ENHANCED VAPOR RECOVERY MODELING OF CONDENSATE TANK FLARING SYSTEM COUPLED WITH EXISTING VAPOR RECOVERY UNIT IN BIBIYANA GAS FIELD.

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APRIL 2012

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A Project

Submitted to the Department of Petroleum & Mineral Resources Engineering

in partial fulfillment of the requirements for the

Degree of Master of Engineering (Petroleum)

By

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CANDIDATE'S DECLARATION

It is hereby declared that this project or any part of it has not been submitted elsewhere for the award of any degree or diploma

Signature of the Candidate

………………………………

(Palash Khanti Saha)

RECOMMENDATION OF THE BOARD OF EXAMINERS

The undersigned certify that they have read and recommended to the department of Petroleum and Mineral Resources Engineering, for acceptance, a project entitled "ENHANCED VAPOR RECOVERY MODELING OF CONDENSATE TANK FLARING SYSTEM COUPLED WITH EXISTING VAPOR RECOVERY UNIT IN BIBIYANA GAS FIELD" submitted by PALASH KHANTI SAHA in partial fulfillment of the requirements for the degree of MASTER OF ENGINEERING in PETROLEUM ENGINEERING.

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ABSTRACT

The Bibiyana gas field is one of the most prolific gas fields in Bangladesh. It stared production in March, 2007 with 200 MMSCFD. Production has steadily increased from this field, and currently it is producing more than 730 MMSCFD. In addition, Bibiyana field also produces about 3,500 barrels of condensate per day. There are six storage tanks for condensate final stabilization and storage. Significant amount of vapor is produced in these tanks due to shrinkage/flash, standing and working effects. This vapor is hydrocarbon, i.e., natural gas, which has heating value and therefore valuable. There is a three-stage vapor recovery system in the field to capture the vapor produced at different stages of production and processing. However, vapor from the storage tanks cannot be recovered by the existing system because its pressure is too low- nearly atmospheric. It is therefore regularly flared through the Low Pressure Flare line.

Flaring of gas is a problem which entails both economical loss and environmental concerns. The hydrocarbon burning produces toxic gases, soot, acid rain, unburned hydrocarbons and a huge amount of CO_2, which contributes to greenhouse effect. The economical impact is the cost of gas that is flared.

This project demonstrates a method for recovering the low pressure vapor from the condensate tanks. This method uses a Gas ejector as a device to compress the low pressure natural gas from the condensate tanks to an intermediate pressure, which can be fed into the intermediated stage of the existing vapor recovery unit. Thus the natural gas will be saved which would have been otherwise flared. The amount of tank vapor is calculated by different methods, which shows a significant amount of gas which is now being flared.

Gas ejector is a device which converts pressure energy of a motive stream into kinetic energy which entrains a secondary stream and discharges the combined stream at an intermediate pressure. This project uses the relatively high pressure gas from the third stage of the existing vapor recovery unit as the motive gas, and the low pressure condensate tank vapor as the suction gas for ejector. The combined discharge stream will be fed into the vapor recovery unit's second stage.

Gas ejector has no moving parts. It involves almost maintenance free operation and no operating cost. It has been demonstrated as a viable technology for recovering condensate tank vapors in different fields in the world. This project describes the system design with the ejector technology, estimation of tank vapor recovered, and economical and environmental benefits. It is estimated that, on average 190 MSCFD tank vapor can be recovered. This should result in to yearly saving of about 68 MMSCF of natural gas. The equivalent to heat energy saving is about $74.55X10^9$ BTU. The simple payback period of this project covers in 7 months.

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NOMENCLATURE

BBLS	Barrels
BCF	Billion Cubic Feet
BPD	Barrels per Day
BTEX	Benzene, Toluene, Ethylbenzene, and Xylene
BTU	British Thermal Unit
EPA	Environmental Protection Agency
EVRU	Environmental Vapor Recovery Unit
GOR	Gas Oil Ratio
HAP	Hazardous Air Pollutants
HP	High Pressure
KO	Knock Out
LP	Low Pressure
MMSCFD	Million Standard Cubic Feet per Day
MSCFD	Thousand Standard Cubic Feet per Day
NGL	Natural Gas Liquid
PCS	Process Control System
PCV	Pressure Control Valve
PI	Pressure Indicating Gauge
PIT	Pressure Indicating Transmitter
PMRE	Petroleum and Mineral Resources Engineering
PSIA	Pounds per Square Inch in Absolute
PSIG	Pounds per Square Inch in Gauge
PSV	Pressure Safety Valve

PV	Pressure Regulating Valve
PVSV	Pressure/Vacuum Safety Valve
SCF	Standard Cubic Feet
SDV	Shutdown Valve
TI	Temperature Indicating Gauge
VRU	Vapor Recovery Unit

A	Area
C	Sound speed
Cp	Specific heat capacity at constant pressure
C_v	Specific heat capacity at constant volume
D	Diameter
ER	Entrainment ratio
γ	Specific heat ratio
h	Enthalpy
L	Length
M	Mach number
m	Mass flow rate
P	Pressure
$\overline{R}$	Gas constant
R	Universal gas constant
ρ	Density

T	Temperature
W	Molecular weight
V	Velocity
v	Specific volume

Subscripts

0	Stagnation condition
1	Ejector section 1
2	Ejector section 2
*	Critical flow
a	Flow station a
b	Flow station b
e	Exit
g	Gas phase
I	Isentropic
l	Liquid phase
m	Ejector mixed flow
p	Ejector primary stream
s	Ejector secondary stream
t	Nozzle throat

Chapter 1

INTRODUCTION

1.1 Overview of Bibiyana Gas Field

The Bibiyana gas field is located in northeastern Bangladesh in Block 12, 150 km northeast of Dhaka. Discovered in 1998, the field is one of the most significant natural gas discoveries in Bangladesh in both quality and size of the reserve. It started gas production in March, 2007 and field is continuously progressing and growing. With 12 producing wells, it currently produces more than 730 MMSCFD of natural gas and approximately 3500 BPD of condensate.

1.2 Condensate Storage System

In Bibiyana, there are 6 condensate storage tanks. The separated condensate from the 4 High-Pressure Inlet Separators is sent to the 2 Flash Gas Separators which operates approximately at 100 psig for a first stage of stabilization. Partially stabilized condensate is then sent to 2 LP Separators operating at 35 psig. Finally, it is sent to the 6 Condensate Storage Tanks for final stabilization at atmospheric pressure.

