LPG COLUMN DESIGN AND SIMULATION

(A Report on Project)

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

By

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LPG COLUMN DESIGN AND SIMULATION

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:

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2. Internal Examiner:
3. Internal Examiner:

Head of Department:

Centre Director:

APRIL - 2009

DECLARATION

This is to declare that the project report entitled "**LPG Column Design and Simulation**" 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.

N. [signature]

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CERTIFICATE

This is to certify that the project work entitled "**LPG Column Design and Simulation**" being submitted by **Mr.Saravana Raj.N** (R030307011), 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

Prof K.V.Rao
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N.

(N.SARAVANA RAJ)

EXTENDED ABSTRACT

A project report to design LPG Column located in Gas Processing Plant which is processing 3 million cubic meters/day for recovering LPG. All the calculations are performed based upon particular composition which is given for this project.

Gas Processing plant has two UNITs, namely Crude Stabilisation Unit and Gas Processing Unit. The associated natural gas contains sour/rich gas as well unsaturated crude, the unsaturated crude is treated in CSU to get saturated crude and the sour/rich gas is processed in gas processing unit to get LPG, Naphtha, C_2C_3 and Lean gas.

Here our focus is mainly LPG Distillation Column which is part of LPG plant, The plant operation is as follows, Sweetened gas from GSU flows to knock out drum, from where liquid present is separated , then the gas is sent through exchangers and precooled. The precooled gas will be sent to KOD where liquefied hydrocarbon and water are separated out and the gas flows to the molecular sieve drier where the moisture is reduced, the dried gas is passed through filters, to retain any foreign or dust particles. Then the dried gas flows through 1st stage chiller. The process fluid next enters in 1st stage liquid separator where the partially condensed hydrocarbon liquids separated out. Gas vapours from the top of the vessel flows through 2nd stage chiller. Then the H/C fluid enters in 2nd stage liquid separator where 2nd stage liquid is separated. The Second Stage Vapours (SSV) after separation of liquid hydrocarbon (Condensate) are passed through and delivered as feed stock to C_2C_3 Recovery Unit. Also SSV can be sent directly to consumers through bypass line if C_2C_3 Recovery Unit is under shut down.

COLD BOX which serves to chill the feed gas by exchanging heat with various cold streams of the plant. Separator liquid from feed gas separator sent to

LEF column to remove the lighter fractions. The bottom liquid goes to LPG column. If ethane propane plant is under shutdown the LEF gases can be sent to consumer line after compressing through residue gas compressor. Also these gases can be used for regeneration of gas dryers.

The bottom liquid of the LEF column is allowed to enter the LPG column. The top product is Liquefied petroleum gas and bottom product is Natural gas liquids.

This particular LPG column is design using manually (SPREADSHEETS) as well as simulator (CHEMCAD 6.0.1). The various Tray columns (sieve – tray column, value tray column and Bubble cap tray column) and also packed column sizing are designed by CHEMCAD 6.0.1 Simulator. Costing for these different columns are prepared in CHEMCAD 6.0.1 Simulator and finally the cost comparison chart is carried out using SPREADSHEETS.

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1. INTRODUCTION

LPG is a mixture of light hydrocarbons which are gaseous at normal temperatures and pressures, and which liquefy readily at moderate pressures or reduced temperature. It is odourless and so, for safety reasons, a pungent compound, ethyl mercaptan, is added to make any leaks easily detectable.

Motor vehicles run on propane or a mixture of propane and butane. Propane and butane are gases at atmospheric pressure and temperature but can easily be liquefied for storage by an increase in pressure. The most common blend by volume is 60 per cent propane to 40 per cent butane. Industry uses propane or butane and the petrochemical industry may use both as a feedstock. Households use only propane.

1.1 History

Liquefied petroleum Gas is a mixture of C_2 – C_3 fractions. It is recovered from refinery Gases and Natural gas streams. It is essentially a twentieth century development long before time. Use has been made of gas compressed containers. Limited amounts of this product were sold in cylinders in England as early as 1810. In 1870, pintsch gas was developed and later used in railway car lighting. Around 1907. blangas formed by craking oil, was liquefied by compressing it to about 1800 psig. Several containers in the United States, despite attendant difficulties such as high transportation cost and complexity of equipment needed for its utilization.

1.2 Scope

Crude oil is a combustible liquid consisting of a mixture of hydrocarbon and compounds such as O_2, S, and N_2. Unstabilised crude oil is received through pipeline from offshore platforms (after two stage separation). A project report to design LPG Column located in Gas Processing Plant, processing 3 million cubic meters/day for recovering LPG. Important scope of this project is design, simulation and cost analysis done by using CHAMCAD 6.0.1 simulator which is capable of simulating any kind of feed composition.

1.3 Limitation

Here design project is for particular composition and feed condition which is mentioned in table 1, this designed column will not be support unless the feed condition and composition ,the basic parameters of distillation column like diameter, height, number of stages, wall thickness etc are vary with feed composition to composition as well as feed conditions. CHEMCAD 6.0.1 simulator is limited upto only 50 number of stages. More than that is not possible in this version.

1.4 Delimitation

Since project is done using CHEMCAD 6.0.1, any kind of feed condition and composition can be specified in simulator to sizing the Distillation column as well as analysis on characteristics of tower, feed stream and product stream. This simulator is capable of designing sieve tray, value tray, bubble cap tray and also packed column, this CHEMCAD 6.0.1 is special to costing of equipment and economics of plant.

1.5 Feed Composition of Gas Processing Plant

Table 1: Feed Composition Gas Processing Plant

Components	Molecular weight	Mole %	Moles	Kgs
N_2	28	1.42	79.3847	2222.7716
CO_2	44	1.75	97.83324	4304.6256
CH_4	16	69.24	3870.842	61933.472
C_2H_6	30	13.05	729.5564	21886.692
C_3H_8	44	9.27	518.2366	22802.4104
i C_4H_{10}	58	1.66	92.8018	5382.5044
n C_4H_{10}	58	2.35	131.3761	7619.8138
i C_5H_{12}	72	0.59	32.9838	2374.8336
n C_5H_{12}	72	0.67	37.4562	2696.8464
Average molecular weight = 23.4728				total = 131224 Kgs
Water in the Feed = 118.46 Kg/ (106Nm3/day)				
Associated gas = 3 x 118.46 Kg = 355.38				

2. LITERATURE REVIEW

2.1 Growth of Gas Processing Plant [8]

Crude oil is a combustible liquid consisting of a mixture hydrocarbon and compound O_2, S, and N_2. Unstabilised crude oil is received through pipeline from offshore platforms (after two stage separation). Stabilisation of crude oil is being done with three stage separation having intermediate dehydration unit.Here Table 2 is explain about growth of Gas Processing Plant.

Table 2: Growth of Gas Processing Plant.

TECHNOLOGICAL SCHEME PHASES	YEAR INSTALLED
Technological Scheme Phase I	1981 (CSU I & LPG I) CSU I - Crude Stabilization Unit I LPG I - Liquefied Petroleum Gas I
Technological Scheme Phase II	1984 (CSU II, LPG II & GT I & II) CSU II - Crude Stabilization Unit I LPG II - Liquefied Petroleum Gas I GT I & II - Gas Turbine I & II
Technological Scheme Phase III	1990(GSU12 &13 ,EPRU,CFU I) GSU12 &13 – Gas Sweetening Unit 12&13 EPRU - Ethane Propane Recovery Plant CFU I - Condensate Fraction Unit I
Minas Plant & CFU II	1994
Gas Turbine III	2001

2.2 Overview of Gas Processing Plant [1]

Following figures (figure 1 & 2) are explaining the overview idea about Gas Processing Plant, and also process description have described.

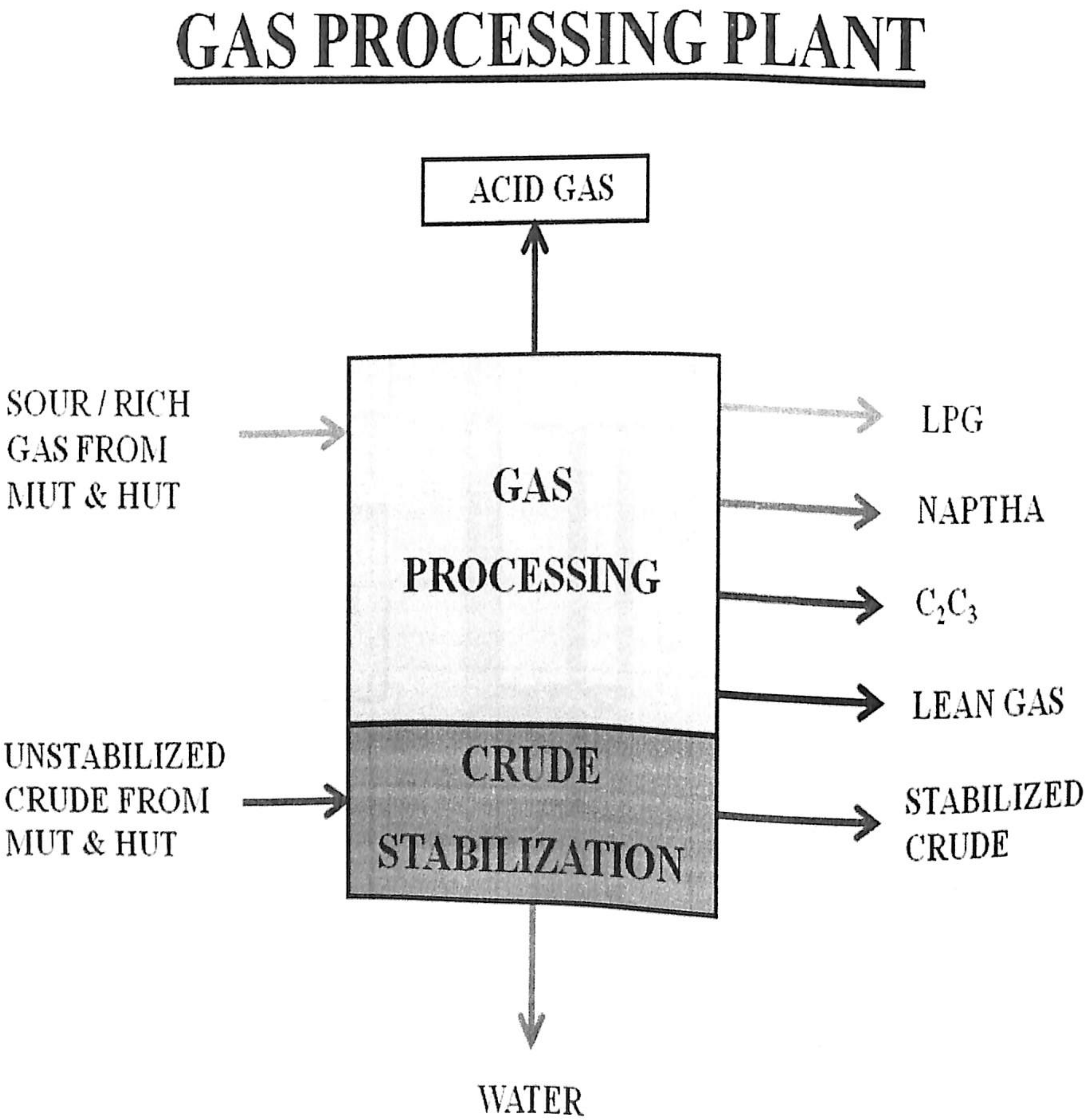


Figure 1: Simple Overview of Gas Processing Plant

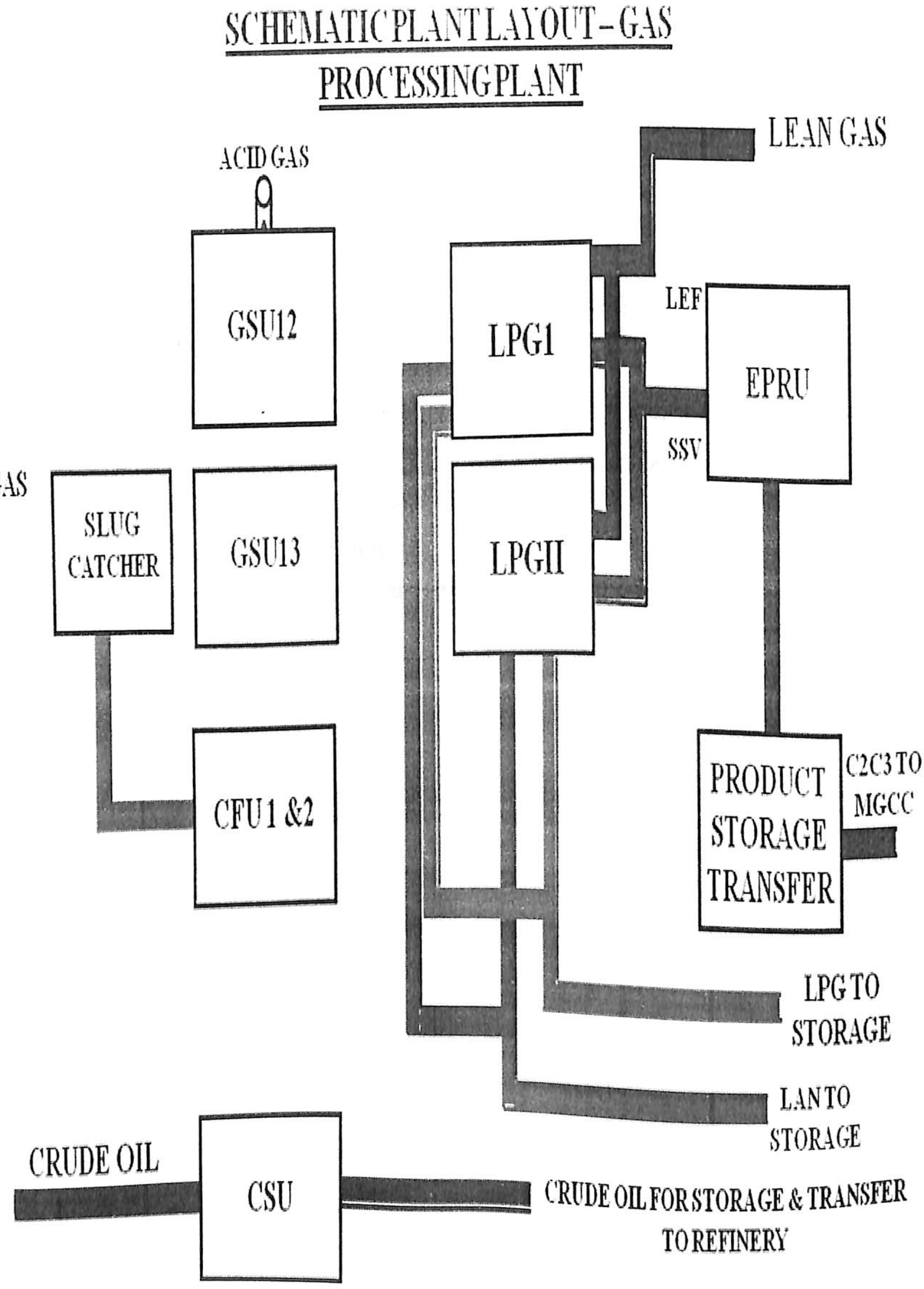


Figure 2: Schematic Plant Layout – Gas Processing Plant

2.2.1 Slug Catcher

It is based on the reducing of the fluid velocity and subsequent gravity separation. To hold the slug liquid reaching URAN at the time of pigging of gas pipeline. Provides room for gas - condensate separation and storage volume of liquid condensate. To continuously send the separated liquid condensate to condensate fractionation unit I & II for further processing.

To partially stabilize the liquid and inject into crude stabilization unit in case of CFU I & II or both in shutdown.

2.2.1 Crude Stabilization Units

Stabilisation of crude oil is required to minimize oil loss through evaporation during storage and transportation for which volatile components are removed. Solids and Salts present in crude oil is removed to minimize corrosive and abrasive wear of equipments. Water content of crude oil is reduced to minimize the cost of pumping and ensure BS& W requirement of refineries. Crude oil is stored in tanks and pumped to refineries as per their requirement.

2.2.3 Crude Sweetening Unit

Removal of undesirable materials such as acid gas includes mainly H_2S and CO_2.Acid gas is highly toxic. It is highly corrosive in presence of water. Forms hydrates and dry ice in cryogenic processes causing serious chocking and corrosion.

Commercial sweetening processes:

1. Chemical absorption

2. Physical absorption

3. Hybrid.

The absorption is at lower temperature and higher pressure and regeneration at high temperature and low pressure. Sulfinol solution is being recycled.

Criteria for process selection is depend on the concentration of impurities in the gas and the degree of removal desired.

GSU plant has two unit GSU 12 and GSU13.the absorption is at lower temperature and higher pressure and regeneration at high temperature and low pressure. Sulfinol solution is being recycled.

2.2.4 Condensate Fraction Unit

Stripping of lighter fraction and separating LPG and LAN with the help of fractionating column from condensate.

2.2.5 LPG Plant

LPG plant is recovering a maximum of butane – propane mixture from gases. The process involves cooling of gases under pressure to condense out LPG & natural gas liquid. The noncondensable gas goes as feed stock of C_2 - C_3 plant propane

2.2.6 Ethane Propane Recovery Plant

C_2 – C_3 feed streams are cooled to partially condensed them. the refrigeration is provided by passing the high pressure feed stream through expander and by a propane refrigeration system. the condensed feeds are fed in demethaniser to separate methane from top and C_2- C_3 mix from bottom. C_2- C_3 stored in spheres and pumped to MGCC NAGOTHANE for their petrochemical use.

2.3 Process Details of LPG – I Plant [1]

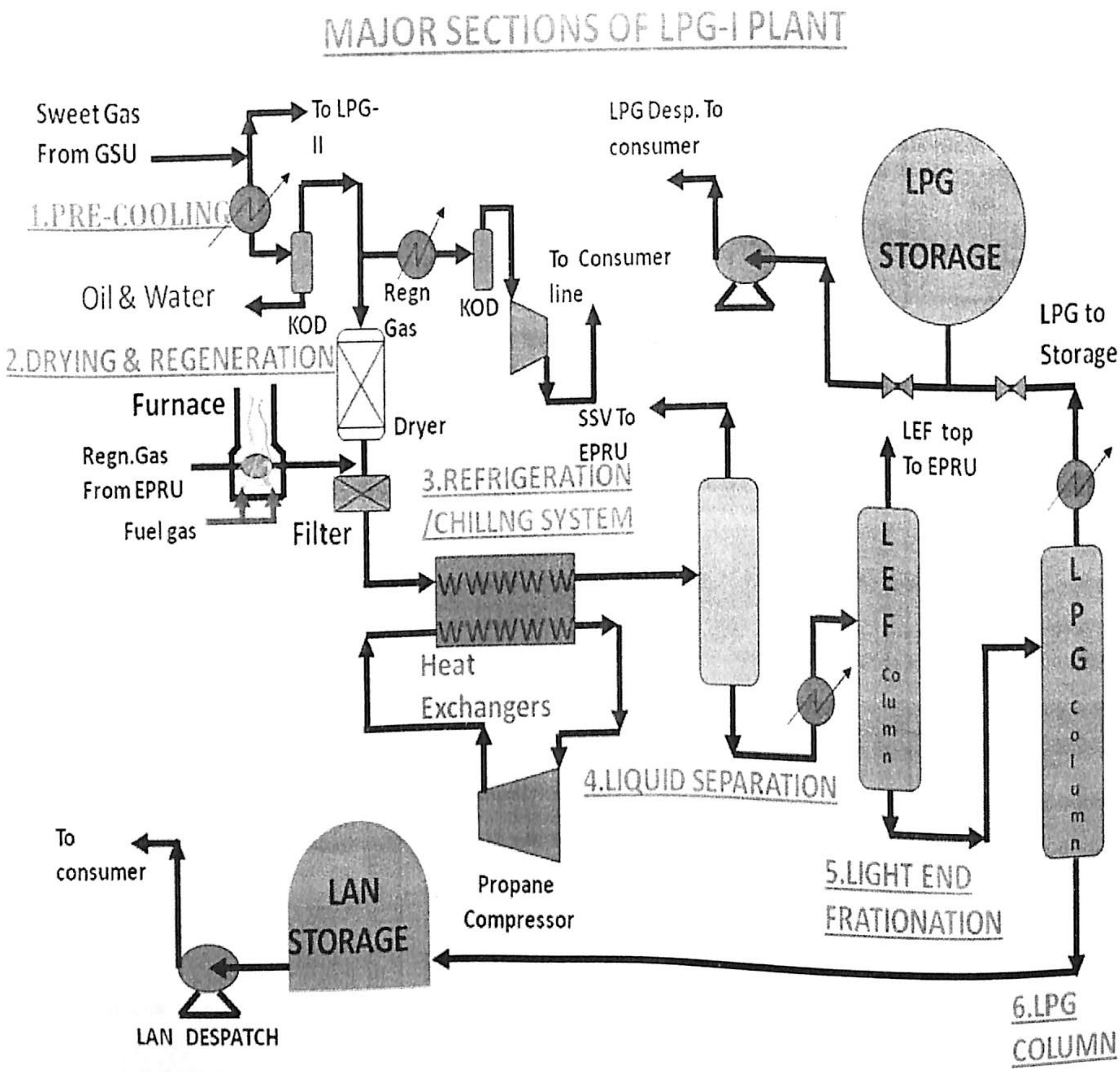


Figure 3: Process Flow Diagram of LPG Plant

2.3.1 Pre cooling & Drying:

Sweetened gas from GSU flows to knock out drum, from where any liquid present is separated , then the gas is sent through exchangers and precooled to 25^0 C . The precooled gas will be sent to KOD where liquefied hydrocarbon and water are separated out and sent to CSU through . The gas then flows to the molecular sieve drier where the moisture is reduced to less than 5ppm or a due point of -60^0 C.

2.3.2 *Refrigeration:*

The dried gas is passed through filters, to retain any foreign or dust particles. Then the dried gas flows through, and cooled to -22^0 C in the 1st stage chiller .The process fluid next enters in 1st stage liquid separator where the partially condensed hydrocarbon liquids separated out. Gas vapours from the top of the vessel flows through and finally cooled to -37^0 C in the 2nd stage chiller. Then the H/C fluid enters in 2nd stage liquid separator where 2nd stage liquid is separated. The Second Stage Vapours (SSV) after separation of liquid hydrocarbon (Condensate) are passed through and delivered as feed stock at a temp of 35^0C, to C_2C_3 Recovery Unit. Also SSV can be sent directly to consumers through bypass line if C_2C_3 Recovery Unit is under shut down.

2.3.3 *Cold Box:*

Chill down section and forms COLD BOX which serves to chill the feed gas by exchanging heat with various cold streams of the plant. These exchangers have aluminum brazed plates. The box is filled with insulation material – Pearlite and sealed with Nitrogen.

2.3.4 *Light End Fractionator:*

Separator liquid from feed gas separator sent to LEF column at around 20°C to remove the lighter fractions. The gas coming out from the top goes to LEF reflux drum and liquid is knocked out by external refrigeration, and the remaining gas is called as LEF top which is sent to ethane propane column after it gets heat from cold box. The bottom liquid goes to LPG column. If ethane propane plant is under shutdown the LEF gases can be sent to consumer line after compressing through residue gas compressor. Also these gases can be used for regeneration of gas dryers.

2.3.5 *LPG Column:*

The bottom liquid of the LEF column can enter the LPG column. The top product is Liquefied petroleum gas and bottom product is Natural gas liquids.

2.3.6 *Propane Column:*

The propane is used as a refrigerant in the refrigeration system. The propane losses which occur during refrigeration of feed gas requires make-up. Therefore a propane column is designed in LPG plant to recover propane taking the LPG as a feed to the column.

2.4 Uses and Properties of LP Gases

LP gases are mainly used for cooking purposes and as a fuel. The motor-vehicle sector continues to be the fastest-growing market for LPG. Motorists account for about 42% of World's domestic LPG consumption and automotive use has been growing at a rate of about 17 per cent per year. This growth has been largely due to the tax-free status of LPG which results in a significant pump price differential compared to petrol. LPG vehicle conversion kits are also exempt from sales-tax. If the production of LPG is abundant in the country, the gases can be used as a petrochemical feed stoke.

The properties of LP Gases are summarized in the ~~following~~ tables(Table 1 & Table 2). The figure for K – Value of several hydrocarbons at various Temperatures and pressures are also presented (De Priester chart – figure 4). The K – Values taken from these figure are used in the design calculations of distillation columns.

