρv / ρl)^.5 = 0.290



Typical design values of Kv for Sieve, Bubble-Cap, and Valve Plates

Fig 8: (Flooding velocity, sieve plates)

Kv, from table = 0.055 Bottom
Kv, from table = 0.05 Top
for tray spacing = 18 In

So. Base Uf = Kv *((ρl - ρv)/ρv)^.5 = 0.189 m/s Base

So. Top Uf = Kv *((ρl - ρv)/ρv)^.5 = 0.209 m/s Top

Design for flooding at maximum velocity of 85 %
Base Uv = Uf * .85 = 0.160 m/s
Top Uv = Uf * .85 = 0.178 m/s

	Top,	1.195	m^2
Area	Base,	1.454	m^2
Diameter	Top,	1.233	m
	Base,	1.361	m

Use column diameter = **1.5** m

TRAY DESIGN OF ABSORBER

COLLECTIVE DATA FROM ABSORBER COLUMN :

Diameter of column :	1.5	m	
Liquid mass velosity :	18.76	kg/s	
Liquid density :	1110.8	kg/m^3	
Liquid viscosity :	0.036	Pa-s	
K, Equilibrium constant for solute :	0.228		
Liquid molecular wt. :	57.37		
Vapour density :	63.744	kg/m^3	
max.vapour velocity at 85% flooding :	0.178	m/s	
max. vapour rate :	0.275	m^3/s	
orifice coefficient Co :	0.84		
Plate spacing :	18	In	
Flv :	0.360		
Ls, Liquid molar flow :	0.327	Kg.moles/s	
Gs, Gas molar flow :	0.755	Kg.moles/s	
yn, fraction of H_2S :	0.000004		TOP
xn, Fraction of H_2S :	0		
L1, Liquid molar flow :	0.401	Kg.moles/s	
G1, Gas molar flow :	0.830	Kg.moles/s	
y1, fraction of H_2S :	0.09		BOTTOM
x1, Fraction of H_2S :	0.185		

Continued Calculations :

Column Area -	1.766	m^2
Downcomer area = 12% of column area =	0.211	m^2
Net area = Ac - Ad =	1.55	m^2
Active area = Ac -2Ad =	1.34	m^2
Hole area = 10% active area =	0.134	m^2
Hole diameter, hd =	5	mm
Plate thickness =	5	mm
wier length, lw = .6 to .85 times Dia of column =	1.14	m
Wier height, hw =	50	mm

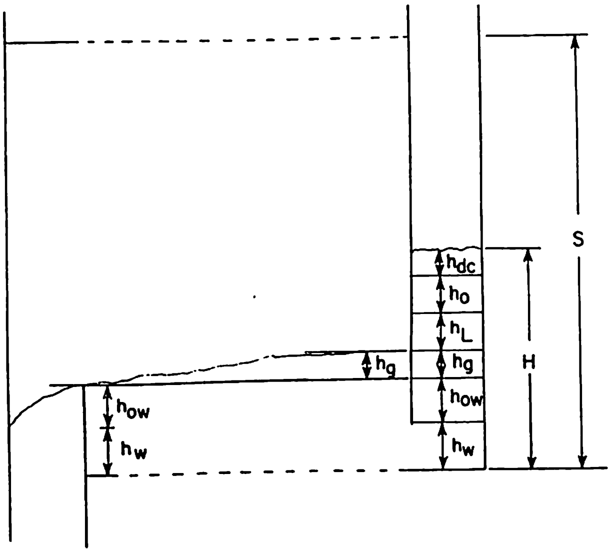


Fig 9 : Single Tray [2]

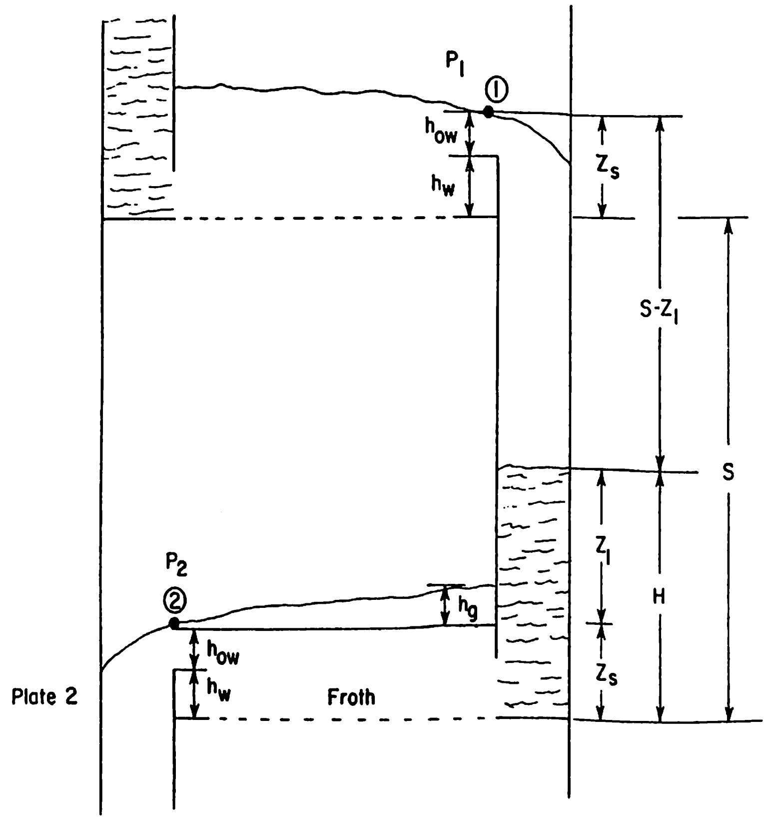


Fig 10 : Sieve Tray arrangement [2]

CHECK WEEPING [2]

Max. liquid rate = 18.768 kg/s

Max, Weir crest, how = 750*(Lw/(ρl*lw))^(2/3) = 45.25 mm

Max. turndown at 70 % liq. Rate = 13.13 kg/s

Min, Weir crest, how = 750*(Lw/(ρl*lw))^(2/3) = 35.67 mm

So, at min. rate (hw + how) = 85.67 mm

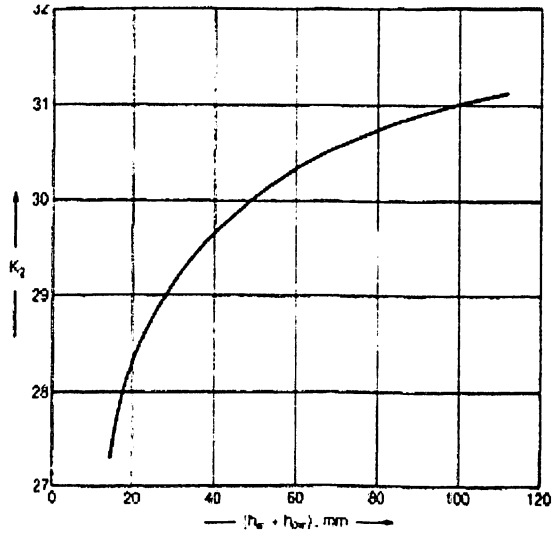


Fig :11 (Weep point correlation)

From the graph K2 = 31

Vhole (min) = (K2 - .9*(25.4-dh))/ρ^2 = 0.0031 m/s

Actual minimum hap velocity, Vhmin = minimum vapour rate/Ah

= 1.435 m/s

PLATE PRESSURE DROP [2]

Minimum vapour velocity through holes,
Vhmax = 2.050 m/s

Plate thickness / hole diameter = 1
So, Ah / Ap = Ah / Aa = 1

Dry plate drop

hd = 51* ((Vhmax / Co)^2)*(ρv / ρl) = 17.441 mm liquid column

Residual head
hr = 12.5* 1000/ ρl = 11.25 mm liq.

Total plate pressure drop
Ht = hd + hr + hw + how = 123.95 mm liquid

DOWNCOMER LIQUID BACKUP [2]

hap = hw - 10 or 5 = 40 mm
Area under apron Aap = hap * lw = 0.045 m^2
As this is less than Ad = 0.21 m^2
So, Ad > Aap
Hd = 166* (Lwd/(ρl * Aap))^2 = 22.791 mm

Downcomer backup measured from plate surface, mm
Hb = hw + how + Hd + Ht
= 241.99 mm

As, Hb < 1/2 * (plate spacing + weir length) = **798.6** mm

So, the diameter and tray spacing -- both are satisfactory

FROTH DENSITY & HEIGHT [2]

Froth density - 0.4 to 0.7 times of clear liquid = 555.4 kg/m^3

Calculation for froth height

Diameter of column :	1.5	m	
Liquid mass velosity :	18.768	kg/s	
Froth density :	555.4	kg/m^3	
Liquid viscosity :	0.036	Pa-s	
K, Equilibrium constant for solute :	0.228		
Liquid molecular wt. :	57.37		
Vapour density :	63.744	kg/m^3	
max.vapour velocity at 85% flooding :	0.178	m/s	
max. vapour rate :	0.275	m^3/s	
orifice coefficient Co :	0.84		
Plate spacing :	18	in	
Flv :	0.360		
Ls, Liquid molar flow :	0.327	Kg.moles/s	
Gs, Gas molar flow :	0.755	Kg.moles/s	
yn, fraction of H_2S :	0.000004		TOP
xn, Fraction of H_2S :	0		
L1, Liquid molar flow :	0.401	Kg.moles/s	
G1, Gas molar flow :	0.830	Kg.moles/s	
y1, fraction of H_2S :	0.09		BOTTOM
x1, Fraction of H_2S :	0.185		

Calculations for froth height :

Column Area -	1.76	m^2
Downcomer area = 12% of column area =	0.21	m^2
Net area = Ac - Ad =	1.55	m^2
Active area = Ac -2Ad =	1.34	m^2
Hole area = 10% active area =	0.134	m^2
Hole diameter, hd =	5	mm
Plate thickness =	5	mm
wier length, lw = .6 to .85 times Dia of column =	1.14	m
Wier height, hw =	50	mm

CHECK WEEPING (with froth consideration) [2]

Max. liquid rate = 18.768 kg/s

Max, Weir crest, how =
750*(Lw/(ρl*lw))^(2/3) = 71.836 mm
Max. turndown at 70 % liq. Rate = 13.138 kg/s
Min, Weir crest, how =
750*(Lw/(ρl*lw))^(2/3) = 56.633 mm

So, at min. rate (hw + how) = 106.633 mm

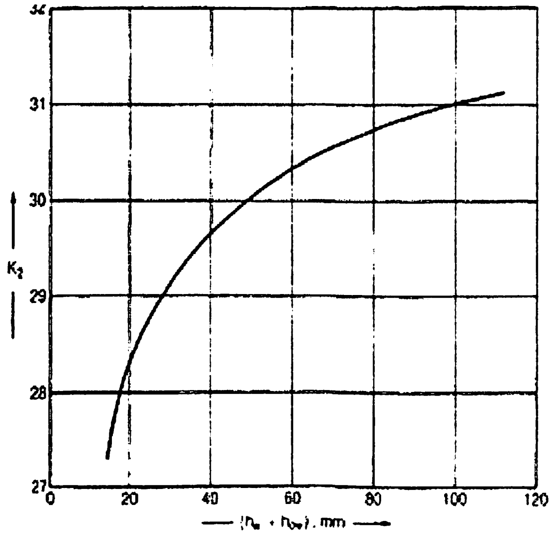


Fig 12: Weep point correlation

From the graph K2 = 31

Vhole (min) = (K2 - .9*(25.4-dh))/ρ^2
= 0.0031 m/s

Actual minimum hap velocity, Vhmin
= minimum vapour rate/Ah
= 1.435 m/s

PLATE PRESSURE DROP (with froth consideration) [2]

Minimum vapour velocity through holes, Vhmax = 2.050 m/s

Plate thickness / hole diameter = 1

So, Ah / Ap = Ah / Aa = 1

Dry plate drop

hd = 51* ((Vhmax / Co)^2)*(ρv / ρl) = 34.88 mm liquid column

Residual head

hr = 12.5* 1000/ ρl = 22.506 mm liq.

Total plate pressure drop

Ht = hd + hr + hw + how = 179.22 mm liquid

DOWNCOMER LIQUID BACKUP (for froth consideration) [2]

hap = hw - 10 or 5 = 40 mm

Area under apron Aap = hap * lw = 0.045 m^2

As this is less than Ad = 0.211 m^2

So, Ad > Aap

Hd = 166* (Lwd/(ρl * Aap))^2 = 91.166 mm

Downcomer backup measured from plate surface, mm

Hb = hw + how + Hd + Ht = 392.228 mm

As, Hb < 1/2 * (plate spacing + weir length) = **798.6** mm

So, the diameter and tray spacing -- both are satisfactory

Residence time

tr = Ad * Hb * ρl / Lwd = 3.03556905 s

Entrainment Check [2]

Uv = (max.vap.rate) Vhmax/ Net area = 0.177 m/s

%age flooding = 85

So, at at Flv = 0.360

we get from graph

ψ, Fractional entrainment = 0.004 < 0.1 (below this value the effect on efficiency will be small)

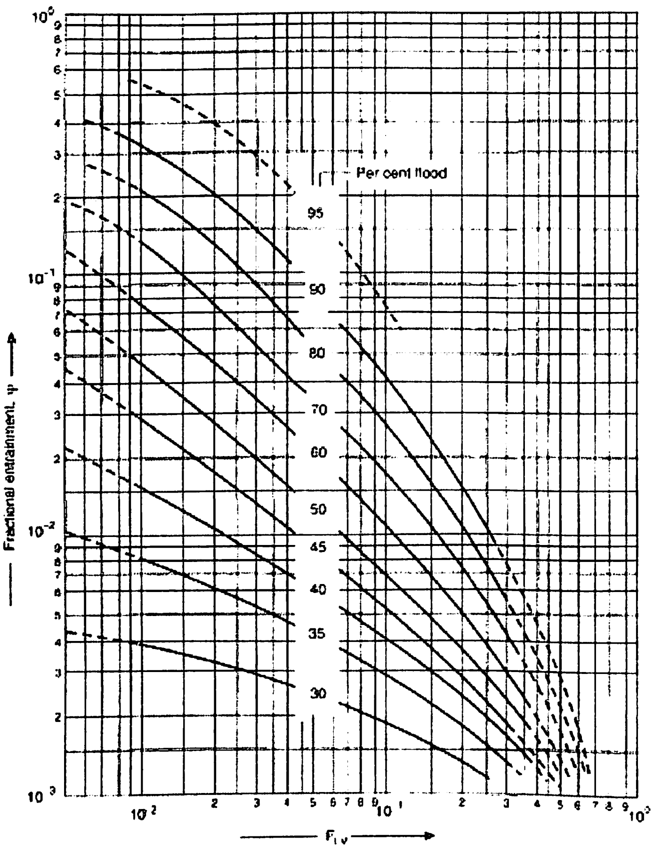


Fig : 13 (Fractional Entrainment factor)

Tray Efficiency [2]

x = .062 * ρs / (μs * K * Ms)= 142.368

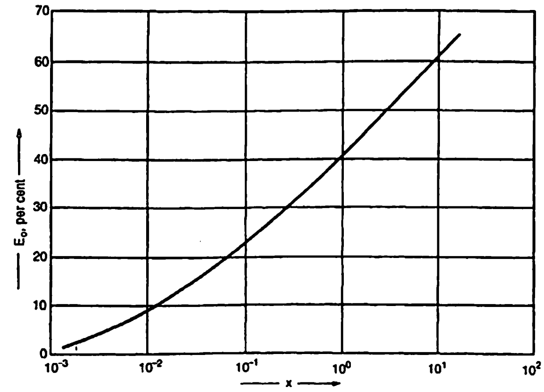


Fig : 14 (Efficiency of tray)
From Graph we get
Ea, efficiency of plate = 70 %

Number of Trays calculation [2]

A, Absorption factor = L/ (K*G)

A1, bottom absorption factor = L1 / (K* G1)
= 2.119

A2, top absorption factor = Ls / (K* Gs)
= 1.896

avg. A = (A1 * A2)^(.5)
= 2.004

Number of theoretical plates = log (((y1 - K*xn) / (yn - K*xn))*(1-1/A) + 1/A) / logA

= 13.414

Number of real trays = Theoretical trays / efficiency
= 19.164
= 19

Height of column

Height of trays = (Total number of trays - 1) * Tray spacing

= 27.246 ft

Height of top & bottom empty spaces

= 8 ft

Total height of column

= 35.246 ft

LOAD BASED SHELL THICKNESS CALCULATION FOR ABSORBER [2]

INPUT DATA :

Shell Material -	SA- 283 Grade C	
Type of shell Plate Joint –	Double welded butt joint with 10% Radiography.	
Skirt Height -	4	m
Weight of Liquid and Tray -	120	kg/m^2
Weight of Attachment - (pipes, ladders & platform)	150	kg/m
Wind Pressure (Pw) -	130	kgf/cm^2
Insulation Thickness t_{ins} -	100	mm
Density of Insulation ρ_{ins} -	500	kg/m^3
Maximum allowable stress of shell plate material at design Temperature, F -	890	kgf/cm^2
Modulus of Elasticity 'E' -	2000000	kgf/cm^2
Poison's Ratio μ -	0.3	
Corrosion Allowance (C.A) -	2	mm
Spgr of material grade choosen -	7.865	
Multiplicative to max allowable stress for, J	0.85	

Design :

From Column dsign calculation we have
Shell O.D at top - 1500 mm
Shell Length Tangent Line to Tangent Line - 10.743 m
Internal Design Pressure - 66.189 kgf/cm^2
Design Temperature - 43 °C
Tray Spacing - 0.457 m
Top Disengaging Space - 1.219 m

For internal design pressure

Internal design pressure = 66.189 kgf/cm^2
O.D of shell at top Do = 1500 mm

ts = (P * Ro / (F * J + .4* P)) + C.A = 65.402 mm

So plate thickness to be used for the fabrication of shell is 66 mm

Inside diameter at top Di = Do - C.A * ts = 1367.195 mm

Design of top head (Type of head : Torispherical)

For head subjected to internal pressure

Rc = Di = 1367.195 mm
Rk = .1* Rc = 136.719 mm
W = .25 * (3 + (Rc / Rk)^.5) = 1.540
th' = (P * Rc * W / (2* F * J -.2 * P)) + C.A 94.956 mm

th = 1.06 * th' = 100.654 mm

So plate thickness used to fabricate top head is 102 mm

Blank diameter = OD + OD/42 +2SF + 2/3 iCr + th for th <>1"

O.D of head = Rc + 2 * th = 1570.503 mm

S.F = 1.5 " = 38.1 mm

Blank diameter = 1876.89 mm
Wt. of top head = 2210.926 Kg

Let X be the distance from the top of up to which we can use calculated thick shell.

(A) Circumferential stress induced in shell plate material at a distance X from the top of shell
(due to internal presssure)

Fcp = P * Di / (2* ts - C.A) = 702.5729993 kgf/cm^2 is tensile in nature

So. Ftmax = J * Fallowable = 756.5 kgf/cm^2

as Fcp < Ftmax
Fcp will remain same for entire length

(B) Various axial stresses induced in the shell plate material at distance X from the top of the shell.

(i) Axial stress induced due to internal pressure, Fap = P * Di/ (4*(ts-C.A)) = 351.286 kgf/cm^2

(ii) Axial stress induced due to dead weights
Fdx = Fdsx +Fdinsx +Fd(ligt + tray)x +Fdattx

Where, Fdsx = stress induced due to weight of shell
Fdsx = ρs * X = 0.786 *X kgf/cm^2

Axial stress due to dead load of insulation, Fdinsx

Dm = (Do + Di)/2 = 1433.597 mm
Dins = Do + tins = 1600 mm (Mean diameter of insulation shell)
Fdinsx =
3.14*Dins*tins*ρins*X/(3.14*Dm*(ts - C.A)) = 0.0866 * X kgf/cm^2

Axial stress induced due to dead load of (liq + trays) up to distance X
from top, Fd(liq + tray)x

Top disengaging space, h = 1.2192
Tray spacing, S = 0.4572
F(liq + tray)x = no. of trays up to X * (wt. of one tray + wt. of liq. On the same tray)
F(liq + tray)x = ((X-h)/S +1) * wt. = 385.128 * (X -0.762) kgf/ m^2

Fd(liq + tray)x = F(liq + tray)x / (3.14*Dm*(ts-C.A) 0.132 * (X -0.762) kgf/cm^2

Fd(att)x = (wt. of top head + wt. of pipe, ladder, platform etc.) / (3.14* Dm * (ts - C.A))

Fd(att)x = 0.76263547 + X * 0.0517409 kgf/cm^2

Fdx = 0.7865 * X + 0.086 * X +0.132* X
0.661406747 + X * 0.051740901
Implies ----------
Fdx = 1.0577 * X + 0.661406747

(iii) Axial stress due to wind load at a distance X from the top of the shell, Fwx

Fwx = (1.4 * Pw * X^2)/ (3.14 *Do * (ts - C.A)) = 599.998 * X^2 kgf/m^2
= 0.0599 * X^2 kgf/cm^2

Maximum ensile stress induced in the shell plate material at a distance X from the top of the shell.

Ftmax = Fap - Fdx + Fwx = 0.059 * X^ 2 - 1.057 * X + -0.66+351.28
= 0.0599 * X^ 2 - 1.057 * X + 350.625
= 756.5

So the equation becomes ------------ 0.0599 * X^ 2 -1.057 * X -405.874 =0
So, X = 91.532 or -0.266

So, $X =$ 91.53264202 m

Maximum compressive stress induced in the shell plate at a distance X from the top of the shell :

$Fcmax = Fwx + Fdx - Fap =$ $0.059 * X^2 + 1.057 * X - 350.62$

$Fcallow = (1/12) * (E/(3*(1-\mu^2))^{.5}) * ((ts - C.A) * 2/Do)$

$=$ 8661.765968 kgf/cm^2

So, $Fcmax =$ 7362.501073 kgf/cm^2

So, equation becomes ------------ $0.0599 * X^2 + 1.057 * X - 7713.12 = 0$

So, $X =$ 349.835914 or -367.4648859

So, $X =$ 349.835914 m

As the lower one to be considered,
therefor up to 349.835914 m

Result : - Absorber shell thickness to be used is 66.402137 mm

11.3 FLASH DRUM [2]

INPUT DATA		
Inlet Temperature :	174.2	^oF
Outlet Pressure :	80.85	psia
Outlet Temperature :	122	^oF
New K at outlet pressure :	0.1	moles H_2S / moles Solvent
Vessel Full %age :	50	%
Residence time :	1	min

(at reduced pressure)
(Assumption : In vapour phase only H_2S present)

Calculation :

Rich Sulfinol calculated data

Inlet Flow rate :	0.4018109	Kg.moles/s

Inlet Composition :

Sulfolane :	0.154	mol. Frac	7.507	Kg/s
DIPA :	0.140	mol. Frac	7.507	Kg/s
Water :	0.519	mol. Frac	3.753	Kg/s
H_2S :	0.185	mol. Frac	7.659	Kg/s
			26.428	Kg/s

Density of rich sulfinol ---	818.586	kg/m^3
Inlet Pressure :	972.99	psia

F, Feed = L, Liquid rate + V, vapor rate
i.e L + V = 0.401

Whereas vapor phase will be according to the reduction in solubility of H_2S in sulfinol due to pressure reduction

So, V = Inlet rich sulfinol - (pure sulfinol * (1 + new K))

= 0.0419 Kg.moles/s

V = Sulfinol releasing = 0.0419 Kg.moles/s

So, L = 0.359 Kg.moles/s
= 0.018 m^3 /s

So, new composition of the outlet rich sulfinol :

		Kg.moles/s	mole. Frac	kg/s	wt. frac
Outlet Composition	Sulfolane :	0.062	0.172	7.50	0.37
	DIPA :	0.056	0.156	7.50	0.37
	Water :	0.208	0.579	3.75	0.18
	H_2S :	0.032	0.090	1.11	0.05
	Total, moles/s :	0.359		19.8	

Consider Length of vessel = 3 * Diameter

as, Vessel capasity = Liquid rate / percentage full
= 2.2303473 m^3

So, (3.14/4) * Diameter^2 * 3* Diameter = 2.23 m^3

So, Diameter = 0.98 m

& Length = 2.94 m

11.4 HEAT EXCHANGER DESIGN

THE DESIGN STEPS OF HEAT EXCHANGER:

1. Process conditions (stream compositions, flow rates, temperatures, pressures) must be specified.

2. Required physical properties over the temperature and pressure ranges of interest must be obtained.

3. The type of heat exchanger to be employed is chosen.

4. A preliminary estimate of the size of the exchanger is made, using a heat-transfer coefficient appropriate to the fluids, the process, and the equipment.