For condensate stabilization and storage, the major items of equipment are the Flash Gas Separators, the LP Gas Separators, and the Condensate Storage Tanks.

The Condensate Storage Tanks are 15,000-bbls each, and are fuel gas blanketed for vacuum protection. Condensate from the 4 HP Separators is routed to an 8-inch header via separate level controls. The Flash Gas Separators are pressure controlled at 100 psig, and the LP Gas Separators are pressure controlled at 35 psig. The Condensate tanks, where the pressure is maintained at a few ounces over atmospheric pressure.

The pressure in the Condensate Storage Tanks is maintained by allowing blanket gas into the tanks. A two step pressure regulator system allows fuel gas into the tanks when the pressure drops below 3 inches water. Blanketing gas first PCV steps down the pressure from 90 to 25 psig and then second PCV decreases the pressure from 25 psig to 1-3 inches water. If the pressure increases above 11 inches water, flare line first PCV vents vapors to the LP Flare and if the pressure increases further above 17 inches water, then second PCV vents additional vapors to the LP Flare. The condensate storage tanks are protected from over and under pressure by a combination of Pressure/Vacuum Safety Valves (PVSV) and Emergency Vents. The PVSV is set at 0.8 psig for overpressure and -1” water for under pressure. The Emergency Vent valve PSV protects the tank from fire and is set at 0.9 psig. [1]

1.3 Vapor Recovery Unit

There is a mechanical Vapor Recovery Unit (VRU) in Bibiyana to collect vapor from Flash gas separators, LP gas separators and LP gas boots.

The VRU consists of a 3 stage reciprocating Vapor Recovery Compressor that compresses Hydrocarbon vapors flashing during the depressuring of the high pressure condensate and water. The vapors are compressed back to 1270 psig to be recycled to the Inlet Gas Manifold (upstream of high pressure separator). The total design capacity of the VRU is 1.7 MMSCFD, but the first stage only handles 0.16 MMSCFD with the remaining gas entering the second stage suction. Each stage has a Suction Scrubber to ensure a dry gas feed, a reciprocating compressor stage (with suction and discharge pulsation dampeners), and a discharge cooler. The suction scrubbers remove entrained liquids to prevent compressor damage. The discharge coolers maintain vapor stream temperatures within operating limits.

The major items of equipment are the 3 stage Vapor Recovery Compressor, the 1st, 2nd, and 3rd Stage Suction Scrubbers, and the 1st, 2nd, and 3rd Stage Discharge Coolers.

Vapors from the 2 LP Gas Boots and the 2 LP Separators are combined in a common first stage suction header. This suction header pressure is controlled at 25 psig by a pressure control valve. The first stage compresses the hydrocarbon vapors to 95 psig before being cooled to 110 °F in the First Stage Discharge Cooler. Vapors that flash from the Flash Gas Separators (in excess of rich fuel gas requirements) are combined with the discharge of the first stage and compressed to 413 psig in the VRU second stage. The second stage discharge is cooled and then compressed to 1270 psig in the third stage. Some of the third stage discharge is spilled back to the second stage suction to maintain the second stage suction at 90 psig. [2]. A schematic diagram of the VRU and Tank connections is given in chapter 4 (figure 4.2).

1.4 Flare System

In Bibiyana south pad, there are two independent flare systems, HP (High Pressure) flare and LP (Low Pressure) flare system.

HP Flare system comprising HP Flare Knockout (KO) Drum, Well Area HP Flare KO Drum, flare stack, flare tip with pilot, ignition panel and the flare network. Pressure relief valves, blowdown valves and manual vent valves on the HP systems discharge into the HP Flare header.

LP Flare system includes a blower-assisted flare tip, LP Flare Knockout (KO) Drum with pumps, and the associated flare tip, pilot, ignition panel and pipe network. The LP Flare KO Drum receives vents from the Condensate Tanks and Produced Water Tanks as well as the Closed Drain System. The major items of equipment are the HP Flare, the LP Flare, a HP KO Drum and Pumps, a LP KO Drum and Pumps, and a (South Pad) Well Area HP Flare KO Drum and

Pumps. Both flare systems are continuously purged with fuel gas to prevent oxygen ingress. Normal operation is no flaring, except for small rates of pilot gas fuel and purge gas. [3]

In existing facility, vapor from the storage tanks is not being recovered by the VRU because its pressure is too low- nearly atmospheric. The Bibiyana gas plant is continuously progressing and growing and present total daily condensate production is about 3500 BPD. As a result significant amount of vapor is produced in these tanks due to shrinkage/flash, standing and working effects, which is eventually flared through LP flare.

1.5 Condensate Tank Vapor Generation

The amount of vapor created in condensate tanks is a function of oil throughput, gravity, and gas-oil separator pressure. The following three effects are mainly responsible for vapor generation in the storage tanks

- **Flash/Shrinkage effects:** It occurs when condensate is transferred from a gas –oil separator at higher pressure to a storage tank at atmospheric pressure. Upon injection of the condensate into the storage tanks, lower molecular weight hydrocarbons dissolved in the condensate come out of solution. This effect is known as flashing. The hydrocarbons that flash include methane, ethane, volatile organic compounds and hazardous air pollutants such as benzene, toluene, ethyl-benzene, and xylene.

- **Working effects:** It occurs when condensate level changes during production and shipping and when condensate in tank is agitated.

- **Standing effects:** Standing effects applicable with daily and seasonal temperature and barometric pressure changes.