2.5 Initial Data for LPG Column [1]

Top temperature	= 60°C	= 333 K
Feed temperature	= 108°C	= 381 K
Bottom temperature	= 157°C	= 430 K

Table 3: Material Balance around LPG Column

Components	Feed (LEF Liquid)Kgs	Vapor (LPG)Kgs	Liquid(Natural gasoline)Kgs
C_2H_6	163.571	163.571	-
C_3H_8	12112.806	12112.806	-
i C_4H_{10}	5025.787	5025.787	-
n C_4H_{10}	7343.117	7342.501	1.756
i C_5H_{12}	2348.522	215.98	2132.861
n C_5H_{12}	2680.197	74.864	2604.883
	29674	24934.5	4739.5

Table 4: Physical property of LP gas [5]

(All values at 60 °F and 14.696 psia unless otherwise stated)

PROPERTIES	PROPANE	ISO – BUTANE	BUTANE
Molecular weight	44.09	58.12	58.12
Boiling point °F	-43.7	+10.9	+31.1
Boiling point °C	-42.1	-11.7	-0.5
Freezing point °F	-305.8	-255.0	-216.9
Density of liquid Specific gravity,(air=1) Degrees, API LB per gallon	0.508 147.2 4.23	0.563 119.8 4.69	0.584 110.6 4.87
Density of vapour (ideal gas) Specific gravity.(air=1) Lb gas per 1000cu ft	1.522 116.2	2.006 153.1	2.006 153.1
Total heat value (after vaporization) Btu per cu ft Btu per lb Btu per gal of liquid	2,563 21,663 91,740	3,269 21,258 99,790	3.39 21.008 103,830
Critical constants Pressure, psia Temperature, °F	617.4 206.2	537.0 272.7	550.1 306.0
Specific heat, Btu/lb °F Cp, vapour Cv , vapour Cp/Cv Cp, liquid 60 °F	0.388 0.343 1.13 0.58	0.387 0.348 1.11 0.56	0.397 0.361 1.10 0.55
Latent heat of vaporization at boiling point, Btu per lb	183.3	157.5	165.6
Vapour pressure. psia 0°F 70°F 100°F 100°F(ASTM10), psig max 130°F	37.8 124.3 188.7 210 274.5	11.5 45.0 71.8 109.5	7.3 31.3 51.6 70 80.8

- Properties are for commercial products and vary with composition

Table 4: Combustion Data for LP – gases [5]

(All values at 60 °F and 14.696 psia unless otherwise stated)

PROPERTIES	PROPANE	ISO - BUTANE	BUTANE
Flash temperature, °F	-156	-117	-101
Ignition temperature, °F	932	950	896
Maximum Flame temperature in air, °F			
Observed	3497	3452	3443
Calculated	3583	3583	3583
Flammability limits, % gas in air			
Lower	2.37	1.30	1.36
Higher	9.50	8.44	8.41
Maximum rate flame propagation in 1 in, tube			
Inches per second	32	33	33
Percentage gas in air	4.6 – 4.8	3.6 – 3.8	3.6 – 3.8
Required for complete combustion (ideal gas)			
Air, cu ft per cu ft gas	23.9	31.1	31.1
lb per lb gas	15.7	15.5	15.5
Oxygen, cu ft per cu ft gas	5.0	6.5	6.5
lb per lb gas	3.63	3.58	3.58
Product of combustion (ideal gas)			
Carbon dioxide, cu ft per cu ft gas	3.0	4.0	4.0
lb per lb gas	2.99	3.03	3.03
Water vapour, cu ft per cu ft gas	4.0	5.0	5.0
Nitrogen, cu ft per cu ft gas	18.9	24.6	24.6

- Properties are for commercial products and vary with composition

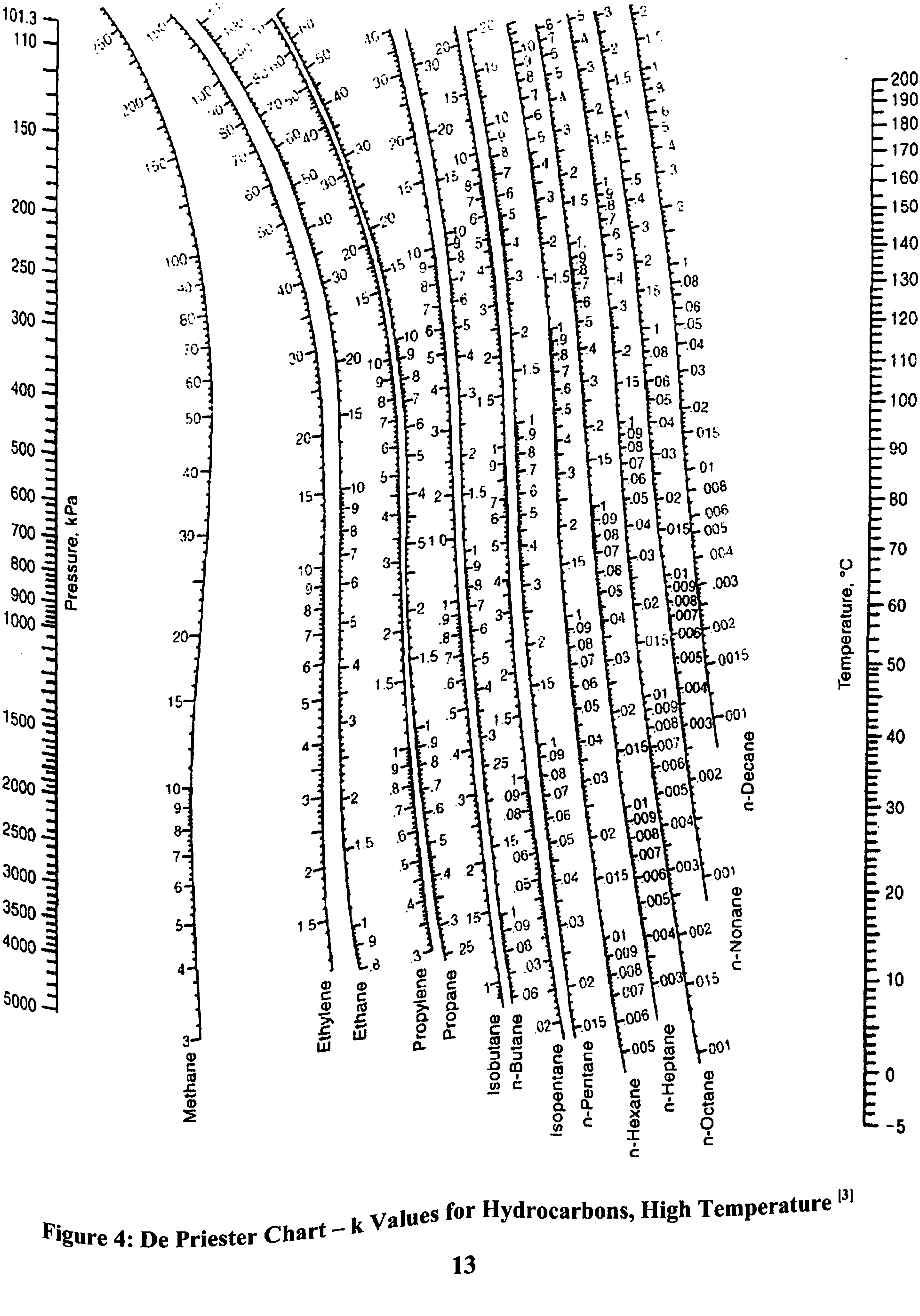


Figure 4: De Priester Chart – k Values for Hydrocarbons, High Temperature [3]

3 PROCESS DESIGN OF DISTILLATION COLUMN – EXCEL – 2007 PROGRAME

To estimate the stage, and the condenser and reboiler temperatures, procedures are required for calculating dew and bubble points. then design the sieve tray column. Which involves following three steps.

Step 1: Dew Points And Bubble Points

Step 2: Mccabe-Thiele Design Method

Step 3: Sieve Tray Design

3.1 Dew Points and Bubble Points [3]

To estimate the stage, and the condenser and reboiler temperatures, procedures are required for calculating dew and bubble points. By definition, a saturated liquid is at its bubble point (any rise in temperature will cause a bubble of vapour to form), and a saturated vapour is at its dew point (any drop in temperature will cause a drop of liquid to form). Dew points and bubble points can be calculated from a knowledge of the vapour-liquid equilibrium for the system. In terms of equilibrium constants, the Bubble point is defined by the equation (1) Dew point is calculating by equation (2).

Bubble point: $$\sum y_i = \sum K_i x_i = 1.0 \qquad (1)$$

[3]

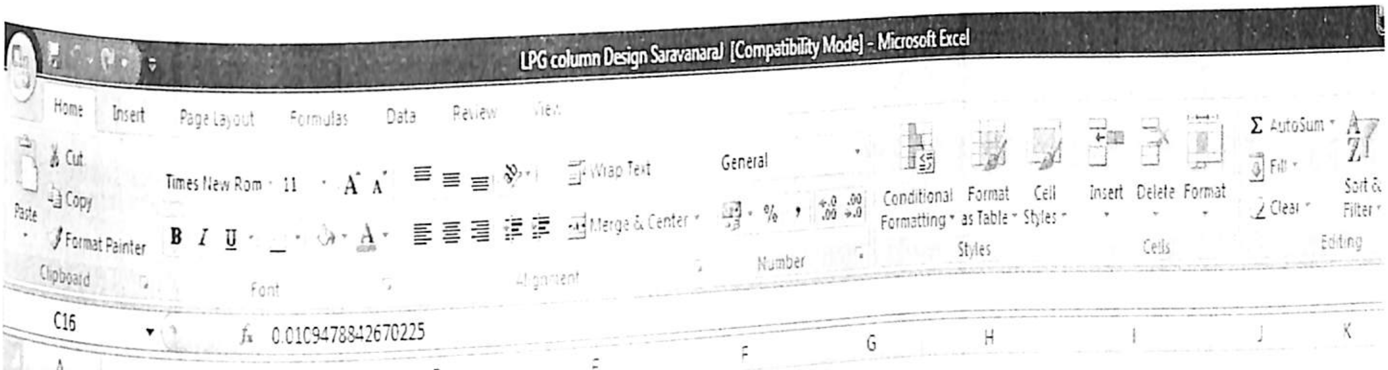
Dew point: $$\sum x_i = \sum (y_i / K_i) = 1.0 \qquad (2)$$ [3]

For multicomponent mixtures the temperature that satisfies these equations (3) & (4), at a given system pressure, must be found by trial and error. For binary systems the equations can be solved more readily because the component compositions are not independent; fixing one fixes the other.

$$y_a = 1 - y_b \qquad (3)$$ [3]

$$x_a = 1 - x_b \qquad (4)$$

[3]



BUBBLE POINT - DEW POINT CALCULATION

Component	"Feed" in Kg	Mol wt	"F" in moles/hr	"Distillate" in Kg	"D_i" in Kg moles/hr	X_{id}	"Bottom" in Kg	"B" in Kg moles/hr	$X_{i,b}$
Eathane	163.571	30	5.452366667	163.571	5.452366667	0.010947884	-	-	-
Propane	12112.806	44	275.2910455	12112.806	275.2910455	0.55276079	-	-	-
i-Butane	5025.787	58	86.6515	5025.787	86.6515	0.173988774	-	-	-
n-Butane	7343.117	58	126.6054655	7342.501	126.5948448	0.254191582	1.756	0.030275862	0.000459894
i-Pentane	2348.522	72	32.61836111	215.98	2.999722222	0.006023185	2132.861	29.62306944	0.449977903
n-Pentane	2680.197	72	37.22495833	74.864	1.039777778	0.002087785	2604.883	36.17893056	0.549562203
			563.8436971	24935.509	498.0292569	1		65.83227586	1

BUBBLE- POINT CALE

Component	"D_i" in Kg moles/hr	$X_{i,d}$	"B" in Kg moles/hr	$X_{i,b}$	Trail 01 (100°C) K_i	Trail 01 (100°C) $K_i X_{i}b$	Trail 02 (130°C) K_i	Trail 02 (130°C) $K_i X_{i}b$	Trail 03 (135°C) K_i	Trail 03 (135°C) $K_i X_{i}b$
Eathane	5.452366667	0.010947884	-	-	-	-	-	-		
Propane	275.2910455	0.55276079	-	-	-	-	-	-		
i-Butane	86.6515	0.173988774	-	-	-	-	-	-		
n-Butane	126.5948448	0.254191582	0.030275862	0.000459894	1.22	0.000561071	1.8	0.000827809	[illegible]	[illegible]
i-Pentane	2.999722222	0.006023185	29.62306944	0.449977903	0.65	0.292485637	1	0.449977903	1.1	[illegible]
n-Pentane	1.039777778	0.002087785	36.17893056	0.549562203	0.55	0.302259212	0.9	0.494605983	0.9	[illegible]
	498.0292569	1	65.83227586	1	$\sum K_i X_{i}b$	0.595305919	$\sum K_i X_{i}b$	0.945411695	$\sum K_i X_{i}b$	[illegible]

DEW - POINT CALE

Component	"D_i" in Kg moles/hr	$X_{i,d}$	"B" in Kg moles/hr	$X_{i,b}$	Trail 01 (50°C) K_i	Trail 01 (50°C) $K_i X_{i}d$	Trail 02 (70°C) K_i	Trail 02 (70°C) $K_i X_{i}d$	Trail 03 (44°C) K_i	Trail 03 (44°C) $K_i X_{i}d$
Eathane	5.452366667	0.010947884	-	-	4.2	0.045981114	5.1	0.05583421	3.7	0.040507
Propane	275.2910455	0.55276079	-	-	1.5	0.829141185	2.1	1.160797659	1.34	[illegible]
i-Butane	86.6515	0.173988774	-	-	0.7	0.121792142	1.05	0.182688213	0.6	[illegible]
n-Butane	126.5948448	0.254191582	0.030275862	0.000459894	0.52	0.132179623	0.76	0.193185603	0.43	[illegible]
i-Pentane	2.999722222	0.006023185	29.62306944	0.449977903	2.6	0.01566028	0.37	0.002228578	[illegible]	[illegible]
n-Pentane	1.039777778	0.002087785	36.17893056	0.549562203	0.19	0.000396679	0.3	0.000626335	0.16	[illegible]
	498.0292569	1	65.83227586	1	$\sum K_i X_{i}d$	1.145151023	$\sum K_i X_{i}d$	1.595360598	$\sum K_i X_{i}d$	[illegible]

Bubble - Point-----------------> 135 °C

Dew - Point-----------------> 44 °C

3.2 Mccabe-Thiele Design Method [3]

The vapour-liquid equilibrium characteristics (indicated by the shape of the equilibrium curve) of the mixture will determine the number of stages, and hence the number of trays, required for the separation. This is illustrated clearly by applying the **McCabe-Thiele** method to design a binary column.

The McCabe-Thiele approach is a graphical one, and uses the VLE plot to determine the theoretical number of stages required to effect the separation of a binary mixture. It assumes ***constant molar overflow*** and this implies that:

- Molal heats of vaporisation of the components are roughly the same
- Heat effects (heats of solution, heat losses to and from column, etc.) are negligible
- For every mole of vapour condensed, 1 mole of liquid is vaporized.

The design procedure is simple. Given the VLE diagram of the mixture, operating lines are drawn first.

- Operating lines define the mass balance relationships between the liquid and vapor phases in the column.
- There is one operating line for the bottom (stripping) section of the column, and on for the top (rectification or enriching) section of the column.
- Use of the constant molar overflow assumption also ensures the operating lines are straight lines

LPG column Design Saravanaraj [Compatibility Mode] - Microsoft Excel

McCABE - THIELE Method

R= 1

Component	"Feed" in Kg	Mol wt	"F" in moles/hr	"Distillate" in Kg	"D_i"in Kg moles/hr	"Bottom" in Kg	"B" in Kg moles/hr
Eathane	163.571	30	5.452366667	163.571	5.452366667	-	-
Propane	12112.806	44	275.2910455	12112.806	275.2910455	-	-
i-Butane	5025.787	58	86.6515	5025.787	86.6515	-	-
n-Butane	7343.117	58	126.6054655	7342.501	126.5948448	1.756	0.030275862
i-Pentane	2348.522	72	32.61836111	215.98	2.999722222	2132.861	29.62306944
n-Pentane	2680.197	72	37.22495833	74.864	1.039777778	2604.883	36.17893056
			563.8436971	24935.509	498.0292569	4739.5	65.83227586

To calculate Maximum Vapour rate (m^3/s) from y_i*Mwt

Component	k Top	K Bott	K_{avg}	α_i
Eathane	3.7	8.5	6.1	9.457364341
Propane	1.34	4.2	2.77	4.294573643
i-Butane	0.6	2.5	1.55	2.403100775
n-Butane	0.43	2	1.215	1.88372093
i-Pentane	0.19	1.1	0.645	1
n-Pentane	0.16	0.9	0.53	0.821705426

yi	yi*Mol wt
0.010947884	0.328436528
0.55276079	24.32147475
0.173988774	10.09134891
0.254191582	14.74311177
0.006023185	0.433669302
0.002087785	0.150320486

Mwa = 50.06836175

Calc of non - Key flows

Component	α_i	"D_i"in Kg moles/hr	$l_i=d_i/(\alpha_i-1)$	vi=li+di
Eathane	9.457364341	5.452366667	0.644688634	6.097055301
Propane	4.294573643	275.2910455	83.55892909	358.8499745
i-Butane	2.403100775	86.6515	61.75714641	148.4086464
		$\sum l_i=$	145.9607641	513.3556763

Component	α_i	"B" in Kg moles/hr	$V'_i=\alpha_i b_i/(\alpha_{LK}-\alpha_i)$	$l'_i=V'_i+b_i$
n-Pentane	0.821705426	36.17893056	27.99245722	64.17138777
		$\sum V'_i=$	27.99245722	64.17138777

Flows of combined key

L_e=	352.0684928
V_e=	482.7028376
V'_e=	968.0660567
L'_e=	997.719402

slope of top operating line

L_e/V_e=	0.729369014

slope of bottom operating line

L'_e/V'_e=	1.030631531

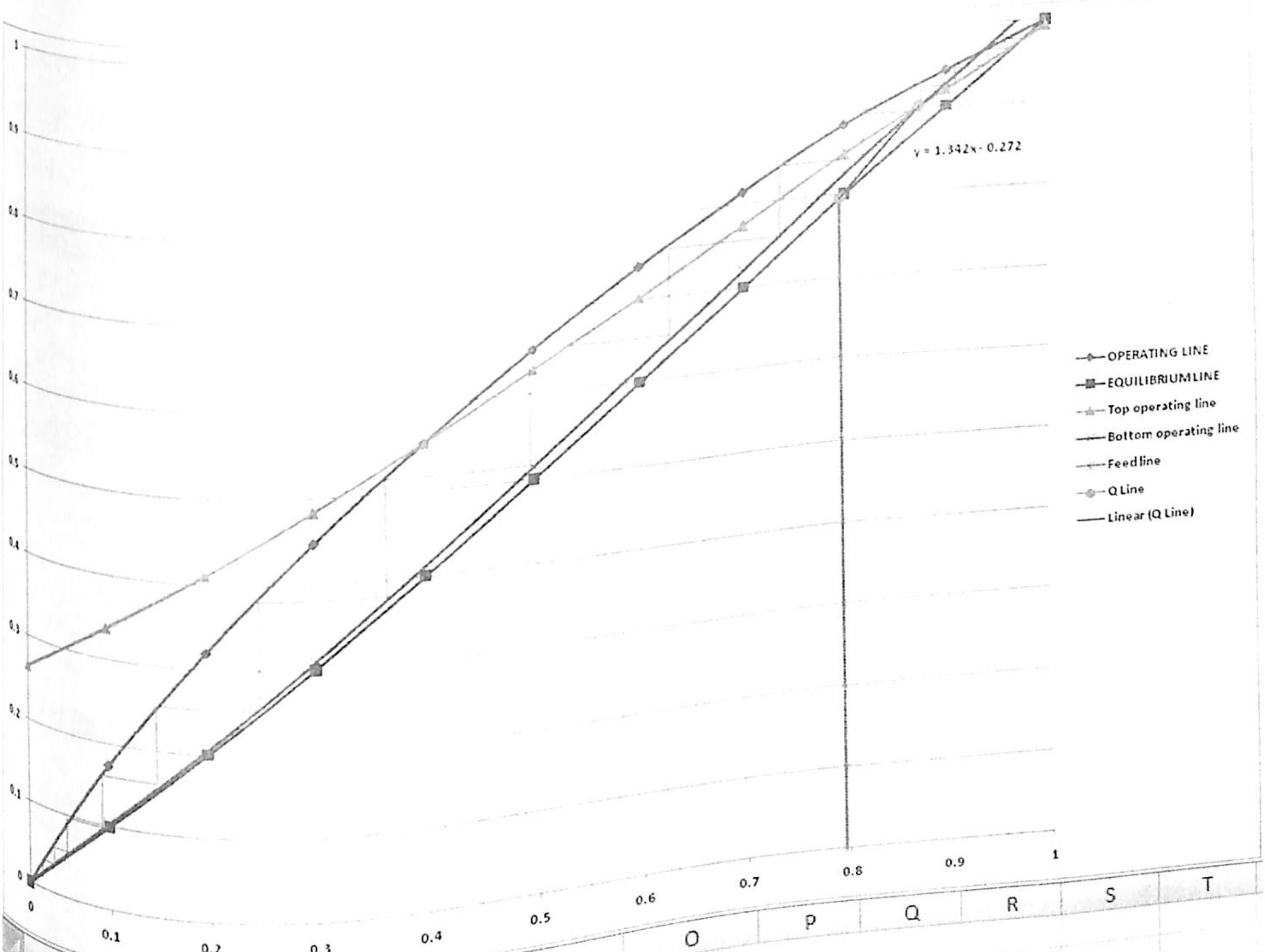
x_b=	0.001020993
x_d=	0.976853025
x_f=	0.795141457
α=	1.88372093

X=		y=
	0	0
	0.1	0.173076923
	0.2	0.320158103
	0.3	0.446691176
	0.4	0.556701031
	0.5	0.653225806
	0.6	0.738601824
	0.7	0.814655172
	0.8	0.882833787
	0.9	0.944300518
	1	1

45°line

X=		Y=
	0	0
	0.1	0.1
	0.2	0.2
	0.3	0.3
	0.4	0.4
	0.5	0.5
	0.6	0.6
	0.7	0.7
	0.8	0.8
	0.9	0.9
	1	1

	A	B	C	D	E	F	G
79	Top operating line		Eqn1---------->Y=0.729X+.264				
80	C=	0.264366697					
81	X=	0	Y=	0.264366697			
82		0.1		0.337303599			
83		0.2		0.4102405			
84		0.3		0.483177401			
85		0.4		0.556114303			
86		0.5		0.629051204			
87		0.6		0.701988106			
88		0.7		0.774925007			
89		0.8		0.847861908			
90		0.9		0.92079881			
91		1		0.993735711			
92						Feed line	
93	Bottom operating line		Eqn2---------->Y=1.031X-3.127E-05			X	Y
94	C=	-3.12746E-05				0.795141457	0
95	X=	0	Y=	-3.12746E-05		0.795141457	0.1
96		0.1		0.103031879		0.795141457	0.2
97		0.2		0.206095032		0.795141457	0.3
98		0.3		0.309158185		0.795141457	0.4
99		0.4		0.412221338		0.795141457	0.5
100		0.5		0.515284491		0.795141457	0.6
101		0.6		0.618347644		0.795141457	0.795141457
102		0.7		0.721410797		Q line	
103		0.8		0.82447395		0.795141457	0.795141457
104		0.9		0.927537103		0.87427	0.90134
105		1		1.030600256			



solve Eqn1,Eqn2-------------------------------> X=0.87427

Y=0.90134

Robinson and gilliland Equation

S' =	1.030632	S =	0.729369
K' =	1.730769	K =	1.049223
x'_r =	0.075	X_r =	0.86
x_b =	0.001021	x_d =	0.976853
		N^*_r =	3.130594
N^*_s =	9.20705		

$$N^*_s = \frac{\log\left[\dfrac{\left(\frac{K'}{s'}-1\right)\left(\frac{x'_r}{x_b}-1\right)}{\frac{1}{s'}(K'-1)}+1\right]}{\log\left(\frac{K'}{s'}\right)}+1$$

$$N^*_r = \frac{\log\left[\dfrac{(1-s)+x_r/x_d(s-K)}{1-K}\right]}{\log\left(\frac{s}{K}\right)}-1$$

From Graph Q Line Slope m=1.34-------------------------> m = 1.34

m=q/(q-1)-----------------------------------> q = 3.941176471

From Graph Top Operating Line Intersect -----------------> Y= 0.264366697

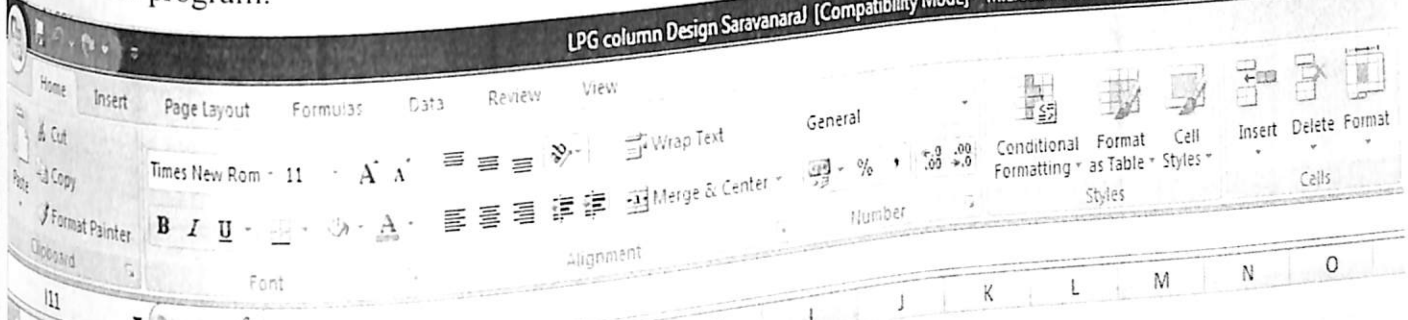
$Y=X_D/(R+1)$--------------------------------> R= 2.695068386

Rectification section	5 Trays
stripping section	20 trays
Feed plate	1 tray
Reboiler	1 tray
Total------------------------------------>	27 trays

Real Stages N=((no of theor st -1)/stage eff) = 43.3333 **44 Trays**

3.3 Sieve Tray Design [13]

Here the stepwise procedure as well as each formulas which involves in that steps, are explain in ExceL - 2007 program.