5. A first design is chosen, complete in all details necessary to carry out the design calculations.

6. The design chosen in step 5 is evaluated, or rated, as to its ability to meet the process specifications with respect to both heat transfer and pressure drop.

7. On the basis of the result of step 6, a new configuration is chosen if necessary and step 6 is repeated. If the first design was inadequate to meet the required heat load, it is usually necessary to increase the size of the exchanger while still remaining within specified or feasible limits of pressure drop, tube length, shell diameter, etc. This will sometimes mean going to multiple-exchanger configurations. If the first design more than meets heat-load requirements or does not use all the allowable pressure drop, a less expensive exchanger can usually be designed to fulfill process requirements.

8. The final design should meet process requirements (within reasonable expectations of error) at lowest cost. The lowest cost should include operation and maintenance costs and credit for ability to meet long-term process changes, as well as installed (capital) cost. Exchangers should not be selected entirely on a lowest-first-cost basis, which frequently results in future penalties.

ASSUMPTIONS:

1. All elements of a given fluid stream have the same thermal history in passing through the exchanger.

2. The exchanger operates at steady state.
3. The specific heat is constant for each stream.
4. The overall heat-transfer coefficient is constant.
5. Heat losses are negligible.

Lean Sulfinol	T1 Hot fluid hot temperature	=	269.6	^{o}F	132	^{o}C
from striper	T2 Hot fluid cold temperature	=	200	^{o}F	93.33	^{o}C
Rich Sulfinol	t1 Cold fluid cold temperature	=	122	^{o}F	50	^{o}C
from flash drum	t2 Cold fluid hot temperature	=	190.356	^{o}F	87.97	^{o}C

Thermal conductivity cold fluid = 0.53 kcal/hr m C
Thermal conductivity hot fluid = 0.5 kcal/hr m C

TUBE DETAILS
BWG 16
OD outside diameter = 0.75 in
ID inside diameter = 0.62 in
L Length = 16 ft
Triangular
Pitch = 1
K1 = 0.319
N1 = 2.142

Calculations

viscosity(cold fluid) at 68.9878 = 0.0021 poise
viscosity(hot fluid) at 112.667 = 0.0026 poise

mk1 mass flow rate hot fluid = 0.327 Kg.moles/s 67567.4 kg/hr Lean Sulfinol
mk2 mass flow rate cold fluid = 0.359 71571.2 kg/hr Rich Sulfinol

ρk1 Density of hot fluid = 1.11 g/cc 1110.8 kg/m^3

ρk2 Density of cold fluid = 0.650 g/cc 650.86 kg/m^3

ck1 Hot fluid specific heat = 0.430 kcal/kg C Lean Sulfinol
ck2 cold fluid specific heat = 0.414 kcal/kg C Rich Sulfinol

Heat Duty Calculation

Temperature ratio, (T1 - T2) / (t2 - t1) = 1.0182

Cold fluid heat absorption = 1125881 kcal/hr
mk2 * ck2 * (t2 - t1)

Hot Fluid heat removal = 1125883 kcal/hr
mk1 * ck1 * (T1 - T2)

mk2 mass flow rate cold fluid = 71571.2 kg/hr

LMTD calculation = 43.678 °C

1st assumption of U = 1500 kcal/hr m^2 °C

A, area required for H.T = 17.184 m^2

With 3/4 " O.D & 16 ft long tube

Nt = At/(3.14 * O.D * L) = 58.908 tubes no. of tubes

Db, Tube bundle Dia = 217.754 mm

Calculation of hi, inside heat transfer coefficient

at = (Nt/Np)*(¶/4)*di^2 = 0.0114 m^2

Gt = m/at = 1733.55 kg/m^2 s

ut, velocity of tube side fluid = 2.663 m/s

Re, = 12938.3

Pr, = 5.866

hi, = 2735.57 kcal/hr m^2 C

ho, = 2261.41 kcal/hr m^2 C

Equipment	H2O percent wt.			[14]
	10	20	25	Btu / hr.ft2.°F
Lean/ Rich exchanger	70	102	119	
Solvent Coolers	37	85	94	
Reboilers	131	107	207	

11.5 COOLER DESIGN

Lean Sulfinol from	T1 Hot fluid hot temperature	=	200	°F	93.3333	°C
heat exchanger R/L	T2 Hot fluid cold temperature	=	119.3	°F	48.5	°C

Required temperature of Lean Solvent for Absorption. Typically 5.5°C above inlet gas temperature

Utility used	t1 Cold fluid cold temperature	=	68	°F	20	°C
is RT water	t2 Cold fluid hot temperature	=	104	°F	40	°C

(water)	Thermal conductivity cold fluid		=	0.54	kcal/hr m C		
(Lean Sulfinol)	Thermal conductivity hot fluid		=	0.5	kcal/hr m C		
(water)	ck1 cold fluid specific heat		=	1.4	kcal/kg C		
(water)	viscosity(cold fluid) at	30	=	0.0006	poise		
(water)	ρk1 Density of cold fluid		=	1	g/cc	1000	kg/m^3

TUBE DETAILS
BWG 16
OD outside diameter = 0.75 in
ID inside diameter = 0.62 in
L Length = 16 ft
Triangular Pitch = 1
K1 = 0.319
N1 = 2.142

Calculations

viscosity(hot fluid) at 70.91666667 = 0.00269 poise
ck2 hot fluid specific heat = 0.43094 kcal/kg C Lean Sulfinol
mk2 mass flow rate hot fluid = 0.35982 moles/s 74324.1 kg/hr Lean Sulfinol

ρk2 Density of hot fluid = 1.1108 g/cc 1110.8 kg/m^3

Heat Duty Calculation

Temperature ratio, (T1 - T2) / (t2 - t1) = 2.24167

Cold fluid heat absorption = 1435986 kcal/hr
mk1 * ck1 * (t2 - t1)

Hot Fluid heat removal = 1435986 kcal/hr
mk2 * ck2 * (T1 - T2)

mk2 mass flow rate cold fluid = 51285.2 kg/hr

LMTD calculation = 39.6282 °C

1st assumption of U = 1500 kcal/hr m^2 °C

A, area required for H.T = 24.1576 m^2

With 3/4 " O.D & 16 ft long tube

Nt = At/(3.14 * O.D * L) = 82.8122 tubes no. of tubes

Db, Tube bundle Dia = 255.282 mm

Calculation of hi, inside heat transfer coefficient

at = $(Nt/Np)*(\P/4)*di^2$ = 0.016 m^2

Gt = m/at = 883.639 $kg/m^2\ s$

ut, velocity of tube side fluid = 0.88 m/s

Re, = 23192.6

Pr, = 1.72

hi, = 2932.33 kcal/hr m^2 C

ho, = 2424.06 kcal/hr m^2 C

11.6 STRIPPING COLUMN DESIGN

y2, Outlet Sulfinol purity possible, wrt.H_2S = 100 %
Steam available at : 120 °C
Steam, H_2S absorbility from table : moles H_2S / moles steam 0.1
Steam density, ρ : 2.7 **kg/m3**
Steam molecular wt. : 18
Design Pressure : 5 **kgf/cm^2**

Sulfinol Properties------

Rich Sulfinol	Molecular wt.	DIPA =	133.2	
		Sulfolane =	120.7	
		Water =	18	
		H_2S =	34	
	Density	DIPA =	989	**kg/m^3**
		Sulfolane =	1288	**kg/m^3**
		Water =	1000	**kg/m^3**
		H_2S =	6	**kg/m^3**
	Viscosity	DIPA =	198	**cp**
		Sulfolane =	6.1	**cp**
		Water =	1	**cp**
		H_2S =	0.3	**cp**
	Specific heat	DIPA =	0.69	**Btu/lb**
		Sulfolane =	0.35	**Btu/lb**
		Water =	1.801	**Btu/lb**
		H_2S =	0.238	**Btu/lb**

Calculation:

Solvent Properties

	by wt.	
DIPA =	37.76	%
Sulfolane =	37.76	%
Water =	18.88	%
H_2S =	5.594	%

	mol. Fraction	**Kg.moles/s**
DIPA =	0.157	0.05637
Sulfolane =	0.173	0.0622
Water =	0.58	0.20854
H_2S =	0.091	0.03271

Avg Viscosity of sulfinol = 33.74 cp

Avg Specific heat = 0.414 kcal/kg °C

Avg Molecular wt. of sulfinol= 55.25

Avg Density of sulfinol = 1049 **kg/m^3**

Avg Viscosity of sulfinol = 3.958 cp

TOP--Rich Solvent (Sulfinol) feed, Ls = 19.88 kg/s

= 0.36 Kg.moles/s

= 0.019 m^3/s

ρl = 1049 **kg/m^3**

xs = 0.091

BOTTOM--Lean Solvent (Sulfinol) out, L1 = Ls - H_2S = 0.327 Kg.moles/s

= 18.77 kg/s

= 0.017 m^3/s

ρl = 1111 **kg/m^3**

x1 = 1E-07

So, H_2S to be absorbed = 0.033 Kg.moles H_2S / s

BOTTOM--G1, Steam flow rate required 3.124 Kg.moles/s

=

= 56.23 kg/s

= 20.83 m^3/s

ρv = 2.7 **kg/m³**

y1 = 0

TOP--Gs, Steam flow rate required = 3.157 Kg.moles/s

= 56.82 kg/s

= 20.78 m^3/s

ρv = 2.734 **kg/m³**

as, ys (H_2S)

= 0.010 mole H_2S / mole gas

Flv bottom = (L/G)*(ρv / ρl)^.5 = 0.005

Flv top = (L/G)*(ρv / ρl)^.5 = 0.006

Typical design values of Kv for Sieve, Bubble-Cap, and Valve Plates

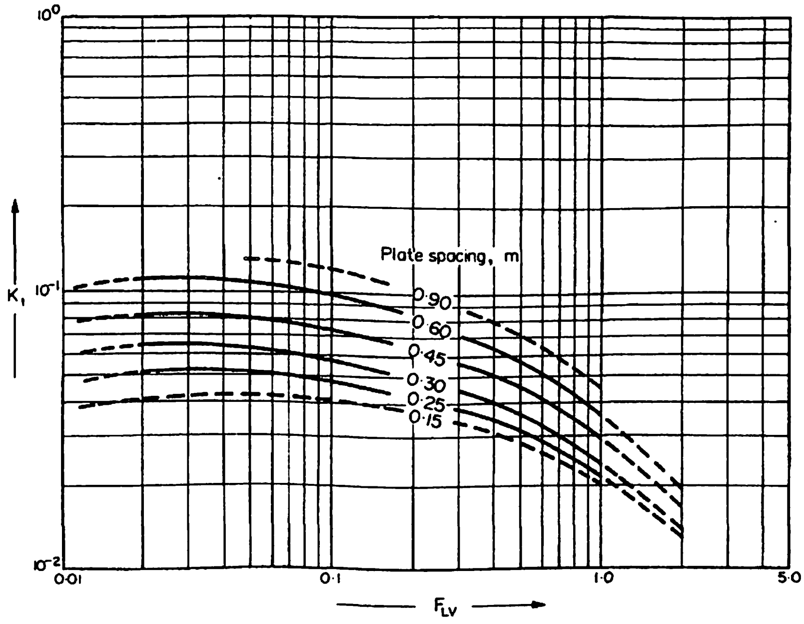


Fig 15: Flooding velocity, sieve plates

Kv, from table =		0.1		
for tray spacing =		30	in	
So. Base Uf = Kv *((ρl - ρv)/ρv)^.5 =		2.026	m/s	Base
So. Top Uf = Kv *((ρl - ρv)/ρv)^.5 =		1.956	m/s	Top
Design for flooding at maximum flow rate of				85 %
Base Uv = Uf * .85 =		1.722	m/s	
Top Uv = Uf * .85 =		1.663	m/s	
	Top,	12.5	m^2	
Area	Base,	12.09	m^2	
Diameter	Top,	3.99	m	
	Base,	3.925	m	
Use column diameter =		4	m	

COLLECTIVE DATA FROM STRIPPER COLUMN

Diameter of column :	4	m	
Liquid mass velocity :	19.880	kg/s	
Liquid density :	1049	kg/m^3	
Liquid viscosity :	0.033	Pa-s	
Liquid molecular wt. :	55.252		
Vapor density :	2.73	kg/m^3	
max.vapor velocity at 85% flooding :	1.721	m/s	
max. vapor rate :	20.78	m^3/s	
orifice coefficient Co :	0.84		
Plate spacing :	30	in	
Flv :	0.0058		
Ls, Liquid molar flow :	0.359	Kg.moles/s	
Gs, Gas molar flow :	3.156	Kg.moles/s	TOP
yn, fraction of H_2S :	0.010		
xn, Fraction of H_2S :	0.090		
L1, Liquid molar flow :	0.327	Kg.moles/s	
G1, Gas molar flow :	3.123	Kg.moles/s	Bottom
y1, fraction of H_2S :	0		
x1, Fraction of H_2S :	0.01		

Tray Efficiency [2]

x = .062 * ρs / (μs * K * Ms) = 34.885

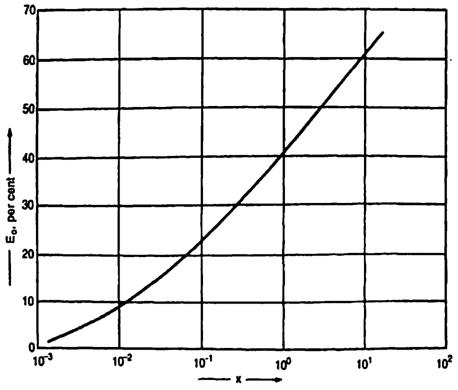


Fig 16 (Efficiency of tray)

From Graph we get

Ea, efficiency of plate = 70 %

Number of Trays calculation [2]

S, Stripping factor = L/ (K*G)

S1, bottom stripping factor = (K* G1) / L1 = 0.955

S2, bottom stripping factor = (K* Gs) / Ls = 0.877

avg. S = (S1 * S2)^(.5) = 0.915

Number of theoretical plates = log (((xn - y1/K) / (x1 - y1/K))*(1-1/S) + 1/S) / logS

= 15.603

Number of real trays = Theoretical trays / efficiency

= 22.29

= 22

Height of column

Height of trays = (Total number of trays - 1) * Tray spacing

= 53.225 ft

Height of top & bottom empty spaces

= 8 ft

Total height of column

= 61.225 ft

SHELL THICKNESS CALCULATION OF STRIPPER [2]

INPUT DATA :

Shell Material -	SA- 283 Grade C	
Type of shell Plate Joint -	Double welded butt joint with 10% Radiography.	
Skirt Height -	4	m
Weight of Liquid and Tray -	120	kg/m^2
Weight of Attachment - (pipes, ladders & platform)	150	kg/m
Wind Pressure (Pw) -	130	kgf/m^2
Insulation Thickness t_{ins} -	100	mm
Density of Insulation ρins -	500	kg/m^3
Maximum allowable stress of shell plate material at design Temperature, F -	890	kgf/cm^2
Modulus of Elasticity 'E' -	2000000	kgf/cm^2
Poison's Ratio μ -	0.3	
Corrosion Allowance (C.A) -	2	mm
Spgr of material grade choosen -	7.865	
Multiplicative to max allowable stress for, J	0.85	

Design :

From Column dsign calculation we have

Shell O.D at top -	4000	mm
Shell Length Tangent Line to Tangent Line -	18.66143603	m
Internal Design Pressure -	5	kgf/cm^2
Design Temperature -	120	°C
Tray Spacing -	0.762	m
Top Disengaging Space -	1.2192	m

For internal design pressure

Internal design pressure = 5 kgf/cm^2
O.D of shell at top Do = 4000 mm

ts = (P * Ro / (F * J + .4* P)) + C.A = 15.183 mm

So plate thickness to be used for the fabrication of shell is 16 mm

Inside diameter at top Di = Do - C.A * ts = 3967.632 mm

Design of top head (Type of head : Torispherical)

For head subjected to internal pressure

Rc = Di = 3967.632 mm
Rk = .1* Rc = 396.763 mm
W = .25 * (3 + (Rc / Rk)^.5) = 1.540
th' = (P * Rc * W / (2* F * J -.2 * P)) + C.A 22.21 mm

th = 1.06 * th' = 23.545 mm

So plate thickness used to fabricate top head is 25 mm

Blank diameter = OD + OD/42 +2SF + 2/3 iCr + th for th <>1"
O.D of head = Rc + 2 * th = 4016.72 mm

S.F = 1.5 " = 38.1 mm

Blank diameter = 4477.614 mm
Wt. of top head = 3038.35 kg

Let X be the distance from the top of up to which we can use calculated thick shell.

(A) Circumferential stress induced in shell plate material at a distance X from the top of shell
(due to internal presssure)

Fcp = P * Di / (2* ts - C.A) = 699.318 kgf/cm^2 is tensile in nature

So. Ftmax = J * Fallowable = 756.5 kgf/cm^2

as Fcp < Ftmax
Fcp will remain same for entire length

(B) Various axial stresses induced in the shell plate material at distance X from the top of the shell.

(i) Axial stress induced due to internal
pressure, Fap = P * Di/ (4*(ts-C.A)) = 349.659 kgf/cm^2

(ii) Axial stress induced due to dead weights
Fdx = Fdsx +Fdinsx +Fd(ligt + tray)x +Fdattx

Where, Fdsx = stress induced due to weight of shell
Fdsx = ρs * X = 0.786 *X kgf/cm^2

Axial stress due to dead load of insulation, Fdinsx

Dm = (Do + Di)/2 = 3983.816 mm
Dins = Do + tins = 4100 mm (Mean diameter of insulation shell)
Fdinsx =
3.14*Dins*tins*ρins*X/(3.14*Dm*(ts - C.A)) = 0.362 * X kgf/cm^2

Axial stress induced due to dead load of (liq + trays) up to distance X from top, Fd(liq + tray)x

Top disengaging space, h = 1.21
Tray spacing, S = 0.762
F(liq + tray)x = no. of trays up to X * (wt. of one tray + wt. of liq. On the same tray)
F(liq + tray)x = ((X-h)/S +1) * wt. = 1946.071 * (X - 0.457) kgf/ m^2
Fd(liq + tray)x = F(liq + tray)x / (
3.14*Dm*(ts-C.A) = 1.096 * (X-0.457) kgf/cm^2

Fd(att)x = (wt. of top head + wt. of pipe, ladder, platform etc.) / (3.14* Dm * (ts - C.A))

Fd(att)x = 1.712433336 + X * 0.084 kgf/cm^2

Fdx = 0.786 * X +0.362* X + 1.096* X 1.21
+ X *0.084

Implies ----------
Fdx = 2.33* X + 1.21

(iii) Axial stress due to wind load at a distance X from the top of the shell, Fwx

Fwx = (1.4 * Pw * X^2)/ (3.14 *Do * (ts - C.A)) = 1021.61 * X^2 kgf/m^2
= 0.102 * X^2 kgf/cm^2

Maximum ensile stress induced in the shell plate material at a distance X from the top of the shell.

Ftmax = Fap - Fdx + Fwx = 0.102 * X^ 2 - 2.33 * X +
-1.21 + 349.659
= 0.102 * X^ 2 - 2.33 * X +
348.448
= 756.5
So the equation becomes ------------ 0.102 * X^ 2 -2.33 * X
-408.051 = 0
So, X = 75.627 or -0.55121
So, X = 75.627 m

Maximum compressive stress induced in the shell plate at a distance X from the top of the shell :

Fcmax = Fwx + Fdx - Fap = 0.102 * X^ 2 + 2.33 * X
-348.448

Fcallow = (1/12) * (E/ (3* (1-μ^2))^.5) * ((ts - C.A) * 2/Do)
= 715.374 kgf/cm^2
So, Fcmax = 608.068 kgf/cm^2
0.102 * X^ 2 + 2.33 * X

So, equation becomes ------------

-956.516 = 0

So, X = 86.024 or -108.838

So, X = 86.024 m

As the lower one to be considered,
therefor up to 86.024 m

Result : Shell thickness to be used -- **16** mm

12. MATERIAL BALANCE

INPUT TO ABSORBER

Gas Composition	mol frac	Kg.moles/s
C_1	0.735	0.61
C_2	0.083	0.068
C_3	0.021	0.017
i-C_4	0.006	0.004
n-C_4	0.002	0.001
i-C_5	0.003	0.002
n-C_5	0.008	0.006
C_6	0.001	0.0008
C_{7+}	0.001	0.0008
N_2	0.05	0.041
CO_2	0	0
H_2S	0.09	0.074
Total		0.830

Sulfinol composition	% by wt.	mol. Fraction	Kg.moles/s
DIPA =	40	0.172	0.056
Sulfolane =	40	0.19	0.062
Water =	20	0.637	0.208
H_2S =	0	0	0
Total			0.327

OUTPUT FROM ABSORBER

Gas Composition	mol frac	Kg.moles/s
C_1	0.807	0.610
C_2	0.091	0.068
C_3	0.023	0.017
i-C_4	0.006	0.004
n-C_4	0.002	0.001
i-C_5	0.003	0.002
n-C_5	0.008	0.006
C_6	0.001	0.0008
C_{7+}	0.001	0.0008
N_2	0.054	0.041
CO_2	0	0
H_2S	4.3956E-06	3.3E-06
Total		0.75537

Sulfinol composition	mol. Fraction	Kg.moles/s
DIPA =	0.140	0.056
Sulfolane =	0.154	0.062
Water =	0.519	0.208
H_2S =	0.185	0.074
Total		0.401

OUTPUT FROM FLASH DRUM (INPUT TO & OUTPUT FROM HEAT EXCHANGER)

Sulfinol composition	**mol. Fraction**	**Kg.moles/s**		**Kg.moles/s**
DIPA =	0.156	0.056	Vapor -- H_2S	0.042
Sulfolane =	0.172	0.06		
Water =	0.579	0.208		
H_2S =	0.090	0.032		
Total		0.359		

OUTPUT FROM STRIPPER

Sulfinol output - 0.327 **Kg.moles/s**

Sulfinol composition	**mol. Fraction**	**Kg.moles/s**
DIPA =	0.172	0.056
Sulfolane =	0.190	0.062
Water =	0.637	0.208
H_2S =	0	0
Total		0.327

13. LEAN SULFINOL TRANSPORT PUMP CALCULATIONS

Pumps can be classified into two general types:

1. Dynamic pumps, such as centrifugal pumps.
2. Positive displacement pumps, such as reciprocating and diaphragm pumps.

The single-stage, horizontal, overhung, centrifugal pump is by far the most commonly used type in the chemical process industry. Other types are used where a high head or other special process considerations are specified. Pump selection is made on the flow rate and head required, together with other process considerations, such as corrosion or the presence of solids in the fluid. Kern (1975) discusses the practical design of pump suction piping, in a series of articles on the practical aspects of piping system design published in the journal Chemical Engineering from December 1973 through to November 1975.