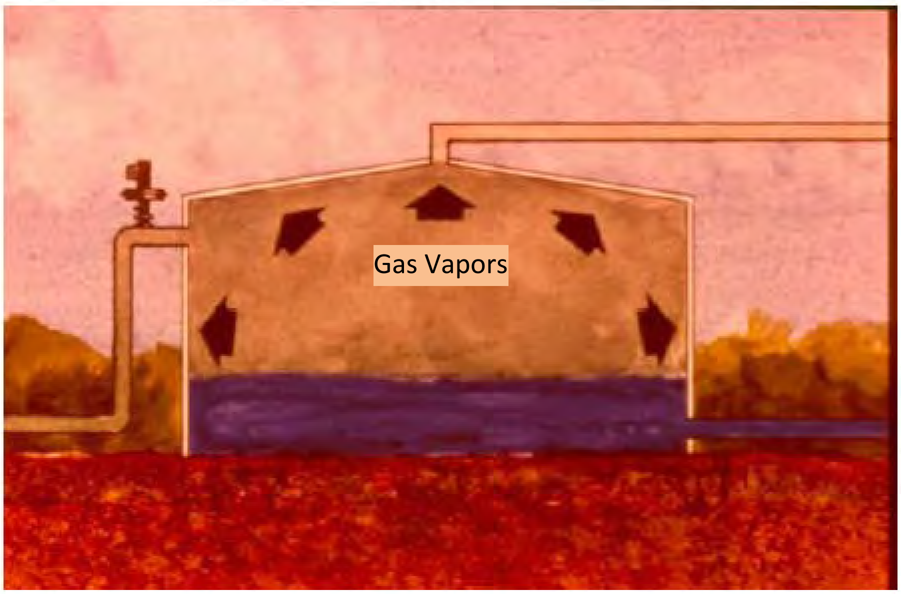


Figure 1.1: Tank Operations, as the condensate resides in the tanks, it gives off vapors,

The volume of gas vapor coming off a storage tank depends on also condensate properties. Lighter condensate (API gravity>36°) flash more hydrocarbon vapors than heavier condensate (API gravity<36°). In storage tanks where the condensate is frequently cycled and the overall throughput is high, more 'working vapors' will be released than in tanks with low throughput and where the condensate is held for longer periods. Finally, the operating temperature and pressure of the condensate in the vessel dumping into the storage tank will also affect the volume of flashed gases coming out of the condensate. [4]

The composition of these vapors varies, but the largest component is methane (between 40% and 60%). Other components include more complex hydrocarbon compounds such as propane, butane, and ethane; natural inert gases such as nitrogen and carbon dioxide; and HAP like benzene, toluene, ethyl-benzene, and xylene collectively these four HAP are referred to as BTEX. [5]

Vapor recovery system can recover over 95% of the hydrocarbon emissions that accumulate in storage tanks. Recovered vapors have a BTU content that is higher than that of pipeline quality natural gas. Depending on the volume of heavier hydrocarbon, NGLs in the vapors, the BTU content can reach as high as 2,000 BTU/ SCF. Therefore, on a volumetric basis, the recovered vapors can be more valuable than methane alone. [5]

Bibiyana condensate API gravity is 43.4 at 60^0 F as shown in chapter 3 (Table 3.2) which is lighter condensate (API gravity>36). Therefore condensate flashing and hydrocarbon vapors will be more compared to heavier condensate (API gravity<36). Condensate production and shipping tanks are frequently changed due to operations requirement that also release more working vapors.

1.6 Vapor Losses from Storage Tanks

In natural gas and oil production sector, according to Inventory of U.S. Greenhouse Gas Emissions and Sinks 1990-2005 over 15 years, data shows the following figures:

- Storage tanks are responsible for 4% of methane emissions in natural gas and oil production sector
- 96% of tank losses occur from tanks without vapor recovery

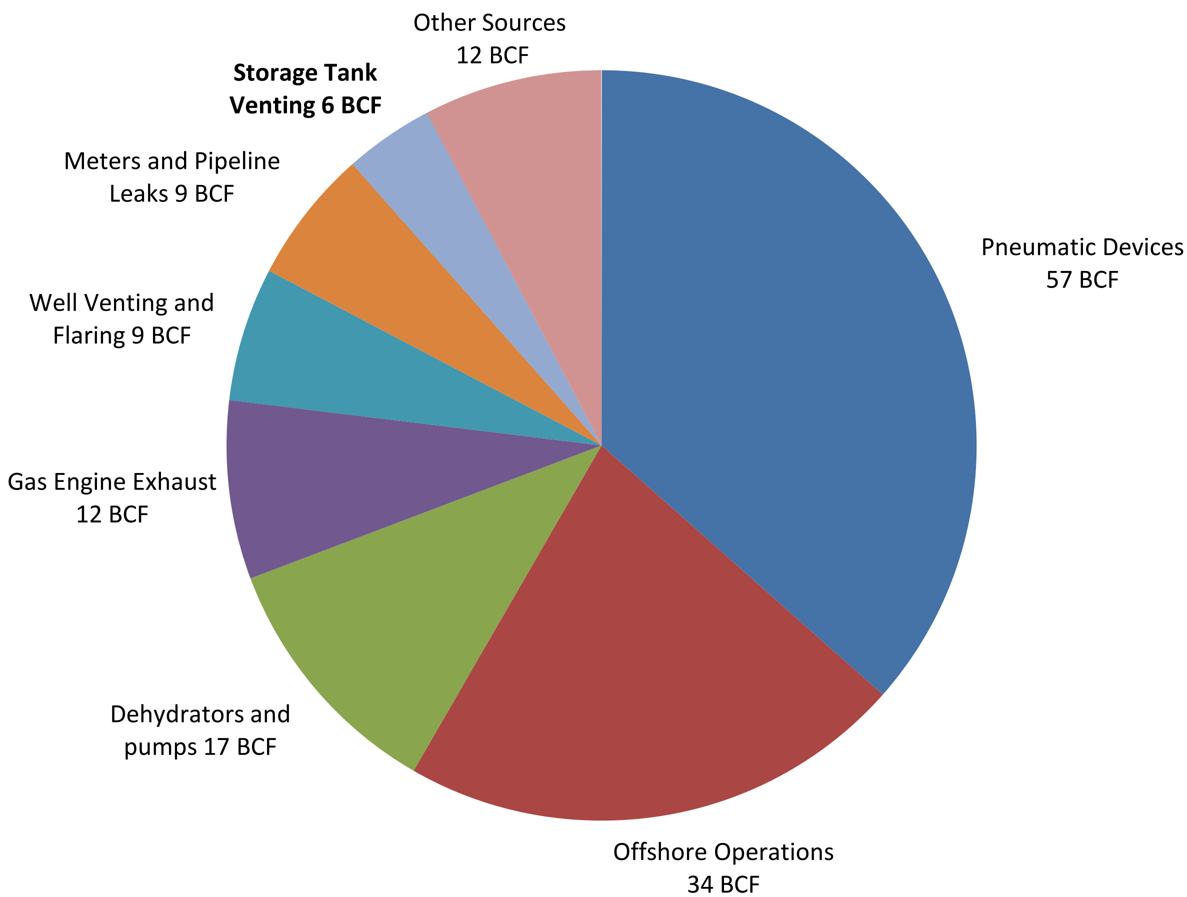


Figure 1.2: Greenhouse gas emissions in natural gas and oil production Sector, 1990-2005