SIEVE – TRAY DESIGN

STEP 01: Flow Rate Of Enriching Section:

G=L+D

R=L/D

RR= 2.6950684

L=RD

L= 1342.2229 Kmol/hr

G= 1840.2522 Kmol/hr

B= 65.832276 Kmol/hr

STEP 02: Density:

Liquid Density---->= 634.4994 Kg/m^3

Vapor Density----->= 4.306843 Kg/m^3

Molecular Weight->= 51

STEP 03: Vapor Load:

$V=D(R+1)*mol.wt/(3600*\rho_V)$

V= 303.07447 m^3/sec

STEP 04: Tray Spacing:

Let us assume the tray spacing: 18 inch

0.457 m

STEP 05: Allowable Vapor Velosity:

$V_t=(L/V)*(\rho_v/\rho_L)$

V_t= 49.507972 m/s

STEP 06: Cross section area of column:

Cross section area of column = vapour flow rate / allowable vapor Velocity

Cross section area of column = 6.12173068 m^2

STEP 07: **Area:**

Total Column Area = Cross Sectional area 0.88

Total Column Area A_c = 6.956512 m^2

Net Area A_n = 6.121731 m^2

Down comer area A_d = 0.834781 m^2 Assume 12 % A_d in A_c

Active area A_a = 5.286949

Hole area A_h = 0.528695 Assume 10 % A_h in A_a

STEP 08: **column Diameter:**

D_c=((4*A)/π)^.5

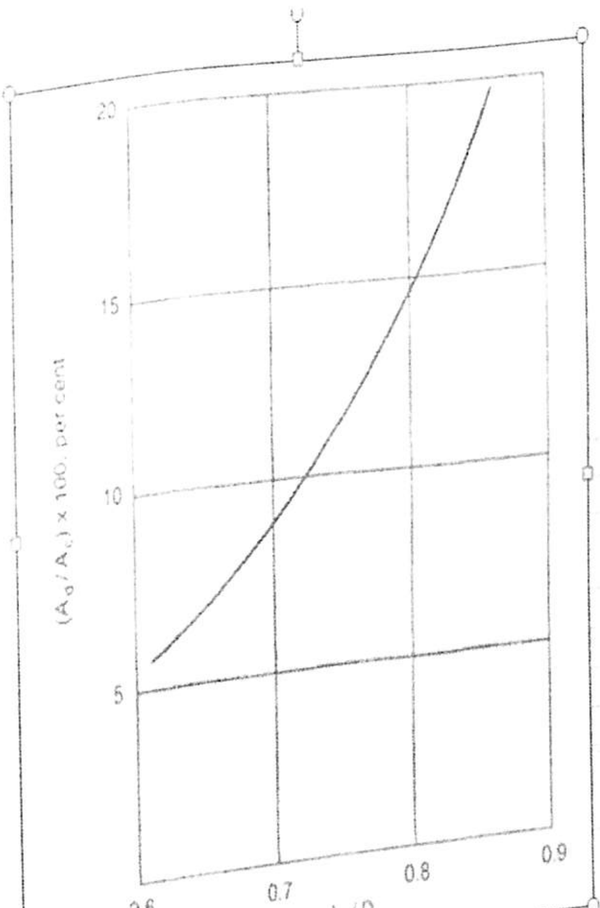
D_C= 2.9768774 m

Column Thinkess is: 0.06 m

OD= 3.0368774 m

The hole size plate varies from 3mm to 12mm hance we assume as 8 mm

0.008 mm



STEP 09: **Thinkness of sieve plate:**

Thinkness = 0.1 to 1.2 hole size

Let us take 0.6 times

Thinkness = 4.8 mm

0.0048 m

STEP 10: **Wire Length (L_w):**

From Graph L_W/D_C = 0.765 By using (Ad/Ac)*100 = 12

Wire Length L_W = 2.277311 m

STEP 11: **Hight of froath in mm (h_{ow}):**

H_{ow} = 43.4 $(Q_L/L_W)^{(2/3)}$

Q_L = flow rate of clear liquid (m^3/min)

Q_L =(Mass flow rate of distillate *R) / (ρ_L * 60)

Q_L = 1.7652473 m^3/min

H_{OW} = 36.622264 mm

STEP 12: **Perforated area (A_p):**

L_W/D_C = 0.765 From graph By using ----> L_w/D_c = 0.765

θ_C = 104 ° 76 °

Angle subtended at plate edge by unperforated strip = 3.880389 m

Mean length, unperforated edge strips = 0.194019 m^2

Area of unperforaded edge strips = 2.887738 m

Mean length of clamping zone = 0.288774 m^2

Area of calaming zone = 4.804156 m^2

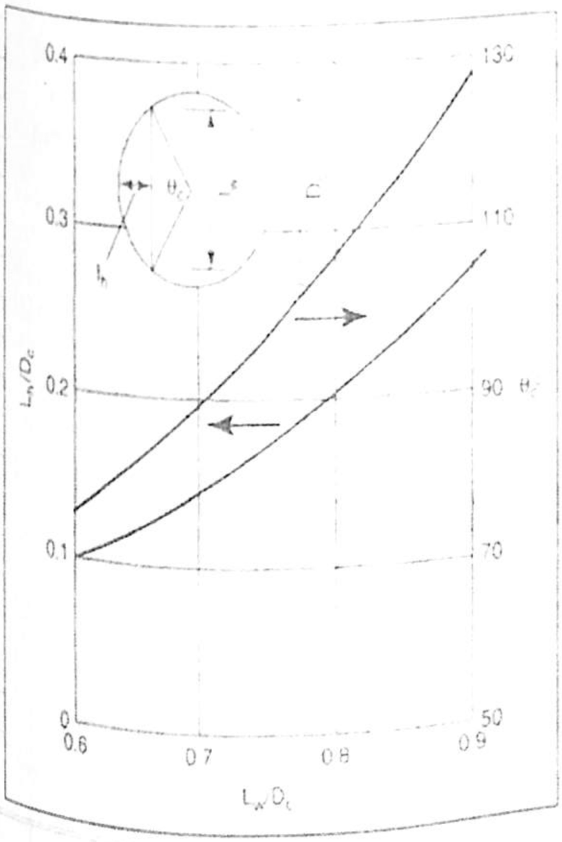
Total area for perforations, A_p =

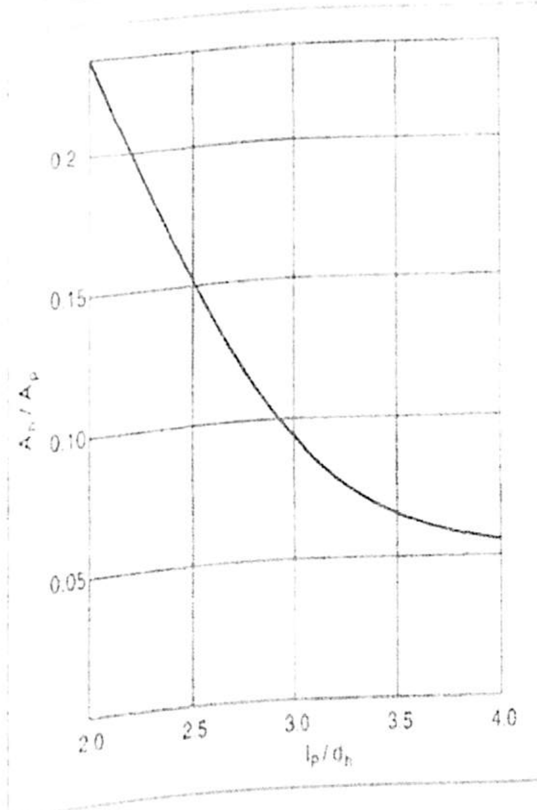
Assume 50 mm unperferated strip round pplate edge

Assume 50 mm wide claming zones

A_h/A_p = 0.110049

L_p/d_h 2.82 From graph By using A_h/A_p 0.110049





STEP 13: Number of holes :

Area of one hole = 0.00005024 m^2

Number of Holes = 10523.3862

STEP 14: Check weeping:

Maiximum liquid rate = 1.31652778 Kg/s

Maximum lqd rate - 70 per cent turn-down = 0.92156944 Kg/s

Assume 50 mm Weir height

h_w+h_{ow} = 86.6222635 mm

K_2 = 30.8 From Graph By using h_w+h_{ow} = 86.62226 mm

$\breve{u}_h$ *(min) = $(K_2-0.90*(25.4-d_h))/(\rho_v)^{0.5}$ m/s*

$\breve{u}_h$ (min) = 7.29535374 m/s

vapour rate in (moles/sec) = 138.34146 moles/s

Maximum Vapor rate (m^3/s) = 11.9000289 m^3/s

Actual minimum vapour velosity = (Min vapour rate/A_h) m/s

Actual minimum vapour velosity 15.7558165 m/s

So minimum operating will be will abow weep point

STEP 15: Plate pressure drop:

Maximum vapour velocity through holes $\hat{u}_h$

$\hat{u}_h$ = 22.5083093 m/s

Co = 0.75 From Graph by using (plate thinkness / plate dia), (A_h / A_p):

Dry plate drop

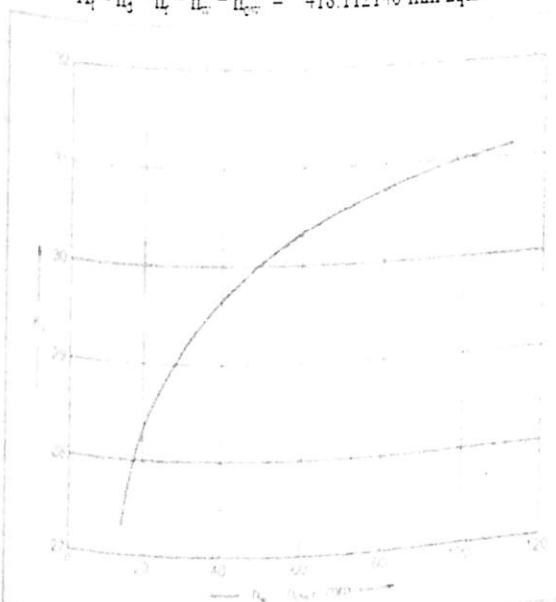
$h_d = 51*((\hat{u}h/Co)^2)*(\rho_v/\rho_l)$ = 311.789313 mm liquid

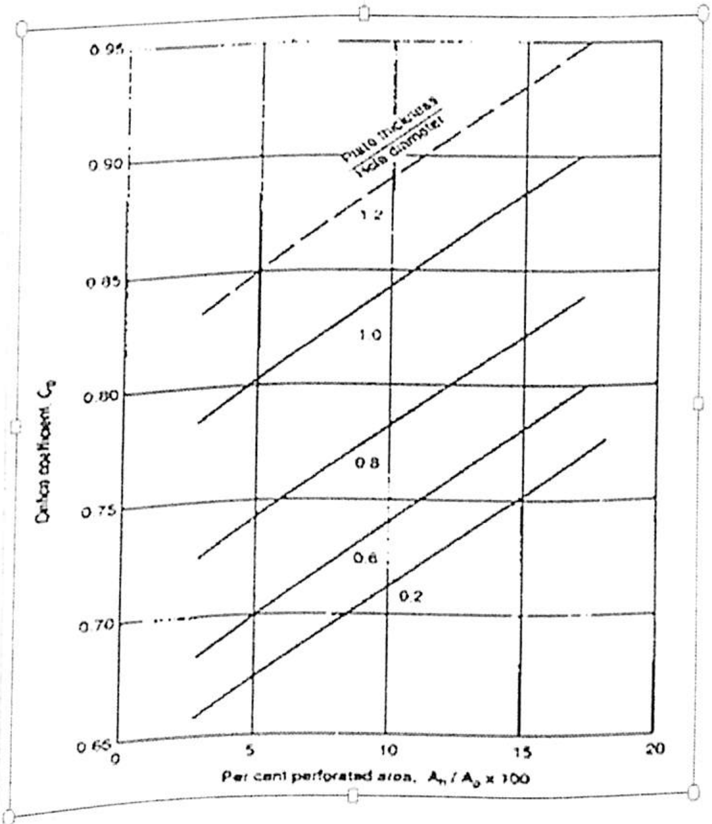
Residual head

$hr = 12.5*1000/\rho_l$ = 19.7005702 mm liquid

Total plate pressure drop

$H_t = h_d + h_r + h_w + h_{ow}$ = 418.112146 mm liquid





STEP 16: Down comer liquid back-up:

Down comer pressure loss

$h_{ap} = h_w - 10$ = 40 mm

Area under apron, A_{ap} = 0.091092 m^2

As this is less than A_d = 0.834781 m^2 so, $A_d > A_{ap}$:

$h_{dc} = 166(L_{wd}/\rho_L A_{ap})^2$ = 0.086127 mm

Downcomer backup measured from plate surface, mm

$H_b = h_w + h_{ow} + H_d + H_t$ = 816.5237 mm liquid

H_b = 0.816524 m

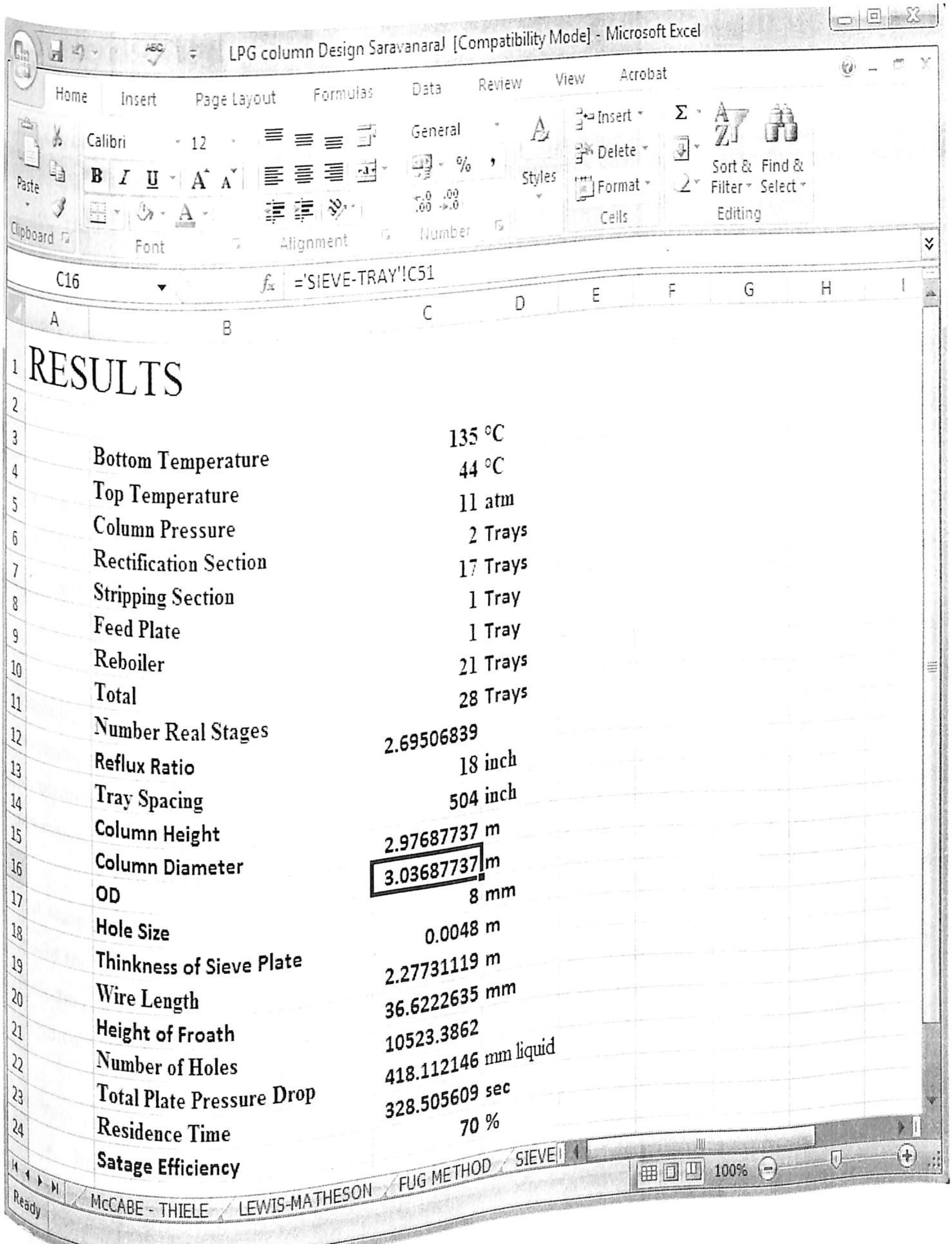
Plate spacing = 18 in assume

816.523723 <(1/2)*(Plate spacing+Weir length): 1367.2556 mm

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

STEP 17: Residence time:

$t_r = (A_d H_{bc} \rho_l)/L_{wd}$ = 328.5056


LPG column Design SaravanaraJ [Compatibility Mode] - Microsoft Excel
Home
Insert
Page Layout
Formulas
Data
Review
View
Acrobat
C16
='SIEVE-TRAY'!C51
RESULTS
135 °C
Bottom Temperature
44 °C
Top Temperature
11 atm
Column Pressure
2 Trays
Rectification Section
17 Trays
Stripping Section
1 Tray
Feed Plate
1 Tray
Reboiler
21 Trays
Total
28 Trays
Number Real Stages
2.69506839
Reflux Ratio
18 inch
Tray Spacing
504 inch
Column Height
2.97687737 m
Column Diameter
3.03687737 m
OD
8 mm
Hole Size
0.0048 m
Thinkness of Sieve Plate
2.27731119 m
Wire Length
36.6222635 mm
Height of Froath
10523.3862
Number of Holes
418.112146 mm liquid
Total Plate Pressure Drop
328.505609 sec
Residence Time
70 %
Satage Efficiency
McCABE - THIELE
LEWIS-MATHESON
FUG METHOD
SIEVE
Ready
100%

4. PROCESS DESIGN OF DISTILLATION COLUMN - CHEMCAD 6.0.1

4.1 Overview of CHEMCAD 6.0.1 and Its Uses [7]

Chemical processing industry (CPI) faces numerous challenges, rising fuel and feedstock costs, reduced engineering staff, shorter product life cycles, increased global competition, and increased regulation. These challenges require that CPI companies seek out and use the best tools to increase productivity and improve engineering decisions.

CHEMCAD is a powerful and flexible chemical process simulation environment, built around three key values of innovation, integration, and open architecture. These values create important advantages for us:

- The latest chemical engineering techniques at our fingertips
- All functionality united in a single software environment
- Seamless connection to the chemical engineering computing environment, with links to tools such as MS Excel and Word and interfaces such as COM, DCOM, OPC, CAPE - OPEN, and XML

CHEMCAD combines a state - of - the - art graphical user interface (GUI), an extensive chemical component database, a large library of thermodynamic data, and a library of the most common unit operations to give users the ability to provide significant and measurable returns on their investment. In addition, the program is customizable to allow custom chemicals, thermodynamics, unit operations, Calculations, and reporting - all ingredients for a powerful user experience.

CHEMCAD is capable of modelling continuous, batch, and semi - batch processes, and it can simulate both steady - state and dynamic systems. This program is used extensively around the world for the design, operation, and maintenance of chemical processes in a wide variety of industries, including oil and gas exploration, production, and refining; gas processing; commodity and specialty chemicals; pharmaceuticals; bio fuels; and process equipment manufacturing.

Within all of these industries, chemical engineers work every day with CHEMCAD to address a variety of challenges:

- Initial design of new processes
- Optimization or de - bottlenecking of existing processes
- Performance monitoring of processes
- Design and rating of process equipment such as vessels, columns, heatexchangers, piping, valves, and instrumentation

- Evaluation of safety relief devices
- Heat exchanger sizing
- Pressure and flow balancing of complex piping networks
- Reconciliation of plant data
- Economic comparisons of process alternatives
- Advanced process control (APC), including model predictive control (MPC), Real - time optimization (RTO), and operator training systems (OTS)
- Binary interaction parameter (BIP) regression from process or lab data

CHEMCAD is capable of delivering the results you need to stay competitive in an increasingly fast and fluid global market. Easy to learn and highly customizable, CHEMCAD can put future - proof solutions within easy reach of our engineering staff

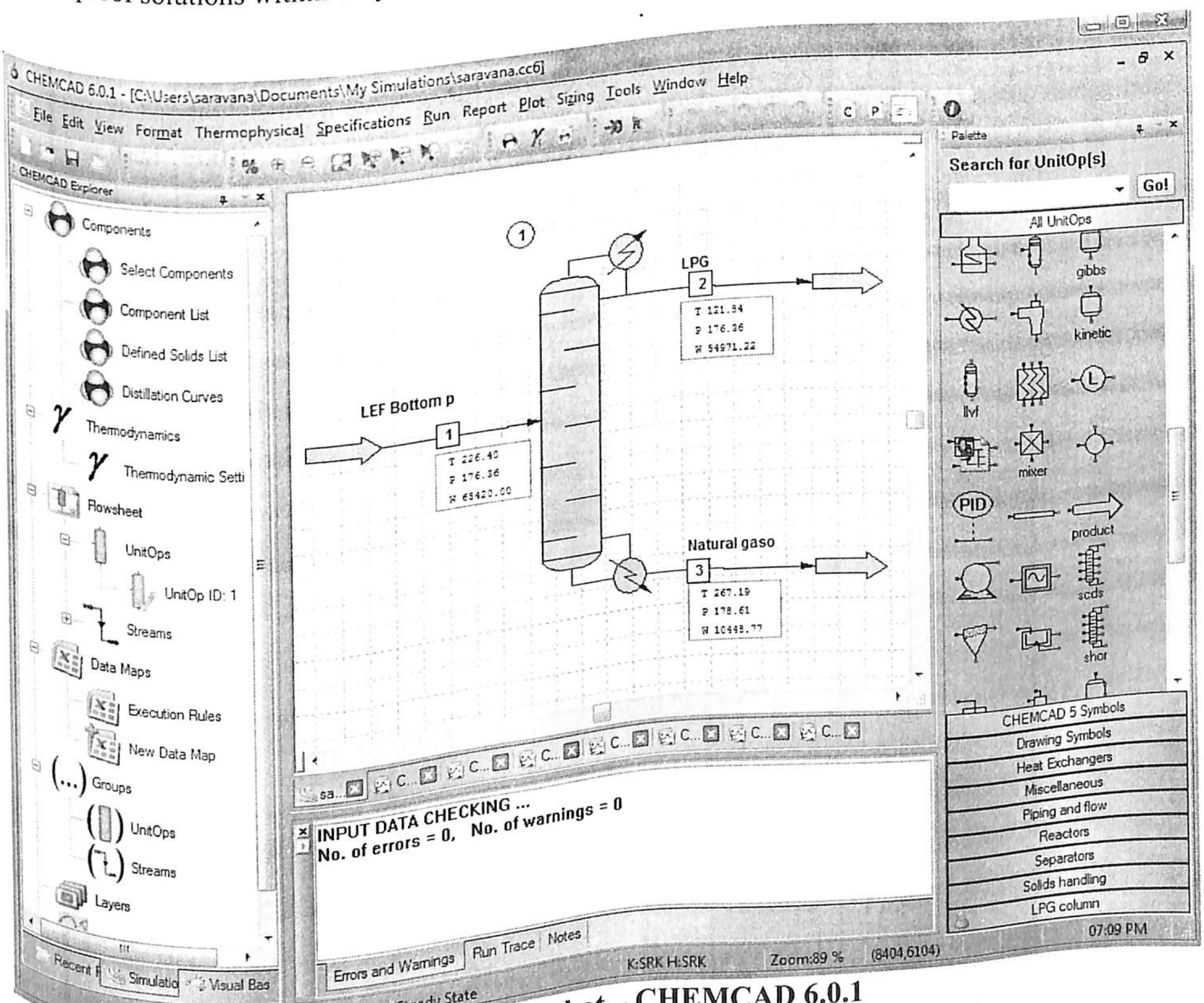


Figure 5: Screen shot – CHEMCAD 6.0.1

4.2 Methodology for Simulation of LPG Column

4.2.1 Creating a New Simulation

Start by creating a new simulation and giving it a name. To do this, launch CHEMCAD and then Select File > Save to open the Save As dialog box. Navigate to the directory where you want to store the simulation (try My Simulations, located under My Documents) and give your simulation a name, leaving the type as CHEMCAD 6 (*.cc6). Then click Save to create the file and return to the main CHEMCAD window.

4.2.2 Selecting Engineering Units

Select Format > Engineering Units to open the Engineering Unit Selection dialog box.The English units option is the default and is currently highlighted. To change the engineering units system, you would click the Alt SI, SI, or Metric button; you could then change any of the individual units as well. For this tutorial, you will use English units, so click Cancel to exit this dialog box without making changes.