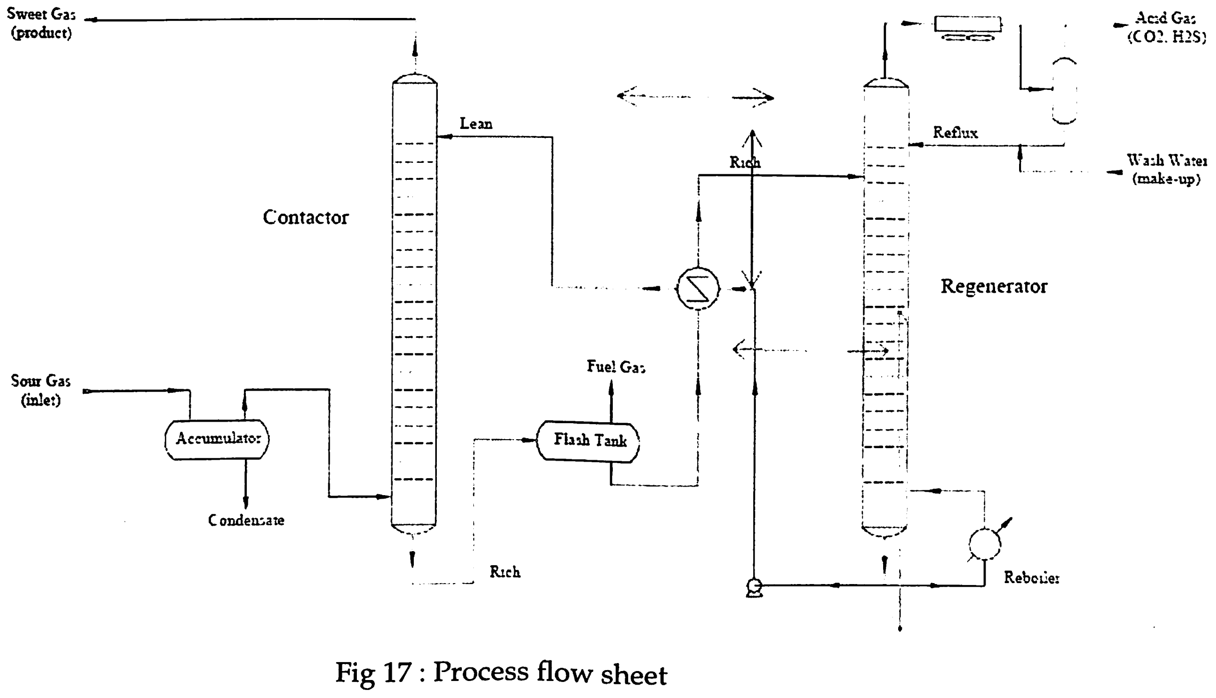


Fig 17 : Process flow sheet

INPUT DATA :		
Piping length to suction :	2	m
Suction static head :	1.5	m
Bends in suction line :	1	
valves in suction line :	1	
Piping length from discharge :	44.5	m
Bends in discharge line :	6	
Valves in discharge line :	3	
Vapor pressure of solvent :	0.1	Kpa
Control valve pressure drop :	140	Kpa
Orifice pressure drop :	15	Kpa
Heat exchanger pressure drop :	70	Kpa

Pipe diameter required :

Lean Sulfinol Density = 1110.8 kg/m^3
Viscosity = 36.916 Cp
Mass flow rate = 18.768 kg/s
Volumetric flow rate = 0.016 m^3 /s
Typical liquid velocity = 2.5 m/s
From table.

	Velocity m/s	ΔP Kpa/m
Liquids, pumped (not viscous)	1 to 3	0.5
Liquids, gravity flow	-	0.05
Gases and vapors	15 - 30	0.02% of line pressure
High-pressure steam, >8 bar	30 - 60	-

Table : 12 (Piping velocity range thumb rule) [2]

Area of pipe = 0.006 m^2
d, Diameter of pipe = (Area * 4 / 3.14) ^.5
= 0.092 m
= 92.788 mm
& with economic diameter formula :

d, optimum = 293 * mass flow ^ .53 * ρ^(-.37)

= 103.491 mm

So, take diameter as = 98 mm

So, cross sectional area = 7593.99 mm^2

= 0.007 m^2

Pressure Drop Calculation :

Fluid velocity with new dia. = 2.224 m/s

Friction loss (Genereaux's formula) :
ΔP = 4.13 *10^10 *mass flow ^1.84 * viscosity ^.16 / (ρ* d^4.84)

= 3.303 Kpa / m

For design criteria a higher value is taken

= 4.293 Kpa/m

Piping Length :

Total length for pump suction = Line Length + (Bend nos. * 30 * d / 1000) + (Valve nos. *18 *d / 1000)

= 6.72 m

Entry loss
= ρ*u^2 / 2

= 3.95 Kpa

Total length from pump discharge = Line Length + (Bend nos. * 30 * d / 1000) + (Valve nos. *18 *d / 1000)

= 71.05 m

Suction Calculation				
Line Size, mm		98		
	Flow	Norm.	Max.	Units
u1	Velocity	2.22	2.66	m/s
ΔP1	Friction Loss	3.303	4.293	Kpa/m
L1	Line length	6.721	-	m
ΔP1 * L1	Line loss	22.206	28.85	Kpa
ρ*u1^2 / 2	Entrance	2.749	3.959	Kpa
(40Kpa)	Strainer	-	-	Kpa
	(1) Sub Total	24.955	32.818	Kpa
z1	Static head	1.5	1.5	m
ρ*g*z1		16.328	16.328	Kpa
	Equip. press.	490.33	490.33	Kpa
	(2) Sub Total	506.661	506.661	Kpa
(2) - (1)	(3) Suction press.	481.705	473.84	Kpa
	(4) Vap. Press	0.1	0.1	Kpa
(3) - (4)	(5) NPSH	481.605	473.742	Kpa
(5) / ρg		44.211	43.489	m

Table 13 :- (Suction calculation of pump)

Discharge Calculation				
Line Size, mm		98		
	Flow	Norm.	Max.	Units
u2	Velocity	2.224	2.669	m/s
ΔP2	Friction Loss	3.303	4.293	Kpa/m
L2	Line length	71.056	-	m
ΔP2 * L2	Line loss	234.766	305.103	Kpa
	Orifice	15	18	Kpa
30%	Control valve	140	168	Kpa
	Heat Exchanger -1	70	84	Kpa
	Heat Exchanger -2	70	84	Kpa
	(6) Dynamic loss	529.766	659.103	Kpa
z2	Static head	11.023	-	m
ρ*g*z1		120.003	120.004	Kpa
	Equip. press.	6706.681	6706.68	Kpa
	(7) Sub Total	6826.685	6826.69	Kpa
(7) + (6)	Discharge press.	7356.451	7485.79	Kpa
(3) suc	Suction pressure	481.705	473.843	Kpa
	(8) Diff. press.	6874.746	7011.95	Kpa
(8) / ρg		631.106	643.701	m

Table 14 :- (Discharge calculation of pump)

Pump Selection :

So, differential head maximum = 646.477806 m

Flowrate of solvent = 102.295251 m³/hr

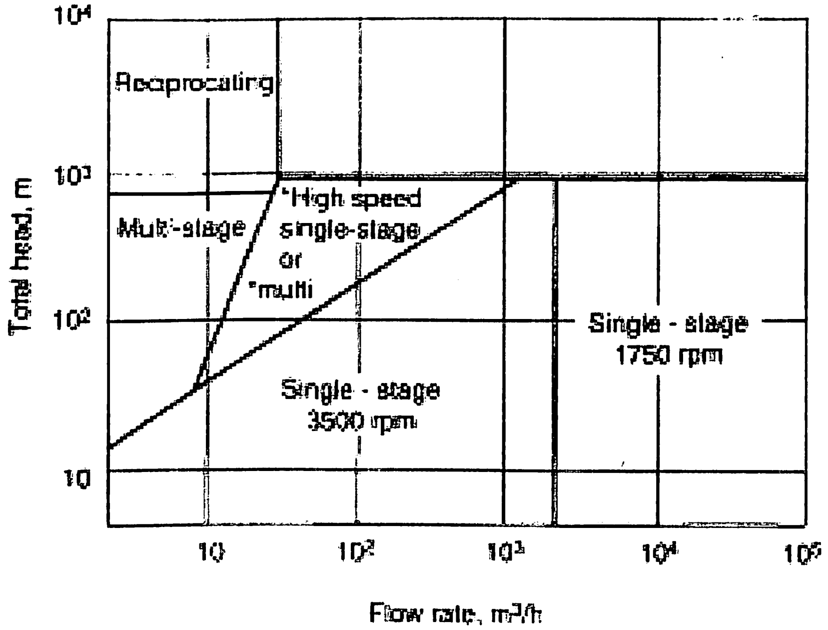


Fig 18 : (Centrifugal pump selection guide. *Single-stage >1750 rpm, multi-stage 1750 rpm)

From the graph we read the type of pump that can be used
High speed single or multistage centrifugal pump.

Power requirements:

$$hp = \left(gpm * \frac{\Delta p(psi)}{1714 * efficiency} \right) \qquad \text{------------ (13.1)}$$

Pressure head:	6874.74	kpa
	997.372	psia
Solvent rate:	267.845	GPM
Pump efficiency:	80	%
Pump Power:	194	HP

14. COMPARISION OF TRAY TYPES

1) For valve tray [20]

Base : $Q_v * (\rho_v/(\rho_l - \rho_v))^{\wedge}.5 =$ 0.079 m^3/s

Top : $Q_v * (\rho_v/(\rho_l - \rho_v))^{\wedge}.5 =$ 0.059 m^3/s

Bottom : Liquid flow = 0.032 m^3/s
Top : Liquid flow = 0.016 m^3/s

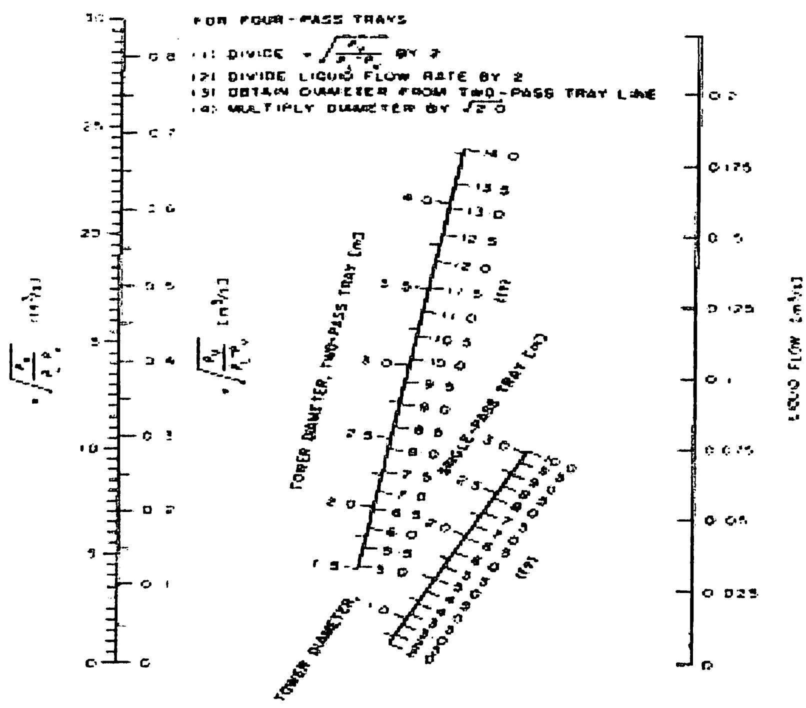


Figure 19. Chart for finding the diameters of valve trays.

Fropm the chart we get :

Bottom :	Single pass tray	**1.52**	m
Top :	Single pass tray	**1.06**	m

2) For Bubble cap tray : Jersey Critical Formula

D = .0956 * (Wv / K*(ρl * ρv)^.5)^.5 ft ----------- (14.1)

W, =vapor flow rate (lb/hr)

Tray Spacing in.--	18	24	30	30+
K	3.2	4.2	4.7	5

K (Spacing 18 in) = 3.2
Bottom, Wv = 139442.9 lb/hr
ρl = 51.16 lb/ft^3
ρv = 3.73 lb/ft^3
Top, Wv = 119210.64 lb/hr
ρl = 69.43 lb/ft^3
ρv = 3.98 lb/ft^3

So, Bottom Diameter = **1.63** m
Top Diameter = **1.37** m

3) For Sieve tray :

Flv bottom = (L/G)*(ρv / ρl)^.5 = 0.36
Flv top = (L/G)*(ρv / ρl)^.5 = 0.29

Typical design values of Kv for Sieve, Bubble-Cap, and Valve Plates

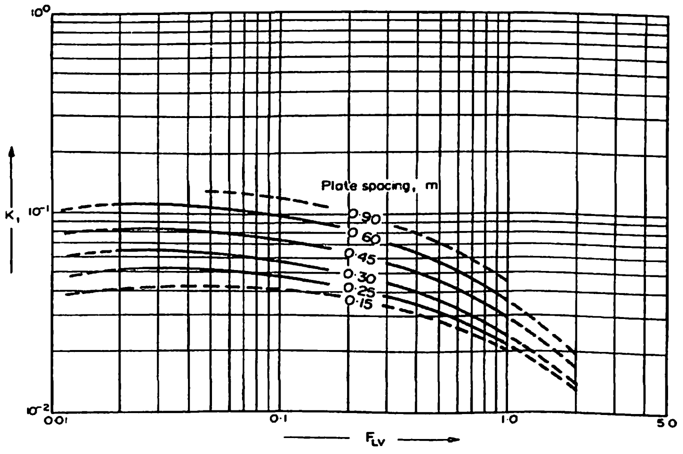


Fig 20 : Flooding velocity, sieve plates

Kv, from table = 0.055 Bottom
Kv, from table = 0.05 Top
for tray spacing = 18 in

So. Base Uf = Kv *((ρl - ρv)/ρv)^.5 = 0.189 m/s

So. Top Uf = Kv *((ρl - ρv)/ρv)^.5 = 0.209 m/s

Design for flooding at maximum velocity of 85 %
Base Uv = Uf * .85 = 0.16 m/s
Top Uv = Uf * .85 = 0.17 m/s

	Top,	1.19	m^2
Area	Base,	1.45	m^2
Diameter	Top,	**1.23**	m
	Base,	**1.36**	m

	Column Diameter, m	
	Top	Bottom
Sieve Tray	**1.23**	**1.36**
Valve Tray	**1.06**	**1.52**
Bubble cap Tray	**1.3**	**1.63**

Table 15 : - (Tray diameter comparison)

15. RESULTS

1) VERTICAL SEPARATOR

Diameter : 36 in
Height : 10 ft
Volume : 22.75 ft^3
Seam to seam length : 9.54 ft
Slenderness ratio : 3.18
Residence Time : 3 min

2) ABSORBER

Diameter = 1.5 m
Top flooding velocity = 0.17 m/s
Top gas flow rate = 0.75 kg.moles/s
Top lean sulfinol flow rate = 0.32 kg.moles/s

Bottom flooding velocity = 0.16 m/s
Bottom gas flow rate = 0.83 kg.moles/s
Bottom rich sulfinol flow rate = 0.40 kg.moles/s

H_2S absorbing cap. of sulfinol= 0.22 moles H_2S / moles solvent
Column Area = 1.76 m^2
Downcomer area = 0.21 m^2
Hole area = 0.13 m^2
Hole diameter = 5 mm
Plate thickness = 5 mm
Wier length, lw = 1.14 m
Wier height, hw = 50 mm
Weir crest, how = 45.25 mm
Total plate pressure drop = 123.94 mm liquid
Downcomer backup measured
From plate surface, mm = 241.99 mm

Residence time = 3.03 s
A, Absorption factor = 2.00
Number of real trays = 19.16
Total height of column = 10.74 m

3) FLASH DRUM

Diameter = 0.98 m
Length = 2.94 m

4) Heat Exchanger

Outside diameter = 0.75 in
Inside diameter = 0.62 in
Length = 16 ft
No. of tubes = 58.90
Tube bundle Dia = 217.75 mm
Re, Renolds no. = 12938.3
Pr, = 5.86
hi, inside heat transfer coeff. = 2735.57 kcal/hr m^2 C
ho,outside heat transfer coeff. = 2261.41 kcal/hr m^2 C

5) Cooler

Outside diameter = 0.75 in
Inside diameter = 0.62 in
Length = 16 ft
No. of tubes = 82.81
Tube bundle Dia = 255.28 mm
Re, Renolds no. = 23192.6
Pr, = 1.72
hi, inside heat transfer coeff. = 2932.33 kcal/hr m^2 C
ho,outside heat transfer coeff. = 2424.06 kcal/hr m^2 C

6) Stripper

Diameter = 4 m
Top flooding velocity = 1.66 m/s
Top steam flow rate = 3.15 kg.moles/s
Top rich sulfinol flow rate = 0.35 kg.moles/s i.e

Bottom flooding velocity = 1.72 m/s
Bottom steam flow rate = 3.12 kg.moles/s
Bottom lean sulfinol flow rate
= 0.32 kg.moles/s i.e

Column Area = 12.56 m^2
Downcomer area = 1.50 m^2
Hole area = 0.95 m^2
Hole diameter = 5 mm
Plate thickness = 5 mm
wier length, lw = 3.2 m
Wier height, hw = 50 mm
Weir crest, how = 24.55 mm
Total plate pressure drop = 175.75 mm liquid
Downcomer backup measured
From plate surface, mm = 253.94 mm
Residence time = 20.19 s
Stripping factor = 0.91
Number of real trays = 22.29
Total height of column = 18.66 m

7) LEAN SULFINOL SUPPLY PUMP

High speed single or multistage centrifugal pump
Differential head maximum = 643.70 m water
Flow rate of solvent = 60.82 m^3 / hr
NPSH = 44.21 m water

16. ENHANCED CORRELATIONS

For the current composition of natural gas to be treated for H_2S removal some correlations are been derived from the plant design calculation for quick reference.

Gas Composition	mol frac
C_1	0.735
C_2	0.083
C_3	0.021
i-C_4	0.006
n-C_4	0.002
i-C_5	0.003
n-C_5	0.008
C_6	0.001
C_{7+}	0.001
N_2	0.050
H_2S	0.090
	1.000

For the case that by the change of operating pressure the partial pressure of the acid gas in natural gas changes, which tends to change in solvent loading. When the solvent loading is changed then the solvent flow rate required also changes. This leads to the change in diameter of the column, number of trays, pump differential head required & pump power required to transport lean solvent.

Therefore some correlations graphs as an extended emphasis are made so as at one look the relative deviation could be determined. This is limited to the give gas composition only.

1) Gas Flow Rate - Sulfinol Flow Rate

9% H2S MMSCF, gas	Sulfinol req, GPM
10	44.641
20	89.282
30	133.923
40	178.56
50	223.204
60	267.845
70	312.486
80	357.127
90	401.768
100	446.408

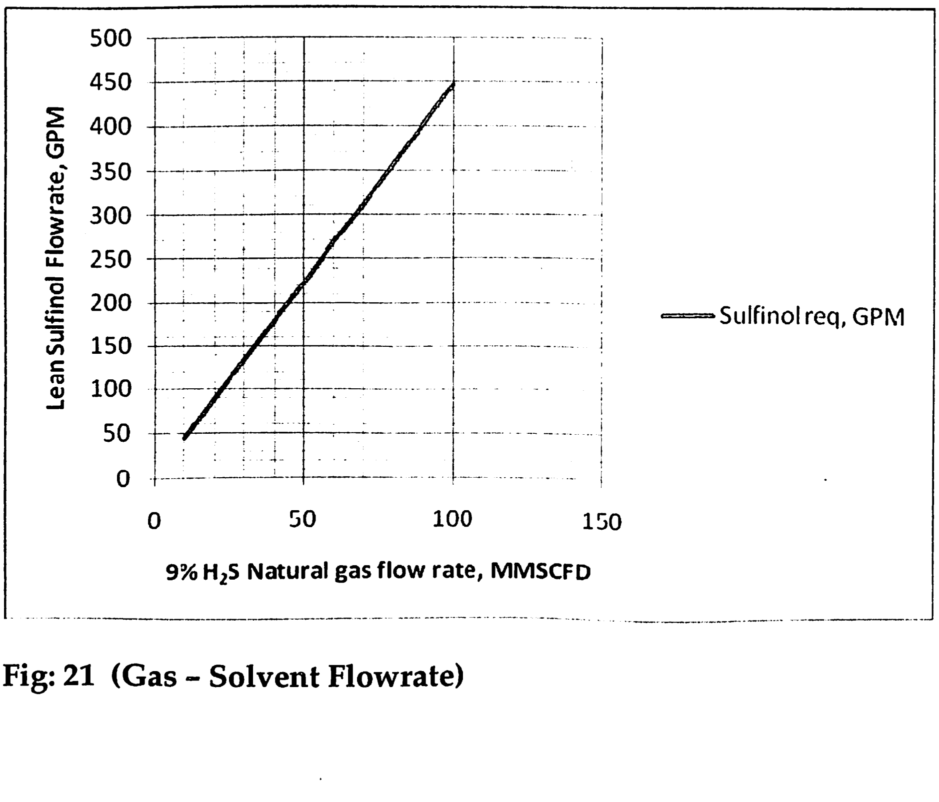


Fig: 21 (Gas - Solvent Flowrate)

2) Operating Pressure - Gas Flow Rate - Sulfinol Flow Rate

Op.	Sulfinol solvent flow rate, GPM							
Press.	10 MMSCFD	20 MMSCFD	30 MMSCFD	40 MMSCFD	50 MMSCFD	60 MMSCFD	70 MMSCFD	80 MMSCFD
100	156.2	312.5	468.73	624.97	781.22	937.46	1093.7	1249.9
200	138.9	277.8	416.65	555.53	694	833.3	972.18	1111.0
300	125	250	374.98	499.98	624.87	749.97	874.96	999.95
400	108	215.8	323.65	431.53	539.41	647.29	755.17	863.05
500	86.31	172.6	258.92	345.22	431.53	517.83	604.14	690.44
600	71.92	143.8	215.76	287.69	359.61	431.53	503.45	575.37
700	61.64	123.3	184.94	246.59	308.23	369.88	431.53	493.17
800	53.94	107.9	161.82	215.76	269.71	323.65	377.59	431.52
900	47.95	95.9	143.84	191.79	239.74	287.69	335.63	383.58
1000	43.15	86.31	129.46	172.61	215.76	258.92	302.07	345.22