1.7 Objectives with Specific Aims and Possible Outcomes

- To estimate the amount of tank vapor that is being flared in Bibiyana gas field
- To design a system to capture this vapor and minimize flaring
- To assess the environmental and economic benefits of recovering tank vapor instead of flaring

1.8 Outline of Methodology

- Study the existing Mechanical vapor recovery system in Bibiyana gas plant. This study includes present capacity, improvement of present capacity and accommodation of tank vapor in existing VRU
- Study ejector technology in regards of gas ejector
- Calculate the amount of gas flaring from condensate storage tanks
- Using HYSYS simulation to determine tank vapor and amount of liquid drop out in motive gas stream while pressure reduction
- Calculate CO_2 gas flaring emissions under Environmental concerns
- Calculate project cost analysis and simple payback period
- Estimate net heat energy saving
- Design of enhanced vapor recovery system which includes piping, controlling and safety system
- Performance study of improved system

Chapter 2

LITERATURE REVIEW

Utilization of low pressure gas like condensate tank vapors often bring both technical and economical challenges. The pressure of this gas is often considerably below that of the inlet of the compressors and as a result, in many cases, it has been flared as waste gas. A cost effective solution to some of these problems is the use of gas ejector which is in most cases of much lower cost than the alternative boosting systems.

The ejector was introduced as an engineering device in the early 20th century. Ejector is a pump-like device that uses the Venturi effect of a converging-diverging nozzle to convert the pressure energy of a motive fluid to velocity energy which creates a low pressure zone that draws in and entrains a suction fluid. After passing through the throat of the ejector, the mixed fluid expands and the velocity is reduced which results in recompressing the mixed fluids by converting velocity energy back into pressure energy.

2.1 Ejector Applications

Ejectors can be operated with incompressible fluids (liquids), and in this application are normally referred to as jet pumps or eductors. On the other hand, when ejectors are operated with compressible fluids (gases and vapors) the terms ejector and injector are generally employed. In gas production, ejectors are commonly referred to as Gas ejectors and in oil production, Jet pumps. A major difference between the two, besides the working fluid states, is the supersonic, choked flow nozzle of the gas ejector system. The supersonic approach allows a greater conversion of primary fluid energy to secondary fluid pressure head increase. However, this occurs with the penalty of considerable thermodynamic complexity in the mixing and diffusion sections. Fig. 2.1 and 2.2 are typical cross sectional views of liquid jet pumps and gas ejectors.