4.2.3 Drawing the Flow sheet

As described in Chapter 5, creating a flowsheet is a matter of placing UnitOp icons on the screen, connecting them with streams, and then adding various graphical objects to enhance the drawing. Placing UnitOps. Click the Distillation column tool (see Figure 6). Drag from the Palette to the workspace

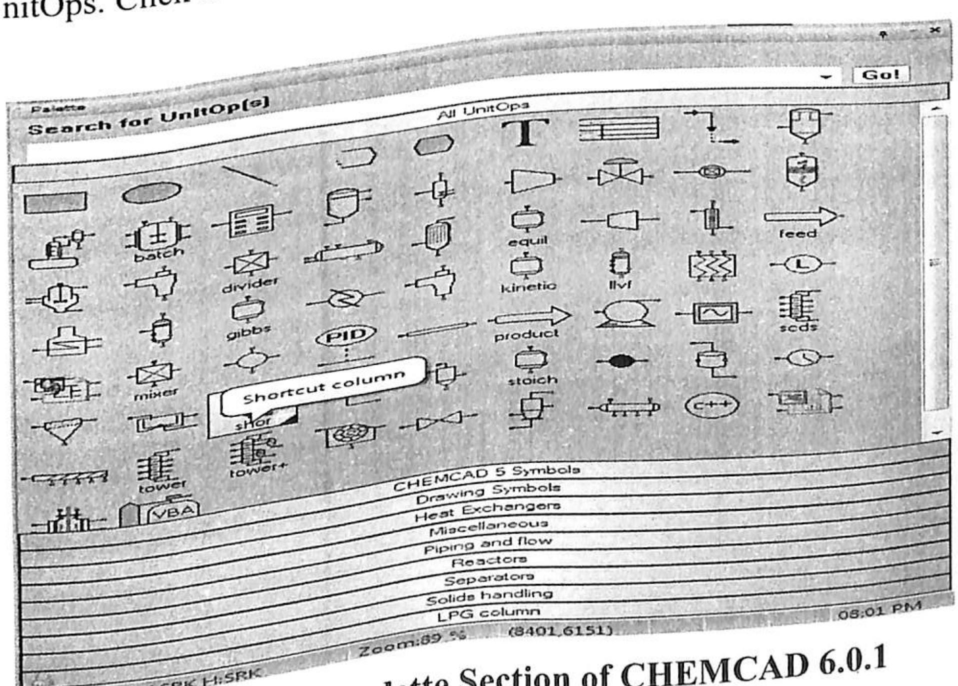


Figure 6: Screen Short view of Palette Section of CHEMCAD 6.0.1

4.2.4 Drawing Streams

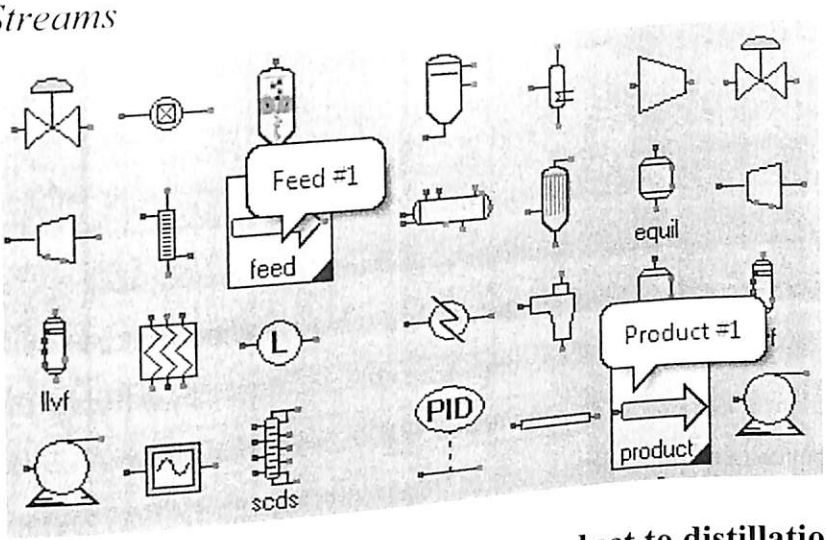


Figure 7: Screen short view of feed and product to distillation column

Click the feed tool (see Figure 7). Drag from the Palette to the workspace place it to feed side of distillation column Then click the product tool (see Figure 7), drag from the palette to the workspace place it to product side.

4.2.5 Selecting Components

Now need to identify the components to be used in this simulation. Start by selecting Thermophysical > Select Components.For this example, choose components from the standard CHEMCAD database. In the Select Components dialog box, find and add each needed component.

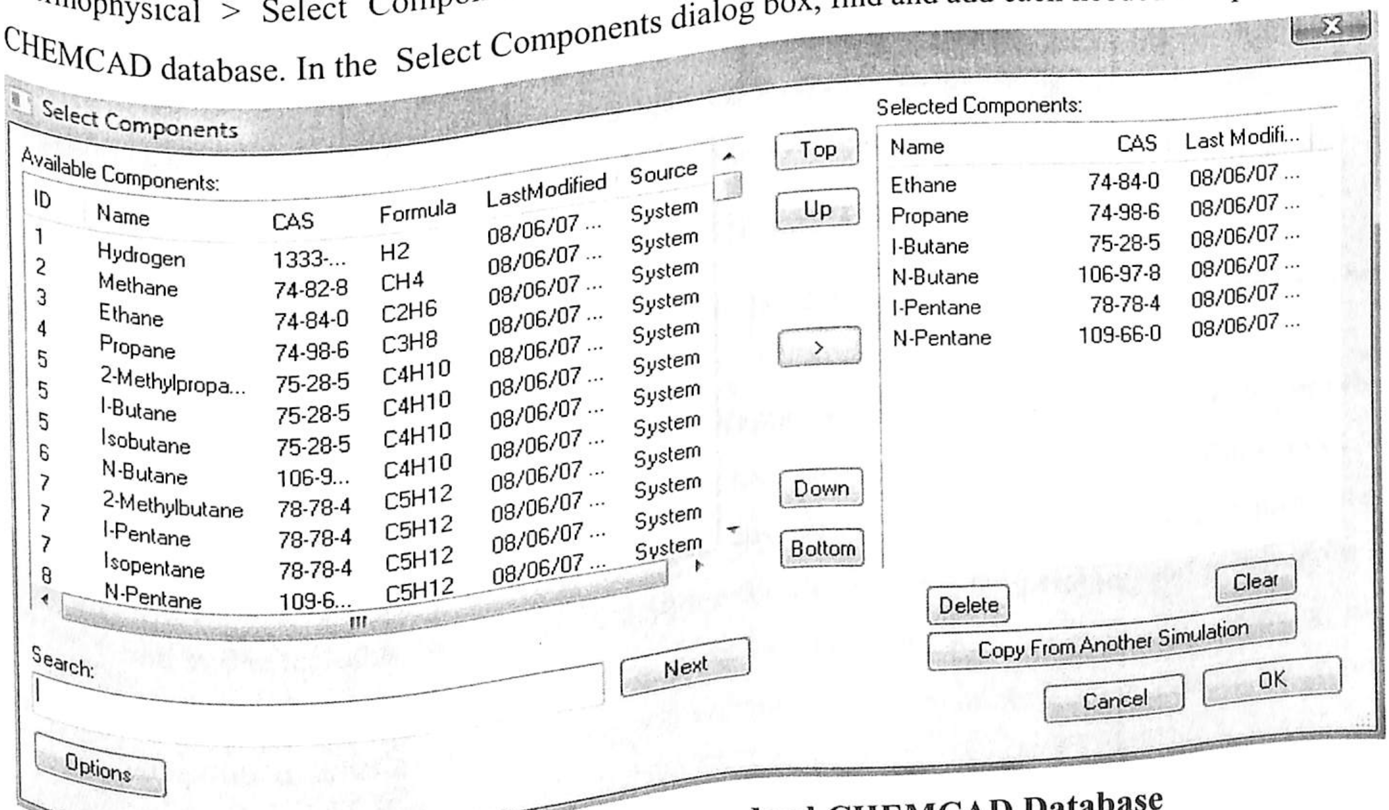


Figure 8: Select Components Standard CHEMCAD Database

4.2.6 *Defining the Feed Streams*

Efficient way to define a single stream is to double - click the stream line. Double - click the line for stream 1, to bring up the Edit Streams dialog box.

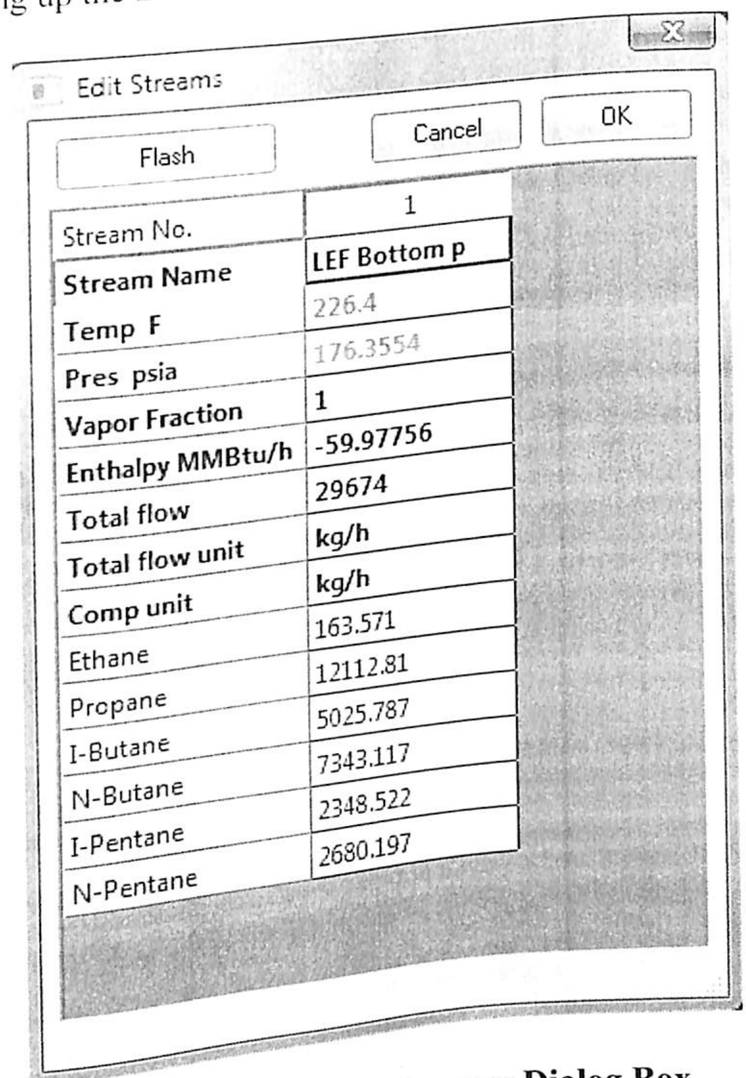


Figure 9: The Edit Streams Dialog Box

1. The Stream Name field can display a stream label of up to 16 alphanumeric characters. This field is optional.
2. The next four fields - Temp F, Pres psia, Vapor Fraction, and Enthalpy MMBtu/h—are the thermodynamic properties of the stream. According to the Gibbs Phase Rule, once a mixture's composition is given, specifying any two of these four thermodynamic properties will define the other two. As such, defining the composition, temperature, and pressure for a mixture uniquely defines its vapor fraction and enthalpy. Alternatively, defining the composition, pressure, and enthalpy will uniquely define the mixture's temperature and vapor fraction.

3. Since enthalpies are calculated relative to a datum, the calculation of any given stream enthalpy is an involved process which is prone to errors. For this reason, CHEMCAD does not permit to enter stream enthalpy as a constraint.
4. In addition to defining the stream's composition, define exactly two of the following properties: temperature, pressure, and vapor fraction. The two variables that specify will display as red text, while the third variable and the value enthalpy will be displayed in black when flash the stream.
5. An exception to this convention allows to add heat duty with an empty stream. If specify a total component flow rate of zero, specify a temperature, pressure, and enthalpy rate. A stream defined this way is treated as a heat duty, and is added to the heat balance of the unit. The temperature and pressure are arbitrary for this situation.
6. The Total flow unit and Comp unit fields work together to provide a variety of ways to define stream compositions. If the selected comp unit is mole, mass, or volume fraction (either globally or locally), then the Total flow unit selection is available. If the selected comp unit is a flow or amount option, then the total flow rate becomes the sum of the component flow rates, and the Total flow unit selection is not available.

4.2.7 Selecting Thermodynamic Options

The Thermodynamics Wizard appears. This tool can suggest thermodynamics options to use with this simulation. CHEMCAD's Thermodynamics Wizard works like as follows.

- *First,* it looks at the component list and decides what general type of model is required,i.e.,equation-of-state, activity model, etc.
- *Second,* it looks at temperature and pressure ranges that you provide and decides which equation within a given category is best at the limits of those ranges.
- If the method is an activity model, the program then looks at the BIP database to see which model has the most data sets for the current problem. It then calculates the fractional completeness of the BIP matrix. If that fraction is greater than the BIP threshold parameter, it uses the chosen activity method; if not, it uses UNIFAC.

The Thermodynamics Wizard is no replacement for engineering judgment. This tool uses an algorithm based on general rules, and is therefore fallible. The suggested model might not always be the best model for the system.

Selecting thermodynamic options basically means selecting a model or method for calculating vapor-liquid (or vapor-liquid-liquid) phase equilibrium (called the *K- value option*) and selecting a method or model for calculating the heat balance (called the *enthalpy option*). The commands for these selections are located on the Thermophysical menu.

CHEMCAD has a library of dozens of K-value models with a variety of options and about 12 enthalpy models. Making the proper selection from these libraries can sometimes be difficult. For the purposes of this tutorial, assume that you want to use the Peng-Robinson method for both the K-value and enthalpy calculations. Follow these steps to select your thermophysical options:

1. Accept the default temperature and pressure ranges in the Thermodynamics Wizard and click **OK**.
2. Click **OK** again to accept the wizard's suggested method of SRK.
3. When the Thermodynamic Settings dialog box opens, find the Global KValue Option selection, in the upper left corner of the K - Value Models tab. The current setting is **SRK**, but for the purposes of the tutorial, you'll need to select the Peng - Robinson model. Click the down arrow at the right end of the selection box to view a long list of K - value choices, then click **Peng - Robinson.**

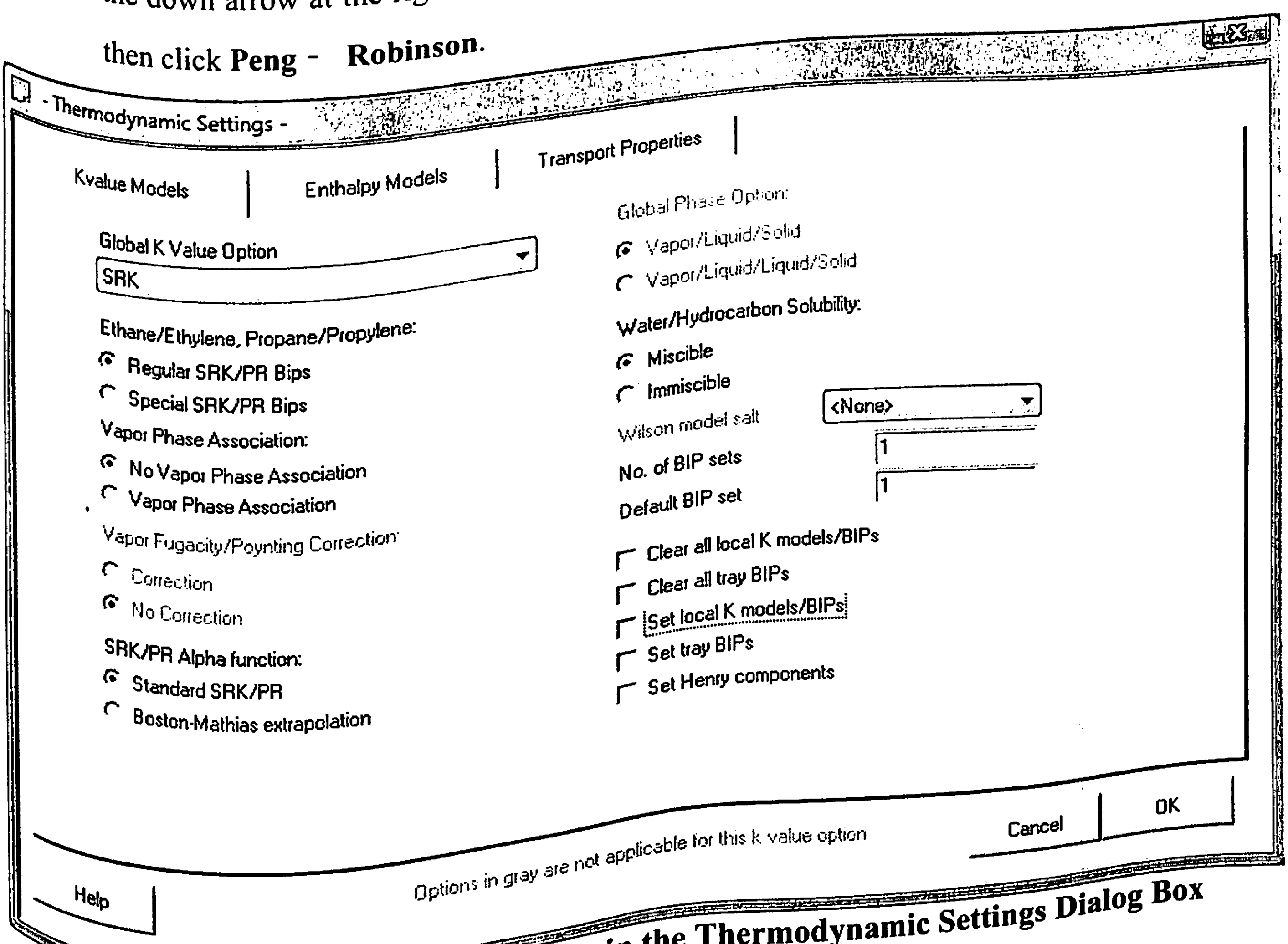


Figure 10: The New K-value Selection in the Thermodynamic Settings Dialog Box

4. Now click the **Enthalpy Models** tab. The Peng - Robinson method has already been entered as the Global Enthalpy Options selection; this was done automatically because you chose Peng - Robinson as your K - value method. While you do have the option to override this choice, in this case you'll need to keep the Peng - Robinson model; leave all settings as they are and click **OK** to return to the main CHEMCAD workspace.

4.2.8 Sizing

After running your simulation, click the UnitOp representing the distillation column and select Sizing > Distillation; choose either Trays or Packing, based on the type of Column. As shone below figure 12.

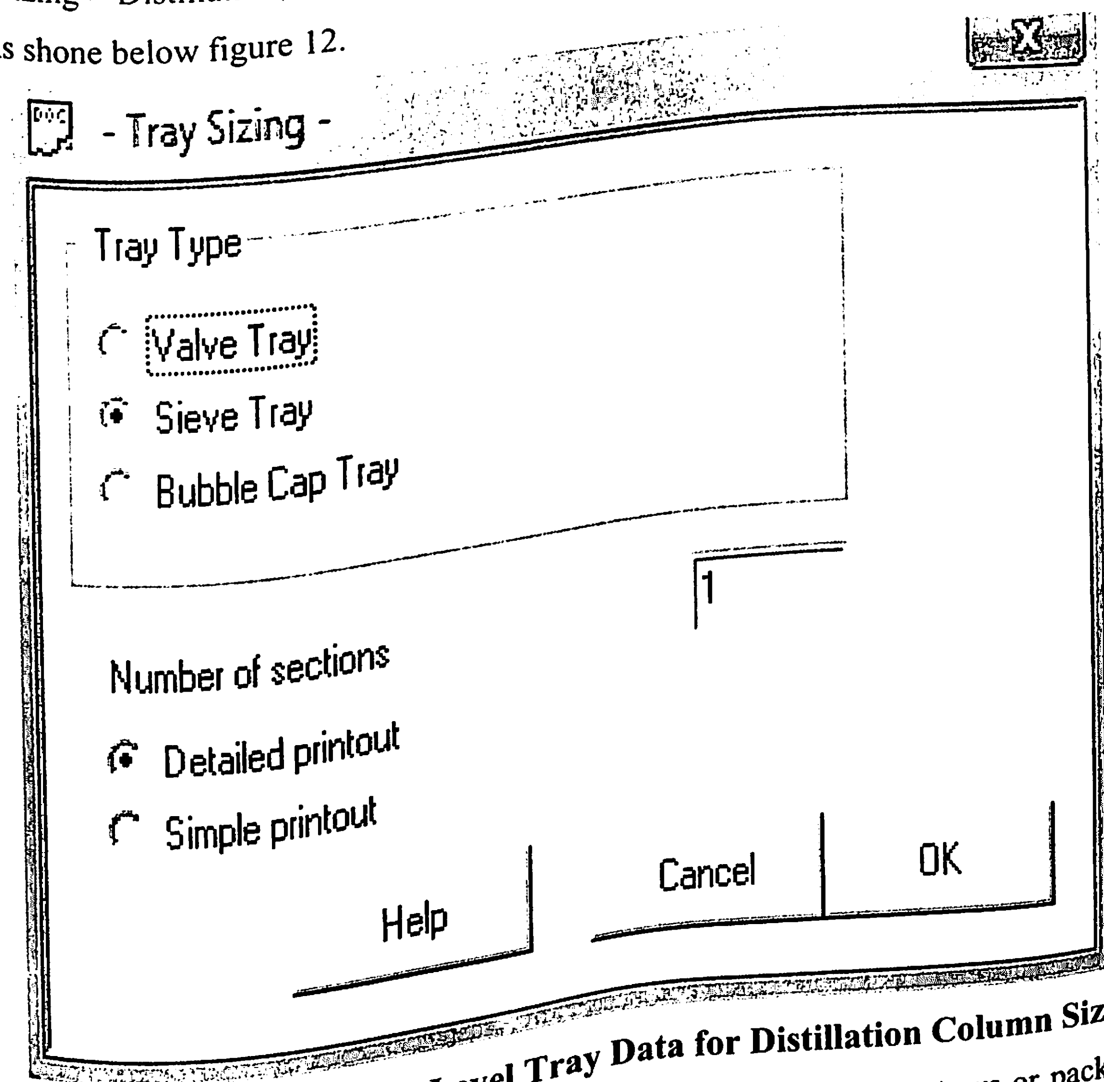


Figure 11: Specifying High-Level Tray Data for Distillation Column Sizing

The resulting dialog boxes prompt you to enter information about trays or packing and the calculation methods. To sizing column, generally need to enter some geometry and hydraulic parameters to complete column sizing. As shown below figure 13.

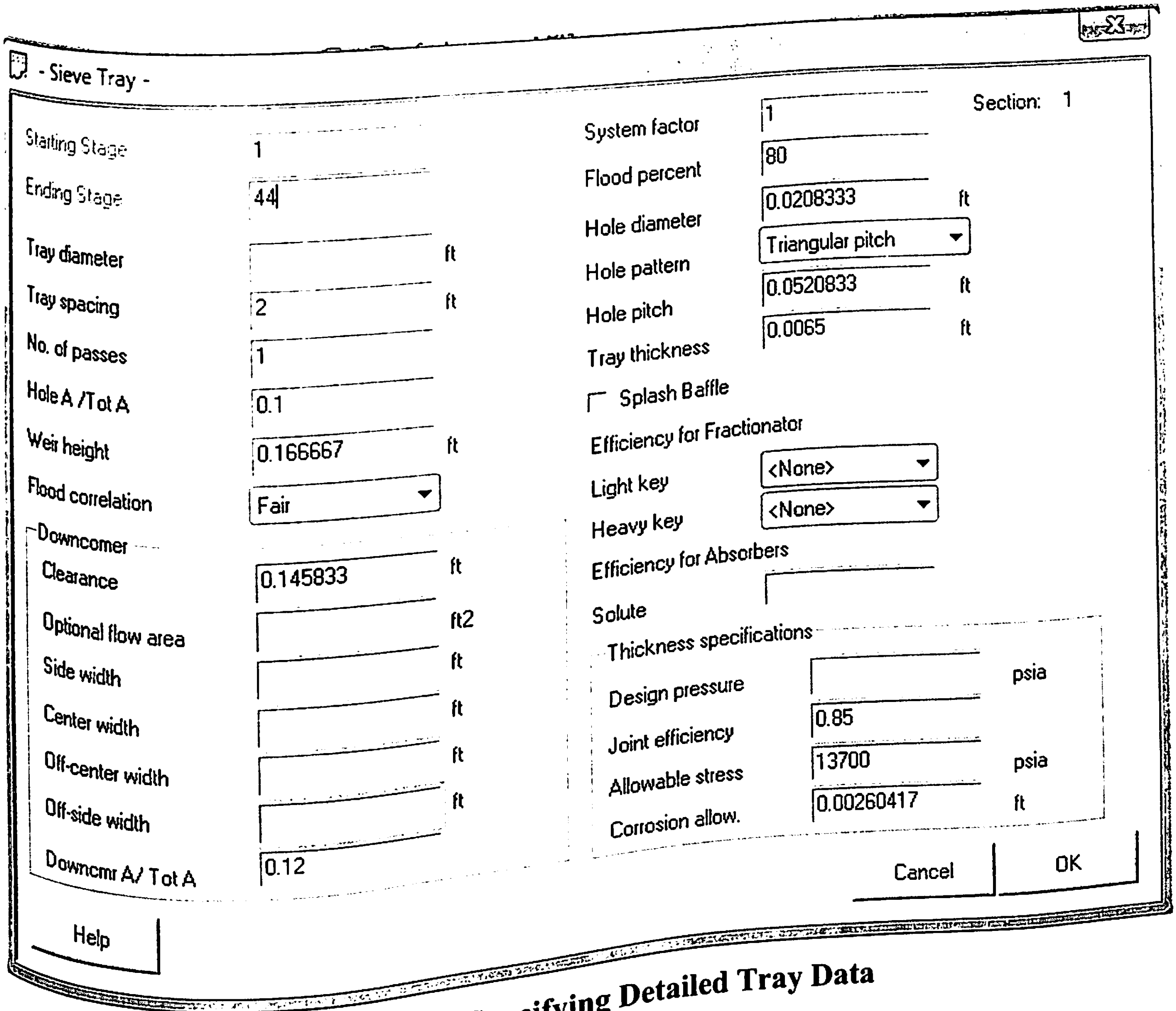


Figure 12: Specifying Detailed Tray Data

Based on flow sheet values and sizing input, CHEMCAD returns column geometry information such as height and diameter. It also provides hydraulic performance information such as predicted amount of flooding.