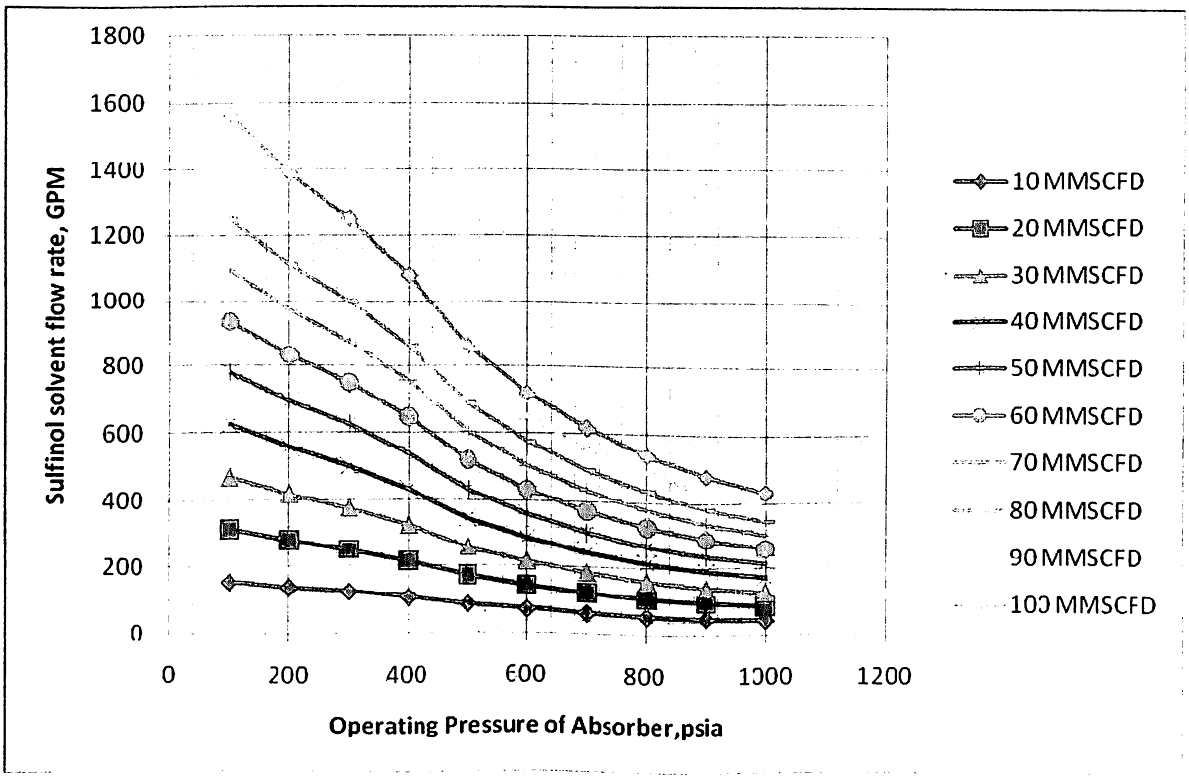


Fig : 21 (Operating Pressure - Gas Flow Rate - Sulfinol Flow Rate)

3) Gas Flow Rate - Sieve Tray Diameter - Bubble Cap Tray Diameter

9% H2S	Sieve Tray Dia. m.			Bubble Tray Dia. m.		
MMSCFD, gas	Top	Bottom	Avg. Sieve Tray Dia.	Top	Bottom	Avg. Bubble Cap Tray Dia.
10	0.50376	0.555686	0.529723	0.56297	0.6677	0.6153215
20	0.71242	0.785858	0.749139	0.79616	0.9442	0.870195
30	0.87253	0.96247	0.9175	0.975095	1.1564	1.0657675
40	1.00751	1.11137	1.05944	1.1259	1.3354	1.230625
50	1.12643	1.24255	1.18449	1.25884	1.493	1.3759
60	1.23394	1.361147	1.2975435	1.37899	1.6355	1.507225
70	1.33281	1.47021	1.40151	1.48948	1.7665	1.62799
80	1.42484	1.57172	1.49828	1.592324	1.8885	1.740392
90	1.51127	1.66705	1.58916	1.68891	2.003	1.845955
100	1.59301	1.757233	1.6751215	1.78027	2.1114	1.94582

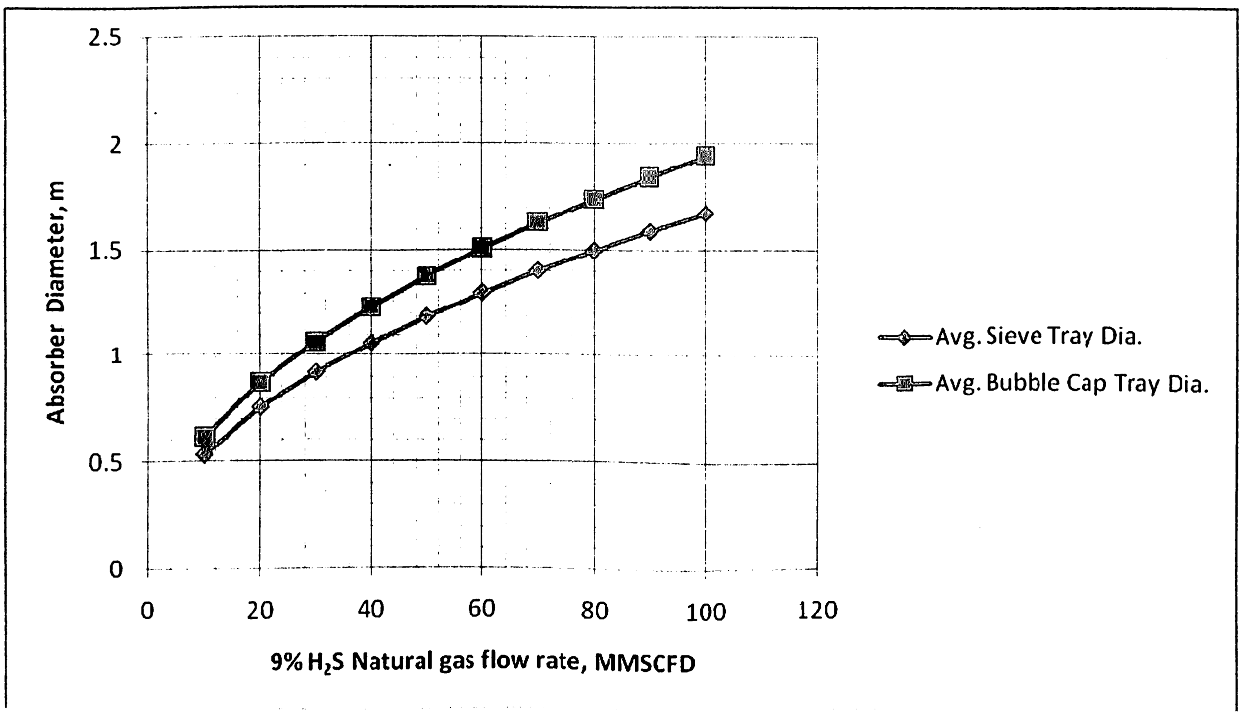


Fig : 22 (Gas Flow Rate - Sieve Tray Diameter - Bubble Cap Tray Diameter)

4) Operation Pressure - Gas Flow Rate - Sieve Tray Diameter

Operating Pressure	Avg. Sieve Tray Dia.									
	10 MMSCFD	20 MMSCFD	30 MMSCFD	40 MMSCFD	50 MMSCFD	60 MMSCFD	70 MMSCFD	80 MMSCFD	90 MMSCFD	100 MMSCFD
100	0.92	1.31	1.60	1.85	2.07	2.26	2.44	2.62	2.77	2.92
200	0.78	1.10	1.35	1.56	1.74	1.91	2.06	2.20	2.34	2.46
300	0.70	1.00	1.22	1.41	1.58	1.73	1.86	1.99	2.11	2.23
400	0.66	0.93	1.14	1.31	1.47	1.61	1.74	1.86	1.97	2.08
500	0.62	0.88	1.08	1.24	1.39	1.52	1.65	1.76	1.87	1.97
600	0.59	0.84	1.03	1.19	1.33	1.46	1.57	1.68	1.78	1.88
700	0.57	0.81	0.99	1.15	1.28	1.40	1.52	1.62	1.72	1.81
800	0.55	0.78	0.96	1.11	1.24	1.36	1.47	1.57	1.66	1.75
900	0.54	0.76	0.93	1.08	1.21	1.32	1.43	1.53	1.62	1.71
1000	0.53	0.74	0.91	1.05	1.18	1.29	1.39	1.49	1.58	1.66

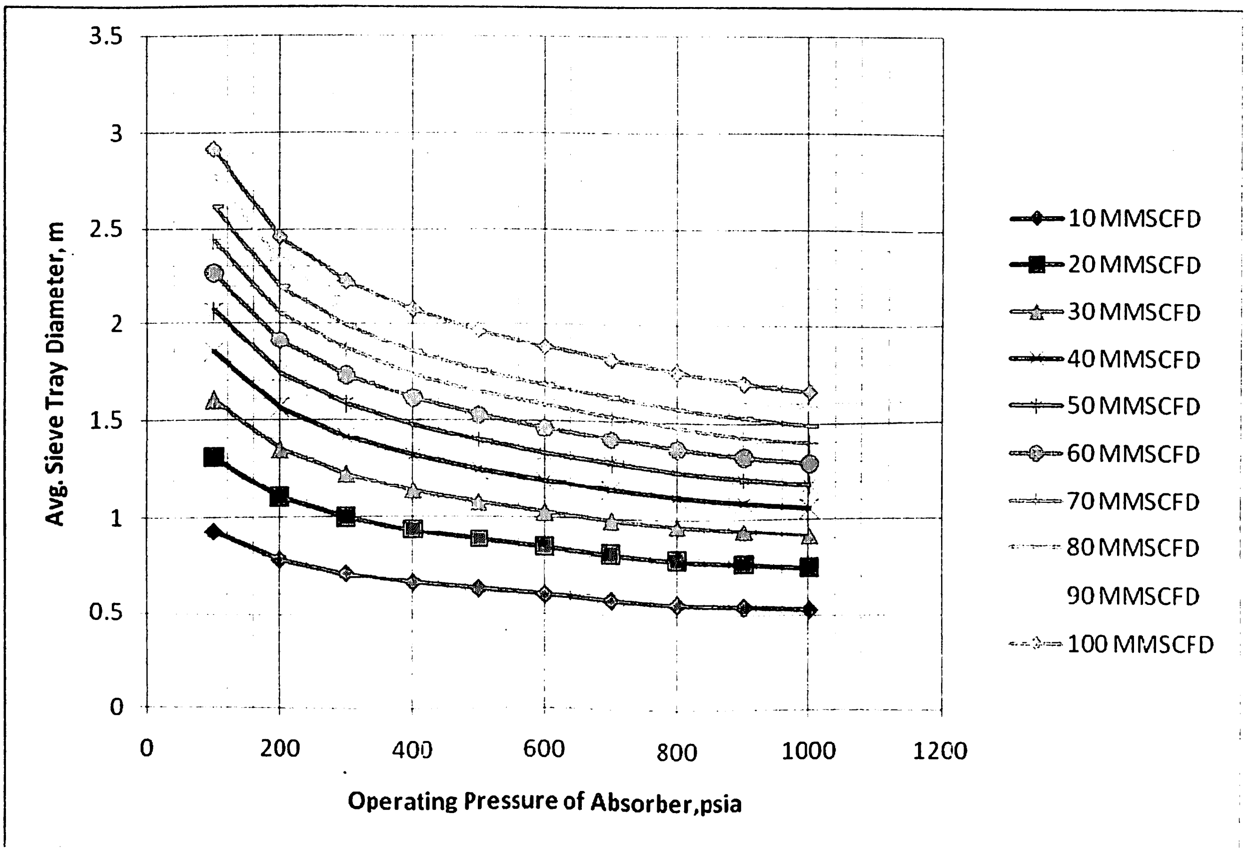


Fig : 23 (Operation Pressure - Gas Flow Rate - Sieve Tray Diameter)

5) Operation Pressure - Gas Flow Rate - Bubble Cap Tray Diameter

Op.	Avg. Bubble Cap Tray Dia.									
Pres	10 MMSCFD	20 MMSCFD	30 MMSCFD	40 MMSCFD	50 MMSCFD	60 MMSCFD	70 MMSCFD	80 MMSCFD	90 MMSCFD	100 MMSCFD
100	1.0892	1.54	1.886	2.178	2.435	2.668	2.882	3.081	3.436	3.444
200	0.9165	1.296	1.587	1.833	2.049	2.245	2.425	2.592	2.903	2.898
300	0.8284	1.172	1.435	1.657	1.852	2.029	2.192	2.343	2.634	2.62
400	0.7709	1.09	1.335	1.542	1.724	1.888	2.04	2.18	2.461	2.438
500	0.729	1.031	1.263	1.458	1.63	1.786	1.929	2.062	2.336	2.305
600	0.6962	0.985	1.206	1.392	1.557	1.705	1.842	1.969	2.239	2.202
700	0.6695	0.947	1.16	1.339	1.497	1.64	1.771	1.893	2.16	2.117
800	0.6471	0.915	1.121	1.294	1.447	1.585	1.712	1.83	2.095	2.046
900	0.6278	0.888	1.087	1.256	1.404	1.538	1.661	1.776	2.039	1.985
1000	0.611	0.864	1.058	1.222	1.366	1.497	1.627	1.728	1.99	1.932

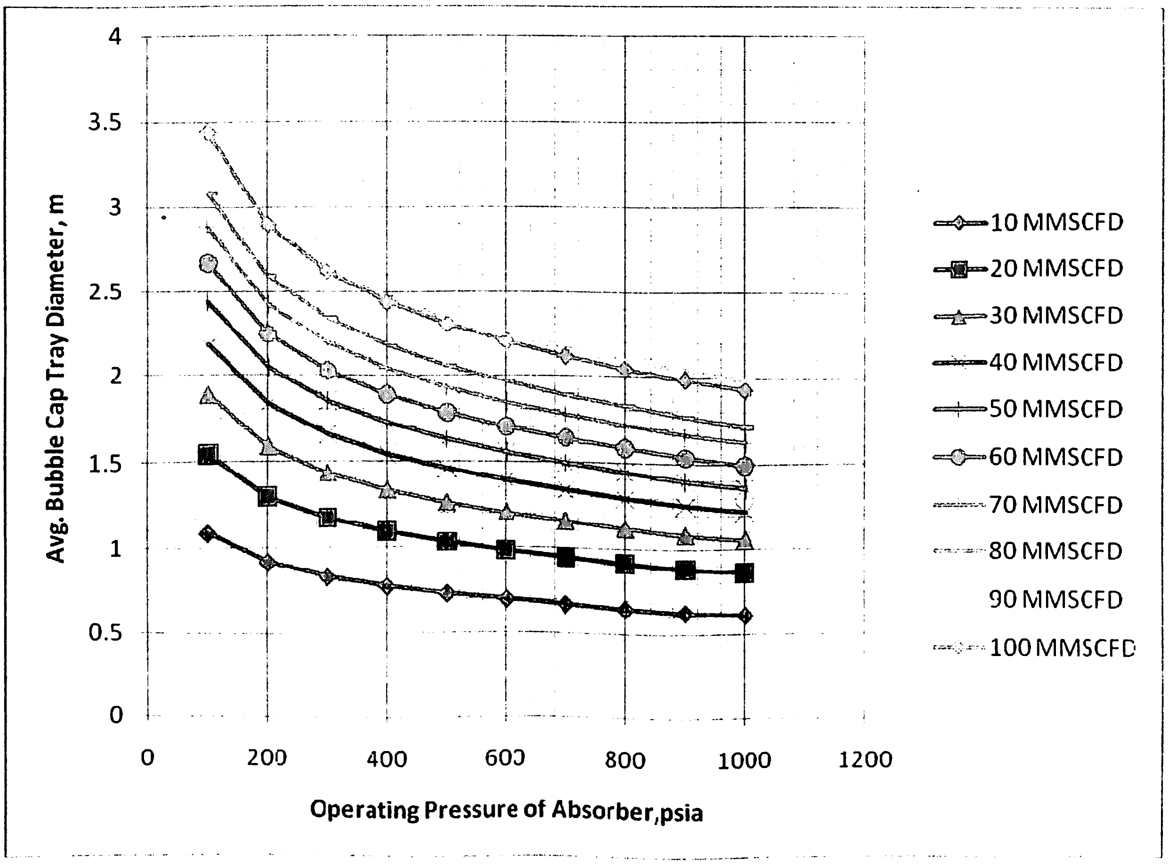


Fig : 24 (Operation Pressure - Gas Flow Rate - Bubble Cap Tray Diameter)

6) Operation Pressure - Gas Flow Rate - Pump Differential Head

Op. Pres	Pump Differential Head									
	10 MMSCFD	20 MMSCFD	30 MMSCFD	40 MMSCFD	50 MMSCFD	60 MMSCFD	70 MMSCFD	80 MMSCFD	90 MMSCFD	100 MMSCFD
100	127	109	101	97	94	91	90	88	87	86
200	190	172	164	160	157	155	153	151	150	149
300	253	235	228	223	220	218	216	215	214	213
400	317	299	291	286	283	281	279	278	277	276
500	380	362	354	350	347	344	343	341	340	339
600	443	425	417	413	410	408	406	405	403	402
700	506	488	481	476	473	471	469	468	467	466
800	570	552	544	540	536	534	533	531	530	529
900	633	615	607	603	600	598	596	594	593	592
1000	696	678	671	666	663	661	659	658	657	656

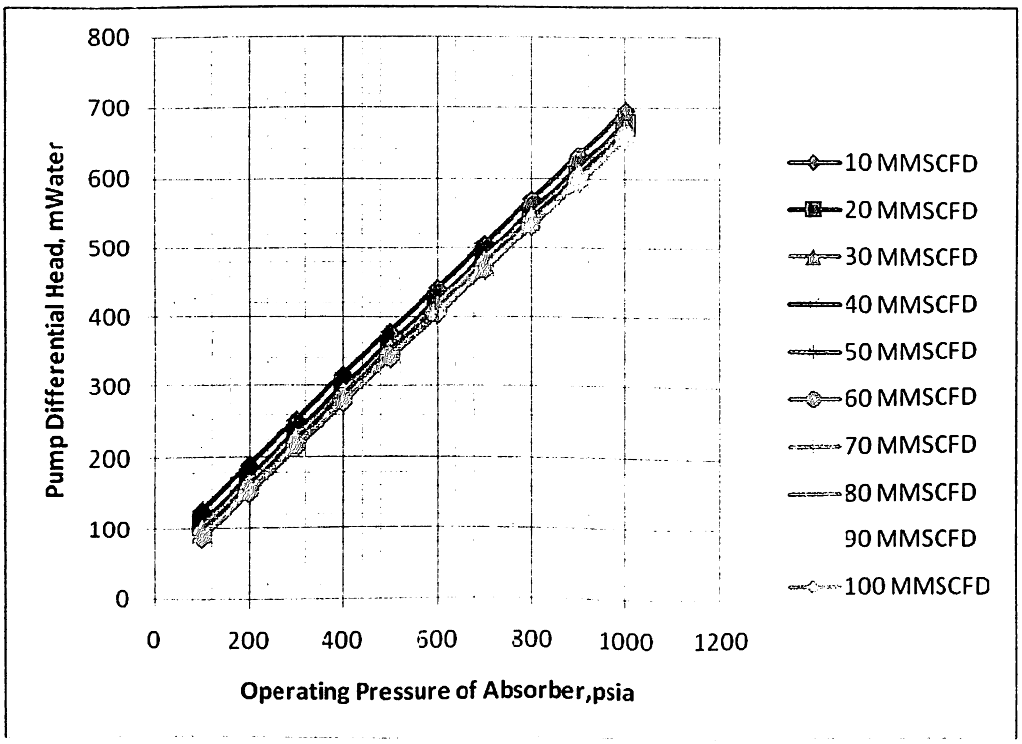


Fig : 25 (Operation Pressure - Gas Flow Rate - Pump Differential Head)

7) Operation Pressure - Gas Flow Rate - Pump Power

Op. Pres	Pump Power in HP									
	10 MMSCFD	20 MMSCFD	30 MMSCFD	40 MMSCFD	50 MMSCFD	60 MMSCFD	70 MMSCFD	80 MMSCFD	90 MMSCFD	100 MMSCFD
100	5.4504	9.482	13.31	17.03	20.69	24.3	27.87	31.41	34.94	38.44
200	8.706	15.99	23.08	30.05	36.97	43.83	50.66	57.46	64.24	70.99
300	11.962	22.5	32.84	43.08	53.24	63.36	73.45	83.5	93.54	103.6
400	15.217	29.02	42.61	56.1	69.52	82.9	96.24	109.5	122.8	136.1
500	18.473	35.53	52.38	69.12	85.8	102.4	119	135.6	152.1	168.7
600	21.728	42.04	62.14	82.14	102.1	122	141.8	161.6	181.4	201.2
700	24.984	48.55	71.91	95.17	118.4	141.5	164.6	187.7	210.7	233.8
800	28.24	55.06	81.68	108.2	134.6	161	187.4	213.7	240	266.3
900	31.495	61.57	91.44	121.2	150.9	180.6	210.2	239.8	269.3	298.9
1000	34.751	68.08	101.2	134.2	167.2	200.1	233	265.8	298.6	331.4

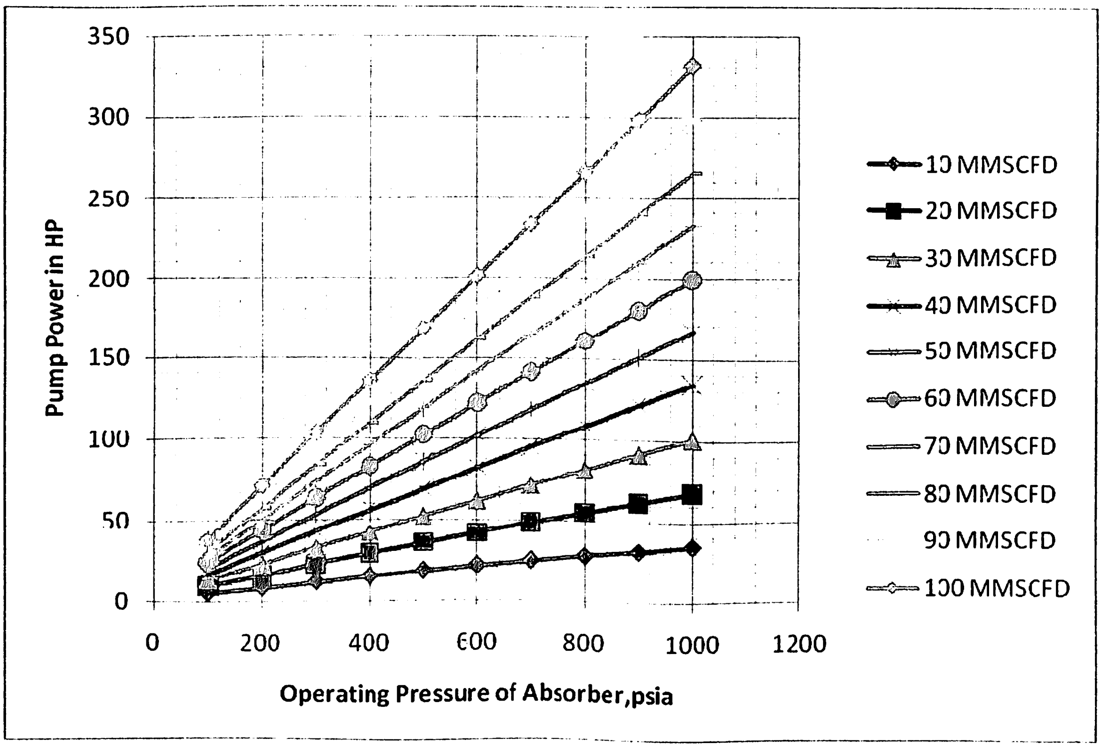


Fig: 26 (Operation Pressure - Gas Flow Rate - Pump Power)

17. SIMULATION OF DIPA PLANT USING HYSYS

This is the part 2 section of the project where the same composition of natural gas at 1000 psia & 43°C flowing at 20.0786 MMSCFD is treated for the removal of H_2S using DIPA as a solvent. But the plant design is carried out by simulation package HYSYS an ASPEN technology product. This is performed for getting a fair comparison between the hand calculation results and simulation result of the plant design. The detailed calculation result sheets are given in the appendix.