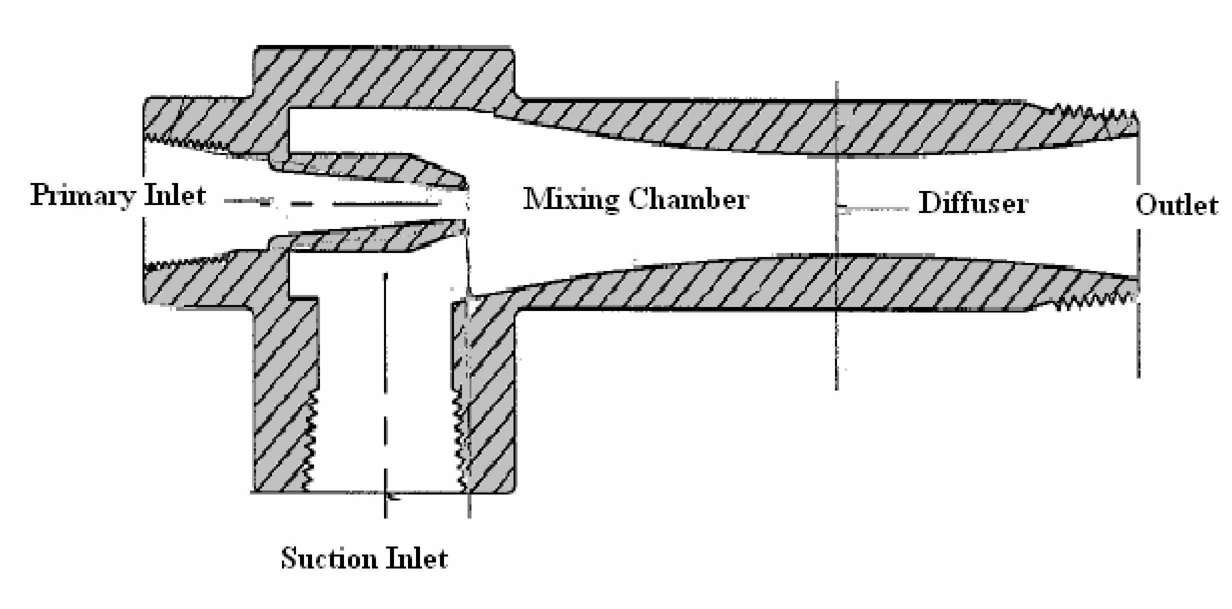


Figure 2.1: Cross sectional view of a typical liquid jet pump

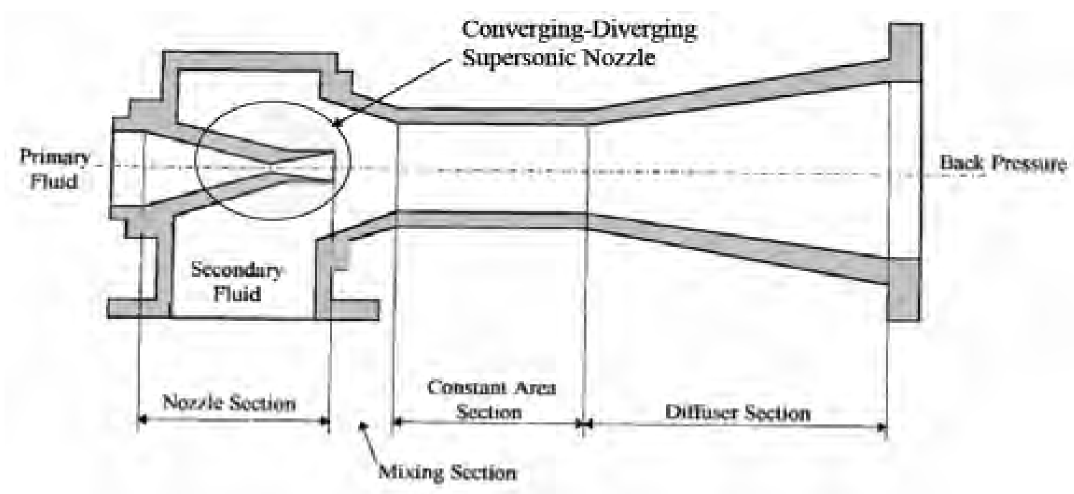


Figure 2.2: Cross sectional view of a typical gas ejector

The working process in a liquid jet pump or in a gas ejector is the same. Ejectors have found wide use in power plant, aerospace, propulsion and refrigeration application. Liquid Jet pumps have very good resistance to cavitations compared to other types of pumps. Thus, jet pumping may be an attractive method for waste heat transport in new generations of spacecraft. Gas ejectors are also found in high-altitude aircraft, thrust augmentation of aircraft and hydrogen fuel cells, etc. However, the research and development of jet pump technology for incompressible fluids (liquid) is more mature than that for ejectors used for compressible fluids (gases).[6]

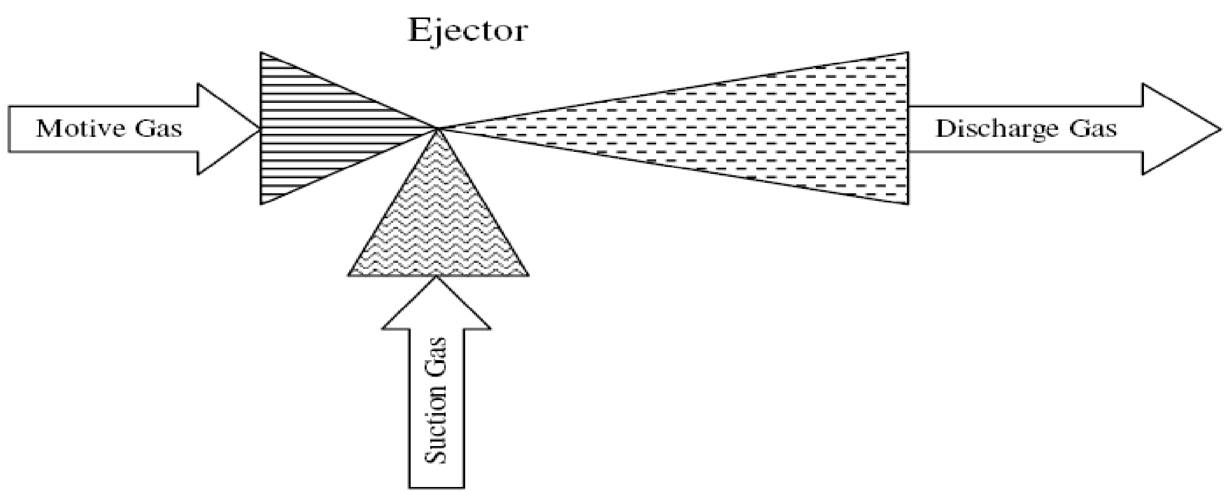


Figure-2.3: Schematic of a Gas Ejector

2.2 Gas Ejector Operations Principle

According to Bernoulli's equation, if no work is done on or by a flowing frictionless fluid, its energy due to pressure and velocity remains constant at all points along the streamline. As a result an increase of velocity is always accompanied by a decrease in pressure. This principle can be used to collect a low pressure natural gas stream with a high pressure motive gas stream for entrainment and compression to an intermediate pressure. Figure 2.4 describes the velocity-pressure changes in a gas ejector.

A gas ejector is a mixing and pressure increasing device which consists of a nozzle and venturi. The nozzle receives the motive fluid (e.g., natural gas) from a high pressure source. As the motive fluid passes through the jet, velocity increases and pressure decreases. The increased velocity, plus the decreased pressure, causes suction around the nozzle. Low pressure around the nozzle is drawn into the motive stream and mixed with it. The venturi consists of piping whose diameter narrows at the throat and then widens at its terminal end. This increased diameter at the terminal end causes the mixed fluids velocity to decrease and the pressure to increase. The venturi converts the high velocity jet stream into an intermediate pressure for delivery to a system for this intermediate pressure. [7]

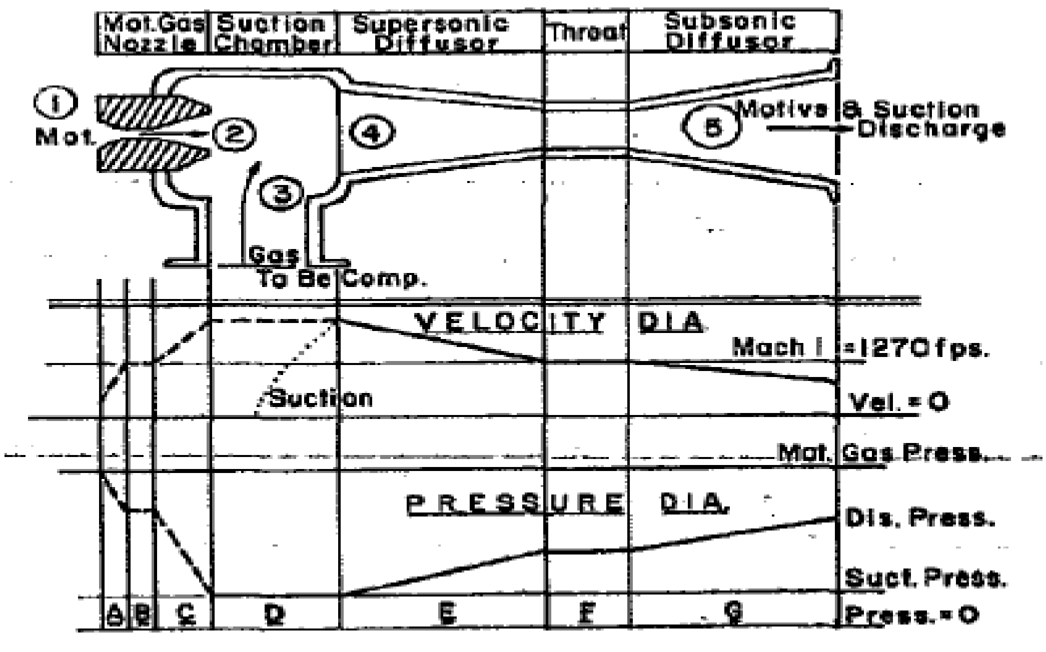


Figure 2.4: Velocity-Pressure changes in a Gas Ejector

It is not uncommon for the flow rate of the flare gas to vary and if not controlled, the suction pressure created by the Gas Ejector will also vary. In order to maintain the desired pressure on the low pressure side of the Gas Ejector, a number of standard control techniques are available. These include:-

- Recycling of gas from the discharge side of the Gas Ejector back into the low pressure side.
- Incorporation of an integral HP gas regulating assembly which varies the motive fluid consumed.