4.2.9 Run the Simulation

To run the simulation, click the Run All button on the toolbar. The program first rechecks the data and lists any errors and warnings in the Messages pane. In this case, you should have no errors, although you will have warnings about estimates you have not given. You can ignore these warnings and proceed by clicking Yes. The calculation will then proceed.

When the run finishes, a message box appears: Recycle calculation has converged. To close this dialog box and clear the screen, click OK.

5. SIMULATION REPORTS FOR LPG COLUMN

5.1 Stream Composition

saravana0 - WordPad

File Edit View Insert Format Help

Courier New 10

CHEMCAD 6.0.1
Page 1

Job Name: LPG Column   Date: 04/30/2009   Time: 10:12:12

Stream No.	1	2	3
Stream Name	LEF Bottom p	LPG	Natural gaso
Temp F	226.4000*	121.8370	267.2344
Pres psia	176.3554*	176.3554	178.6497
Enth MMBtu/h	-59.978	-61.321	-9.6754
Vapor mole fraction	1.0000	0.00000	0.00000
Total lbmol/h	1240.3984	1095.5776	144.8203
Total lb/h	65420.0000	54971.1758	10448.7793
Total std L ft3/hr	1900.9370	1634.0311	266.9051
Total std V scfh	470704.69	415748.28	54956.19
Flowrates in lb/h			
Ethane	360.6123	360.6125	0.0000
Propane	26704.1738	26704.1680	0.0000
I-Butane	11079.9639	11079.9619	0.0000
N-Butane	16188.8027	16188.7773	0.0089
I-Pentane	5177.6050	474.8668	4702.7437
N-Pentane	5908.8232	162.7892	5746.0269

For Help, press F1

5.2 Feed Stream Property

saravana5 - WordPad

File Edit View Insert Format Help

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Page 1

Job Name: feed stream property Date:
04/27/2009 Time: 21:15:22

Stream No.	1
Name	LEF Bottom P
- - Overall - -	
Molar flow lbmol/h	1240.3984
Mass flow lb/h	65420.0000
Temp F	226.4000
Pres psia	176.3554
Vapor mole fraction	1.000
Enth MMBtu/h	-59.978
Tc F	269.3230
Pc psia	627.7147
Std. sp gr. wtr = 1	0.551
Std. sp gr. air = 1	1.821
Degree API	125.1802
Average mol wt	52.7411
Actual dens lb/ft3	1.4645
Actual vol ft3/hr	44671.3594
Std liq ft3/hr	1900.9370
Std vap 60F scfh	470704.6875
- - Vapor only - -	
Molar flow lbmol/h	1240.3984
Mass flow lb/h	65420.0000
Average mol wt	52.7411
Actual dens lb/ft3	1.4645
Actual vol ft3/hr	44671.3594
Std liq ft3/hr	1900.9370
Std vap 60F scfh	470704.6875
Cp Btu/lbmol-F	27.8714
Z factor	0.8628
Visc cP	0.01047
Th cond Btu/hr-ft-F	0.0159

For Help, press F1

5.3 Product Stream Property

saravana6 - WordPad

File Edit View Insert Format Help

CHEMCAD 6.0.1
Page 1

Job Name: saravana Date: 04/27/2009 Time: 21:24:38

Stream No.	2	3
Name	LPG	Natural gaso
- - Overall - -		
Molar flow lbmol/h	1095.5781	144.8200
Mass flow lb/h	54971.2031	10448.7607
Temp F	121.8376	267.2919
Pres psia	176.3554	178.7765
Vapor mole fraction	0.0000	0.0000
Enth MMBtu/h	-61.321	-9.6751
Tc F	249.5892	378.2989
Pc psia	618.6827	489.8495
Std. sp gr. wtr = 1	0.539	0.627
Std. sp gr. air = 1	1.732	2.491
Degree API	131.0788	94.1471
Average mol wt	50.1755	72.1500
Actual dens lb/ft3	30.2482	30.3757
Actual vol ft3/hr	1817.3358	343.9847
Std liq ft3/hr	1634.0298	266.9068
Std vap 60F scfh	415748.4688	54956.0898
- - Liquid only - -		
Molar flow lbmol/h	1095.5781	144.8200
Mass flow lb/h	54971.2031	10448.7607
Average mol wt	50.1755	72.1500
Actual dens lb/ft3	30.2482	30.3757
Actual vol ft3/hr	1817.3358	343.9847
Std liq ft3/hr	1634.0298	266.9068
Std vap 60F scfh	415748.4688	54956.0898
Cp Btu/lbmol-F	35.7113	56.0904
Z factor	0.0509	0.0622
Visc cP	0.09573	0.1092
Th cond Btu/hr-ft-F	0.0488	0.0444
Surf tens dyne/cm	5.5020	4.4801

For Help, press F1

5.4 Topology Report

CHEMCAD 6.0.1

Job Name: LPG Column Date: 04/28/2009 Time: 06:20:59

FLOWSHEET SUMMARY

Equipment	Label	Stream Numbers
1	TOWR	1 -2 -3

Stream Connections

Stream	Equipment From	Equipment To
1		1
2	1	
3	1	

5.5 Mass and Energy Balance

CHEMCAD 6.0.1

Job Name: saravana Date: 04/28/2009 Time: 06:32:51

Calculation mode : Sequential
Flash algorithm : Normal

Equipment Calculation Sequence
1

No recycle loops in the flowsheet.

CHEMCAD 6.0.1

Job Name: saravana Date: 04/28/2009 Time: 06:32:51

Overall Mass Balance

	lbmol/h Input	lbmol/h Output	lb/h Input	lb/h Output
Ethane	11.992	11.992	360.612	360.612
Propane	605.592	605.592	26704.174	26704.179
I-Butane	190.630	190.630	11079.964	11079.966
N-Butane	278.527	278.527	16188.803	16188.813
I-Pentane	71.762	71.761	5177.605	5177.582
N-Pentane	81.896	81.896	5908.823	5908.816
Total	1240.398	1240.398	65420.000	65420.000

5.6 Thermodynamics Report

CHEMCAD 6.0.1

Job Name: LPG Column   Date: 04/28/2009   Time: 06:27:49

COMPONENTS

	ID #	Name	Formula
1	3	Ethane	C2H6
2	4	Propane	C3H8
3	5	I-Butane	C4H10
4	6	N-Butane	C4H10
5	7	I-Pentane	C5H12
6	8	N-Pentane	C5H12

THERMODYNAMICS

K-value model : SRK
Enthalpy model : SRK
Liquid density : Library

Std vapor rate reference temperature is 60 F.
Atmospheric pressure is 14.6959 psi.

5.7 Net Flow Report

Page 1

CHEMCAD 6.0.1

Job Name: **LPG Column** Date: 04/27/2009 Time: 22:29:37

Unit type : TOWR Unit name: Eqp # 1

Stg	Temp F	Pres psia	* Net Flows * Liquid lb/h	Vapor lb/h	Feeds lb/h	Product lb/h	Duties MMBtu/h
1	121.8	176.36	79650.22			54971.28	-19.01
2	142.8	176.41	82165.73	134621.36			
3	154.8	176.46	83566.66	137137.00			
4	161.9	176.51	83843.00	138537.81			
5	167.0	176.56	83478.83	138814.08			
6	171.6	176.62	82868.21	138450.00			
7	176.2	176.67	67658.55	137839.34	65420.00		
8	189.4	176.72	70358.48	57209.70			
9	198.6	176.77	72185.00	59909.64			
10	205.4	176.82	73357.06	61736.09			
11	211.3	176.88	74207.29	62908.22			
12	217.3	176.93	75013.21	63758.44			
13	223.6	176.98	75954.45	64564.36			
14	230.2	177.04	77103.13	65505.59			
15	236.7	177.09	78420.73	66654.29			
16	242.7	177.14	79805.21	67971.89			
17	247.9	177.20	81127.31	69356.38			
18	252.1	177.25	82290.44	70678.48			
19	255.3	177.30	83247.00	71841.59			
20	257.7	177.35	83993.29	72798.16			
21	259.5	177.40	84553.60	73544.45			
22	260.8	177.46	84963.10	74104.77			
23	261.7	177.51	85257.00	74514.27			
24	262.3	177.56	85462.12	74808.13			
25	262.8	177.61	85605.76	75013.29			
26	263.1	177.67	85706.00	75156.92			
27	263.3	177.72	85778.03	75257.15			
28	263.5	177.77	85828.55	75329.21			
29	263.7	177.83	85862.41	75379.72			
30	263.8	177.88	85875.47	75413.58			
31	264.0	177.93	85880.85	75426.64			
32	264.1	177.98	85879.31	75432.02			
33	264.2	178.04	85872.14	75430.48			
34	264.4	178.09	85859.55	75423.32			
35	264.5	178.14	85848.24	75410.73			
36	264.7	178.19	85831.90	75399.42			
37	264.9	178.25	85810.28	75383.08			
38	265.1	178.30	85781.81	75361.45			
39	265.4	178.35	85748.26	75333.00			
40	265.6	178.41	85708.31	75299.43			
41	266.0	178.46	85661.77	75259.49			
42	266.3	178.51	85607.39	75213.00			
43	266.7	178.56	85546.59	75158.57			
44	267.2	178.62		75097.77		10448.82	7.991

Mass Reflux ratio 1.449

Total liquid entering stage 7 at 171.555 F, 82868.211 lb/h.

CHEMCAD 6.0.1
Page 1

Job Name: **LPG Column**   Date: 04/27/2009   Time: 21:51:04

Unit type : TOWR   Unit name: Eqp # 1

Stage # 1	121.84 F	176.36 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	522.50787	0.00000
Propane	0.00000	38692.90234	0.00000
I-Butane	0.00000	16054.26855	0.00000
N-Butane	0.00000	23456.65430	0.00000
I-Pentane	0.00000	660.30176	0.00000
N-Pentane	0.00000	263.57697	0.00000
Total lb/h	0.0000	79650.2188	

Stage # 2	142.75 F	176.41 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	883.12030	134.31738	3.77018
Propane	65397.07813	24196.55078	1.54981
I-Butane	27134.23242	19231.89453	0.80904
N-Butane	39645.42969	35692.10938	0.63694
I-Pentane	1116.01355	1965.92102	0.32552
N-Pentane	445.48642	944.93927	0.27034
Total lb/h	134621.3594	82165.7344	

Stage # 3	154.84 F	176.46 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	494.92981	70.98138	4.00734
Propane	50900.72656	17205.81641	1.70022
I-Butane	30311.86133	19200.07617	0.90733
N-Butane	51880.87891	41348.32422	0.72112
I-Pentane	2421.63306	3689.88452	0.37718
N-Pentane	1126.84875	2051.58130	0.31567
Total lb/h	137137.0000	83566.6641	

Stage # 4	161.90 F	176.51 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	431.59375	59.32879	4.14907
Propane	43910.00000	13983.90625	1.79092
I-Butane	30280.04102	17852.83594	0.96736
N-Butane	57537.09375	42467.87891	0.77273
I-Pentane	4145.59668	5776.14014	0.40935
N-Pentane	2233.49097	3702.85010	0.34402
Total lb/h	138537.8125	83843.0000	

Stage # 5	166.97 F	176.56 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	419.94116	55.65730	4.25607
Propane	40688.07422	12353.48535	1.85789
I-Butane	28932.79688	16133.34863	1.01160
N-Butane	58656.63672	40814.53516	0.81067
I-Pentane	6231.85254	8117.31641	0.43306
N-Pentane	3884.75977	6004.47852	0.36495
Total lb/h	138814.0781	83478.8281	

Stage # 6	171.56 F	176.62 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	416.26962	53.28003	4.35621

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Job Name: **LPG Column** Date: 04/27/2009 Time: 21:51:04

Propane	39057.66016	11342.87793	1.91991
I-Butane	27213.31055	14415.66211	1.05256
N-Butane	57003.30078	37578.71094	0.84578
I-Pentane	8573.02734	10504.08691	0.45507
N-Pentane	6186.38770	8973.59375	0.38439
Total lb/h	138450.0000	82868.2109	

Stage # 7	176.17 F	176.67 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	413.89233	42.13249	4.45814
Propane	38047.05078	8705.40039	1.98343
I-Butane	25495.62500	10569.36035	1.09471
N-Butane	53767.47656	27665.23828	0.88200
I-Pentane	10959.79883	10407.34375	0.47791
N-Pentane	9155.50293	10269.07227	0.40461
Total lb/h	137839.3438	67658.5469	

Stage # 8	189.43 F	176.72 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	42.13249	10.44367	4.68214
Propane	8705.40039	4688.07568	2.15514
I-Butane	10569.36035	10073.24609	1.21776
N-Butane	27665.19531	32386.89648	0.99139
I-Pentane	5685.45020	11983.77148	0.55062
N-Pentane	4542.15674	11216.05371	0.47001
Total lb/h	57209.6953	70358.4844	

Stage # 9	198.59 F	176.77 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	10.44367	2.49081	4.82948
Propane	4688.07520	2374.89941	2.27372
I-Butane	10073.24414	8890.53320	1.30506
N-Butane	32386.85742	34871.79688	1.06975
I-Pentane	7261.87842	13848.28516	0.60401
N-Pentane	5489.13818	12196.93164	0.51837
Total lb/h	59909.6367	72185.0000	

Stage # 10	205.40 F	176.82 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	2.49081	0.57677	4.93798
Propane	2374.89966	1149.52332	2.36233
I-Butane	8890.53223	7414.10498	1.37114
N-Butane	34871.75781	35307.54688	1.12933
I-Pentane	9126.39453	16175.00488	0.64516
N-Pentane	6470.01709	13310.30566	0.55582
Total lb/h	61736.0898	73357.0625	

Stage # 11	211.33 F	176.88 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.57677	0.13046	5.03240
Propane	1149.52332	536.30585	2.43988
I-Butane	7414.10498	5904.14648	1.42944
N-Butane	35307.50781	34001.51172	1.18204
I-Pentane	11453.11328	19118.39063	0.68192
N-Pentane	7583.38965	14646.80371	0.58937
Total lb/h	62908.2188	74207.2891	

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Job Name: **LPG Column** Date: 04/27/2009 Time: 21:51:04

Stage # 12	217.27 F	176.93 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.13046	0.02893	5.12536
Propane	536.30591	242.12886	2.51757
I-Butane	5904.14648	4508.57324	1.48845
N-Butane	34001.47656	31276.75977	1.23564
I-Pentane	14396.49707	22736.66602	0.71969
N-Pentane	8919.88770	16249.05664	0.62395
Total lb/h	63758.4414	75013.2109	

Stage # 13	223.57 F	176.98 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.02893	0.00631	5.22074
Propane	242.12888	105.97120	2.59961
I-Butane	4508.57275	3306.00000	1.55162
N-Butane	31276.72070	27513.86914	1.29336
I-Pentane	18014.77539	26938.60156	0.76086
N-Pentane	10522.13867	18090.00000	0.66178
Total lb/h	64564.3633	75954.4453	

Stage # 14	230.17 F	177.04 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00631	0.00135	5.31363
Propane	105.97121	45.04261	2.68383
I-Butane	3306.00000	2331.09106	1.61783
N-Butane	27513.83398	23174.35156	1.35436
I-Pentane	22216.70703	31480.18164	0.80507
N-Pentane	12363.07813	20072.46875	0.70261
Total lb/h	65505.5938	77103.1328	

Stage # 15	236.69 F	177.09 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00135	0.00029	5.40477
Propane	45.04261	18.62324	2.76719
I-Butane	2331.09082	1583.81543	1.68393
N-Butane	23174.31445	18731.77148	1.41546
I-Pentane	26758.29102	36026.55859	0.84978
N-Pentane	14345.55176	22059.96680	0.74402
Total lb/h	66654.2891	78420.7344	

Stage # 16	242.70 F	177.14 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00029	0.00006	5.48478
Propane	18.62323	7.51175	2.84302
I-Butane	1583.81543	1040.81995	1.74500
N-Butane	18731.73438	14590.12598	1.47226
I-Pentane	31304.66992	40250.59375	0.89187
N-Pentane	16333.04980	23916.16016	0.78315
Total lb/h	67971.8906	79805.2109	

Stage # 17	247.88 F	177.20 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00006	0.00001	5.55148
Propane	7.51175	2.96535	2.90770
I-Butane	1040.81995	664.58356	1.79767

N-Butane	14590.08789	11007.28418	1.52147
I-Pentane	35528.71094	43912.32031	0.92870
N-Pentane	18189.24609	25540.16211	0.81748
Total lb/h	69356.3750	81127.3125	

Stage # 18	252.07 F	177.25 psia	Y/X
	Vap lb/h	Liq lb/h	
Ethane	0.00001	0.00000	5.60428
Propane	2.96535	1.14975	2.95974
I-Butane	664.58356	414.39511	1.84041
N-Butane	11007.24512	8089.31299	1.56152
I-Pentane	39190.43750	46900.29297	0.95893
N-Pentane	19813.24805	26885.28320	0.84571
Total lb/h	70678.4766	82290.4375	

Stage # 19	255.32 F	177.30 psia	Y/X
	Vap lb/h	Liq lb/h	
Ethane	0.00000	0.00000	5.64442
Propane	1.14975	0.43941	2.99971
I-Butane	414.39511	253.57959	1.87345
N-Butane	8089.27441	5823.16406	1.59255
I-Pentane	42178.41016	49216.55469	0.98247
N-Pentane	21158.37109	27953.26563	0.86774
Total lb/h	71841.5938	83247.0000	

Stage # 20	257.73 F	177.35 psia	Y/X
	Vap lb/h	Liq lb/h	
Ethane	0.00000	0.00000	5.67410
Propane	0.43941	0.16607	3.02938
I-Butane	253.57953	152.95741	1.89807
N-Butane	5823.12500	4126.33545	1.61570
I-Pentane	44494.66797	50936.29688	1.00011
N-Pentane	22226.35352	28777.53516	0.88427
Total lb/h	72798.1641	83993.2891	

Stage # 21	259.50 F	177.40 psia	Y/X
	Vap lb/h	Liq lb/h	
Ethane	0.00000	0.00000	5.69561
Propane	0.16607	0.06224	3.05089
I-Butane	152.95740	91.28358	1.91595
N-Butane	4126.29639	2890.07373	1.63252
I-Pentane	46214.41406	52166.44922	1.01296
N-Pentane	23050.61914	29405.73438	0.89631
Total lb/h	73544.4531	84553.6016	

Stage # 22	260.76 F	177.46 psia	Y/X
	Vap lb/h	Liq lb/h	
Ethane	0.00000	0.00000	5.71095
Propane	0.06224	0.02318	3.06620
I-Butane	91.28356	54.05696	1.92870
N-Butane	2890.03467	2007.19495	1.64451
I-Pentane	47444.57031	53015.18750	1.02214
N-Pentane	23678.82031	29886.64258	0.90491
Total lb/h	74104.7734	84963.1016	

Stage # 23	261.65 F	177.51 psia	

CHEMCAD 6.0.1

Job Name: **LPG Column** Date: 04/27/2009 Time: 21:51:04

	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	5.72167
Propane	0.02318	0.00860	3.07697
I-Butane	54.05692	31.83530	1.93768
N-Butane	2007.15527	1385.66040	1.65297
I-Pentane	48293.30859	53576.10938	1.02862
N-Pentane	24159.72852	30263.35352	0.91099
Total lb/h	74514.2734	85257.0000	

Stage # 24	262.29 F	177.56 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	5.72913
Propane	0.00860	0.00318	3.08449
I-Butane	31.83526	18.67446	1.94399
N-Butane	1385.62085	952.48297	1.65890
I-Pentane	48854.23047	53921.01172	1.03318
N-Pentane	24536.43945	30569.94727	0.91527
Total lb/h	74808.1328	85462.1172	

Stage # 25	262.75 F	177.61 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00318	0.00117	3.08991
I-Butane	18.67442	10.92378	1.94848
N-Butane	952.44330	652.74176	1.66310
I-Pentane	49199.13281	54106.72656	1.03640
N-Pentane	24843.03320	30835.35938	0.91828
Total lb/h	75013.2891	85605.7578	

Stage # 26	263.09 F	177.67 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00117	0.00043	3.09352
I-Butane	10.92375	6.37756	1.95159
N-Butane	652.70221	446.37421	1.66605
I-Pentane	49384.85547	54172.27734	1.03869
N-Pentane	25108.44141	31080.93750	0.92044
Total lb/h	75156.9219	85706.0000	

Stage # 27	263.34 F	177.72 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00043	0.00016	3.09544
I-Butane	6.37752	3.71877	1.95357
N-Butane	446.33463	304.81308	1.66802
I-Pentane	49450.41016	54145.66016	1.04035
N-Pentane	25354.02344	31323.83984	0.92203
Total lb/h	75257.1484	85778.0313	

Stage # 28	263.53 F	177.77 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00016	0.00006	3.09720
I-Butane	3.71873	2.16622	1.95518
N-Butane	304.77350	207.90465	1.66959
I-Pentane	49423.78906	54041.21875	1.04161

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N-Pentane	25596.92773	31577.25977	0.92323
Total lb/h	75329.2109	85828.5469	

Stage # 29	263.70 F	177.83 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00006	0.00002	3.09897
I-Butane	2.16619	1.26072	1.95663
N-Butane	207.86508	141.66353	1.67092
I-Pentane	49319.33984	53866.78125	1.04262
N-Pentane	25850.34961	31852.70117	0.92417
Total lb/h	75379.7188	85862.4063	

Stage # 30	263.84 F	177.88 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00002	0.00001	3.10231
I-Butane	1.26068	0.73282	1.95861
N-Butane	141.62398	96.40909	1.67247
I-Pentane	49144.90234	53619.09375	1.04351
N-Pentane	26125.79102	32159.22656	0.92492
Total lb/h	75413.5781	85875.4688	

Stage # 31	263.97 F	177.93 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00001	0.00000	3.10270
I-Butane	0.73278	0.42578	1.95935
N-Butane	96.36951	65.56772	1.67327
I-Pentane	48897.22266	53306.71094	1.04429
N-Pentane	26432.31445	32508.14844	0.92568
Total lb/h	75426.6406	85880.8516	

Stage # 32	264.10 F	177.98 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.10311
I-Butane	0.42574	0.24727	1.96007
N-Butane	65.52815	44.56070	1.67406
I-Pentane	48584.83203	52925.24609	1.04504
N-Pentane	26781.24023	32909.25781	0.92642
Total lb/h	75432.0234	85879.3125	

Stage # 33	264.24 F	178.04 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.10360
I-Butane	0.24723	0.14353	1.96085
N-Butane	44.52113	30.25929	1.67490
I-Pentane	48203.37109	52468.48047	1.04583
N-Pentane	27182.34766	33373.25781	0.92719
Total lb/h	75430.4844	85872.1406	

Stage # 34	264.38 F	178.09 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000

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Propane	0.00000	0.00000	3.10430
I-Butane	0.14349	0.08326	1.96177
N-Butane	30.21972	20.52693	1.67584
I-Pentane	47746.60547	51927.35938	1.04668
N-Pentane	27646.35352	33911.57813	0.92802
Total lb/h	75423.3203	85859.5547	

Stage # 35	264.54 F	178.14 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.10423
I-Butane	0.08323	0.04828	1.96243
N-Butane	20.48736	13.90978	1.67669
I-Pentane	47205.48438	51295.76563	1.04761
N-Pentane	28184.67383	34538.51172	0.92896
Total lb/h	75410.7266	85848.2422	

Stage # 36	264.71 F	178.19 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.10578
I-Butane	0.04824	0.02796	1.96385
N-Butane	13.87020	9.40925	1.67803
I-Pentane	46573.89453	50555.64844	1.04869
N-Pentane	28811.60938	35266.81250	0.92998
Total lb/h	75399.4219	85831.8984	

Stage # 37	264.90 F	178.25 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.10766
I-Butane	0.02793	0.01617	1.96551
N-Butane	9.36967	6.35008	1.67960
I-Pentane	45833.77734	49692.08594	1.04993
N-Pentane	29539.90625	36111.82422	0.93115
Total lb/h	75383.0781	85810.2813	

Stage # 38	265.12 F	178.30 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.11007
I-Butane	0.01614	0.00934	1.96752
N-Butane	6.31050	4.27190	1.68146
I-Pentane	44970.21094	48687.52344	1.05136
N-Pentane	30384.91406	37090.00781	0.93249
Total lb/h	75361.4453	85781.8125	

Stage # 39	265.37 F	178.35 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.11270
I-Butane	0.00930	0.00537	1.96977
N-Butane	4.23232	2.86146	1.68356
I-Pentane	43965.65234	47525.09766	1.05300
N-Pentane	31363.09961	38220.29688	0.93403
Total lb/h	75333.0000	85748.2578	

CHEMCAD 6.0.1

Job Name: **LPG Column** Date: [illegible] 27/2009 Time: 21:51:04

Stage # 40	265.65 F	178.41 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.11585
I-Butane	0.00534	0.00308	1.97241
N-Butane	2.82188	1.90505	1.68601
I-Pentane	42803.21875	46184.59375	1.05489
N-Pentane	32493.39063	39521.81641	0.93581
Total lb/h	75299.4297	85708.3125	

Stage # 41	265.97 F	178.46 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	3.11955
I-Butane	0.00305	0.00175	1.97548
N-Butane	1.86548	1.25725	1.68885
I-Pentane	41462.71875	44645.50000	1.05707
N-Pentane	33794.91016	41015.01172	0.93785
Total lb/h	75259.4922	85661.7734	