Basis of calculation by HYSYS :

Amine property package is a special property package designed to aid in the modeling of alkanolamine treating units in which H2S & CO2 are removed from gas streams. The property package contains data to model the absorption/desorption process where aqueous solution of amines, like MEA, DEA, MDEA, DIPA are used.

The following considerations are taken for design of the treatment plant :

- The process configuration
- The amine concentration
- The solution circulation rate
- The reboiler heat requirements
- The operating pressure & temperature

The AMSIM uses technology developed by DB Robinson & Associates ltd. To model the equilibrium solubility of acid gas in aqueous amine solutions.

Stage efficiency :

As defined by the amine property package is given by :-

$$\eta = \frac{(V_j + SV_j)Y_j - V_{j+1}Y_{ij+1}}{(V_j + SV_j)K_{ij}X_{ij} - V_{j+1}Y_{ij-1}} \quad \text{----------------------(17.1)}$$

Where:

η = Stage efficiency
i = Component number
j = Stage no.
K = Equilibrium ratio
V = Molar flow rate of vapor
X = Mole fraction in liquid phase
Y = Mole fraction in vapor phase

Equilibrium solubility of acid gas in amine solution :

- Kent & Eisenberg Model – This is used to correlate the equilibrium solubility of acid gas in amine solution. Improvement are also made to the model to extend the reliable range to mole loading between .0001 and 1.2. A proprietary model was developed to predict the solubility of acid gas mixtures in tertiary amine solutions. Solubility of inert component such as hydrocarbon are modeled using a Henry's constant adjusted for ionic strength effects.
- Li- Mather Electrolyte model – Amine property package is modified to simulate 3-phase behavior. For the three phase simulation. The K values from the Peng - Robinson property package were combined with the K values from the amines LLE & VLE package. This model shows a strong predictive capability over a wide range of temperatures, pressures, acid gas loadings & amine concentrations.

The vapor liquid equilibrium of the molecular species is given by:

$$y_i \Phi_i^{V} P = H_i x_i \gamma_i^{L}$$ ------------------(17.2)

Where; Hi = Henry's constant

P = System Pressure

Xi, yi = mole fraction of molecular species i

Φ_i^v = fugacity coefficient on the gas phase

γ_i^L = activity coefficient in the liquid phase

Fugacity coefficient is calculated by the Peng- Robinson equation of states, 1976 :

$$P = \frac{RT}{V-b} - \frac{a(T)}{V(V+b)+b(V-b)}$$ -------------------(17.3)

Brief comparison of results :

	Hand Cal.	Simulation
Absorber		
Diameter =	1.5 m	1.5 m
Tray Spacing =	0.45 m	0.5 m
Wier length, lw =	1.14 m	1.2 m
Wier height, hw =	50 mm	50 mm
Plate pressure drop =	1.21 kPa	4.7 kPa
Number of real trays =	19	20
Heat exchanger		
Tube Outside diameter =	19 mm	20 mm
Tube Thickness =	3.3 mm	2mm
Tube Length =	4.8 m	6 m
No. of tubes =	59 nos.	160 nos.
Stripper		
Diameter =	4 m	1.5 m
Tray Spacing =	0.76 m	0.55 m
wier length, lw =	3.2 m	1.2 m
Wier height, hw =	50 mm	50 mm
Number of real trays =	22 nos.	18 nos.

18. BRIEF DISCUSTION

When natural gas is produced from any reservoir, it contains some undesirable contents. H_2S is one of those, & natural gas containing it is called sour. Therefore before selling natural gas to the market it is treated to an acceptable limit for H_2S concentration. Now on the basis of the partial pressure of H_2S in natural gas an economical method is identified for the treatment process. Here we have chosen Sulfinol process for the treatment process.

To start with the plant design, 1st the aim of removal is identified so as the amount of H_2S to be removed is calculated. Now both the sour & sweet gas properties are calculated for different composition i.e density, viscosity, compressibility factor. After that the solvent loading is calculated from the amount of H_2S to be removed, implying to the solvent flow rate both lean & rich. Now the properties of the same solvent at different composition is looked from the data book. When both the top & bottom, liquid & gas flow rate is shown then the Flv is calculated for the diameter of column with a selected tray spacing. After the diameter calculation from the Souder & Brown's equation with relevant consideration of flooding velocity, the tray sizing is done. Different considerations are taken for safe operation in tray viz. weeping correction, entrainment check & pressure drop calculation. Now the downcomer liquid backup is calculated for proper operation i.e under satisfactory range, otherwise tray spacing is changed and checked again. Again the calculation is repeated for froth density consideration which comes about .4 to .7 times of the liquid density. This calculation is again checked for the downcomer liquid backup to be under satisfactory limits. Next tray (Murphy) efficiency is calculated and eventually the number of trays are even calculated by application of suitable absorption factors.

A material of construction of shell is chosen which is suitable for the given operation. Due all consideration of the material properties and load considerations the thickness of the shell is calculated.

Next the knockout drum dimensions are calculated for a given amount of pressure reduction, wherein a good amount of H_2S is removed (Solvent contains Sulfolane). Here an assumption is taken that no hydrocarbon is released in the knockout drum.

Lean/Rich heat exchanger is designed for the amount of heat duty calculated and a non corrosive material is chosen. In the same consideration cooler is also designed as the lean sulfinol out from the bottom of the stripper needs to be cooled before letting it into the absorber top (10°C bellow the absorber temperature).

Stripper calculation is done for the diameter, height, number of trays, check for weeping, entrainment check, pressure drop, downcomer liquid backup, shell thickness etc . As for a situation of low pressure in it, the diameter which is calculated comes higher than the absorber.

19. CONCLUSIONS

- In any acid gas treatment plant the main criteria of process selection depends on the location of the plant, giving the knowledge of the H_2S content of natural gas which is going to be available to the plant.

- For high concentration of H_2S in natural gas (i.e more than 60 psia partial pressure of H_2S) Sulfinol process is the most economical (especially in absence of CO_2).

- It is more economical to use a packed column as an absorber if gas flow rate is within 10 to 40 MMSCFD.

- It is more economical to use a bubble cap tray column for gas flow rate more than 60 MMSCFD.

- Account for corrosion allowance in shell thickness calculation is to be taken higher for bottom of the column.

- With low operating pressure in absorber, use of packed tower is economical above 200psia.

- At high operating pressure of absorber, differential head of the pump required do not deviate much for different gas flow rate whereas the power required by the same pump deviate big for different flow rates.

- In part - 2 of the project, simulating the whole plant gives an accurate idea for process convergence which is comparable to sieve tray, bubble cap tray, valve tray & packing.

20. REFERENCES

[1] Arthur L. Kohl and Richard Nielsen, 'Gas purification. - 5th ed.' , Gulf Publishing Company, 1997, Houston, TX, ISBN: 0-88415-220-0

[2] R.K. Sinnott, 'Coulson & Richardson's Chemical Engineering Design. - 3rd ed. Vol-6', Published by Butterworth Heinemann, 2003, Oxford ISBN 0 7506 41428.

[3] H.K. Abdel-Aal and Mohamed Aggour , 'Petroleum & Gas Field Processing', Published by Marcel Dekker Inc, 2003, New York, ISBN: 0-8247-0962-4

[4] Design & Operation of a Selective Sweetening Plant Using MDEA, Technical Papers, Bryan Research and Engineering, Inc, 2008

[5] Boyun Gou & Ali Ghalambor, 'Natural Gas Engineering Handbook', Gulf Publishing Company, 2005, Houston, TX ISBN 0-9765113-3-9.

[6] Robert E. Treybal, 'Mass Transfer Operations. - 3rd ed', 2002 McGraw-Hill Publications.

[7] Donald Q. Kern, 'Process Heat Transfer', McGraw-Hill Publications, 1950, Singapore, ISBN 0-07-0853533

[8] Design and Operations of Natural Gas Sweetening Facilities Gas Sweetening Facilities. - National Iranian Gas Company workshop, June-2008.

[9] Solari, R. B., and R.L Bell, AICHE Journal, 1986.

[10] E. R Freitas et al., Method of separating acid gases from gaseous mixtures, U.S Patent - 3463603, Aug 26, 1969

[11] Erik S. E. Werner et al, Process for the removal of hydrogen sulfide from gas stream, U.S Patent - 4241032, Dec 23, 1980

[12] James D Yearout, Gas treating process & system, U.S Patent - 3594983, Jul 27, 1969.

[13] Kenneth H Bacon, Natural gas treatment, U.S Patent - 4128410, Dec 5, 1978.

[14] Michael N. Papadopoulos et al., Method of separating acid gases from gaseous mixture, U.S Patent - 3347621, Oct17, 1967

[15] Charles R Ferrin et.al, Physical solvent for gas sweetening, U.S Patent - 4360363, Nov 23, 1982

[16] Thomas H Baize, Method and apparatus for sweetening natural gas, U.S Patent - 4748011, May 31, 1988.

[17] Paul J Capobianco et.al, Mixed solvent system for treating acid gas, U.S Patent - 4705673, Nov 10, 1988

[18] Robert D Fish et. al, Configuration & method of acid gas removal, U.S Patent Application - US 2006 /0110300 A1, May 25, 2006

[19] Kaaeid A Lokhandwala et.al, Natural gas treatment process for stimulated well., U.S Patent Application - US 2007 /0125537 A1, Jun 7, 2007

[20] Bulletin 4900, Glitsch Inc., Dallas, TX,1974.

[21] Kh. Mohamadbigy, M. Bazmi. R, Amine absorption column design using mass transfer rate simulation, Tehran , November 11, 2005, ISSN 1337-7027

[22] HYSYS 3.2 - Simulation basis, Hyprotech, Aspen technology Inc., Cambridge, U.S.A, 2003

21. APPENDIX

HYSYS simulation results for the following DIPA plant using the following gas ----

Gas Composition	mol frac
C_1	0.73500
C_2	0.08300
C_3	0.02100
$i\text{-}C_4$	0.00600
$n\text{-}C_4$	0.00200
$i\text{-}C_5$	0.00300
$n\text{-}C_5$	0.00800
C_6	0.00100
C_{7+}	0.00100
N_2	0.05000
H_2S	0.09000
	1.00000

Pressure: (psia)	**1000 psia**
TEMPERATURE(°F)	**109.4°F**
Flow rate : (MMSCFD)	**20.0786**

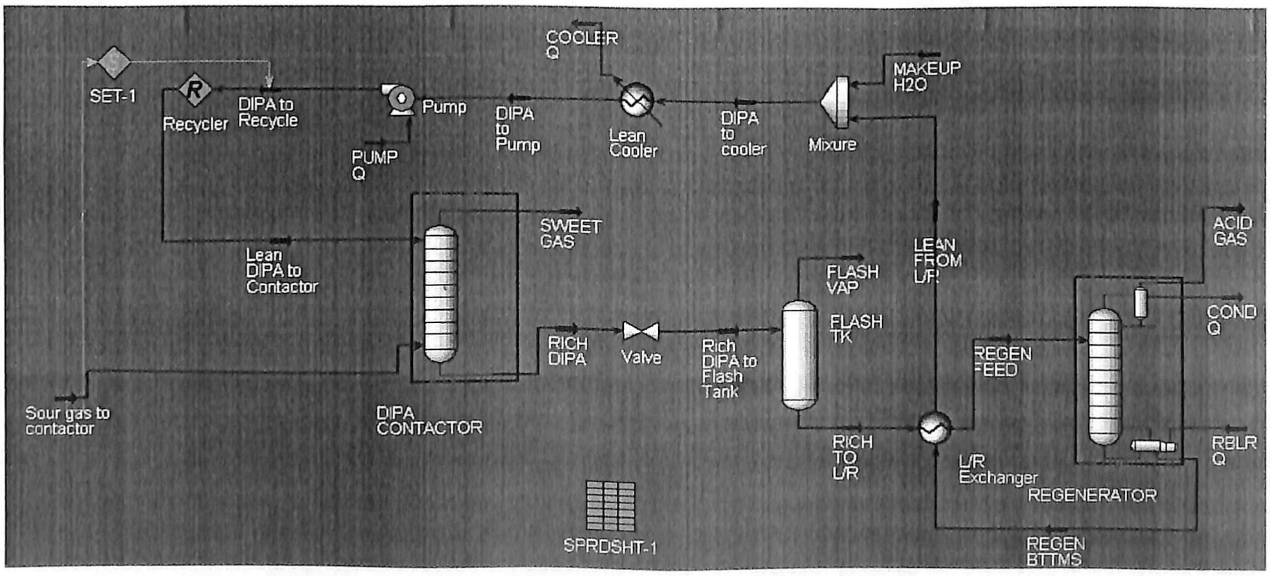


Fig 28 :- Developed PFD for DIPA plant using HYSYS

HYPROTECH LIFECYCLE INNOVATION	TEAM LND Calgary, Alberta CANADA	Case Name:	E:\PRAKASH MUKHOPADHYAY -- DIPA.HSC
		Unit Set:	SI
		Date/Time:	Mon May 11 08:30:11 2009

Absorber: DIPA CONTACTOR @Main

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION	
SOUR GAS TO CONT	20__TS-1		
DIPA TO CONT	1__TS-1	Recycle	Recycler

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION	
SWEET GAS	1__TS-1		
RICH DIPA FROM CONT	20 TS-1	Valve	Valve

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used

SPECS

Column Specification Parameters

Fixed / Ranged:	Primary / Alternate:	Lower Bound:	Upper Bound:

SUBCOOLING

Degrees of Subcooling		
Subcool to		

User Variables

PROFILES

General Parameters

Sub-Flow Sheet:	DIPA CONTACTOR (COL1)	Number of Stages:	20 *

Profile Estimates

	Temperature (C)	Net Liquid (kgmole/h)	Net Vapour (kgmole/h)
1__TS-1	37.78	1649	909.6
2__TS-1	34.97	1649	911.5
3__TS-1	34.97	1649	911.5
4__TS-1	34.97	1649	911.5
5__TS-1	34.97	1649	911.5
6__TS-1	34.98	1649	911.5
7__TS-1	34.98	1649	911.5
8__TS-1	34.99	1649	911.6
9__TS-1	35.00	1649	911.6
10__TS-1	35.04	1649	911.6
11__TS-1	35.11	1649	911.7
12__TS-1	35.26	1650	911.8
13__TS-1	35.57	1650	912.1
14__TS-1	36.23	1651	912.6
15__TS-1	37.54	1654	913.8
16__TS-1	40.10	1659	916.3
17__TS-1	44.70	1670	921.5
18__TS-1	51.71	1689	932.1
19__TS-1	59.00	1717	951.6
20__TS-1	71.11	1738	978.8

EFFICIENCIES

Stage Efficiencies

Stages	Overall Efficiency	Nitrogen	CO2	H2S	Methane	Ethane
1__TS-1	1.000	1.000	0.1401	0.5261	1.000	1.000
2__TS-1	1.000	1.000	0.1402	0.5279	1.000	1.000
3__TS-1	1.000	1.000	0.1402	0.5279	1.000	1.000
4__TS-1	1.000	1.000	0.1403	0.5279	1.000	1.000
5__TS-1	1.000	1.000	0.1403	0.5279	1.000	1.000
6__TS-1	1.000	1.000	0.1403	0.5279	1.000	1.000

Absorber: DIPA CONTACTOR @Main (continued)

Stage Efficiencies

Stages	Overall Efficiency	Nitrogen	CO2	H2S	Methane	Ethane
7__TS-1	1.000	1.000	0.1403	0.5279	1.000	1.000
8__TS-1	1.000	1.000	0.1403	0.5279	1.000	1.000
9__TS-1	1.000	1.000	0.1404	0.5279	1.000	1.000
10__TS-1	1.000	1.000	0.1405	0.5280	1.000	1.000
11__TS-1	1.000	1.000	0.1407	0.5280	1.000	1.000
12__TS-1	1.000	1.000	0.1412	0.5280	1.000	1.000
13__TS-1	1.000	1.000	0.1421	0.5281	1.000	1.000
14__TS-1	1.000	1.000	0.1439	0.5283	1.000	1.000
15__TS-1	1.000	1.000	0.1475	0.5286	1.000	1.000
16__TS-1	1.000	1.000	0.1543	0.5291	1.000	1.000
17__TS-1	1.000	1.000	0.1656	0.5297	1.000	1.000
18__TS-1	1.000	1.000	0.1791	0.5297	1.000	1.000
19__TS-1	1.000	1.000	0.1887	0.5284	1.000	1.000
20__TS-1	1.000	1.000	0.1547	0.5259	1.000	1.000

Stages	Overall Efficiency	Propane	i-Butane	n-Butane	i-Pentane	n-Pentane
1__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
2__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
3__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
4__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
5__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
6__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
7__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
8__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
9__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
10__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
11__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
12__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
13__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
14__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
15__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
16__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
17__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
18__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
19__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
20__TS-1	1.000	1.000	1.000	1.000	1.000	1.000

Stages	Overall Efficiency	n-Hexane	n-Heptane	H2O	DIsoPAmine	
1__TS-1	1.000	1.000	1.000	1.000	1.000	
2__TS-1	1.000	1.000	1.000	1.000	1.000	
3__TS-1	1.000	1.000	1.000	1.000	1.000	
4__TS-1	1.000	1.000	1.000	1.000	1.000	
5__TS-1	1.000	1.000	1.000	1.000	1.000	
6__TS-1	1.000	1.000	1.000	1.000	1.000	
7__TS-1	1.000	1.000	1.000	1.000	1.000	
8__TS-1	1.000	1.000	1.000	1.000	1.000	
9__TS-1	1.000	1.000	1.000	1.000	1.000	
10__TS-1	1.000	1.000	1.000	1.000	1.000	
11__TS-1	1.000	1.000	1.000	1.000	1.000	
12__TS-1	1.000	1.000	1.000	1.000	1.000	
13__TS-1	1.000	1.000	1.000	1.000	1.000	
14__TS-1	1.000	1.000	1.000	1.000	1.000	
15__TS-1	1.000	1.000	1.000	1.000	1.000	
16__TS-1	1.000	1.000	1.000	1.000	1.000	
17__TS-1	1.000	1.000	1.000	1.000	1.000	
18__TS-1	1.000	1.000	1.000	1.000	1.000	
19__TS-1	1.000	1.000	1.000	1.000	1.000	
20__TS-1	1.000	1.000	1.000	1.000	1.000	

SOLVER

Absorber: DIPA CONTACTOR @Main (continued)

Column Solving Algorithm: HYSIM Inside-Out

Save Solutions as Initial Estimate:	On
Super Critical Handling Model:	Simple K
Trace Level:	Low
Init from Ideal K's:	Off

Initial Estimate Generator Parameters

Iterative IEG (Good for Chemicals):	Off

Damping Parameters

Azeotrope Check:	Off
Fixed Damping Factor:	1

RATING

Tray Sections

Tray Section		TS-1			
Tray Diameter	(m)	1.500 *			
Weir Height	(m)	5.000e-002 *			
Weir Length	(m)	1.200 *			
Tray Space	(m)	0.5000 *			
Tray Volume	(m3)	0.8836			
Disable Heat Loss Calculations		No			
Heat Model		None			
Rating Calculations		No			
Tray Hold Up	(m3)	8.836e-002			

Vessels

Vessel				
Diameter				
Length				
Volume				
Orientation				
Vessel has a Boot				
Boot Diameter				
Boot Length				
Hold Up				

Other Equipment In Column Flowsheet

Pressure Profile

	Pressure (kPa)	Pressure Drop (kPa)
1__TS-1	6619 kPa *	4.717 kPa
2__TS-1	6624 kPa	4.717 kPa
3__TS-1	6628 kPa	4.717 kPa
4__TS-1	6633 kPa	4.717 kPa
5__TS-1	6638 kPa	4.717 kPa
6__TS-1	6643 kPa	4.717 kPa
7__TS-1	6647 kPa	4.717 kPa
8__TS-1	6652 kPa	4.717 kPa
9__TS-1	6657 kPa	4.717 kPa
10__TS-1	6661 kPa	4.717 kPa
11__TS-1	6666 kPa	4.717 kPa
12__TS-1	6671 kPa	4.717 kPa
13__TS-1	6676 kPa	4.717 kPa
14__TS-1	6680 kPa	4.717 kPa
15__TS-1	6685 kPa	4.717 kPa
16__TS-1	6690 kPa	4.717 kPa
17__TS-1	6694 kPa	4.717 kPa
18__TS-1	6699 kPa	4.717 kPa
19__TS-1	6704 kPa	4.717 kPa
20__TS-1	6709 kPa *	---

Pressure Solving Options

Pressure Tolerance	1.000e-004 *	Pressure Drop Tolerance	1.000e-004 *	Damping Factor	1.000 *	Max Press Iterations	100 *

PROPERTIES

Absorber: DIPA CONTACTOR @Main (continued)

Properties : DIPA TO CONT

		Overall	Aqueous Phase			
Vapour/Phase Fraction		0.0000	1.0000			
Temperature:	(C)	35.00	35.00			
Pressure:	(kPa)	6860	6860			
Molar Flow	(kgmole/h)	1647	1647			
Mass Flow	(kg/h)	4.319e+004	4.319e+004			
Std Ideal Liq Vol Flow	(m3/h)	43.27	43.27			
Molar Enthalpy	(kJ/kgmole)	-2.536e+004	-2.536e+004			
Mass Enthalpy	(kJ/kg)	-967.2	-967.2			
Molar Entropy	(kJ/kgmole-C)	94.17	94.17			
Mass Entropy	(kJ/kg-C)	3.592	3.592			
Heat Flow	(kJ/h)	-4.178e+007	-4.178e+007			
Molar Density	(kgmole/m3)	38.81	38.81			
Mass Density	(kg/m3)	1018	1018			
Std Ideal Liq Mass Density	(kg/m3)	998.3	998.3			
Liq Mass Density @Std Cond	(kg/m3)	1022	1022			
Molar Heat Capacity	(kJ/kgmole-C)	94.17	94.17			
Mass Heat Capacity	(kJ/kg-C)	3.592	3.592			
Thermal Conductivity	(W/m-K)	0.4369	0.4369			
Viscosity	(cP)	3.076	3.076			
Surface Tension	(dyne/cm)	59.43	59.43			
Molecular Weight		26.22	26.22			
Z Factor		6.899e-002	6.899e-002			