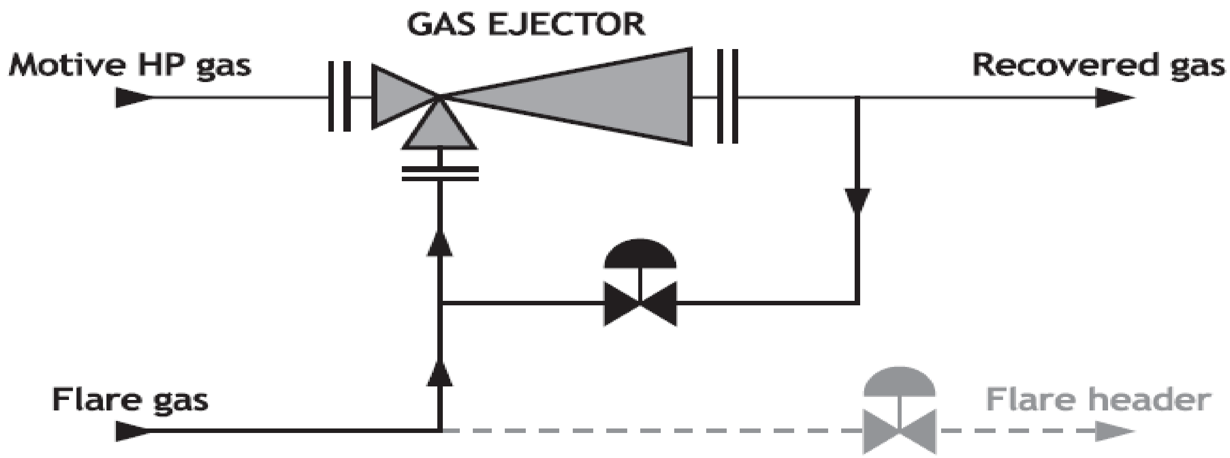


Figure 2.5: Ejector flare gas recovery system

2.3 Advantages of Gas ejector over Mechanical Compressor

The gas ejector system is an alternative to conventional vapor recovery technology for the recovery of hydrocarbon vapors from storage tanks. Mechanical vapor recovery system uses compressors and other moving parts. Ejector based vapor recovery or boosting low pressure gas is called the EVRU (Environmental Vapor Recovery unit) or vapor recovery of natural gas using non mechanical technology. This ejector technology gives following advantages over other boosting systems. [8]

1. No moving parts, therefore less wear and considerably less maintenance
2. Built in performance flexibility to suit changing well conditions
3. Significantly lower discharge temperatures
4. Can handle 2-phase flow and liquid slugs without mechanical damage
5. Tailor made to suit customers individual requirements
6. Very simple in operation
7. Very simple to commission with no specialist skills required

8. Suitable for remote operation
9. Less noisy
10. Can be manufactured from surface hardened materials for sand producing applications
11. Over-pressure protection devices not required
12. Very simple to install
13. Suitable for sub-sea installation
14. Does not need consumables to operate
15. Spares holding costs are significantly less
16. Significantly cheaper project costs
17. Significantly shorter payback times

Since Ejector system offers above advantages, compared with other boosting systems, it was chosen for recovering condensate vapors. But gas ejector also has some limitations. The main disadvantages are due to the lack of reliable design methods and to the sharp decrease of the performance when operating conditions change. In fact when the ejector is fed with multiphase fluids, significant modeling problems arise both on the physical side and in the numerical solution due to the complexity of the fluid dynamics. On the other hand, the efficiency of an ejector is very sensitive to fluid properties, which rapidly change when the operative conditions change from the design conditions. [9]. In this project, motive and suction gas used for ejector is almost dry and operating conditions are also nearly constant. Therefore, liquid and operating conditions would not create any problem for ejector performance. The effect of liquid in gas phase and ejector design boundary conditions have been discussed in chapter 4.

Chapter 3

CONDENSATE TANK VAPOR ESTIMATION

Calculating total vapor emissions from oil tanks is complicated because many factors affect the amount of gas that will be released from a crude oil tank, including:

- Operating pressure and temperature of the separator dumping the oil to the tank and the pressure in the tank
- Oil composition and API gravity
- Tank operating characteristics (e.g., sales flow rates, size of tank), and
- Ambient temperatures.

In this project, three methods were used to estimate the quantity of vapor emissions from condensate tanks:

3.1 Condensate Shrinkage Volume approach/Flash calculations

3.2 API Gravity and separator pressure to determine Gas-Oil Ratio(GOR)

3.3 HYSYS simulation approach

3.1. Condensate Shrinkage Volume Approach/Flash Calculations

Condensate shrinkage when condensate is transferred from a gas –oil separator at higher pressure to a storage tank at atmospheric pressure. Upon injection of the oil into the storage tanks, lower molecular weight hydrocarbons dissolved in the crude oil come out of solutions or “flashes” when the pressure drops as the oil enters the tank. The hydrocarbons that flash include methane, ethane, volatile organic compounds and hazardous air pollutants.

Condensate shrinkage volume by pressure reduction at different stages is shown in the following Table. Condensate from High-Pressure Inlet Separator is sent to Flash Gas Separator which operates approximately at 100 psig for a first stage of stabilization. Partially stabilized condensate is then sent to LP Separator operating at about 35 psig and, finally, to the Condensate Storage Tanks for final stabilization at atmospheric pressure.

Table 3.1 and Table 3.2 data found from Bibiyana gas plant laboratory for condensate analysis. For shrinkage result, Shrinkage Tester/Machine was used at sample point of high pressure separator downstream condensate. And instrument of auto system simulated distillation GC were used for sales condensate to obtain physical properties.