Stage # 42	266.32 F	178.51 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	0.00000
I-Butane	0.00172	0.00099	1.97905
N-Butane	1.21768	0.81906	1.69214
I-Pentane	39923.62500	42886.24609	1.05957
N-Pentane	35288.10547	42720.32422	0.94018
Total lb/h	75213.0000	85607.3906	

Stage # 43	266.73 F	178.56 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	0.00000
I-Butane	0.00095	0.00055	1.98312
N-Butane	0.77948	0.52316	1.69589
I-Pentane	38164.37500	40887.06250	1.06242
N-Pentane	36993.41797	44659.00000	0.94284
Total lb/h	75158.5703	85546.5859	

Stage # 44	267.19 F	178.62 psia	
	Vap lb/h	Liq lb/h	Y/X
Ethane	0.00000	0.00000	0.00000
Propane	0.00000	0.00000	0.00000
I-Butane	0.00051	0.00004	1.98779
N-Butane	0.48358	0.03957	1.70018
I-Pentane	36165.19531	4721.87012	1.06565
N-Pentane	38932.08984	5726.90625	0.94586
Total lb/h	75097.7734	10448.8164	

5.8 Tray Property Report

CHEMCAD 6.0.1

Job Name: LPG COLUMN Date: 04/27/2009 Time: 22:31:24

Unit type : TOWR Unit name: Eqp # 1

LIQUID

Stg	lb/h	Average mol wt	Actual vol rate ft3/hr	Actual density lb/ft3	viscosity cP	Thermal conduct. Btu/hr-ft-F	Surface tension dyne/cm
1	79650	50.18	2633.22	30.25	0.0957	0.049	5.502
2	82166	53.41	2702.20	30.41	0.0958	0.048	5.391
3	83567	55.20	2747.20	30.42	0.0957	0.047	5.247
4	83843	56.34	2754.63	30.44	0.0956	0.047	5.171
5	83479	57.27	2739.84	30.47	0.0958	0.047	5.126
6	82868	58.21	2716.83	30.50	0.0962	0.046	5.083
7	67659	59.18	2216.51	30.52	0.0967	0.046	5.030
8	70358	60.72	2313.03	30.42	0.0961	0.045	4.837
9	72185	61.81	2375.91	30.38	0.0958	0.045	4.751
10	73357	62.71	2418.22	30.34	0.0957	0.044	4.650
11	74207	63.60	2442.45	30.38	0.0960	0.044	4.660
12	75013	64.58	2460.88	30.48	0.0965	0.044	4.746
13	75954	65.66	2488.39	30.52	0.0974	0.044	4.753
14	77103	66.79	2525.33	30.53	0.0985	0.044	4.721
15	78421	67.89	2569.26	30.52	0.0999	0.044	4.674
16	79805	68.89	2616.02	30.51	0.1015	0.044	4.626
17	81127	69.73	2660.76	30.49	0.1030	0.044	4.586
18	82290	70.39	2700.00	30.48	0.1044	0.044	4.558
19	83247	70.90	2732.12	30.47	0.1056	0.044	4.539
20	83993	71.27	2757.07	30.46	0.1065	0.044	4.527
21	84554	71.54	2775.75	30.46	0.1072	0.045	4.520
22	84963	71.73	2789.39	30.46	0.1077	0.045	4.515
23	85257	71.86	2799.21	30.46	0.1081	0.045	4.512
24	85462	71.95	2806.11	30.46	0.1084	0.045	4.509
25	85606	72.02	2810.99	30.45	0.1086	0.045	4.507
26	85706	72.06	2814.45	30.45	0.1087	0.045	4.505
27	85778	72.09	2817.01	30.45	0.1088	0.045	4.504
28	85829	72.11	2818.86	30.45	0.1089	0.045	4.502
29	85862	72.12	2820.18	30.45	0.1089	0.045	4.501
30	85875	72.13	2820.83	30.44	0.1089	0.045	4.499
31	85881	72.14	2821.24	30.44	0.1090	0.045	4.498
32	85879	72.14	2821.43	30.44	0.1090	0.045	4.497
33	85872	72.14	2821.46	30.44	0.1090	0.045	4.495
34	85860	72.15	2821.33	30.43	0.1090	0.045	4.494
35	85848	72.15	2821.27	30.43	0.1090	0.045	4.493
36	85832	72.15	2821.07	30.43	0.1090	0.045	4.492
38	85782	72.15	2820.18	30.42	0.1091	0.045	4.489
39	85748	72.15	2819.51	30.41	0.1091	0.045	4.488
40	85708	72.15	2818.68	30.41	0.1091	0.045	4.487
41	85662	72.15	2817.67	30.40	0.1092	0.045	4.486
42	85607	72.15	2816.47	30.40	0.1092	0.045	4.486
43	85547	72.15	2815.10	30.39	0.1092	0.044	4.485
44	10449	72.15	343.93	30.38	0.1093	0.044	4.484

Stg	Liq H MMBtu/h
1	-88.85
2	-89.13
3	-89.213

CHEMCAD 6.0.1

Job Name: LPG COLUMN Date: 04/27/2009 Time: 22:31:24

4	-88.62[illegible]
5	-87.575
6	-86.3[illegible]
7	-69.9[illegible]8
8	-71.647
9	-72.703
10	-73.26
11	-73.556
12	-73.806
13	-74.163
14	-74.705
15	-75.42
16	-76.241
17	-77.068
18	-77.82
19	-78.453
20	-78.954
21	-79.332
22	-79.61
23	-79.809
24	-79.945
25	-80.039
26	-80.101
27	-80.144
28	-80.17
29	-80.182
30	-80.176
31	-80.162
32	-80.14
33	-80.111
34	-80.074
35	-80.035
36	-79.987
38	-79.861
39	-79.78
40	-79.687
41	-79.579
42	-79.455
43	-79.316
44	-9.6763

VAPOR

Stg	lb/h	Average mol wt	Actual vol rate ft3/hr	Actual density lb/ft3	viscosity cP	Thermal conduct. Btu/hr-ft-F	Compr. factor
1	0	0.00	0	0.0000	0.0000	0.000	0.000
2	134621	50.18	78850	1.7073	0.0096	0.013	0.802
3	137137	52.06	78677	1.7430	0.0097	0.013	0.799
4	138538	53.09	78721	1.7599	0.0097	0.014	0.798
5	138814	53.72	78588	1.7664	0.0098	0.014	0.799
6	138450	54.23	78282	1.7686	0.0098	0.014	0.800
7	137839	54.72	77887	1.7697	0.0098	0.014	0.801
8	57210	57.30	31219	1.8325	0.0099	0.014	0.793
9	59910	59.09	31937	1.8759	0.0100	0.014	0.788
10	61736	60.35	32418	1.9044	0.0100	0.015	0.785
11	62908	61.38	32676	1.9252	0.0101	0.015	0.783

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Job Name: LPG COLUMN Date: [illegible] 2009 Time: 22:31:24

12	63758	62.39	[illegible]	1.9443	0.0101	0.015	0.782
13	64564	63.50	[illegible]	1.9654	0.0102	0.015	0.780
14	65506	64.73	32918	1.9900	0.0102	0.015	0.778
15	66654	66.02	33038	2.0175	0.0103	0.015	0.776
16	67972	67.29	33221	2.0461	0.0103	0.015	0.773
17	69356	68.42	33454	2.0732	0.0103	0.015	0.770
18	70678	69.38	33705	2.0970	0.0104	0.015	0.768
19	71842	70.15	33943	2.1165	0.0104	0.016	0.766
20	72798	70.72	34147	2.1319	0.0104	0.016	0.764
21	73544	71.15	34311	2.1435	0.0104	0.016	0.763
22	74105	71.46	34434	2.1521	0.0104	0.016	0.762
23	74514	71.67	34523	2.1584	0.0104	0.016	0.762
24	74808	71.82	34586	2.1630	0.0104	0.016	0.761
25	75013	71.93	34626	2.1664	0.0104	0.016	0.761
26	75157	72.00	34652	2.1689	0.0104	0.016	0.760
27	75257	72.05	34668	2.1708	0.0104	0.016	0.760
28	75329	72.08	34678	2.1723	0.0104	0.016	0.760
29	75380	72.10	34682	2.1735	0.0104	0.016	0.760
30	75414	72.12	34682	2.1744	0.0104	0.016	0.760
31	75427	72.13	34675	2.1752	0.0104	0.016	0.760
32	75432	72.13	34667	2.1759	0.0104	0.016	0.760
33	75430	72.14	34657	2.1765	0.0104	0.016	0.760
34	75423	72.14	34646	2.1770	0.0104	0.016	0.760
35	75411	72.15	34633	2.1774	0.0104	0.016	0.760
36	75399	72.15	34623	2.1777	0.0104	0.016	0.760
37	75383	72.15	34611	2.1780	0.0104	0.016	0.759
38	75361	72.15	34599	2.1782	0.0104	0.016	0.759
39	75333	72.15	34584	2.1782	0.0104	0.016	0.759
40	75299	72.15	34569	2.1782	0.0104	0.016	0.759
41	75259	72.15	34553	2.1781	0.0104	0.016	0.759
42	75213	72.15	34536	2.1778	0.0104	0.016	0.759
43	75159	72.15	34518	2.1774	0.0104	0.016	0.759
44	75098	72.15	34498	2.1769	0.0104	0.016	0.759

	Vap H MMBtu/h
1	0
2	-131.16
3	-131.44
4	-131.52
5	-130.94
6	-129.89
7	-128.63
8	-52.301
9	-53.98
10	-55.036
11	-55.593
12	-55.889
13	-56.138
14	-56.495
15	-57.037
16	-57.752
17	-58.573
18	-59.4
19	-60.153
20	-60.785

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Job Name: LPG COLUMN Date: [illegible] Time: 22:31:24

21
22 -61.[illegible]
23 -61.664
24 -61.94
25 -62.139
26 -62.275
27 -62.369
28 -62.431
29 -62.473
30 -62.499
31 -62.511
32 -62.505
33 -62.492
34 -62.471
35 -62.442
36 -62.406
37 -62.367
38 -62.319
39 -62.262
40 -62.193
41 -62.113
42 -62.019
43 -61.912
44 -61.788
-61.649

5.9 Column Diagnosis Report

CHEMCAD 6.0.1

Job Name: LPG COLUMN Date: 04/28/2009 Time: 06:30:07

Column 1 has convereged

6. ANALYSIS ON CHARACTERISTICS OF TOWER AND FEED STREAM - CHEMCAD 6.0.1

Following charts are generated by CHEMCAD 6.0.1 which are explain about characteristics of tower as well as stream in varies conditions.

1. Tray temperature profile - Tower
2. Pseudo component curve - Tower
3. Heat curve - Tower
4. Phase envelope – Feed stream
5. Liquid viscosity Vs Temperature - Feed stream
6. Vapor viscosity Vs Temperature – Feed stream
7. Vapor pressure Vs Temperature – Feed stream
8. Liquid density Vs Temperature – Feed stream
9. Vapor density Vs Temperature – Feed stream

Tray temperature profile - Tower

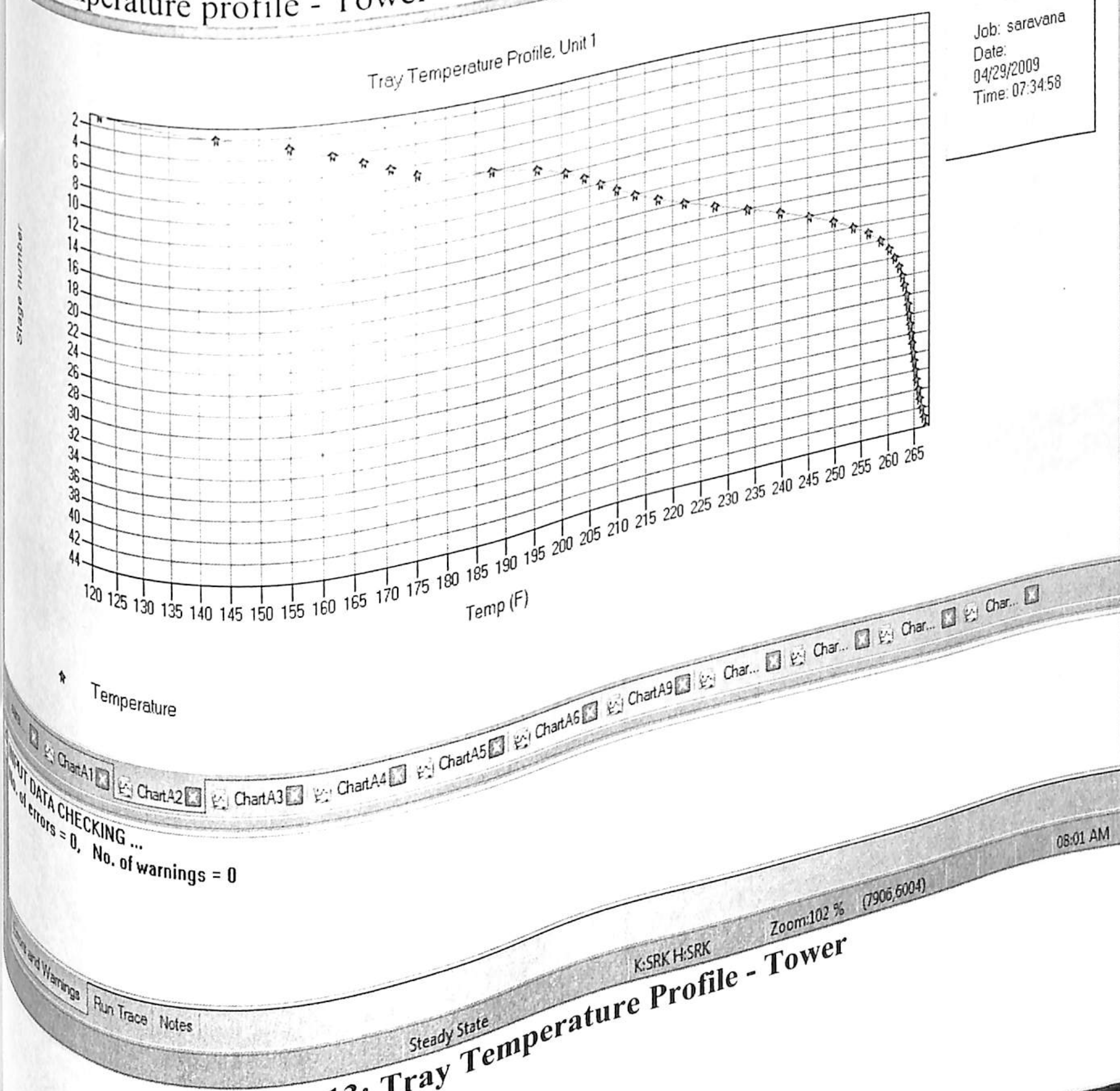

Figure 13: Tray Temperature Profile - Tower

6.2 Pseudocomponent Curve - Tower

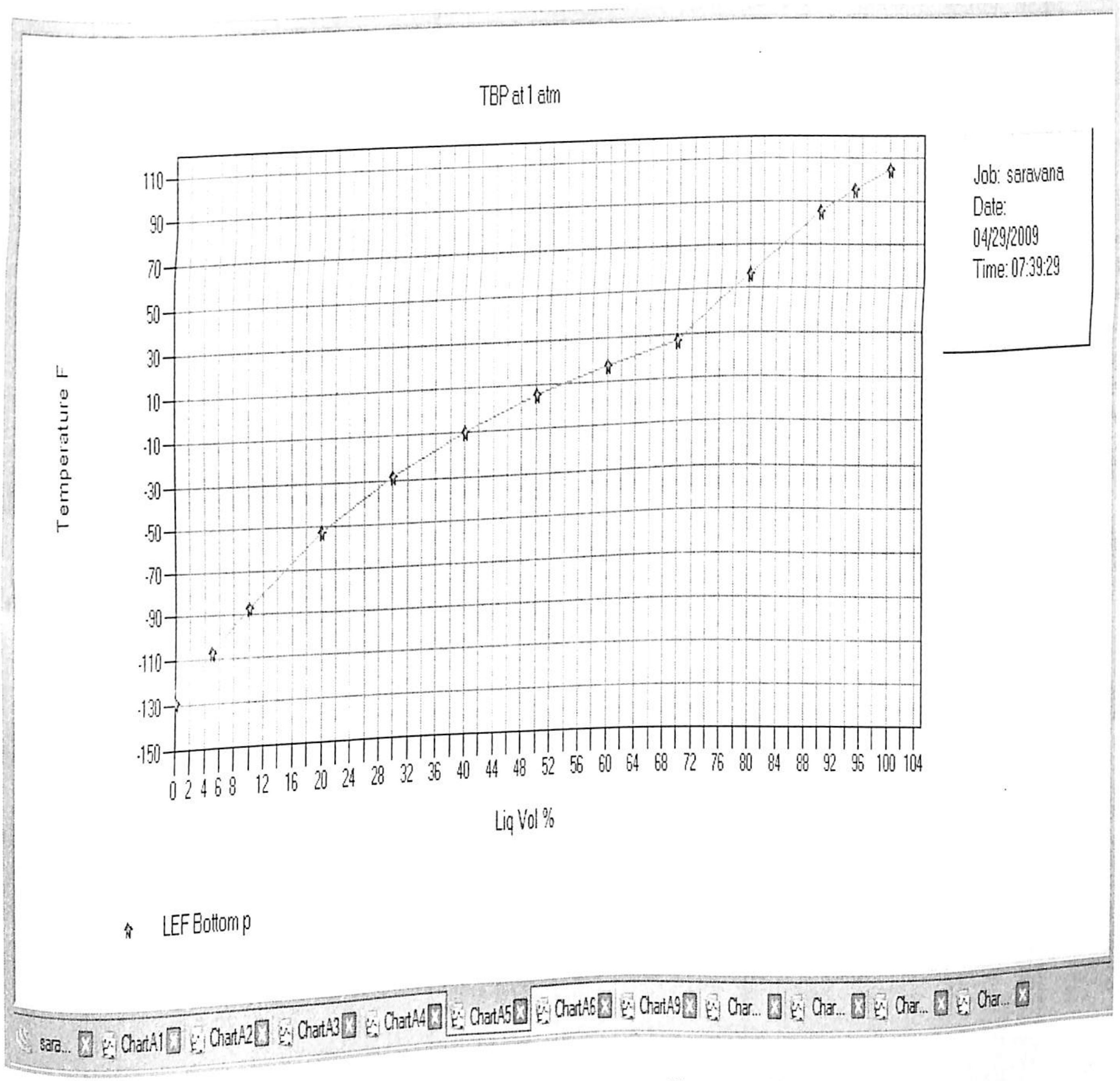

Figure 14: Pseudo Component Curve - Tower

6.3 Heat Curve - Tower

CHEMCAD 6.0.1 Page 1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 16:09:09

Eqp # 1 Unit type : TOWR Unit name:

Condenser

NP	Temp F	Pres psia	Del H MMBtu/h	Vapor lb/h	Liquid lb/h	Vap mole frac.	Vap mass frac.	
1	142.8	176.4	19.1	135286	0	1.0000	1.0000	Dew
2	140.6	176.4	17.2	121386	13900	0.9029	0.8973	
3	138.5	176.4	15.3	107556	27730	0.8051	0.7950	
4	136.4	176.4	13.4	93808	41478	0.7066	0.6934	
5	134.3	176.4	11.5	80143	55143	0.6075	0.5924	
6	132.3	176.4	9.55	66562	68725	0.5078	0.4920	
7	130.2	176.4	7.64	53063	82223	0.4074	0.3922	
8	128.1	176.4	5.73	39647	95639	0.3064	0.2931	
9	126.1	176.4	3.82	26319	108968	0.2047	0.1945	
10	124.0	176.4	1.91	13091	122195	0.1025	0.0968	
11	121.8	176.4	0.000	1	135285	0.0000	0.0000	Bub

Reboiler

NP	Temp F	Pres psia	Del H MMBtu/h	Vapor lb/h	Liquid lb/h	Vap mole frac.	Vap mass frac.	
1	266.7	178.6	0.000	0	86422	0.0000	0.0000	Bub
2	266.8	178.6	0.808	7603	78818	0.0880	0.0880	
3	266.8	178.6	1.62	15204	71217	0.1759	0.1759	
4	266.9	178.6	2.43	22805	63617	0.2639	0.2639	
5	266.9	178.6	3.23	30404	56018	0.3518	0.3518	
6	267.0	178.6	4.04	38002	48420	0.4397	0.4397	
7	267.0	178.6	4.85	45600	40822	0.5276	0.5276	
8	267.1	178.6	5.66	53194	33227	0.6155	0.6155	
9	267.1	178.6	6.47	60788	25634	0.7034	0.7034	
10	267.1	178.6	7.28	68381	18041	0.7912	0.7912	
11	267.2	178.6	8.08	75972	10449	0.8791	0.8791	

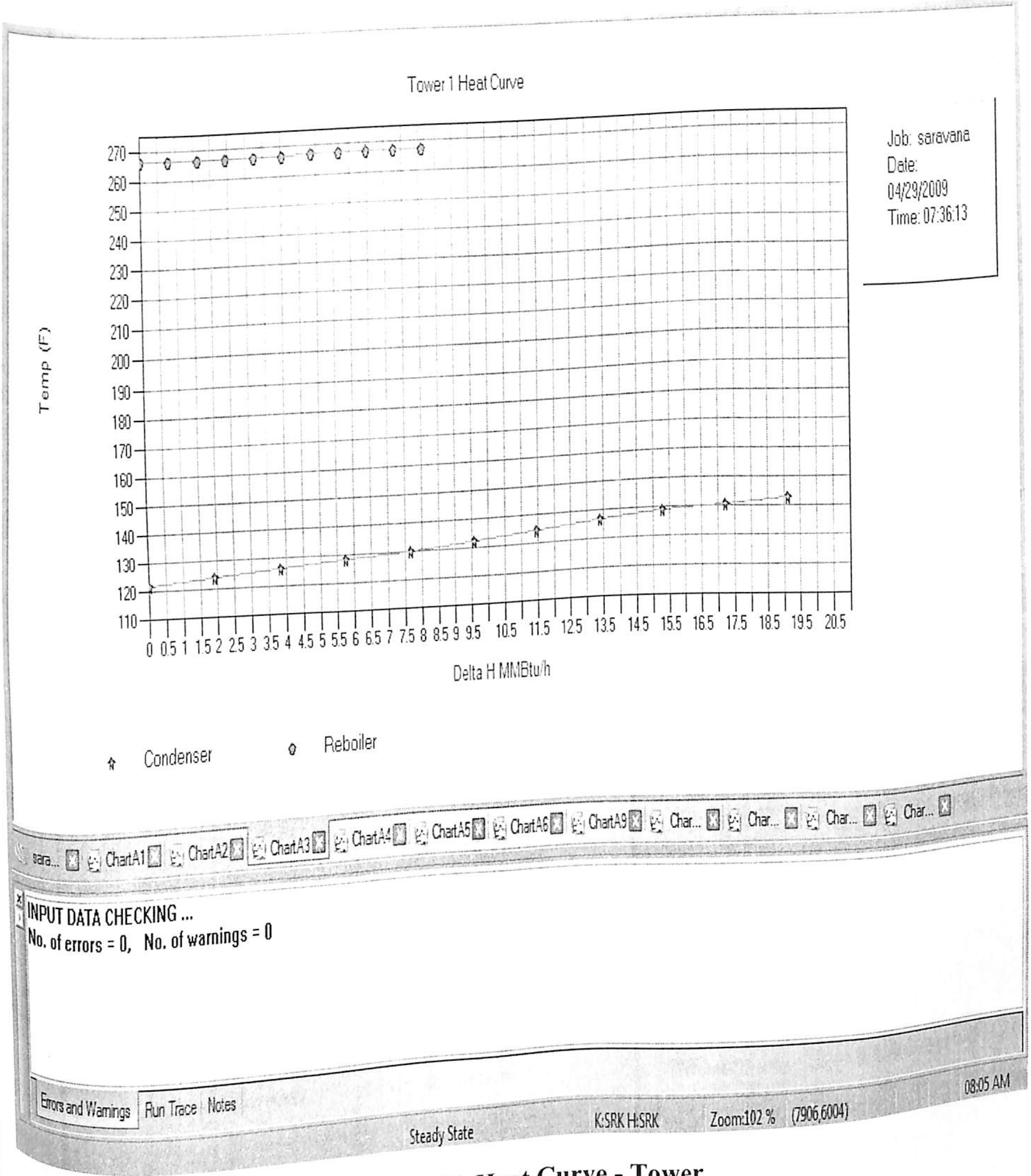

Figure 15: Heat Curve - Tower

6.4 Phase Envelope – Feed Stream

CHEMCAD 6.0.1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 08:17:35

Phase Envelope for Stream 1

No.	Temp F	Press psia	Vfrac	Zv	Zl
1	56.866	64.339	0.00000	0.911	0.019
2	66.866	74.866	0.00000	0.900	0.022
3	76.866	86.602	0.00000	0.889	0.025
4	86.866	99.622	0.00000	0.876	0.029
5	96.866	113.999	0.00000	0.863	0.033
6	106.866	129.807	0.00000	0.849	0.038
7	116.866	147.116	0.00000	0.835	0.043
8	126.866	165.994	0.00000	0.819	0.049
9	136.866	186.509	0.00000	0.802	0.055
10	146.866	208.721	0.00000	0.785	0.062
11	156.866	232.687	0.00000	0.766	0.070
12	166.866	258.458	0.00000	0.746	0.078
13	176.866	286.075	0.00000	0.725	0.088
14	186.866	315.570	0.00000	0.702	0.099
15	196.866	346.958	0.00000	0.678	0.111
16	206.866	380.233	0.00000	0.652	0.124
17	216.866	415.354	0.00000	0.625	0.140
18	226.866	452.228	0.00000	0.594	0.158
19	231.866	471.271	0.00000	0.578	0.168
20	234.366	480.930	0.00000	0.569	0.173
21	235.616	485.791	0.00000	0.565	0.176
22	236.616	489.694	0.00000	0.561	0.178
23	237.616	493.610	0.00000	0.557	0.181
24	238.616	497.538	0.00000	0.554	0.183
25	239.616	501.476	0.00000	0.550	0.186
26	240.616	505.426	0.00000	0.546	0.188
27	241.616	509.386	0.00000	0.542	0.191
28	242.616	513.356	0.00000	0.539	0.193
29	243.616	517.334	0.00000	0.535	0.196
30	244.616	521.320	0.00000	0.531	0.199
31	245.616	525.313	0.00000	0.527	0.202
32	246.616	529.313	0.00000	0.523	0.204
33	247.616	533.317	0.00000	0.518	0.207
34	248.616	537.325	0.00000	0.514	0.210
35	249.616	541.335	0.00000	0.510	0.213
36	250.616	545.347	0.00000	0.505	0.217
37	251.616	549.358	0.00000	0.501	0.220
38	252.616	553.366	0.00000	0.496	0.223
39	253.616	557.369	0.00000	0.491	0.227
40	254.616	561.367	0.00000	0.487	0.231
41	255.616	565.355	0.00000	0.482	0.234
42	256.616	569.330	0.00000	0.476	0.238
43	257.616	573.290	0.00000	0.471	0.242
45	259.616	581.146	0.00000	0.460	0.251
46	260.616	585.031	0.00000	0.454	0.256
47	261.616	588.881	0.00000	0.448	0.260
48	262.616	592.685	0.00000	0.442	0.266
49	263.616	596.432	0.00000	0.436	0.271
50	264.616	600.111	0.00000	0.429	0.277
51	265.616	603.702	0.00000	0.421	0.283