Properties : SOUR GAS TO CONT

		Overall	Vapour Phase			
Vapour/Phase Fraction		1.0000	1.0000			
Temperature:	(C)	43.00	43.00			
Pressure:	(kPa)	6895	6895			
Molar Flow	(kgmole/h)	1000	1000			
Mass Flow	(kg/h)	2.113e+004	2.113e+004			
Std Ideal Liq Vol Flow	(m3/h)	56.22	56.22			
Molar Enthalpy	(kJ/kgmole)	1.489e+004	1.489e+004			
Mass Enthalpy	(kJ/kg)	704.9	704.9			
Molar Entropy	(kJ/kgmole-C)	196.0	196.0			
Mass Entropy	(kJ/kg-C)	9.276	9.276			
Heat Flow	(kJ/h)	1.489e+007	1.489e+007			
Molar Density	(kgmole/m3)	3.164	3.164			
Mass Density	(kg/m3)	66.85	66.85			
Std Ideal Liq Mass Density	(kg/m3)	375.8	375.8			
Liq Mass Density @Std Cond	(kg/m3)	---	---			
Molar Heat Capacity	(kJ/kgmole-C)	42.07	42.07			
Mass Heat Capacity	(kJ/kg-C)	1.991	1.991			
Thermal Conductivity	(W/m-K)	3.114e-002	3.114e-002			
Viscosity	(cP)	1.181e-002	1.181e-002			
Surface Tension	(dyne/cm)	---	---			
Molecular Weight		21.13	21.13			
Z Factor		0.8289	0.8289			

Properties : RICH DIPA FROM CONT

		Overall	Vapour Phase	Aqueous Phase		
Vapour/Phase Fraction		0.0000	0.0000	1.0000		
Temperature:	(C)	59.95	59.95	59.95		
Pressure:	(kPa)	6709	6709	6709		
Molar Flow	(kgmole/h)	1738	0.0000	1738		
Mass Flow	(kg/h)	4.627e+004	0.0000	4.627e+004		
Std Ideal Liq Vol Flow	(m3/h)	47.22	0.0000	47.22		
Molar Enthalpy	(kJ/kgmole)	-2.331e+004	1.571e+004	-2.331e+004		
Mass Enthalpy	(kJ/kg)	-875.6	756.0	-875.6		
Molar Entropy	(kJ/kgmole-C)	94.49	198.0	94.49		
Mass Entropy	(kJ/kg-C)	3.549	9.527	3.549		
Heat Flow	(kJ/h)	-4.051e+007	0.0000	-4.051e+007		
Molar Density	(kgmole/m3)	39.91	2.794	39.91		
Mass Density	(kg/m3)	1063	58.06	1063		
Std Ideal Liq Mass Density	(kg/m3)	979.8	368.5	979.8		
Liq Mass Density @Std Cond	(kg/m3)	1078	---	1078		

Absorber: DIPA CONTACTOR @Main (continued)

Properties : RICH DIPA FROM CONT

	Overall	Vapour Phase	Aqueous Phase		
Molar Heat Capacity (kJ/kgmole-C)	94.49	43.29	94.49		
Mass Heat Capacity (kJ/kg-C)	3.549	2.083	3.549		
Thermal Conductivity (W/m-K)	0.4623	3.350e-002	0.4623		
Viscosity (cP)	1.561	1.228e-002	1.561		
Surface Tension (dyne/cm)	55.31	---	55.31		
Molecular Weight	26.63	20.78	26.63		
Z Factor	6.069e-002	0.8670	6.069e-002		

Properties : SWEET GAS

	Overall	Vapour Phase			
Vapour/Phase Fraction	1.0000	1.0000			
Temperature: (C)	34.97	34.97			
Pressure: (kPa)	6619	6619			
Molar Flow (kgmole/h)	909.6	909.6			
Mass Flow (kg/h)	1.805e+004	1.805e+004			
Std Ideal Liq Vol Flow (m3/h)	52.26	52.26			
Molar Enthalpy (kJ/kgmole)	1.498e+004	1.498e+004			
Mass Enthalpy (kJ/kg)	754.7	754.7			
Molar Entropy (kJ/kgmole-C)	192.8	192.8			
Mass Entropy (kJ/kg-C)	9.715	9.715			
Heat Flow (kJ/h)	1.362e+007	1.362e+007			
Molar Density (kgmole/m3)	3.080	3.080			
Mass Density (kg/m3)	61.13	61.13			
Std Ideal Liq Mass Density (kg/m3)	345.4	345.4			
Liq Mass Density @Std Cond (kg/m3)	---	---			
Molar Heat Capacity (kJ/kgmole-C)	42.30	42.30			
Mass Heat Capacity (kJ/kg-C)	2.131	2.131			
Thermal Conductivity (W/m-K)	3.163e-002	3.163e-002			
Viscosity (cP)	1.129e-002	1.129e-002			
Surface Tension (dyne/cm)	---	---			
Molecular Weight	19.85	19.85			
Z Factor	0.8389	0.8389			

SUMMARY

Flow Basis: Molar The composition option is selected

Feed Composition

	DIPA TO CONT	SOUR GAS TO CONT			
Flow Rate (kgmole/h)	1.647285e+03	1.000000e+03			
	---	---			
Nitrogen	0.0000	0.0500			
CO2	0.0000	0.0000			
H2S	0.0002	0.0900			
Methane	0.0000	0.7350			
Ethane	0.0000	0.0830			
Propane	0.0000	0.0210			
i-Butane	0.0000	0.0060			
n-Butane	0.0000	0.0020			
i-Pentane	0.0000	0.0030			
n-Pentane	0.0000	0.0080			
n-Hexane	0.0000	0.0010			
n-Heptane	0.0000	0.0010			
H2O	0.9286	0.0000			
DIsoPAmine	0.0712	0.0000			

Flow Basis: Molar The composition option is selected

Feed Flows

	DIPA TO CONT	SOUR GAS TO CONT			
Flow Rate (kgmole/h)	1.647285e+03	1.000000e+03			
	---	---			
Nitrogen (kgmole/h)	0.0000	50.0000			
CO2 (kgmole/h)	0.0000	0.0000			
H2S (kgmole/h)	0.3977	90.0000			
Methane (kgmole/h)	0.0000	735.0000			
Ethane (kgmole/h)	0.0000	83.0000			
Propane (kgmole/h)	0.0000	21.0000			

Absorber: DIPA CONTACTOR @Main (continued)

SUMMARY

	DIPA TO CONT	SOUR GAS TO CONT			
i-Butane (kgmole/h)	0.0000	6.0000			
n-Butane (kgmole/h)	0.0000	2.0000			
i-Pentane (kgmole/h)	0.0000	3.0000			
n-Pentane (kgmole/h)	0.0000	8.0000			
n-Hexane (kgmole/h)	0.0000	1.0000			
n-Heptane (kgmole/h)	0.0000	1.0000			
H2O (kgmole/h)	1.529587e+03	0.0000			
DIsoPAmine (kgmole/h)	117.3003	0.0000			

Products

Flow Basis: Molar — The composition option is selected

Product Compositions

	SWEET GAS	RICH DIPA FROM CONT			
Flow Rate (kgmole/h)	909.6114	1.737674e+03			
	---	---			
Nitrogen	0.0549	0.0000			
CO2	0.0000	0.0000			
H2S	0.0000	0.0520			
Methane	0.8066	0.0008			
Ethane	0.0911	0.0001			
Propane	0.0231	0.0000			
i-Butane	0.0066	0.0000			
n-Butane	0.0022	0.0000			
i-Pentane	0.0033	0.0000			
n-Pentane	0.0088	0.0000			
n-Hexane	0.0011	0.0000			
n-Heptane	0.0011	0.0000			
H2O	0.0012	0.8796			
DIsoPAmine	0.0000	0.0675			

Flow Basis: Molar — The composition option is selected

Product Flows

	SWEET GAS	RICH DIPA FROM CONT			
Flow Rate (kgmole/h)	909.6114 *	1.737674e+03 *			
	---	---			
Nitrogen (kgmole/h)	49.9496 *	0.0504 *			
CO2 (kgmole/h)	0.0000 *	0.0000 *			
H2S (kgmole/h)	0.0004 *	90.3972 *			
Methane (kgmole/h)	733.6823 *	1.3177 *			
Ethane (kgmole/h)	82.8879 *	0.1121 *			
Propane (kgmole/h)	20.9807 *	0.0193 *			
i-Butane (kgmole/h)	5.9996 *	0.0004 *			
n-Butane (kgmole/h)	1.9999 *	0.0001 *			
i-Pentane (kgmole/h)	2.9998 *	0.0002 *			
n-Pentane (kgmole/h)	7.9995 *	0.0005 *			
n-Hexane (kgmole/h)	0.9995 *	0.0005 *			
n-Heptane (kgmole/h)	0.9999 *	0.0001 *			
H2O (kgmole/h)	1.1118 *	1.528475e+03 *			
DIsoPAmine (kgmole/h)	0.0004 *	117.2999 *			

Flow Basis: Molar — The composition option is selected

Product Recoveries

	SWEET GAS	RICH DIPA FROM CONT			
Flow Rate (kgmole/h)	909.6114	1.737674e+03			
	---	---			
Nitrogen (%)	99.8992	0.1008			
CO2 (%)	0.3393	99.6607			
H2S (%)	0.0005	99.9995			
Methane (%)	99.8207	0.1793			
Ethane (%)	99.8649	0.1351			
Propane (%)	99.9082	0.0918			
i-Butane (%)	99.9941	0.0059			
n-Butane (%)	99.9944	0.0056			
i-Pentane (%)	99.9932	0.0068			
n-Pentane (%)	99.9934	0.0066			

Absorber: DIPA CONTACTOR @Main (continued)

SUMMARY

	SWEET GAS	RICH DIPA FROM CONT			
n-Hexane (%)	99.9532	0.0468			
n-Heptane (%)	99.9939	0.0061			
H2O (%)	0.0727	99.9273			
DIsoPAmine (%)	0.0003	99.9997			

COLUMN PROFILES

Reflux Ratio:	1.813	Reboil Ratio:	0.5515	The Flows Option is Selected	Flow Basis: Molar

Column Profiles Flows

	Temperature (C)	Pressure (kPa)	Net Liq (kgmole/h)	Net Vap (kgmole/h)	Net Feed (kgmole/h)	Net Draws (kgmole/h)
1__TS-1	34.97	6619	1649	---	1647	909.6
2__TS-1	34.97	6624	1649	911.5	---	---
3__TS-1	34.97	6628	1649	911.5	---	---
4__TS-1	34.97	6633	1649	911.5	---	---
5__TS-1	34.97	6638	1649	911.5	---	---
6__TS-1	34.98	6643	1649	911.5	---	---
7__TS-1	34.98	6647	1649	911.5	---	---
8__TS-1	34.99	6652	1649	911.6	---	---
9__TS-1	35.00	6657	1649	911.6	---	---
10__TS-1	35.04	6661	1649	911.6	---	---
11__TS-1	35.11	6666	1649	911.7	---	---
12__TS-1	35.26	6671	1650	911.8	---	---
13__TS-1	35.57	6676	1650	912.1	---	---
14__TS-1	36.23	6680	1651	912.6	---	---
15__TS-1	37.54	6685	1654	913.8	---	---
16__TS-1	40.10	6690	1659	916.3	---	---
17__TS-1	44.70	6694	1670	921.5	---	---
18__TS-1	51.71	6699	1689	932.1	---	---
19__TS-1	59.00	6704	1717	951.6	---	---
20__TS-1	59.95	6709	---	978.8	1000	1738

Column Profiles Energy

	Temperature (C)	Liquid Enthalpy (kJ/kgmole)	Vapour Enthalpy (kJ/kgmole)	Heat Loss (kJ/h)
1__TS-1	34.97	-2.532e+004	1.498e+004	---
2__TS-1	34.97	-2.532e+004	1.498e+004	---
3__TS-1	34.97	-2.532e+004	1.498e+004	---
4__TS-1	34.97	-2.532e+004	1.498e+004	---
5__TS-1	34.97	-2.532e+004	1.498e+004	---
6__TS-1	34.98	-2.532e+004	1.498e+004	---
7__TS-1	34.98	-2.532e+004	1.498e+004	---
8__TS-1	34.99	-2.532e+004	1.498e+004	---
9__TS-1	35.00	-2.531e+004	1.498e+004	---
10__TS-1	35.04	-2.531e+004	1.498e+004	---
11__TS-1	35.11	-2.531e+004	1.498e+004	---
12__TS-1	35.26	-2.529e+004	1.499e+004	---
13__TS-1	35.57	-2.526e+004	1.500e+004	---
14__TS-1	36.23	-2.521e+004	1.502e+004	---
15__TS-1	37.54	-2.509e+004	1.507e+004	---
16__TS-1	40.10	-2.487e+004	1.517e+004	---
17__TS-1	44.70	-2.448e+004	1.534e+004	---
18__TS-1	51.71	-2.390e+004	1.559e+004	---
19__TS-1	59.00	-2.332e+004	1.580e+004	---
20__TS-1	59.95	-2.331e+004	1.571e+004	---

FEEDS / PRODUCTS

Flow Basis: Molar

	Stream	Type	Duty (kJ/h)	State	Flows (kgmole/h)	Enthalpy (kJ/kgmole)	Temp (C)
1__TS-1	DIPA TO CONT	Feed	---	Liquid	1647 *	-2.536e+004 *	35.00 *
	SWEET GAS	Draw	---	Vapour	909.6 *	1.498e+004 *	34.97 *
2__TS-1							
3__TS-1							
4__TS-1							
5__TS-1							
6__TS-1							

Absorber: DIPA CONTACTOR @Main (continued)

FEEDS / PRODUCTS

7__TS-1							
8__TS-1							
9__TS-1							
10__TS-1							
11__TS-1							
12__TS-1							
13__TS-1							
14__TS-1							
15__TS-1							
16__TS-1							
17__TS-1							
18__TS-1							
19__TS-1							
20__TS-1	SOUR GAS TO CO	Feed	---	Vapour	1000 *	1.489e+004 *	43.00 *
	RICH DIPA FROM C	Draw	---	Liquid	1738 *	-2.331e+004 *	59.95 *

SETUP

Sub-Flowsheet

Feed Streams			Product Streams		
Internal Stream	External Stream	Transfer Basis	Internal Stream	External Stream	Transfer Basis
SOUR GAS TO CONT	Sour gas to contactor @Main	P-H Flash	SWEET GAS	SWEET GAS @Main	P-H Flash
DIPA TO CONT	ean DIPA to Contactor @Mai	P-H Flash	RICH DIPA FROM CONT	RICH DIPA @Main	P-H Flash

VARIABLES

Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
SOUR GAS TO CONT		
DIPA TO CONT		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
SWEET GAS		
RICH DIPA FROM CONT		

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name:	E:\PRAKASH MUKHOPADHYAY -- DIPA.HSC
Unit Set:	SI
Date/Time:	Mon May 11 08:47:16 2009

Separator: FLASH TK

CONNECTIONS

Inlet Stream

Stream Name	From Unit Operation	
Rich DIPA to Flash Tank	Valve	Valve

Outlet Stream

Stream Name	To Unit Operation	
FLASH VAP		
RICH TO L/R	Heat Exchanger:	L/R Exchanger

Energy Stream

Stream Name	From Unit Operation

PARAMETERS

Vessel Volume:	---	Level SP:	50.00 %	Liquid Volume:	---		
Vessel Pressure:	620.5 kPa	Pressure Drop:	0.0000 kPa *	Duty:	0.0000 kJ/h	Heat Transfer Mode:	Heating

User Variables

RATING

Sizing

Cylinder		Vertical		Separator has a Boot:	No
Volume	---	Diameter	---	Height	---

Level Taps: Level Tap Specification

Level Tap	PV High	PV Low	OP High	OP Low

Level Taps: Calculated Level Tap Values

Level Tap	Liquid Level	Aqueous Level

Options

PV Work Term Contribution	(%)	100.00 *		

PROPERTIES

Rich DIPA to Flash Tank

		Overall	Vapour Phase	Aqueous Phase		
Vapour/Phase Fraction		0.0013	0.0013	0.9987		
Temperature:	(C)	59.74	59.74	59.74		
Pressure:	(kPa)	620.5 *	620.5	620.5		
Molar Flow	(kgmole/h)	1738	2.209	1735		
Mass Flow	(kg/h)	4.627e+004	51.72	4.621e+004		
Std Ideal Liq Vol Flow	(m3/h)	47.22	0.1113	47.11		
Molar Enthalpy	(kJ/kgmole)	-2.331e+004	1.354e+004	-2.336e+004		
Mass Enthalpy	(kJ/kg)	-875.6	578.0	-877.3		
Molar Entropy	(kJ/kgmole-C)	94.67	220.6	94.51		
Mass Entropy	(kJ/kg-C)	3.556	9.423	3.549		
Heat Flow	(kJ/h)	-4.051e+007	2.990e+004	-4.054e+007		
Molar Density	(kgmole/m3)	32.69	0.2283	39.91		
Mass Density	(kg/m3)	870.4	5.345	1063		
Std Ideal Liq Mass Density	(kg/m3)	979.8	464.9	981.0		
Liq Mass Density @Std Cond	(kg/m3)	1078	---	1078		
Molar Heat Capacity	(kJ/kgmole-C)	94.46	38.91	94.53		
Mass Heat Capacity	(kJ/kg-C)	3.548	1.662	3.550		
Thermal Conductivity	(W/m-K)	---	2.976e-002	0.4621		
Viscosity	(cP)	---	1.318e-002	1.569		
Surface Tension	(dyne/cm)	55.34	---	55.34		
Molecular Weight		26.63	23.42	26.63		
Z Factor		---	0.9822	5.617e-003		

Separator: FLASH TK (continued)

FLASH VAP

		Overall	Vapour Phase	Aqueous Phase		
Vapour/Phase Fraction		1.0000	1.0000	0.0000		
Temperature:	(C)	59.74	59.74	59.74		
Pressure:	(kPa)	620.5	620.5	620.5		
Molar Flow	(kgmole/h)	2.209	2.209	0.0000		
Mass Flow	(kg/h)	51.72	51.72	0.0000		
Std Ideal Liq Vol Flow	(m3/h)	0.1113	0.1113	0.0000		
Molar Enthalpy	(kJ/kgmole)	1.354e+004	1.354e+004	-2.336e+004		
Mass Enthalpy	(kJ/kg)	578.0	578.0	-877.3		
Molar Entropy	(kJ/kgmole-C)	220.6	220.6	94.51		
Mass Entropy	(kJ/kg-C)	9.423	9.423	3.549		
Heat Flow	(kJ/h)	2.990e+004	2.990e+004	0.0000		
Molar Density	(kgmole/m3)	0.2283	0.2283	39.91		
Mass Density	(kg/m3)	5.345	5.345	1063		
Std Ideal Liq Mass Density	(kg/m3)	464.9	464.9	981.0		
Liq Mass Density @Std Cond	(kg/m3)	---	---	1078		
Molar Heat Capacity	(kJ/kgmole-C)	38.91	38.91	94.53		
Mass Heat Capacity	(kJ/kg-C)	1.662	1.662	3.550		
Thermal Conductivity	(W/m-K)	2.976e-002	2.976e-002	0.4621		
Viscosity	(cP)	1.318e-002	1.318e-002	1.569		
Surface Tension	(dyne/cm)	---	---	55.34		
Molecular Weight		23.42	23.42	26.63		
Z Factor		0.9822	0.9822	5.617e-003		

RICH TO L/R

		Overall	Vapour Phase	Aqueous Phase		
Vapour/Phase Fraction		0.0000	0.0000	1.0000		
Temperature:	(C)	59.74	59.74	59.74		
Pressure:	(kPa)	620.5	620.5	620.5		
Molar Flow	(kgmole/h)	1735	0.0000	1735		
Mass Flow	(kg/h)	4.621e+004	0.0000	4.621e+004		
Std Ideal Liq Vol Flow	(m3/h)	47.11	0.0000	47.11		
Molar Enthalpy	(kJ/kgmole)	-2.336e+004	1.354e+004	-2.336e+004		
Mass Enthalpy	(kJ/kg)	-877.3	578.0	-877.3		
Molar Entropy	(kJ/kgmole-C)	94.51	220.6	94.51		
Mass Entropy	(kJ/kg-C)	3.549	9.423	3.549		
Heat Flow	(kJ/h)	-4.054e+007	0.0000	-4.054e+007		
Molar Density	(kgmole/m3)	39.91	0.2283	39.91		
Mass Density	(kg/m3)	1063	5.345	1063		
Std Ideal Liq Mass Density	(kg/m3)	981.0	464.9	981.0		
Liq Mass Density @Std Cond	(kg/m3)	1078	---	1078		
Molar Heat Capacity	(kJ/kgmole-C)	94.53	38.91	94.53		
Mass Heat Capacity	(kJ/kg-C)	3.550	1.662	3.550		
Thermal Conductivity	(W/m-K)	0.4621	2.976e-002	0.4621		
Viscosity	(cP)	1.569	1.318e-002	1.569		
Surface Tension	(dyne/cm)	55.34	---	55.34		
Molecular Weight		26.63	23.42	26.63		
Z Factor		5.617e-003	0.9822	5.617e-003		

Inlet Stream

		Rich DIPA to Flash Tank		
Vapour Fraction		0.0013		
Temperature	(C)	59.74		
Pressure	(kPa)	620.5 *		
Molar Flow	(kgmole/h)	1738		
MassFlow	(kg/h)	4.627e+004		
Std Ideal Liq Vol Flow	(m3/h)	47.22		
HeatFlow	(kJ/h)	-4.051e+007		

Outlet Stream

		RICH TO L/R	FLASH VAP		
Vapour Fraction		0.0000	1.0000		
Temperature	(C)	59.74	59.74		

Separator: FLASH TK (continued)

		RICH TO L/R	FLASH VAP		
Pressure	(kPa)	620.5	620.5		
Molar Flow	(kgmole/h)	1735	2.209		
MassFlow	(kg/h)	4.621e+004	51.72		
Std Ideal Liq Vol Flow	(m3/h)	47.11	0.1113		
HeatFlow	(kJ/h)	-4.054e+007	2.990e+004		

HYPROTECH LIFECYCLE INNOVATION	TEAM LND Calgary, Alberta CANADA

Case Name:	E:\PRAKASH MUKHOPADHYAY -- DIPA.HSC
Unit Set:	SI
Date/Time:	Mon May 11 08:53:10 2009