Table-3.1: Condensate Shrinkage Report

Location	Pressure Changes	Shrinkage Volume %
HP Separator	1220 psi to 100 psi	9.4
	100 psi to 35 psi	0.8
	35 psi to Ambient Pressure at 82.0 ^{0}F	1.0

Table-3.2: Condensate Physical Properties

Parameters	Results	Test Methods
API Gravity	43.4 at 60 ^{0}F	ASTM D 1298 - 90
Specific Gravity	0.809034 at 60 ^{0}F	ASTM D 1298 - 90
Reid Vapor Pressure	6.6 psia	ASTM D 323 - 94
Basic Sediment and Water	0 %	ASTM D 4007 - 95

Calculations:

From table 3.1 & 3.2,

Condensate shrinkage factor at Flash Gas Separator = 0.094

Condensate shrinkage factor at LP Gas Separator = 0.008

Condensate shrinkage factor at Condensate tank = 0.01

Gas Specific Gravity = 0.61028

Condensate Specific Gravity = 0.809034

Gross Heating value of Gas = 1075 BTU/SCF

The conversion factor of gas volume to liquid amount, i.e. Million Standard Cubic Feet gas to amount of condensate, Barrel [10]

3178 (MMSCFD)* S.G. of Gas = 14.6 (BBLD)* S.G of Condensate

MMSCF = {14.6 * S.G. of Condensate} * Number of BBL / {3178 * S.G. of Gas}

= {14.6 * 0.809034}/ {3178 * 0.61028 } Number of BBL

MMSCF = 0.00609 * Number of BBL ……………………………………….. (3.1)

Let,

Bibiyana total Condensate Production = A

Condensate from LP Gas Separator = L

Condensate from Flash Gas Separator = F

Condensate from HP Separator = H

Bibiyana total Condensate Production, A

= Condensate from LP gas separator - Shrinkage volume at Condensate tank

= L - 0.01L [from table 3.1, shrinkage volume 1% from LP sep. to tank]

= 0.99L

Condensate coming from LP Gas Separator, L = A/0.99

Condensate shrinkage at Storage Tank, = 0.01L

= 0.01*A / 0.99 BBL

= 0.010101A BBL

From Appendix-I, average condensate Production, A= 3510 BPD

Condensate Shrinkage volume at Storage Tank,

= 0.010101*3510

= 35.45 BPD

= (0.00609*35.45) MMSCFD(from equation 3.1)

= 0.2158 MMSCFD

= 215.8 MSCFD

The estimated Gas from condensate storage tanks is 215.8 MSCFD

3. 2 Estimated Tank Vapors using API Gravity and Separator Pressure

The volume of gas vapor coming off a storage tank depends on also crude oil properties. Lighter crude oils (API gravity>36°) flash more hydrocarbon vapors than heavier crudes (API gravity<36°).Condensate storage tanks are located downstream of LP Separators. LP Gas Separators are pressure controlled at about 35 psig. The condensate outlet from the LP Gas Separators is directed to the Condensate Storage Tanks for the final step of stabilization, where the pressure is maintained at a few ounces over atmospheric pressure.

Figure 3.1 analyzes API gravity and separator pressure to determine GOR. These curves were constructed using empirical flash data from laboratory studies and field measurements for Natural gas, EPA pollution preventer in USA. As illustrated, this graph can be used to approximate total potential vapor emissions from a barrel of condensate. For example, given a certain condensate API gravity (e.g., 38°) and vessel dumping pressure (e.g., 40 psi), the total volume of vapors can be estimated per barrel of condensate (e.g., 43 scf per bbl). Once the rate per barrel is estimated, the total quantity of vapors from the tank can be determined by multiplying the per barrel estimate by the total amount of oil cycled through the tank.

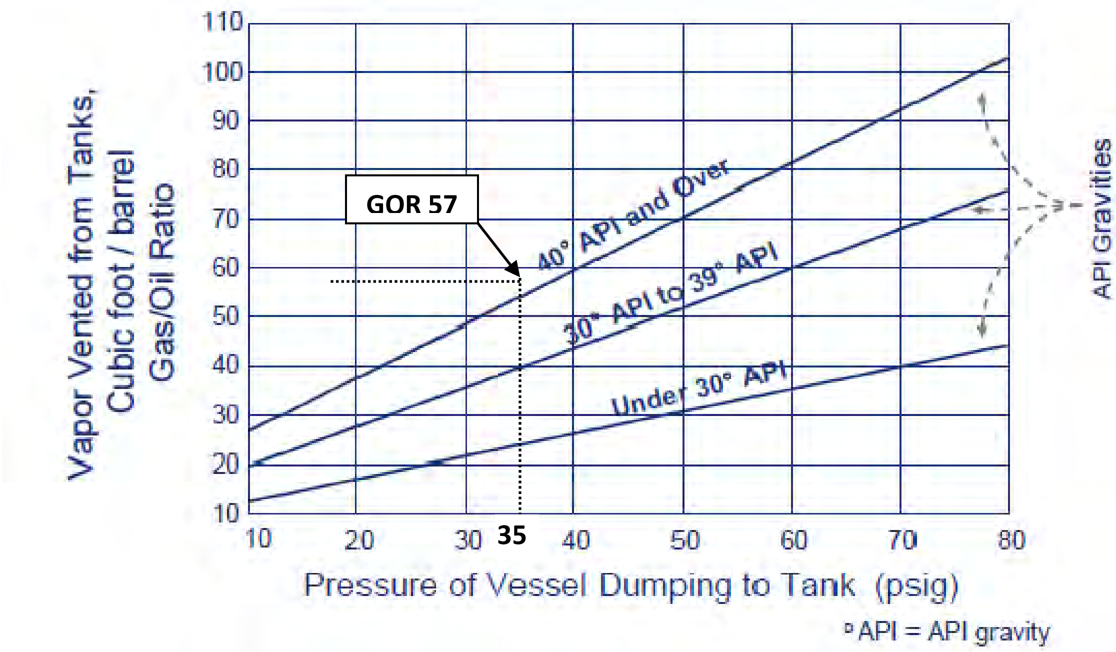


Figure-3.1: Calculating GOR from API chart

Calculations:

Condensate API gravity from Table-3.2 found 43.4

Pressure of vessel dumping to tank (LP Separator, upstream of tank) is 35 psig

From figure-3.1, Gas/Oil ratio (GOR) found 57 SCF/barrel

From Appendix-A, average condensate Production, 3510 BPD

Vapor from tanks = GOR x No. of barrels condensate production

= 57 x 3510

= 199500 SCF/day

= 199.5 MSCFD

The estimated Gas from condensate storage tanks is 199.5 MSCFD

3.3 HYSYS Simulation Approach

HYSYS, an engineering simulation tool, is widely used in universities and industries for research, development, modeling and design. HYSYS serves as the engineering platform for modeling processes in Gas processing, Refining and Chemical processes. Here HYSYS was used to find the vapor outlet streams from condensate storage tanks due to flashing and pressure drop between LP gas separators and condensate tanks. In steady state mode, the separator divides the vessel contents into its constituent vapor and liquid phases. The vapor and liquid in the vessel are allowed to reach equilibrium, before they are separated.