Job Name: LPG COLUMN Date: 04/29/2009 Time: 08:17:35

52	266.616	607.183	0.00000	0.414	0.290
53	267.616	610.524	0.00000	0.405	0.297
54	268.616	613.666	0.00000	0.397	0.305
55	269.616	616.524	0.00000	0.386	0.315
56	270.616	618.916	0.00000	0.378	0.326
57	271.616	614.160	0.00000	0.382	0.366
1	98.617	62.771	1.00000	0.912	0.020
2	121.585	91.019	1.00000	0.884	0.028
3	139.635	119.266	1.00000	0.858	0.037
4	154.685	147.513	1.00000	0.833	0.045
5	167.685	175.760	1.00000	0.809	0.054
6	179.179	204.007	1.00000	0.786	0.063
7	189.513	232.254	1.00000	0.764	0.073
8	198.919	260.502	1.00000	0.742	0.082
9	207.560	288.749	1.00000	0.720	0.092
10	215.557	316.996	1.00000	0.698	0.103
11	222.999	345.243	1.00000	0.676	0.114
12	229.956	373.490	1.00000	0.654	0.126
13	236.479	401.737	1.00000	0.632	0.138
14	242.609	429.985	1.00000	0.608	0.151
15	248.373	458.232	1.00000	0.584	0.166
16	253.770	486.479	1.00000	0.559	0.182
17	258.859	514.726	1.00000	0.532	0.200
18	263.564	542.973	1.00000	0.503	0.220
19	265.762	557.097	1.00000	0.486	0.232
20	266.817	564.159	1.00000	0.478	0.239
21	267.329	567.689	1.00000	0.473	0.242
22	267.760	570.689	1.00000	0.469	0.245
23	268.182	573.689	1.00000	0.465	0.248
24	268.598	576.689	1.00000	0.461	0.252
25	269.006	579.689	1.00000	0.457	0.255
26	269.405	582.689	1.00000	0.452	0.258
27	269.795	585.689	1.00000	0.448	0.262
28	270.174	588.689	1.00000	0.443	0.266
29	270.542	591.689	1.00000	0.438	0.270
30	270.896	594.689	1.00000	0.433	0.274
31	271.235	597.689	1.00000	0.428	0.278
32	271.555	600.689	1.00000	0.422	0.283
33	271.852	603.689	1.00000	0.416	0.288
34	272.119	606.689	1.00000	0.409	0.293
35	272.348	609.689	1.00000	0.401	0.299
36	272.526	612.689	1.00000	0.393	0.306
37	272.729	615.689	1.00000	0.384	0.320
38	272.116	618.689	1.00000	0.359	0.318

T_c = 271.116 F
P_c = 616.538 psia
Z_c = 0.344

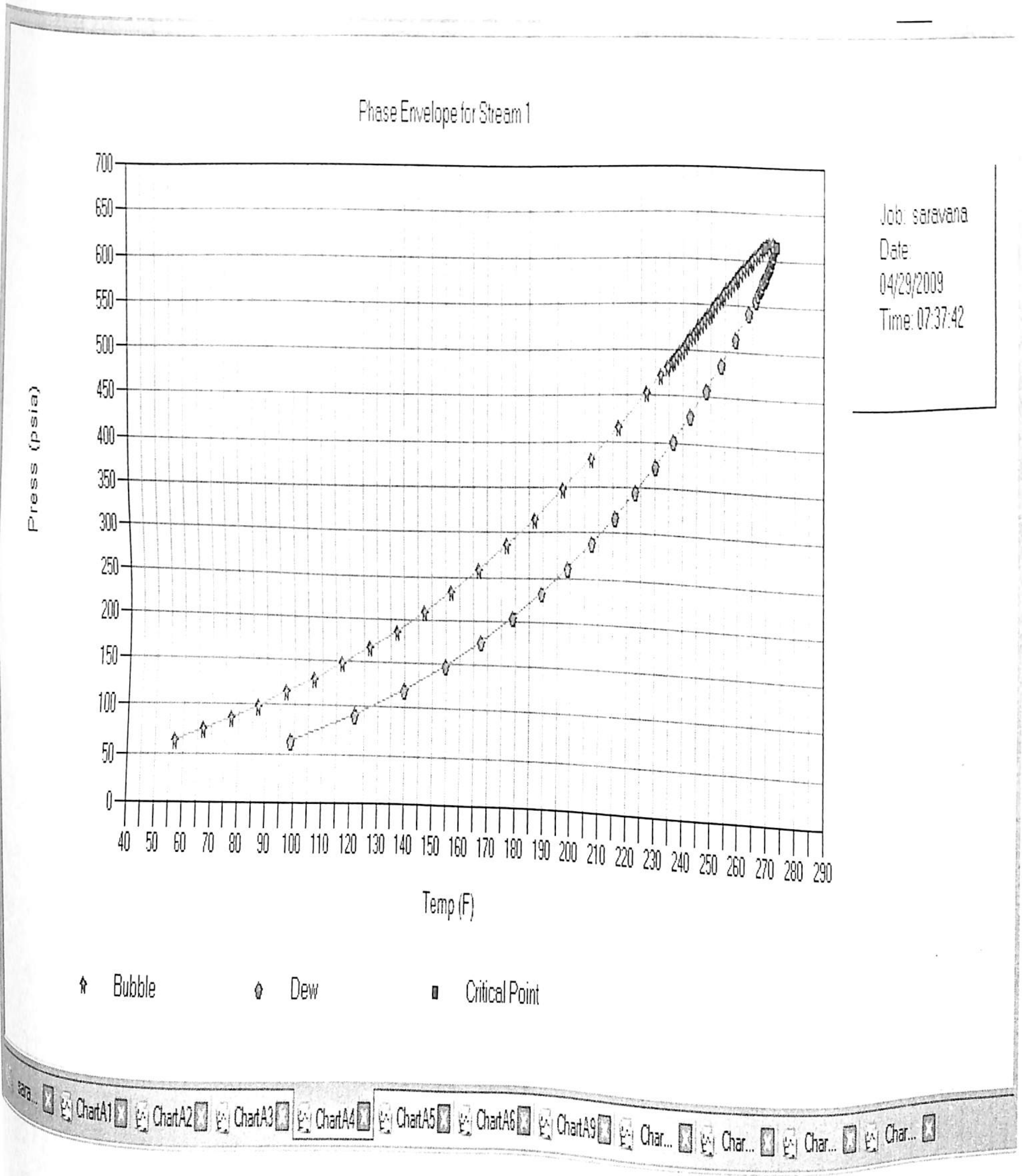

Figure 16: Phase Envelope – Feed Stream

6.5 Liquid Viscosity Vs Temperature – Feed Stream

CHEMCAD 6.0.1 Page 1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 16:17:23

Stream 1 Pressure = 176.355 psia

Temperature F	Liquid Viscosity cP
121.000	1.03931e-001
128.842	9.91624e-002
136.684	9.45123e-002
144.526	8.99762e-002
152.368	8.55469e-002
160.210	8.12150e-002
168.053	7.69820e-002
175.895	7.28721e-002
183.737	6.88536e-002
191.579	6.49237e-002
199.421	6.11194e-002
207.263	5.77802e-002
215.105	5.64272e-002
222.947	5.50461e-002
230.789	5.36338e-002
238.632	5.21867e-002
246.474	5.07012e-002
254.316	4.91739e-002
262.158	4.76057e-002
270.000	4.59962e-002

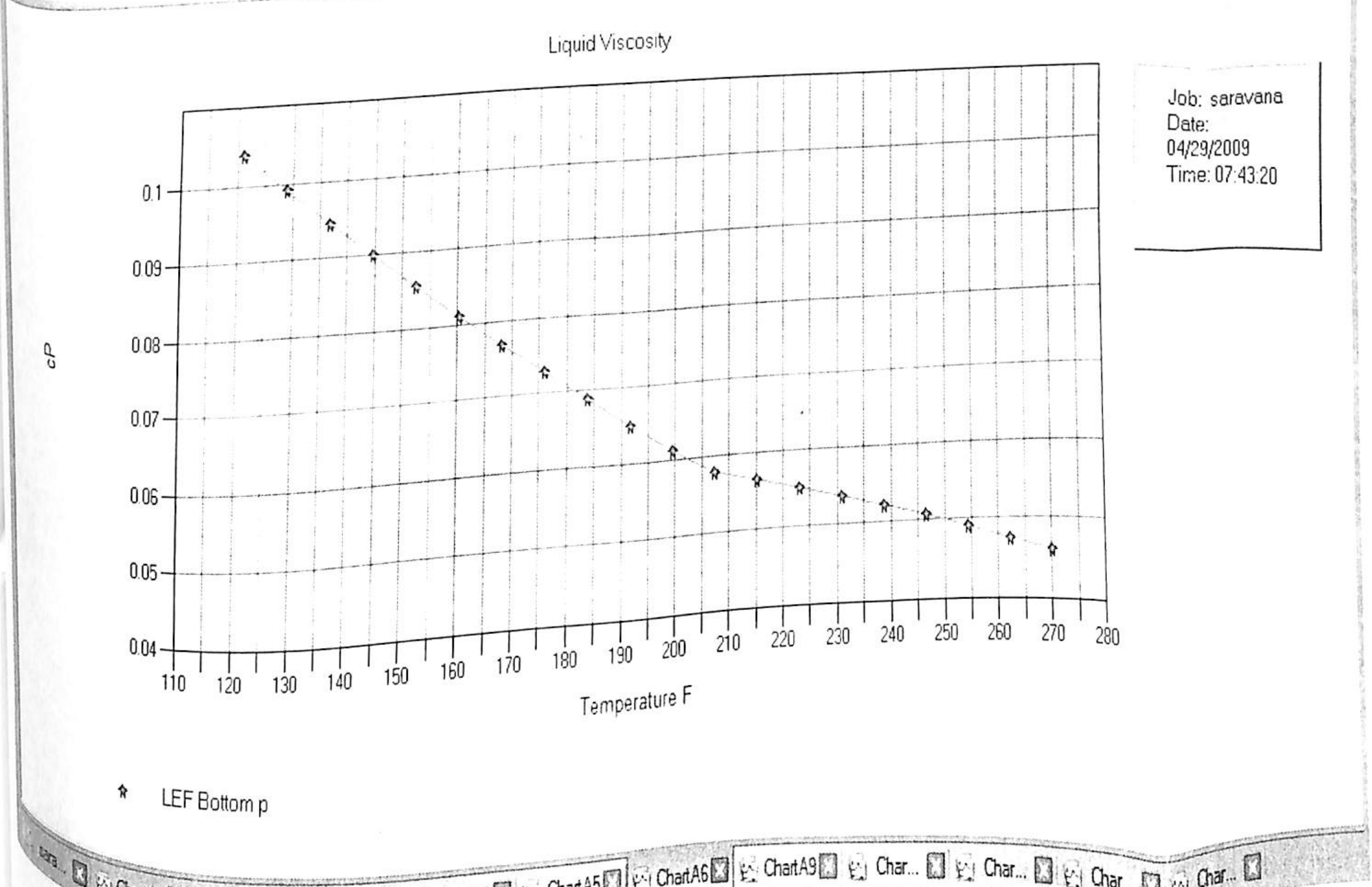

Figure 17: Liquid Viscosity Vs Temperature – Feed Stream

6.6 Vapor Viscosity Vs Temperature – Feed Stream

CHEMCAD 6.0.1 Page 1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 16:18:23

Stream 1 Pressure = 176.355 psia

Temperature F	Vapor Viscosity cP
121.000	9.31369e-003
128.842	9.39012e-003
136.684	9.46959e-003
144.526	9.55145e-003
152.368	9.63523e-003
160.210	9.72057e-003
168.053	9.80718e-003
175.895	9.89484e-003
183.737	9.98336e-003
191.579	1.00726e-002
199.421	1.01624e-002
207.263	1.02527e-002
215.105	1.03434e-002
222.947	1.04344e-002
230.789	1.05256e-002
238.632	1.06171e-002
246.474	1.07087e-002
254.316	1.08003e-002
262.158	1.08921e-002
270.000	1.09839e-002

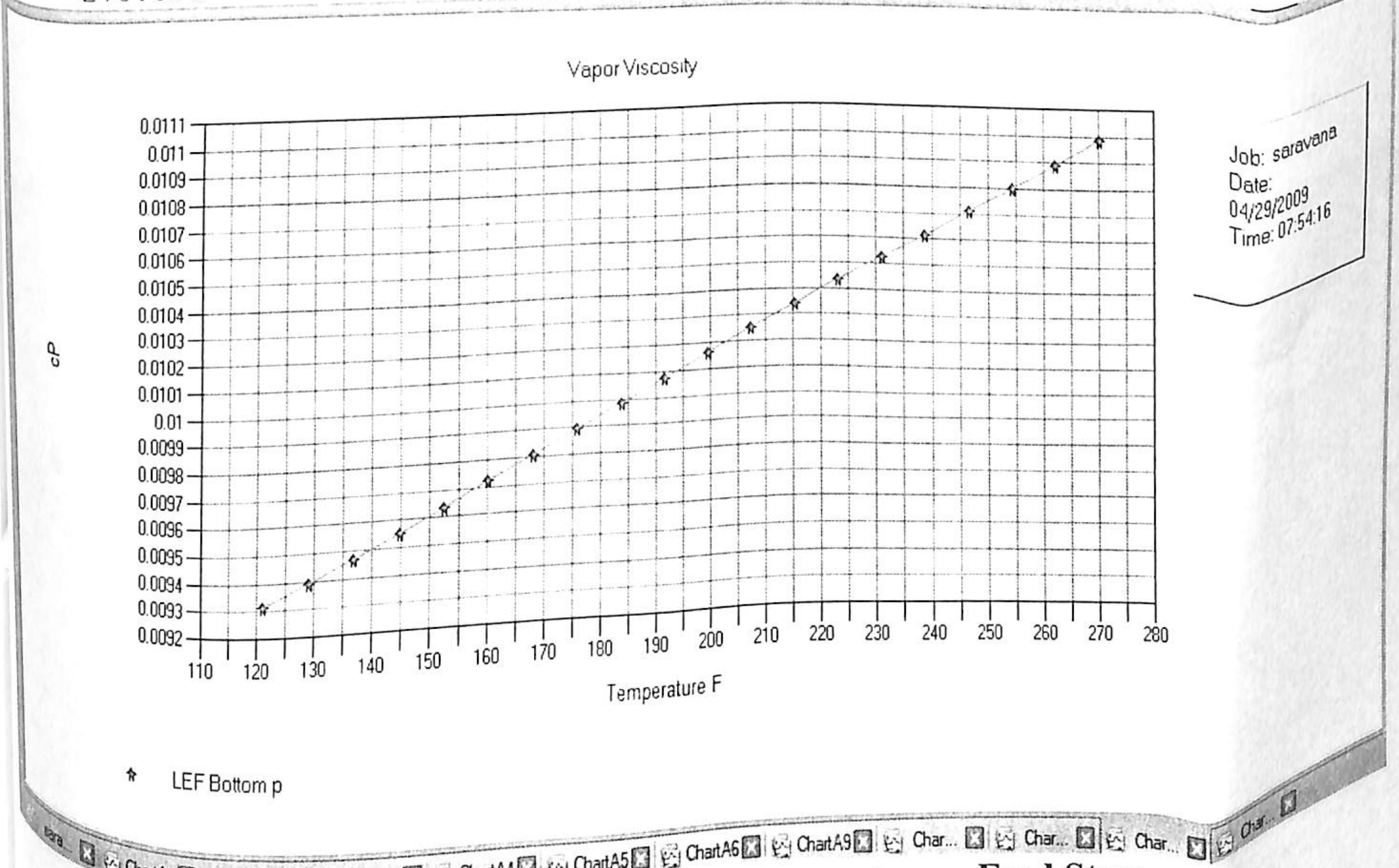

Figure 18: Vapor Viscosity Vs Temperature – Feed Stream

6.7 Vapor Pressure Vs Temperature – Feed Stream

```
CHEMCAD 6.0.1                                                    Page 1

Job Name: LPG COLUMN   Date: 04/29/2009  Time: 08:20:38

Stream 1  Pressure =      176.355 psia

                          Vapor Pressure
Temperature F                 psia
   121.000               6.79130e+005
   128.842               5.88018e+007
   136.684               5.09130e+009
   144.526               4.40825e+011
   152.368               1.50150e+005
   160.210               1.30006e+007
   168.053               1.12564e+009
   175.895               9.74629e+010
   183.737               8.43873e+012
 Warning: K values = 1.
   191.579               1.00000e+005
 Warning: K values = 1.
   199.421               1.00000e+005
 Warning: K values = 1.
   207.263               1.00000e+005
   215.105               8.65841e+006
 Warning: K values = 1.
   222.947               1.00000e+005
 Warning: K values = 1.
   230.789               1.00000e+005
 Warning: K values = 1.
   238.632               1.00000e+005
 Warning: K values = 1.
   246.474               1.00000e+005
   254.316               8.65841e+006
   262.158               7.49680e+008
   270.000               1.00100e+005
```

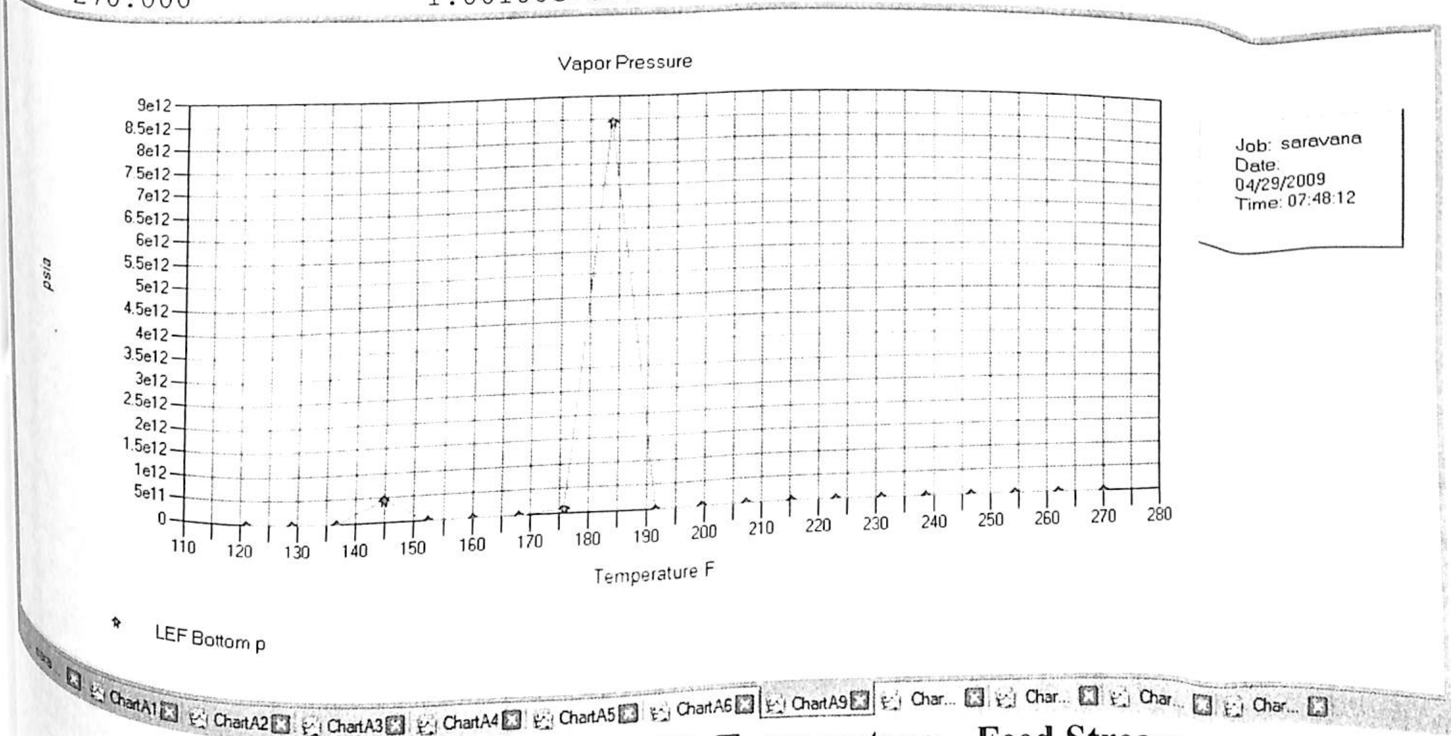

Figure 19: Vapor Pressure Vs Temperature – Feed Stream

6.8 Liquid Density Vs Temperature – Feed Stream

CHEMCAD 6.0.1 Page 1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 16:20:25

Stream 1 Pressure = 176.355 psia

Temperature F	Liquid Density lb/ft3
121.000	3.12027e+001
128.842	3.07572e+001
136.684	3.02916e+001
144.526	2.98021e+001
152.368	2.92837e+001
160.210	2.87293e+001
168.053	2.81285e+001
175.895	2.74641e+001
183.737	2.67057e+001
191.579	2.57880e+001
199.421	2.45099e+001
207.263	2.01451e+001
215.105	2.00079e+001
222.947	1.98598e+001
230.789	1.96983e+001
238.632	1.95198e+001
246.474	1.93186e+001
254.316	1.90845e+001
262.158	1.87956e+001
270.000	1.83774e+001

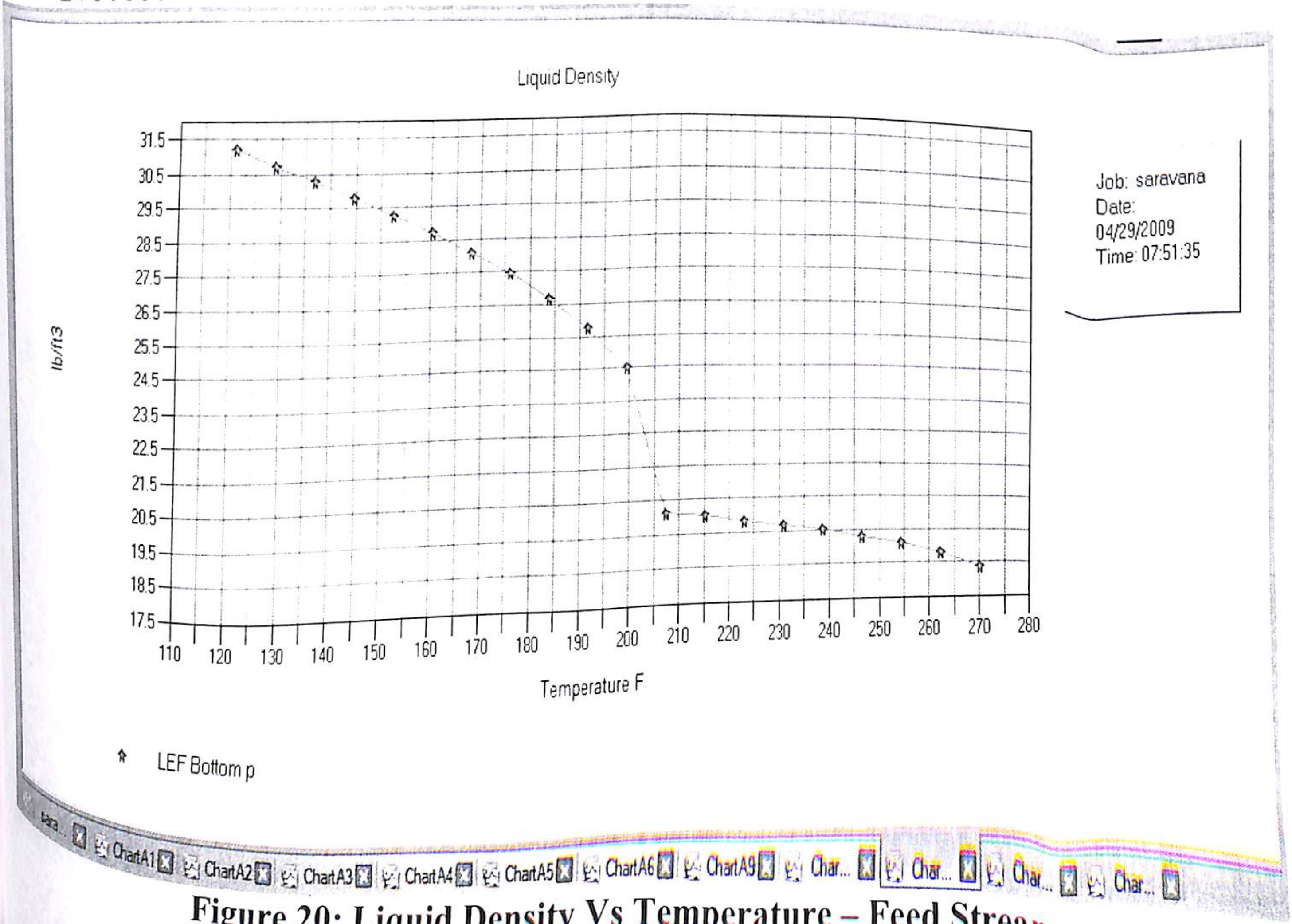

Figure 20: Liquid Density Vs Temperature – Feed Stream

6.9 Vapor Density Vs Temperature – Feed Stream

CHEMCAD 6.0.1

Page [illegible]