Heat Exchanger: L/R Exchanger

CONNECTIONS

Tube Side				Shell Side			
Inlet		Outlet		Inlet		Outlet	
Name	RICH TO L/R	Name	REGEN FEED	Name	REGEN BTTMS	Name	LEAN FROM L/R
From Op.	FLASH TK Separator	To Op.	DistillatioREGENERATOR	From Op.	DistillatioREGENERATOR	To Op.	Mixer Mixure
Temp	59.74 C	Temp	93.33 C *	Temp	124.76 C	Temp	88.02 C

PARAMETERS

Exchanger Design (End Point)

Tube Side DeltaP:	68.95 kPa *	Shell Side DeltaP:	68.95 kPa *	Passes:	---
UA:	2.090e+005 kJ/C-h	Tolerance:	1.0000e-04 *		

Tube Side Data		Shell Side Data	
Heat Transfer Coefficient	---	Heat Transfer Coefficient	---
Tube Pressure Drop	68.95 kPa *	Shell Pressure Drop	68.95 kPa *
Fouling	0.00000 C-h-m2/kJ *	Fouling	0.00000 C-h-m2/kJ *
Tube Length	6.00 m *	Shell Passes	1
Tube O.D.	20.00 mm *	Shell Series	1 *
Tube Thickness	2.0000 mm	Shell Parallel	1 *
Tube Pitch	50.0000 mm *	Baffle Type	Single
Orientation	Horizontal	Baffle Cut(%Area)	20.00 *
Passes Per Shell	1 *	Baffle Orientation	Horizontal
Tubes Per Shell	160 *	Spacing	800.0000 mm *
Layout Angle	Triangular (30 degrees)	Diameter	739.0488 mm *
TEMA Type	A E L	Area	60.32 m2

SPECS

	Specified Value	Current Value	Relative Error	Active	Estimate
E-100 Heat Balance	0.0000 kJ/h	-2.456e-009 kJ/h	6.821e-013	On	Off
E-100 UA	---	2.090e+005 kJ/C-h	---	On	Off

Detailed Specifications

E-100 Heat Balance

Type: Duty	Pass: Error	Spec Value: 0.0000 kJ/h

E-100 UA

Type: UA	Pass: Overall	Spec Value: ---

User Variables

RATING

Sizing

Overall Data

Configuration

# of Shells in Series	1 *	Tube Passes per Shell	1 *	Elevation (Base)	0.0000 m
# of Shells in Parallel	1 *	Exchange Orientation	Horizontal	First Tube Pass Flow Direction	Counter

TEMA Type:	A	E	L

Calculated Information

Shell HT Coeff	---	Overall U	166 kJ/h-m2-C	Shell DP	68.95 kPa *	Shell Vol per Shell	2.272 m3	HT Area per Shell	60.32 m2
Tube HT Coeff	---	Overall UA	)e+005 kJ/C-h	Tube DP	68.95 kPa *	Tube Vol per Shell	0.1930 m3		

Shell Data

Shell and Tube Bundle

Shell Diameter	739.0 mm *	Tube Pitch	50.00 mm *	Shell Fouling	0.0000 C-h-m2/kJ *
# of Tubes per Shell	160 *	Tube Layout Angle	Triangular (30 degrees)		

Shell Baffles

Shell Baffle Type	Single	Shell Baffle Orientation	Horizontal	Baffle Cut (%Area)	20.00 *	Baffle Spacing	800.0 mm *

Tube Data

Dimensions

OD	20.00 mm *	ID	16.00 mm *	Tube Thickness	2.000 mm	Tube Length	6.000 m *

Tube Properties

Tube Fouling	0.0000 C-h-m2/kJ *	Thermal Cond.	45.00 W/m-K *	Wall Cp	---	Wall Density	---

PROPERTIES

Heat Exchanger: L/R Exchanger (continued)

RICH TO L/R

	Overall	Vapour Phase	Aqueous Phase		
Vapour/Phase Fraction	0.0000	0.0000	1.0000		
Temperature: (C)	59.74	59.74	59.74		
Pressure: (kPa)	620.5	620.5	620.5		
Molar Flow (kgmole/h)	1735	0.0000	1735		
Mass Flow (kg/h)	4.621e+004	0.0000	4.621e+004		
Std Ideal Liq Vol Flow (m3/h)	47.11	0.0000	47.11		
Molar Enthalpy (kJ/kgmole)	-2.336e+004	1.354e+004	-2.336e+004		
Mass Enthalpy (kJ/kg)	-877.3	578.0	-877.3		
Molar Entropy (kJ/kgmole-C)	94.51	220.6	94.51		
Mass Entropy (kJ/kg-C)	3.549	9.423	3.549		
Heat Flow (kJ/h)	-4.054e+007	0.0000	-4.054e+007		
Molar Density (kgmole/m3)	39.91	0.2283	39.91		
Mass Density (kg/m3)	1063	5.345	1063		
Std Ideal Liq Mass Density (kg/m3)	981.0	464.9	981.0		
Liq Mass Density @Std Cond (kg/m3)	1078	---	1078		
Molar Heat Capacity (kJ/kgmole-C)	94.53	38.91	94.53		
Mass Heat Capacity (kJ/kg-C)	3.550	1.662	3.550		
Thermal Conductivity (W/m-K)	0.4621	2.976e-002	0.4621		
Viscosity (cP)	1.569	1.318e-002	1.569		
Surface Tension (dyne/cm)	55.34	---	55.34		
Molecular Weight	26.63	23.42	26.63		
Z Factor	5.617e-003	0.9822	5.617e-003		

REGEN FEED

	Overall	Vapour Phase	Aqueous Phase		
Vapour/Phase Fraction	0.0082	0.0082	0.9918		
Temperature: (C)	93.33 *	93.33	93.33		
Pressure: (kPa)	551.6	551.6	551.6		
Molar Flow (kgmole/h)	1735	14.22	1721		
Mass Flow (kg/h)	4.621e+004	450.9	4.576e+004		
Std Ideal Liq Vol Flow (m3/h)	47.11	0.5660	46.54		
Molar Enthalpy (kJ/kgmole)	-1.977e+004	1.230e+004	-2.003e+004		
Mass Enthalpy (kJ/kg)	-742.4	387.6	-753.5		
Molar Entropy (kJ/kgmole-C)	100.3	221.1	99.33		
Mass Entropy (kJ/kg-C)	3.768	6.972	3.736		
Heat Flow (kJ/h)	-3.431e+007	1.748e+005	-3.448e+007		
Molar Density (kgmole/m3)	14.42	0.1860	39.22		
Mass Density (kg/m3)	384.1	5.899	1043		
Std Ideal Liq Mass Density (kg/m3)	981.0	796.6	983.2		
Liq Mass Density @Std Cond (kg/m3)	1078	1055	1078		
Molar Heat Capacity (kJ/kgmole-C)	99.28	35.14	99.81		
Mass Heat Capacity (kJ/kg-C)	3.728	1.108	3.754		
Thermal Conductivity (W/m-K)	---	2.290e-002	0.4850		
Viscosity (cP)	---	1.548e-002	0.7652		
Surface Tension (dyne/cm)	49.78	---	49.78		
Molecular Weight	26.63	31.72	26.59		
Z Factor	---	0.9733	4.616e-003		

REGEN BTTMS

	Overall	Aqueous Phase			
Vapour/Phase Fraction	0.0000	1.0000			
Temperature: (C)	124.8	124.8			
Pressure: (kPa)	217.2	217.2			
Molar Flow (kgmole/h)	1612	1612			
Mass Flow (kg/h)	4.256e+004	4.256e+004			
Std Ideal Liq Vol Flow (m3/h)	42.63	42.63			
Molar Enthalpy (kJ/kgmole)	-1.610e+004	-1.610e+004			
Mass Enthalpy (kJ/kg)	-609.8	-609.8			
Molar Entropy (kJ/kgmole-C)	108.1	108.1			
Mass Entropy (kJ/kg-C)	4.097	4.097			
Heat Flow (kJ/h)	-2.595e+007	-2.595e+007			
Molar Density (kgmole/m3)	36.50	36.50			

Heat Exchanger: L/R Exchanger (continued)

REGEN BTTMS

	Overall	Aqueous Phase			
Mass Density (kg/m3)	963.5	963.5			
Std Ideal Liq Mass Density (kg/m3)	998.3	998.3			
Liq Mass Density @Std Cond (kg/m3)	1023	1023			
Molar Heat Capacity (kJ/kgmole-C)	108.1	108.1			
Mass Heat Capacity (kJ/kg-C)	4.097	4.097			
Thermal Conductivity (W/m-K)	0.4927	0.4927			
Viscosity (cP)	0.4512	0.4512			
Surface Tension (dyne/cm)	44.47	44.47			
Molecular Weight	26.40	26.40			
Z Factor	1.799e-003	1.799e-003			

LEAN FROM L/R

	Overall	Aqueous Phase			
Vapour/Phase Fraction	0.0000	1.0000			
Temperature: (C)	88.02	88.02			
Pressure: (kPa)	148.2	148.2			
Molar Flow (kgmole/h)	1612	1612			
Mass Flow (kg/h)	4.256e+004	4.256e+004			
Std Ideal Liq Vol Flow (m3/h)	42.63	42.63			
Molar Enthalpy (kJ/kgmole)	-1.996e+004	-1.996e+004			
Mass Enthalpy (kJ/kg)	-756.3	-756.3			
Molar Entropy (kJ/kgmole-C)	102.4	102.4			
Mass Entropy (kJ/kg-C)	3.881	3.881			
Heat Flow (kJ/h)	-3.219e+007	-3.219e+007			
Molar Density (kgmole/m3)	37.54	37.54			
Mass Density (kg/m3)	990.9	990.9			
Std Ideal Liq Mass Density (kg/m3)	998.3	998.3			
Liq Mass Density @Std Cond (kg/m3)	1023	1023			
Molar Heat Capacity (kJ/kgmole-C)	102.4	102.4			
Mass Heat Capacity (kJ/kg-C)	3.881	3.881			
Thermal Conductivity (W/m-K)	0.4798	0.4798			
Viscosity (cP)	0.8113	0.8113			
Surface Tension (dyne/cm)	50.53	50.53			
Molecular Weight	26.40	26.40			
Z Factor	1.315e-003	1.315e-003			

Stream Properties

	RICH TO L/R	REGEN FEED	REGEN BTTMS	LEAN FROM L/R
Vapour Fraction	0.0000	0.0082	0.0000	0.0000
Temperature	59.74	93.33 *	124.8	88.02
Pressure	620.5	551.6	217.2	148.2
Enthalpy	-2.336e+004	-1.977e+004	-1.610e+004	-1.996e+004
Molar Flow	1735	1735	1612	1612
MassFlow	4.621e+004	4.621e+004	4.256e+004	4.256e+004
Std Ideal Liq Vol Flow	47.11	47.11	42.63	42.63
HeatFlow	-4.054e+007	-3.431e+007	-2.595e+007	-3.219e+007

DETAILS

Overall/Detailed Performance

Duty:	6.235e+06 kJ/h	UA:	2.090e+05 kJ/C-h	UA Curv. Error:	0.00e-01 kJ/C-h	Ft Factor:	1
Heat Leak:	0.000e-01 kJ/h	Min. Approach:	28.28 C	Hot Pinch Temp:	88.02 C	Uncorrected Lmtd:	29.83 C
Heat Loss:	0.000e-01 kJ/h	Lmtd:	29.83 C	Cold Pinch Temp:	59.74 C		

TABLES

Shell Side - Overall Phase

Temperature (C)	Pressure (kPa)	Heat Flow (kJ/h)	Enthalpy (kJ/kgmole)	Molar Vap Frac	Mass Vap Frac
88.02	148.24	0.00	-19964.69	0.0000	0.0000
124.76	217.18	6235243.72	-16097.15	0.0000	0.0000

Heat Exchanger: L/R Exchanger (continued)

Shell Side - Vapour Phase

Mass Flow (kg/h)	Molecular Weight	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)
---	---	---	---	---	---
---	---	---	---	---	---

Shell Side - Light Liquid Phase

Mass Flow (kg/h)	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)	Surface Tension (dyne/cm)
---	---	---	---	---	---
---	---	---	---	---	---

Shell Side - Heavy Liquid Phase

Mass Flow (kg/h)	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)	Surface Tension (dyne/cm)
42560.25	37.54	3.88	0.81	---	---
42560.25	36.50	4.10	0.45	---	---

Tube Side - Overall Phase

Temperature (C)	Pressure (kPa)	Heat Flow (kJ/h)	Enthalpy (kJ/kgmole)	Molar Vap Frac	Mass Vap Frac
59.74	620.53	0.00	-23361.31	0.0000	0.0000
93.33	551.58	6235243.72	-19768.47	0.0082	0.0098

Tube Side - Vapour Phase

Mass Flow (kg/h)	Molecular Weight	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)
0.00	23.42	0.23	1.66	0.01	---
450.90	31.72	0.19	1.11	0.02	---

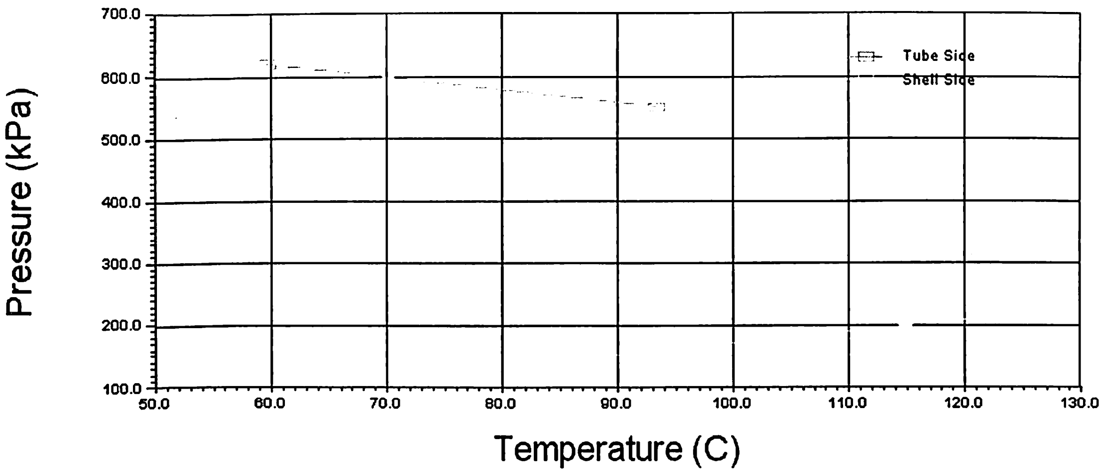
Tube Side - Light Liquid Phase

Mass Flow (kg/h)	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)	Surface Tension (dyne/cm)
---	---	---	---	---	---
---	---	---	---	---	---

Tube Side - Heavy Liquid Phase

Mass Flow (kg/h)	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)	Surface Tension (dyne/cm)
46214.37	39.91	3.55	1.57	---	---
45763.47	39.21	3.75	0.77	---	---

Heat Exchanger: L/R Exchanger (continued)



HYPROTECH	TEAM LND Calgary, Alberta CANADA	Case Name:	E:\PRAKASH MUKHOPADHYAY -- DIPA.HSC
		Unit Set:	SI
		Date/Time:	Mon May 11 09:01:43 2009

Distillation: REGENERATOR @Main

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION	
RBLR Q	Reboiler		
REGEN FEED	4__Main TS	Heat Exchanger	L/R Exchanger

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION	
COND Q	Condenser		
ACID GAS	Condenser		
REGEN BTTMS	Reboiler	Heat Exchanger	L/R Exchanger

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
Reflux Ratio	0.5000 *	0.8422	0.6845	1.000e-002 *	1.000e-002 *	Off	On	Off
Ovhd Vap Rate	99.61 kgmole/h *	123.3 kgmole/h	0.2375	1.000e-002 *	1.000 kgmole/h *	Off	On	Off
Reboiler Duty	1.431e+007 kJ/h *	1.431e+007 kJ/h	-0.1205	1.000e-002 *	10.00 kJ/h *	On	On	On
T Top	82.00 C *	82.00 C	7.340e-008	1.000e-002 *	1.000 C *	On	On	On

SPECS

Column Specification Parameters

Reflux Ratio

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stage:	Condenser	Flow Basis:	Molar	Liquid Specification:	---		

Ovhd Vap Rate

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	ACID GAS	Flow Basis:	Molar				

Reboiler Duty

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Energy Stream:	RBLR Q						

T Top

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stage:	Condenser						

SUBCOOLING

	Condenser	
Degrees of Subcooling	---	
Subcool to	---	

User Variables

PROFILES

General Parameters

Sub-Flow Sheet:	REGENERATOR (COL2)	Number of Stages:	18 *

Profile Estimates

	Temperature (C)	Net Liquid (kgmole/h)	Net Vapour (kgmole/h)
Condenser	82.00	103.8	123.3
1__Main TS	105.6	108.6	227.1
2__Main TS	106.1	108.7	231.9
3__Main TS	106.2	108.2	232.0
4__Main TS	[illegible]	[illegible]	[illegible]
5__Main TS	119.6	1949	280.4
6__Main TS	122.2	1962	336.5
7__Main TS	122.9	1965	349.6
8__Main TS	123.2	1967	353.2
9__Main TS	123.3	1968	354.7
10__Main TS	123.5	1968	355.6
11__Main TS	123.6	1969	356.3
12__Main TS	123.7	1970	356.9

Distillation: REGENERATOR @Main (continued)

Profile Estimates

	Temperature (C)	Net Liquid (kgmole/h)	Net Vapour (kgmole/h)
13__Main TS	123.8	1970	357.4
14__Main TS	123.9	1971	357.9
15__Main TS	124.0	1971	358.4
16__Main TS	124.1	1972	358.9
17__Main TS	124.2	1972	359.4
18__Main TS	124.3	1972	359.8
Reboiler	124.8	1612	360.2

EFFICIENCIES

Stage Efficiencies

Stages	Overall Efficiency	Nitrogen	CO2	H2S	Methane	Ethane
Condenser	1.000	1.000	1.000	1.000	1.000	1.000
1__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
2__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
3__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
4__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
5__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
6__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
7__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
8__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
9__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
10__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
11__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
12__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
13__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
14__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
15__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
16__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
17__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
18__Main TS	1.000	1.000	0.1500 *	0.8000 *	1.000	1.000
Reboiler	1.000	1.000	1.000	1.000	1.000	1.000

Stages	Overall Efficiency	Propane	i-Butane	n-Butane	i-Pentane	n-Pentane
Condenser	1.000	1.000	1.000	1.000	1.000	1.000
1__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
2__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
3__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
4__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
5__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
6__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
7__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
8__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
9__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
10__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
11__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
12__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
13__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
14__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
15__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
16__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
17__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
18__Main TS	1.000	1.000	1.000	1.000	1.000	1.000
Reboiler	1.000	1.000	1.000	1.000	1.000	1.000

Stages	Overall Efficiency	n-Hexane	n-Heptane	H2O	DIsoPAmine	
Condenser	1.000	1.000	1.000	1.000	1.000	
1__Main TS	1.000	1.000	1.000	1.000	1.000	
2__Main TS	1.000	1.000	1.000	1.000	1.000	
3__Main TS	1.000	1.000	1.000	1.000	1.000	
4__Main TS	1.000	1.000	1.000	1.000	1.000	
5__Main TS	1.000	1.000	1.000	1.000	1.000	
6__Main TS	1.000	1.000	1.000	1.000	1.000	
7__Main TS	1.000	1.000	1.000	1.000	1.000	
8__Main TS	1.000	1.000	1.000	1.000	1.000	

Distillation: REGENERATOR @Main (continued)

Stage Efficiencies

Stages	Overall Efficiency	n-Hexane	n-Heptane	H2O	DIsoPAmine	
9__Main TS	1.000	1.000	1.000	1.000	1.000	
10__Main TS	1.000	1.000	1.000	1.000	1.000	
11__Main TS	1.000	1.000	1.000	1.000	1.000	
12__Main TS	1.000	1.000	1.000	1.000	1.000	
13__Main TS	1.000	1.000	1.000	1.000	1.000	
14__Main TS	1.000	1.000	1.000	1.000	1.000	
15__Main TS	1.000	1.000	1.000	1.000	1.000	
16__Main TS	1.000	1.000	1.000	1.000	1.000	
17__Main TS	1.000	1.000	1.000	1.000	1.000	
18__Main TS	1.000	1.000	1.000	1.000	1.000	
Reboiler	1.000	1.000	1.000	1.000	1.000	

SOLVER

Column Solving Algorithm: HYSIM Inside-Out

Solving Options

Maximum Iterations:	10000
Equilibrium Error Tolerance:	1.000e-05
Heat/Spec Error Tolerance:	5.000e-004
Save Solutions as Initial Estimate:	On
Super Critical Handling Model:	Simple K
Trace Level:	Low
Init from Ideal K's:	Off

Acceleration Parameters

Accelerate K Value & H Model Parameters:	Off

Initial Estimate Generator Parameters

Iterative IEG (Good for Chemicals):	Off

Damping Parameters

Azeotrope Check:	Off
Fixed Damping Factor:	0 *

RATING

Tray Sections

Tray Section		Main TS			
Tray Diameter	(m)	1.500 *			
Weir Height	(m)	5.000e-002 *			
Weir Length	(m)	1.200 *			
Tray Space	(m)	0.5500 *			
Tray Volume	(m3)	0.9719			
Disable Heat Loss Calculations		No			
Heat Model		None			
Rating Calculations		No			
Tray Hold Up	(m3)	8.836e-002			

Vessels

Vessel		Condenser	Reboiler		
Diameter	(m)	1.193	1.193		
Length	(m)	1.789	1.789		
Volume	(m3)	2.000 *	2.000 *		
Orientation		Horizontal	Horizontal		
Vessel has a Boot		No	No		
Boot Diameter	(m)	---	---		
Boot Length	(m)	---	---		
Hold Up	(m3)	1.000	1.000		

Other Equipment In Column Flowsheet

Pressure Profile

	Pressure (kPa)	Pressure Drop (kPa)
Condenser	189.6 kPa *	17.24 kPa
1__Main TS	206.8 kPa	0.6084 kPa
2__Main TS	207.5 kPa	0.6084 kPa
3__Main TS	208.1 kPa	0.6084 kPa
4__Main TS	208.7 kPa	0.6084 kPa
5__Main TS	209.3 kPa	0.6084 kPa

Distillation: REGENERATOR @Main (continued)

6__Main TS	209.9 kPa	0.6084 kPa
7__Main TS	210.5 kPa	0.6084 kPa
8__Main TS	211.1 kPa	0.6084 kPa
9__Main TS	211.7 kPa	0.6084 kPa
10__Main TS	212.3 kPa	0.6084 kPa
11__Main TS	212.9 kPa	0.6084 kPa
12__Main TS	213.5 kPa	0.6084 kPa
13__Main TS	214.1 kPa	0.6084 kPa
14__Main TS	214.8 kPa	0.6084 kPa
15__Main TS	215.4 kPa	0.6084 kPa
16__Main TS	216.0 kPa	0.6084 kPa
17__Main TS	216.6 kPa	0.6084 kPa
18__Main TS	217.2 kPa	---
Reboiler	217.2 kPa *	0.0000 kPa

Pressure Solving Options

Pressure Tolerance	1.000e-004 *	Pressure Drop Tolerance	1.000e-004 *	Damping Factor	1.000 *	Max Press Iterations	100 *