Job Name: LPG COLUMN Date: 04/29/2009 Time: [illegible]

Stream 1 Pressure = 176.355 psia

Temperature F	Vapor Density lb/ft3
121.000	2.02028e+000
128.842	1.95529e+000
136.684	1.89665e+000
144.526	1.84325e+000
152.368	1.79424e+000
160.210	1.74898e+000
168.053	1.70695e+000
175.895	1.66774e+000
183.737	1.63102e+000
191.579	1.59649e+000
199.421	1.56394e+000
207.263	1.53316e+000
215.105	1.50398e+000
222.947	1.47625e+000
230.789	1.44985e+000
238.632	1.42467e+000
246.474	1.40060e+000
254.316	1.37757e+000
262.158	1.35549e+000
270.000	1.33430e+000

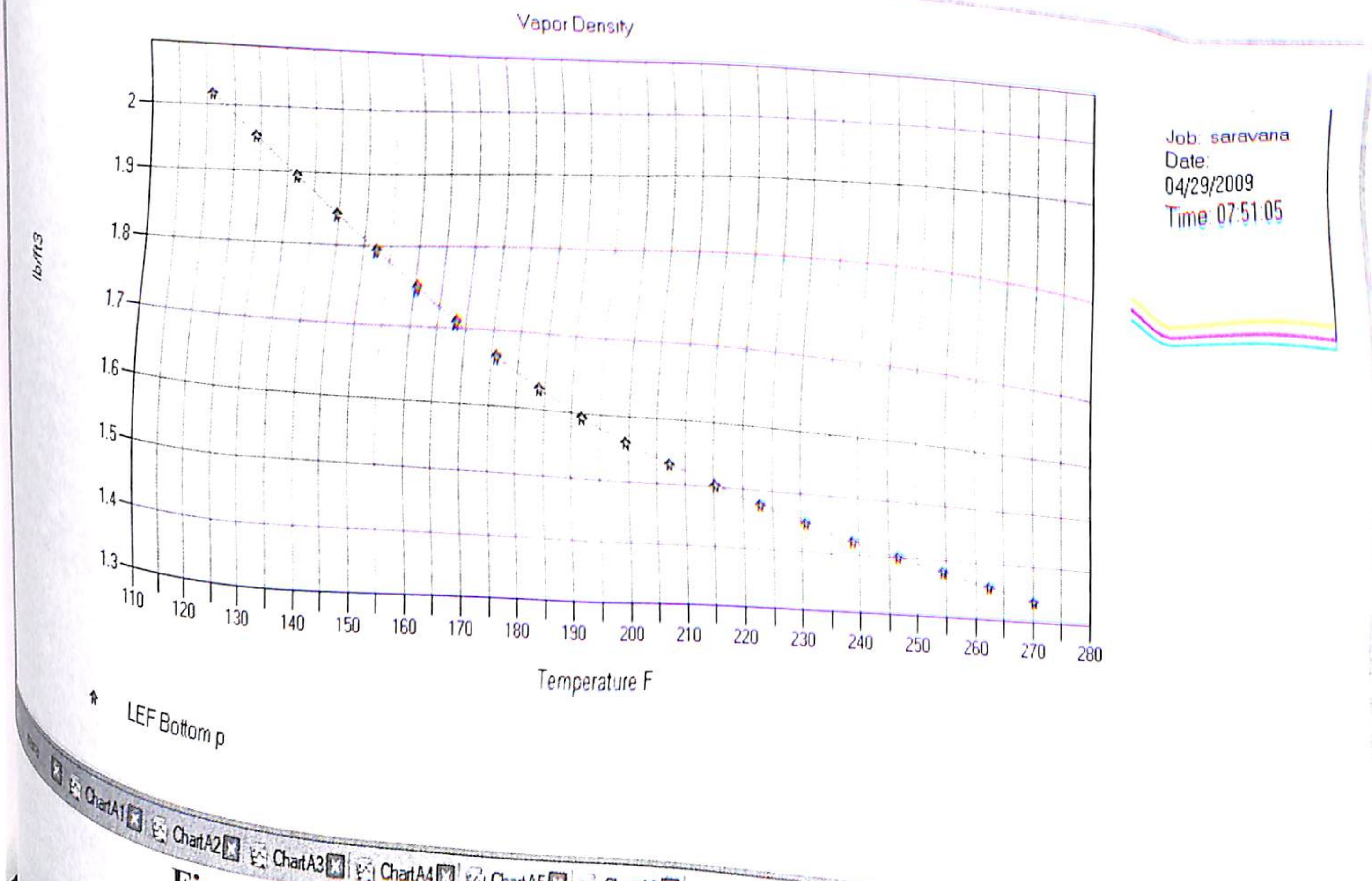

Figure 21: Vapor Density Vs Temperature – Feed Stream

7. LPG COLUMN SIZING

7.1 Column Sizing Sieve Tray

Page 1

CHEMCAD 6.0.1

Job Name: LPG COLUMN   Date: 04/28/2009   Time: 19:51:51

Vapor load is defined as the vapor from the tray below.
Liquid load is defined as the liquid on the tray.

Section: 1
Flood correlation: Fair

Equip. 1   Tray No. 1

		Vapor	Liquid
Tray Loadings		104119.000 lb/h	49147.703 lb/h
		61186.832 ft3/hr	1622.310 ft3/hr
Density		1.702 lb/ft3	30.295 lb/ft3
Tower internal diameter, ft	....		6.000
Tray spacing, ft	....		2.000
No. of tray liquid passes	....		1
Downcomer dimension	Width ft	Length ft	Area ft2
Side	1.083	4.616	3.476
Avg. weir length ft	....		4.616
Weir height, ft	....		0.167
Flow path length ft	....		3.833
Flow path width ft	....		5.562
Tray area, ft2	....		28.274
Tray active area ft2	....		21.322
% flood	....		68.868
Fractional entrainment	....		0.015
Aeration factor	....		0.646
Minimum (Weeping) vapor flow lb/h	....		102956.078
Tray press loss, ft	....		0.230
Tray press loss, psi	....		0.048
Downcomer clearance ft	....		0.146
Downcomer backup ft	....		0.516
Downcomer residence time sec	....		3.982
Downcomer apparent residence time sec	....		15.428
Downcomer velocity ft/sec	....		0.130
Liquid holdup ft3	....		5.432
Liquid holdup lb	....		164.570
Design pressure psia	....		176.355
Joint efficiency	....		0.850
Allowable stress psia	....		13700.000
Corrosion allowance ft	....		0.003
Column thickness ft	....		0.047
Bottom thickness ft	....		0.049

7.2 Costing Sieve Tray


```
CHEMCAD 6.0.1

Job Name: LPG COLUMN     Date: 04/28/2009   Time: 20:00:45

Preliminary Distillation Towers Cost Estimation

Distillation Tower Cost for Equip. 1
Column material = Carbon steel
Sieve tray
Tray material = Carbon steel
Base cost index = 347.5
Current cost index = 610.4

Calculated cost:                  = $        118284
Shell cost                        = $         45320
Tray cost                         = $         24500
Platform & ladders                = $        330414
Column cost (purchase)            = $        991241
Column cost (installed)           = $             0
Condenser cost (purchase)         = $             0
Condenser cost (installed)        = $             0
Reboiler cost (purchase)          = $             0
Reboiler cost (installed)         = $        330414
Total cost (purchase)             = $        991241
Total cost (installed)
```

7.3 Column Sizing – Value Tray

CHEMCAD 6.0.1

Job Name: LPG COLUMN   Date: 04/28/2009   Time: 19:55:01

Vapor load is defined as the vapor from the tray below.
Liquid load is defined as the liquid on the tray.

Section: 1
Flood correlation: Glistch

Equip. 1   Tray No. 2

	Vapor		Liquid	
Tray Loadings	105157.141	lb/h	50185.875	lb/h
	60794.498	ft3/hr	1644.608	ft3/hr
Density	1.730	lb/ft3	30.515	lb/ft3

System factor			1.000
Valve type : V-1			
Valve material : S.S.			
Valve thickness gauge			12.000
Deck thickness gauge			14.000
Tower internal diameter, ft			5.000
Tray spacing, ft			2.000
No. of tray liquid passes			1
Downcomer dimension	Width ft	Length ft	Area ft2
Side	0.771	3.611	1.922
Avg. weir length ft			3.611
Weir height, ft			0.167
Flow path length ft			3.458
Flow path width ft			4.566
Tray area, ft2			19.635
Tray active area ft2			15.792
% flood			70.723
Hole area ft2			3.000
Approx # of valves			235
Tray press loss, ft			0.386
Tray press loss, psi			0.082
Dry press drop, ft			0.225
Downcomer clearance ft			0.146
Downcomer head loss ft			0.041
Downcomer backup ft			0.713
Downcomer residence time sec			3.000
Downcomer apparent residence time sec			8.413
Downcomer velocity ft/sec			0.238
Liquid holdup ft3			3.907
Liquid holdup lb			119.217
Design pressure psia			176.409
Joint efficiency			0.850
Allowable stress psia			13700.000
Corrosion allowance ft			0.003
Column thickness ft			0.039
Bottom thickness ft			0.052

7.4 Costing - Value Tray

CHEMCAD 6.0.1

Job Name: LPG COLUMN   Date: 04/28/2009   Time: 20:01:55

Preliminary Distillation Towers Cost Estimation

Distillation Tower Cost for Equip. 1
Column material = Carbon steel
Valve tray
Tray material = Carbon steel
Base cost index = 347.5
Current cost index = 610.4

Calculated cost:		
Shell cost	= $	110639
Tray cost	= $	44807
Platform & ladders	= $	22221
Column cost (purchase)	= $	312082
Column cost (installed)	= $	936245
Condenser cost (purchase)	= $	0
Condenser cost (installed)	= $	0
Reboiler cost (purchase)	= $	0
Reboiler cost (installed)	= $	0
Total cost (purchase)	= $	312082
Total cost (installed)	= $	936245

7.6 Costing - Bubble Cap Tray

CHEMCAD 6.0.1

Job Name: LPG COLUMN   Date: 04/28/2009   Time: 19:59:33

Preliminary Distillation Towers Cost Estimation

Distillation Tower Cost for Equip. 1
Column material = Carbon steel
Bubble cap tray
Tray material = Carbon steel
Base cost index = 347.5
Current cost index = 610.4

Calculated cost:

Shell cost	= $	108430
Tray cost	= $	77716
Platform & ladders	= $	23377
Column cost (purchase)	= $	368035
Column cost (installed)	= $	1104104
Condenser cost (purchase)	= $	0
Condenser cost (installed)	= $	0
Reboiler cost (purchase)	= $	0
Reboiler cost (installed)	= $	0
Total cost (purchase)	= $	368035
Total cost (installed)	= $	1104104

7.7 Packed Column Sizing

Page 1

CHEMCAD 6.0.1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 00:06:06

Billet-Schultes Correlation

Packing Parameters

dN = 50.00000 mm
A = 105.00000
VoidFrac. = 0.96000

Packed Tower Design for Tower 1

Stg	PDrop psi	%Flood	VapLoad lb/(ft2*sec)	LiqLoad lb/(ft2*sec)	Diam ft	HTUov ft
2	0.011	75.00000	1.634	0.780	4.772	1.536
3	0.011	75.00000	1.627	0.779	4.791	1.593
4	0.011	75.00000	1.624	0.777	4.792	1.628
5	0.011	75.00000	1.623	0.774	4.787	1.654
6	0.011	75.00000	1.623	0.772	4.783	1.673
7	0.012	75.00000	1.346	1.906	2.570	0.747
8	0.012	75.00000	1.355	1.888	2.634	0.744
9	0.012	75.00000	1.361	1.875	2.682	0.744
10	0.012	75.00000	1.365	1.867	2.714	0.746
11	0.012	75.00000	1.367	1.862	2.734	0.748
12	0.012	75.00000	1.368	1.859	2.746	0.746
13	0.012	75.00000	1.372	1.861	2.751	0.761
14	0.012	75.00000	1.374	1.862	2.753	0.770
15	0.012	75.00000	1.375	1.862	2.755	0.776
16	0.012	75.00000	1.376	1.863	2.755	0.779
17	0.012	75.00000	1.376	1.863	2.756	0.782
18	0.012	75.00000	1.377	1.863	2.757	0.783
19	0.012	75.00000	1.377	1.864	2.757	0.785
20	0.012	75.00000	1.377	1.864	2.757	0.786
21	0.012	75.00000	1.378	1.864	2.757	0.787
22	0.012	75.00000	1.378	1.864	2.758	0.788
23	0.012	75.00000	1.378	1.864	2.758	0.789
24	0.012	75.00000	1.378	1.864	2.758	0.789
25	0.012	75.00000	1.379	1.865	2.758	0.790
26	0.012	75.00000	1.379	1.865	2.759	0.791
27	0.012	75.00000	1.379	1.865	2.759	0.791
28	0.012	75.00000	1.379	1.865	2.759	0.791
29	0.012	75.00000	1.380	1.865	2.759	0.792
30	0.012	75.00000	1.380	1.866	2.759	0.792
31	0.012	75.00000	1.380	1.866	2.759	0.792
32	0.012	75.00000	1.381	1.867	2.759	0.792
33	0.012	75.00000	1.382	1.867	2.760	0.792
34	0.012	75.00000	1.383	1.868	2.760	0.792
35	0.012	75.00000	1.384	1.869	2.760	0.792
36	0.012	75.00000	1.386	1.871	2.760	0.792
37	0.012	75.00000	1.389	1.874	2.761	0.792
38	0.012	75.00000	1.392	1.877	2.762	0.791
39	0.012	75.00000	1.398	1.881	2.765	0.790
40	0.012	75.00000	1.404	1.886	2.770	0.790
41	0.012	75.00000	1.413	1.891	2.779	0.790
42	0.012	75.00000	1.422	1.896	2.794	0.790
43	0.012	75.00000	1.431	1.898	2.815	0.792

Page 2

CHEMCAD 6.0.1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 00:06:07

Overall :		
Height ft	....	88.000
HETP ft	....	2.095
Pressure drop psi	....	0.496
Vapor Load at Loading lb/(ft2*sec)	....	1.266
Vapor load at Flooding lb/(ft2*sec)	......	1.880
Liquid Load at Loading lb/(ft2*sec)	......	1.562
Liquid Load at Flooding lb/(ft2*sec)	.....	2.321
Diameter at Loading ft	....	3.157
Diameter at Flooding ft	....	2.590
Design pressure psia	....	178.580
Joint efficiency	....	0.850
Allowable stress psia	....	13700.000
Corrosion allowance ft	....	0.003
Column thickness ft	....	0.036
Bottom thickness ft	....	0.089

7.8 Costing Packed Tower

Page 1

CHEMCAD 6.0.1

Job Name: LPG COLUMN Date: 04/29/2009 Time: 00:08:35

Preliminary Distillation Towers Cost Estimation

Distillation Tower Cost for Equip. 1

Column material = Carbon steel

Packed tower

Base cost index = 347.5

Current cost index = 610.4

Calculated cost:		
Shell cost	= $	127057
Packing cost	= $	0
Platform & ladders	= $	19272
Column cost (purchase)	= $	257034
Column cost (installed)	= $	771103
Condenser cost (purchase)	= $	0
Condenser cost (installed)	= $	0
Reboiler cost (purchase)	= $	0
Reboiler cost (installed)	= $	0
Total cost (purchase)	= $	257034
Total cost (installed)	= $	771103

8. DISCUSSION ON RESULTS

From calculation which are done by manually as well as simulator seams the packing tower is a best in cases of purchasing, installation, place occupation and also efficiency. Following Figures 23 & 24 are screen shots of Excel 2007, here cost comparison is worked out by using Excel 2007.

8.1 Purchase Cost Comparison

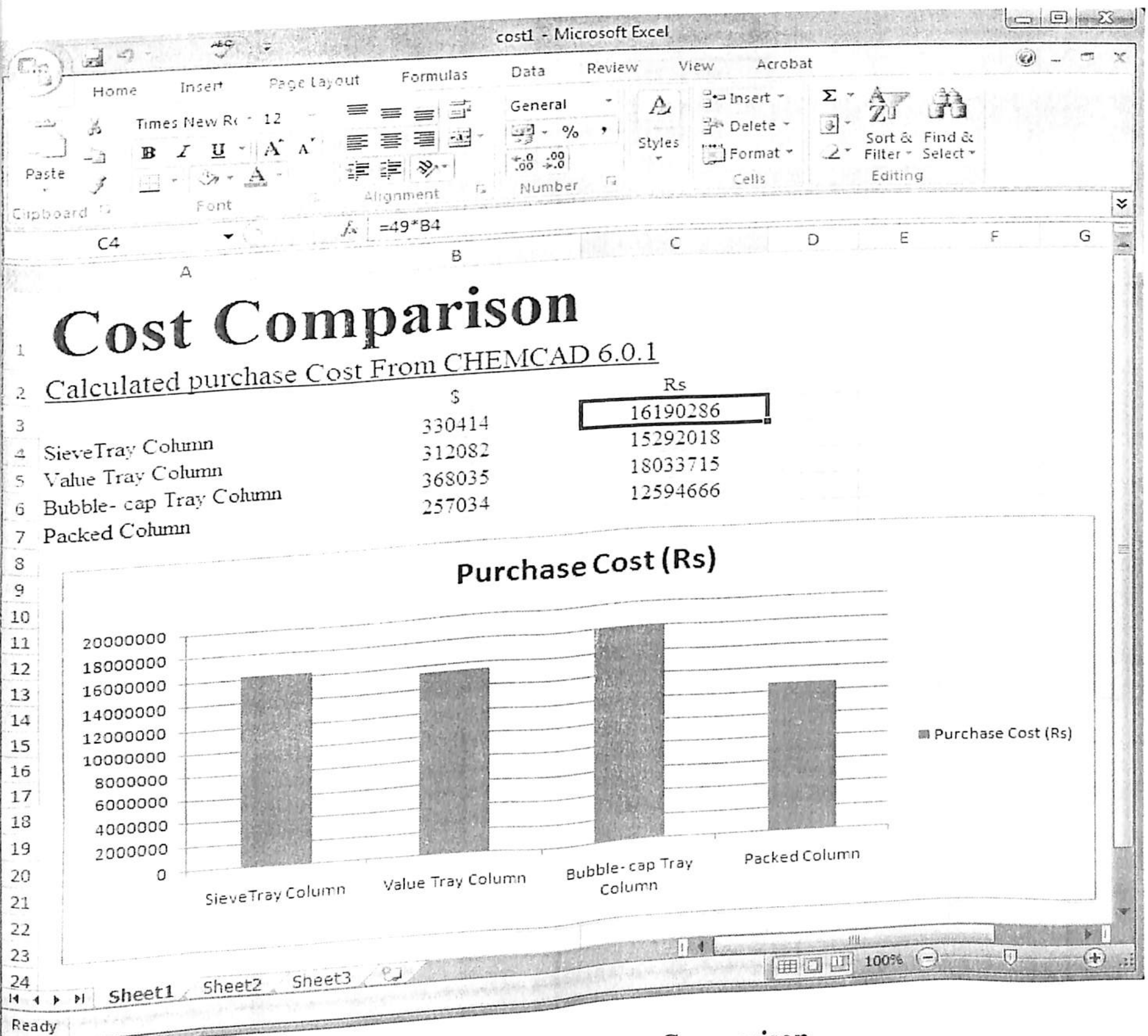

Figure 21: Purchase Cost Comparison

8.2 Installed Cost Comparison

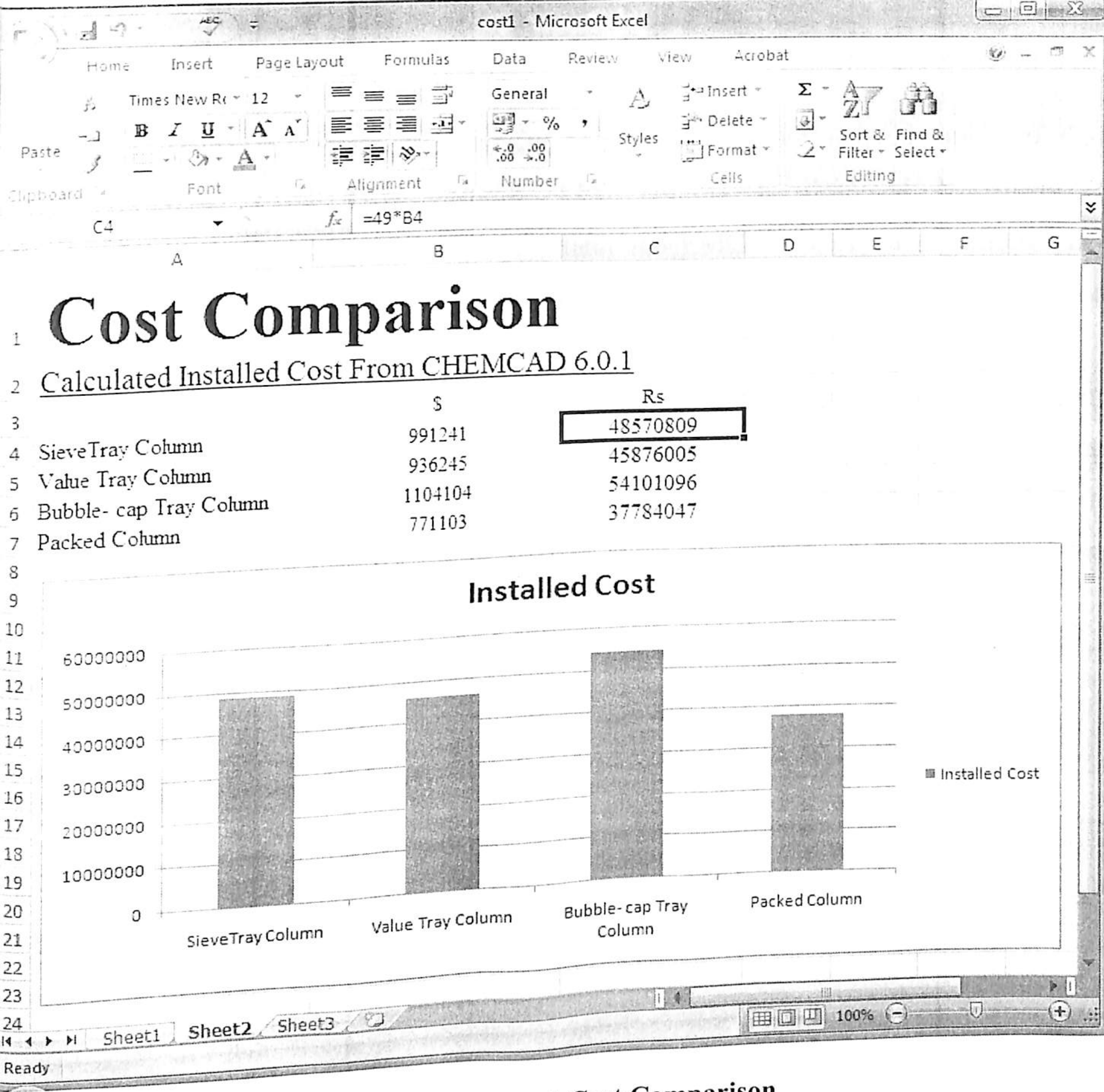

Figure 23: Installed Cost Comparison

9. CONCLUSIONS

- The project has been successfully focused on Design & Simulation of Gas Processing plant LPG column. LPG column is designed by manually (SPREADSHEATS) as well as simulator (CHEMCAD 6.0.1). The various Tray columns (sieve tray column, value tray column and Bubble cap tray column) and also packed column sizing are designed by CHEMCAD 6.0.1 Simulator.
- Before sizing the tower all the stream composition and property are specified then analysis on characteristics of feed and product streams and also tower are worked out in CHEMCAD 6.0.1. Costing for these different columns are prepared in CHEMCAD 6.0.1 Simulator and finally the cost comparison chart is carried out using SPREADSHEET.
- The project work has enhanced my fundamental knowledge based on " Gas Processing & Handling System Design " as well as " Simulation & Modeling " of M.Tech Gas Engineering.

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