PROPERTIES

Properties : REGEN FEED

		Overall	Vapour Phase	Aqueous Phase		
Vapour/Phase Fraction		0.0082	0.0082	0.9918		
Temperature:	(C)	93.33	93.33	93.33		
Pressure:	(kPa)	551.6	551.6	551.6		
Molar Flow	(kgmole/h)	1735	14.22	1721		
Mass Flow	(kg/h)	4.621e+004	450.9	4.576e+004		
Std Ideal Liq Vol Flow	(m3/h)	47.11	0.5660	46.54		
Molar Enthalpy	(kJ/kgmole)	-1.977e+004	1.230e+004	-2.003e+004		
Mass Enthalpy	(kJ/kg)	-742.4	387.6	-753.5		
Molar Entropy	(kJ/kgmole-C)	100.3	221.1	99.33		
Mass Entropy	(kJ/kg-C)	3.768	6.972	3.736		
Heat Flow	(kJ/h)	-3.431e+007	1.748e+005	-3.448e+007		
Molar Density	(kgmole/m3)	14.42	0.1860	39.22		
Mass Density	(kg/m3)	384.1	5.899	1043		
Std Ideal Liq Mass Density	(kg/m3)	981.0	796.6	983.2		
Liq Mass Density @Std Cond	(kg/m3)	1078	1055	1078		
Molar Heat Capacity	(kJ/kgmole-C)	99.28	35.14	99.81		
Mass Heat Capacity	(kJ/kg-C)	3.728	1.108	3.754		
Thermal Conductivity	(W/m-K)	---	2.290e-002	0.4850		
Viscosity	(cP)	---	1.548e-002	0.7652		
Surface Tension	(dyne/cm)	49.78	---	49.78		
Molecular Weight		26.63	31.72	26.59		
Z Factor		---	0.9733	4.616e-003		

Properties : REGEN BTTMS

		Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction		0.0000	0.0000	1.0000		
Temperature:	(C)	124.8	124.8	124.8		
Pressure:	(kPa)	217.2	217.2	217.2		
Molar Flow	(kgmole/h)	1612	0.0000	1612		
Mass Flow	(kg/h)	4.256e+004	0.0000	4.256e+004		
Std Ideal Liq Vol Flow	(m3/h)	42.63	0.0000	42.63		
Molar Enthalpy	(kJ/kgmole)	-1.610e+004	1.332e+004	-1.610e+004		
Mass Enthalpy	(kJ/kg)	-609.8	737.4	-609.8		
Molar Entropy	(kJ/kgmole-C)	108.1	215.7	108.1		
Mass Entropy	(kJ/kg-C)	4.097	11.94	4.097		
Heat Flow	(kJ/h)	-2.595e+007	0.0000	-2.595e+007		
Molar Density	(kgmole/m3)	36.50	6.669e-002	36.50		
Mass Density	(kg/m3)	963.5	1.205	963.5		
Std Ideal Liq Mass Density	(kg/m3)	998.3	997.7	998.3		
Liq Mass Density @Std Cond	(kg/m3)	1023	1055	1023		
Molar Heat Capacity	(kJ/kgmole-C)	108.1	34.40	108.1		
Mass Heat Capacity	(kJ/kg-C)	4.097	1.905	4.097		
Thermal Conductivity	(W/m-K)	0.4927	3.533e-002	0.4927		
Viscosity	(cP)	0.4963	1.408e-002	0.4963		
Surface Tension	(dyne/cm)	0.5174	---	0.5174		

Distillation: REGENERATOR @Main (continued)

Properties : REGEN BTTMS

	Overall	Vapour Phase	Liquid Phase		
Molecular Weight	26.40	18.06	26.40		
Z Factor	1.799e-003	0.9843	1.799e-003		

Properties : ACID GAS

	Overall	Vapour Phase			
Vapour/Phase Fraction	1.0000	1.0000			
Temperature: (C)	82.00	82.00			
Pressure: (kPa)	189.6	189.6			
Molar Flow (kgmole/h)	123.3	123.3			
Mass Flow (kg/h)	3654	3654			
Std Ideal Liq Vol Flow (m3/h)	4.476	4.476			
Molar Enthalpy (kJ/kgmole)	1.186e+004	1.186e+004			
Mass Enthalpy (kJ/kg)	400.0	400.0			
Molar Entropy (kJ/kgmole-C)	228.2	228.2			
Mass Entropy (kJ/kg-C)	7.698	7.698			
Heat Flow (kJ/h)	1.462e+006	1.462e+006			
Molar Density (kgmole/m3)	6.492e-002	6.492e-002			
Mass Density (kg/m3)	1.924	1.924			
Std Ideal Liq Mass Density (kg/m3)	816.4	816.4			
Liq Mass Density @Std Cond (kg/m3)	998.6	998.6			
Molar Heat Capacity (kJ/kgmole-C)	34.69	34.69			
Mass Heat Capacity (kJ/kg-C)	1.170	1.170			
Thermal Conductivity (W/m-K)	2.338e-002	2.338e-002			
Viscosity (cP)	1.476e-002	1.476e-002			
Surface Tension (dyne/cm)	---	---			
Molecular Weight	29.64	29.64			
Z Factor	0.9891	0.9891			

SUMMARY

Flow Basis: Molar | The composition option is selected

Feed Composition

	REGEN FEED				
Flow Rate (kgmole/h)	1.735465e+03				

Nitrogen	0.0000				
CO2	0.0000				
H2S	0.0517				
Methane	0.0001				
Ethane	0.0000				
Propane	0.0000				
i-Butane	0.0000				
n-Butane	0.0000				
i-Pentane	0.0000				
n-Pentane	0.0000				
n-Hexane	0.0000				
n-Heptane	0.0000				
H2O	0.8807				
DIsoPAmine	0.0676				

Flow Basis: Molar | The composition option is selected

Feed Flows

	REGEN FEED				
Flow Rate (kgmole/h)	1.735465e+03				

Nitrogen (kgmole/h)	0.0019				
CO2 (kgmole/h)	0.0000				
H2S (kgmole/h)	89.6483				
Methane (kgmole/h)	0.0972				
Ethane (kgmole/h)	0.0079				
Propane (kgmole/h)	0.0011				
i-Butane (kgmole/h)	0.0000				
n-Butane (kgmole/h)	0.0000				
i-Pentane (kgmole/h)	0.0000				
n-Pentane (kgmole/h)	0.0000				
n-Hexane (kgmole/h)	0.0000				

Distillation: REGENERATOR @Main (continued)

SUMMARY

	REGEN FEED				
n-Heptane (kgmole/h)	0.0000				
H2O (kgmole/h)	1.528408e+03				
DIsoPAmine (kgmole/h)	117.2999				

Products

Flow Basis: Molar | The composition option is selected

Product Compositions

	ACID GAS	REGEN BTTMS			
Flow Rate (kgmole/h)	123.2689	1.612196e+03			
	---	---			
Nitrogen	0.0000	0.0000			
CO2	0.0000	0.0000			
H2S	0.7240	0.0002			
Methane	0.0008	0.0000			
Ethane	0.0001	0.0000			
Propane	0.0000	0.0000			
i-Butane	0.0000	0.0000			
n-Butane	0.0000	0.0000			
i-Pentane	0.0000	0.0000			
n-Pentane	0.0000	0.0000			
n-Hexane	0.0000	0.0000			
n-Heptane	0.0000	0.0000			
H2O	0.2751	0.9270			
DIsoPAmine	0.0000	0.0728			

Flow Basis: Molar | The composition option is selected

Product Flows

	ACID GAS	REGEN BTTMS			
Flow Rate (kgmole/h)	123.2689 *	1.612196e+03 *			
	---	---			
Nitrogen (kgmole/h)	0.0019 *	0.0000 *			
CO2 (kgmole/h)	0.0000 *	0.0000 *			
H2S (kgmole/h)	89.2516 *	0.3968 *			
Methane (kgmole/h)	0.0972 *	0.0000 *			
Ethane (kgmole/h)	0.0079 *	0.0000 *			
Propane (kgmole/h)	0.0011 *	0.0000 *			
i-Butane (kgmole/h)	0.0000 *	0.0000 *			
n-Butane (kgmole/h)	0.0000 *	0.0000 *			
i-Pentane (kgmole/h)	0.0000 *	0.0000 *			
n-Pentane (kgmole/h)	0.0000 *	0.0000 *			
n-Hexane (kgmole/h)	0.0000 *	0.0000 *			
n-Heptane (kgmole/h)	0.0000 *	0.0000 *			
H2O (kgmole/h)	33.9091 *	1.494499e+03 *			
DIsoPAmine (kgmole/h)	0.0000 *	117.2999 *			

Flow Basis: Molar | The composition option is selected

Product Recoveries

	ACID GAS	REGEN BTTMS			
Flow Rate (kgmole/h)	123.2689	1.612196e+03			
	---	---			
Nitrogen (%)	100.0000	0.0000			
CO2 (%)	83.9876	16.0124			
H2S (%)	99.5574	0.4426			
Methane (%)	100.0000	0.0000			
Ethane (%)	100.0000	0.0000			
Propane (%)	100.0000	0.0000			
i-Butane (%)	100.0000	0.0000			
n-Butane (%)	100.0000	0.0000			
i-Pentane (%)	100.0000	0.0000			
n-Pentane (%)	100.0000	0.0000			
n-Hexane (%)	100.0000	0.0000			
n-Heptane (%)	100.0000	0.0000			
H2O (%)	2.2186	97.7814			
DIsoPAmine (%)	0.0000	100.0000			

Distillation: REGENERATOR @Main (continued)

COLUMN PROFILES

Reflux Ratio:	0.8422	Reboil Ratio:	0.2234	The Flows Option is Selected	Flow Basis:	Molar

Column Profiles Flows

	Temperature (C)	Pressure (kPa)	Net Liq (kgmole/h)	Net Vap (kgmole/h)	Net Feed (kgmole/h)	Net Draws (kgmole/h)
Condenser	82.00	189.6	103.8	---	---	123.3
1__Main TS	105.6	206.8	108.6	227.1	---	---
2__Main TS	106.1	207.5	108.7	231.9	---	---
3__Main TS	106.2	208.1	108.2	232.0	---	---
4__Main TS	108.1	208.7	1893	231.5	1735	---
5__Main TS	119.6	209.3	1949	280.4	---	---
6__Main TS	122.2	209.9	1962	336.5	---	---
7__Main TS	122.9	210.5	1965	349.6	---	---
8__Main TS	123.2	211.1	1967	353.2	---	---
9__Main TS	123.3	211.7	1968	354.7	---	---
10__Main TS	123.5	212.3	1968	355.6	---	---
11__Main TS	123.6	212.9	1969	356.3	---	---
12__Main TS	123.7	213.5	1970	356.9	---	---
13__Main TS	123.8	214.1	1970	357.4	---	---
14__Main TS	123.9	214.8	1971	357.9	---	---
15__Main TS	124.0	215.4	1971	358.4	---	---
16__Main TS	124.1	216.0	1972	358.9	---	---
17__Main TS	124.2	216.6	1972	359.4	---	---
18__Main TS	124.3	217.2	1972	359.8	---	---
Reboiler	124.8	217.2	---	360.2	---	1612

Column Profiles Energy

	Temperature (C)	Liquid Enthalpy (kJ/kgmole)	Vapour Enthalpy (kJ/kgmole)	Heat Loss (kJ/h)
Condenser	82.00	-2.966e+004	1.186e+004	---
1__Main TS	105.6	-2.776e+004	1.264e+004	---
2__Main TS	106.1	-2.772e+004	1.266e+004	---
3__Main TS	106.2	-2.770e+004	1.266e+004	---
4__Main TS	108.1	-1.932e+004	1.275e+004	---
5__Main TS	119.6	-1.838e+004	1.313e+004	---
6__Main TS	122.2	-1.816e+004	1.322e+004	---
7__Main TS	122.9	-1.810e+004	1.324e+004	---
8__Main TS	123.2	-1.808e+004	1.325e+004	---
9__Main TS	123.3	-1.806e+004	1.326e+004	---
10__Main TS	123.5	-1.805e+004	1.326e+004	---
11__Main TS	123.6	-1.804e+004	1.327e+004	---
12__Main TS	123.7	-1.803e+004	1.327e+004	---
13__Main TS	123.8	-1.802e+004	1.327e+004	---
14__Main TS	123.9	-1.801e+004	1.328e+004	---
15__Main TS	124.0	-1.801e+004	1.328e+004	---
16__Main TS	124.1	-1.800e+004	1.328e+004	---
17__Main TS	124.2	-1.799e+004	1.329e+004	---
18__Main TS	124.3	-1.798e+004	1.329e+004	---
Reboiler	124.8	-1.610e+004	1.332e+004	---

FEEDS / PRODUCTS

Flow Basis:	Molar

	Stream	Type	Duty (kJ/h)	State	Flows (kgmole/h)	Enthalpy (kJ/kgmole)	Temp (C)
Condenser	COND Q	Energy	4.489e+006 *		---	---	---
	ACID GAS	Draw	---	Vapour	123.3 *	1.186e+004 *	82.00 *
1__Main TS							
2__Main TS							
3__Main TS							
4__Main TS	REGEN FEED	Feed	---	Mixed	1735 *	-1.977e+004 *	93.33 *
5__Main TS							
6__Main TS							
7__Main TS							
8__Main TS							
9__Main TS							
10__Main TS							
11__Main TS							
12__Main TS							

Distillation: REGENERATOR @Main (continued)

FEEDS / PRODUCTS

13__Main TS							
14__Main TS							
15__Main TS							
16__Main TS							
17__Main TS							
18__Main TS							
Reboiler	RBLR Q	Energy	1.431e+007 *		---	---	---
	REGEN BTTMS	Draw	---	Liquid	1612 *	-1.610e+004 *	124.8 *

SETUP

Sub-Flowsheet

Feed Streams			Product Streams		
Internal Stream	External Stream	Transfer Basis	Internal Stream	External Stream	Transfer Basis
RBLR Q	RBLR Q @Main	None Req'd	COND Q	COND Q @Main	None Req'd
REGEN FEED	REGEN FEED @Main	P-H Flash	ACID GAS	ACID GAS @Main	P-H Flash
			REGEN BTTMS	REGEN BTTMS @Main	P-H Flash

VARIABLES

Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
RBLR Q		
REGEN FEED		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
COND Q		
ACID GAS		
REGEN BTTMS		

Cooler: Lean Cooler

CONNECTIONS

Inlet Stream

STREAM NAME	FROM UNIT OPERATION	
DIPA to cooler	Mixer	Mixure

Outlet Stream

STREAM NAME	TO UNIT OPERATION	
DIPA to Pump	Pump	Pump

Energy Stream

STREAM NAME	TO UNIT OPERATION
COOLER Q	

PARAMETERS

Pressure Drop:	34.47 kPa *	Duty:	8.774e+006 kJ/h	Volume:	0.1000 m3
Function:	Not Selected	Zones:	1		

User Variables

PROPERTIES

DIPA to cooler

		Overall	Aqueous Phase			
Vapour/Phase Fraction		0.0000	1.0000			
Temperature:	(C)	87.00	87.00			
Pressure:	(kPa)	148.2	148.2			
Molar Flow	(kgmole/h)	1647	1647			
Mass Flow	(kg/h)	4.319e+004	4.319e+004			
Std Ideal Liq Vol Flow	(m3/h)	43.27 *	43.27			
Molar Enthalpy	(kJ/kgmole)	-2.027e+004	-2.027e+004			
Mass Entropy	(kJ/kg-C)	3.884	3.884			
Heat Flow	(kJ/h)	-3.338e+007	-3.338e+007			
Molar Density	(kgmole/m3)	37.80	37.80			
Mass Density	(kg/m3)	991.3	991.3			
Std Ideal Liq Mass Density	(kg/m3)	998.3	998.3			
Liq Mass Density @Std Cond	(kg/m3)	1022	1022			
Molar Heat Capacity	(kJ/kgmole-C)	101.8	101.8			
Mass Heat Capacity	(kJ/kg-C)	3.884	3.884			
Thermal Conductivity	(W/m-K)	0.4818	0.4818			
Viscosity	(cP)	0.8131	0.8131			
Surface Tension	(dyne/cm)	50.84	50.84			
Molecular Weight		26.22	26.22			
Z Factor		1.309e-003	1.309e-003			

DIPA to Pump

		Overall	Aqueous Phase			
Vapour/Phase Fraction		0.0000	1.0000			
Temperature:	(C)	32.53	32.53			
Pressure:	(kPa)	113.8	113.8			
Molar Flow	(kgmole/h)	1647	1647			
Mass Flow	(kg/h)	4.319e+004	4.319e+004			
Std Ideal Liq Vol Flow	(m3/h)	43.27	43.27			
Molar Enthalpy	(kJ/kgmole)	-2.559e+004	-2.559e+004			
Mass Enthalpy	(kJ/kg)	-976.1	-976.1			
Molar Entropy	(kJ/kgmole-C)	93.82	93.82			
Mass Entropy	(kJ/kg-C)	3.578	3.578			
Heat Flow	(kJ/h)	-4.216e+007	-4.216e+007			
Molar Density	(kgmole/m3)	38.85	38.85			
Mass Density	(kg/m3)	1019	1019			
Std Ideal Liq Mass Density	(kg/m3)	998.3	998.3			
Liq Mass Density @Std Cond	(kg/m3)	1022	1022			

Cooler: Lean Cooler (continued)

PROPERTIES

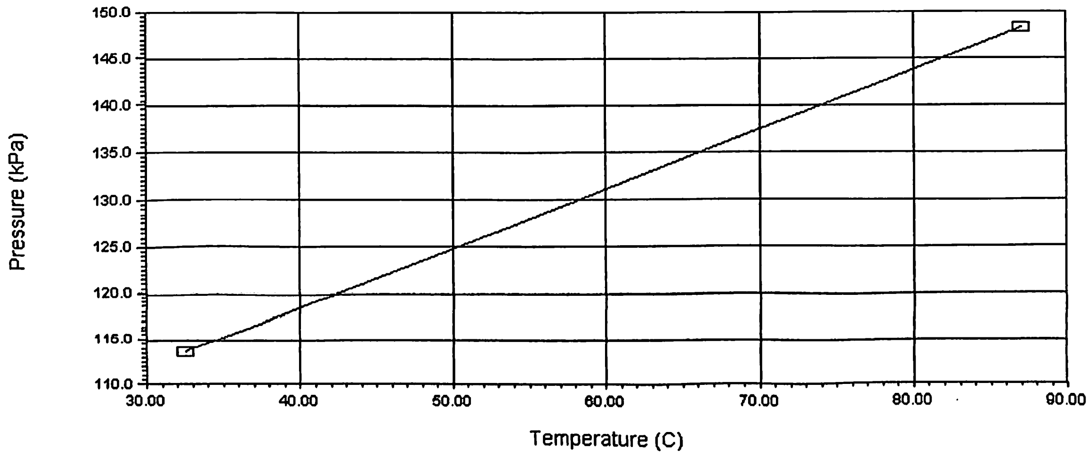
DIPA to Pump

	Overall	Aqueous Phase			
Molar Heat Capacity (kJ/kgmole-C)	93.82	93.82			
Mass Heat Capacity (kJ/kg-C)	3.578	3.578			
Thermal Conductivity (W/m-K)	0.4340	0.4340			
Viscosity (cP)	3.347	3.347			
Surface Tension (dyne/cm)	59.84	59.84			
Molecular Weight	26.22	26.22			
Z Factor	1.152e-003	1.152e-003			

PERFORMANCE PROFILES

Zone	Pressure (kPa)	Temperature (C)	Vapour Fraction	Enthalpy (kJ/kgmole)
Inlet	148.24	87.00	0.0000	-20265.95
0	113.76	32.53	0.0000	-25592.45

PERFORMANCE PLOT



PERFORMANCE TABLE

Overall Phase

Temperature (C)	Pressure (kPa)	Heat Flow (kJ/h)	Enthalpy (kJ/kgmole)	Vapour Fraction
87.00	148.24	0.00	-20265.95	0.0000
32.53	113.76	-8774284.06	-25592.45	0.0000

Vapour Phase

Mass Flow (kg/h)	Molecular Weight	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)
43192.45	26.22	37.80	3.88	---	---
43192.45	26.22	38.84	3.58	3.35	0.43

Light Liquid Phase

Mass Flow (kg/h)	Density (kg/m3)	Mass Specific Heat (kJ/kg-C)	Viscosity (cP)	Thermal Conductivity (W/m-K)	Surface Tension (dyne/cm)
---	---	---	---	---	---
---	---	---	---	---	---

HYPROTECH LIFECYCLE INNOVATION	TEAM LND Calgary, Alberta CANADA	Case Name:	E:\PRAKASH MUKHOPADHYAY -- DIPA.HSC
		Unit Set:	SI
		Date/Time:	Mon May 11 09:14:00 2009

Spreadsheet: SPRDSHT-1

Units Set: User Field

CONNECTIONS

Imported Variables

Cell	Object	Variable Description	Value
B1	Material Stream: RICH DIPA	Comp Molar Flow (DEAmine)	258.5994 lbmole/hr
B2	Material Stream: RICH DIPA	Comp Molar Flow (H2S)	199.2897 lbmole/hr
B3	Material Stream: RICH DIPA	Comp Molar Flow (CO2)	0.0000 lbmole/hr
B4	Material Stream: Lean DIPA to Contactor	Comp Molar Flow (DEAmine)	258.6003 lbmole/hr
B5	Material Stream: Lean DIPA to Contactor	Comp Molar Flow (H2S)	0.8767 lbmole/hr
B6	Material Stream: Lean DIPA to Contactor	Comp Molar Flow (CO2)	0.0000 lbmole/hr
D7	Material Stream: Lean DIPA to Contactor	Comp Volume Frac (H2S)	0.0004
D9	Material Stream: Lean DIPA to Contactor	Comp Mole Frac (H2S)	0.0002

Exported Variables' Formula Results

Cell	Object	Variable Description	Value

PARAMETERS

Exportable Variables

Cell	Visible Name	Variable Description	Variable Type	Value
D8	D8:		Vapour Fraction	2.4140e+02
D1	D1:		---	0.7707
D2	D2:		---	1.052e-007
D4	D4:		---	3.390e-003
D5	D5:		---	1.055e-007

User Variables

FORMULAS

Cell	Formula	Result
D1	=b2/b1	0.7707
D2	=b3/b1	1.052e-007
D4	=b5/b4	3.390e-003
D5	=b6/b4	1.055e-007
D8	=d9*1e6	2.4140e+02

Spreadsheet

	A	B	C	D
1	Rich DEA *	258.5994 lbmole/hr *	Rich H2S Loading *	0.7707 *
2	Rich H2S *	199.2897 lbmole/hr *	Rich CO2 Loading *	1.052e-007 *
3	Rich CO2 *	0.0000 lbmole/hr *		
4	Lean DEA *	258.6003 lbmole/hr *	Lean H2S Loading *	3.390e-003 *
5	Lean H2S *	0.8767 lbmole/hr *	Lean CO2 Loading *	1.055e-007 *
6	Lean CO2 *	0.0000 lbmole/hr *		
7			H2S Vol Frac *	0.0004 *
8			H2S ppm *	2.4140e+02 *
9				0.0002 *
10				