ASPEN^{TM} HYSYS software is widely accepted and used for oil and gas industries. The whole simulation study and analysis was done on ASPEN^{TM} HYSYS 7.1.For simulation, Fluid package was selected to be Peng-Robinson. The main reason behind this, for oil, gas and petrochemical applications, the Peng-Robinson is generally the recommended and widely accepted property package. ASPEN HYSYS contains an oil manager which organizes the data and HYSYS properties were used for property generation of the streams. [11].

Figure 3.2 is here to show the condensate tank inlet and outlet connections. Main purpose is to estimate outlet vapor through HYSYS simulation.

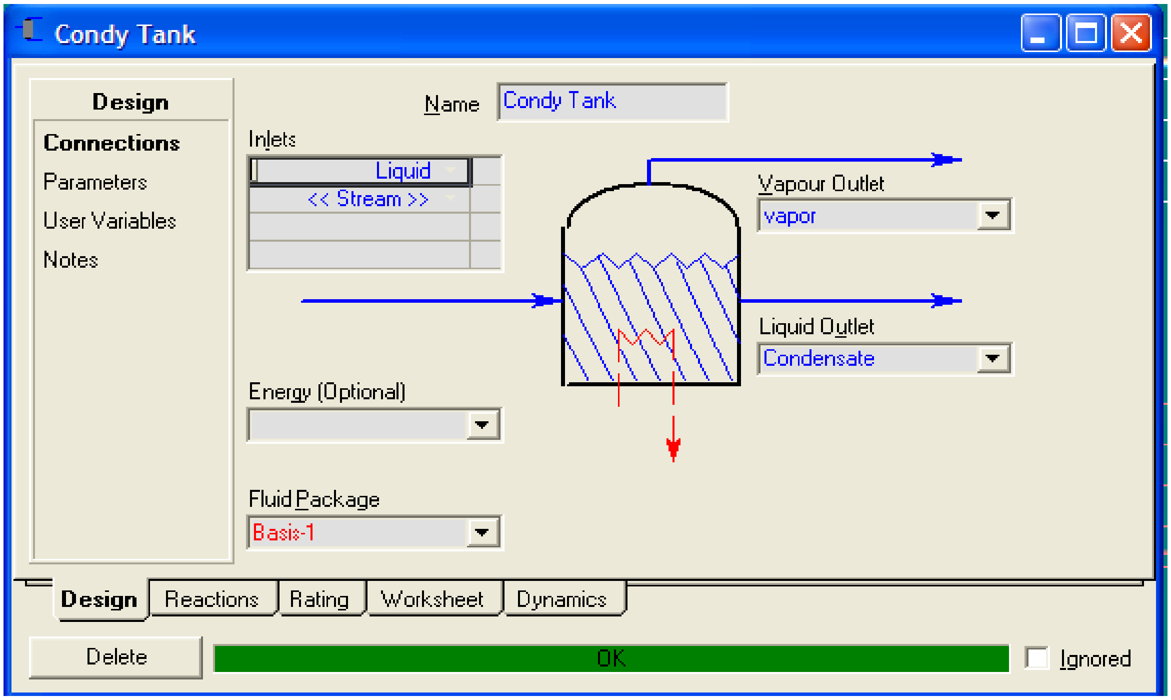


Figure-3.2: Condensate tank inlet and outlet streams

Control valve (Pressure reducing element) is located upstream of condensate tank is shown in figure 3.3 in which control valve inlet stream pressure is 35 psig and outlet stream pressure is near about atmospheric

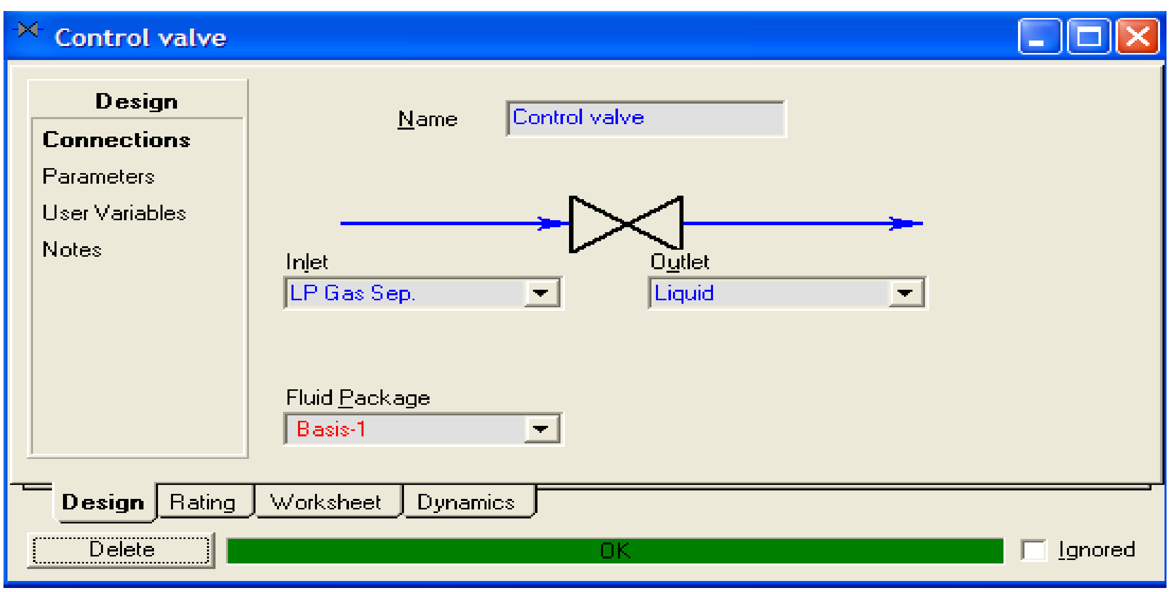
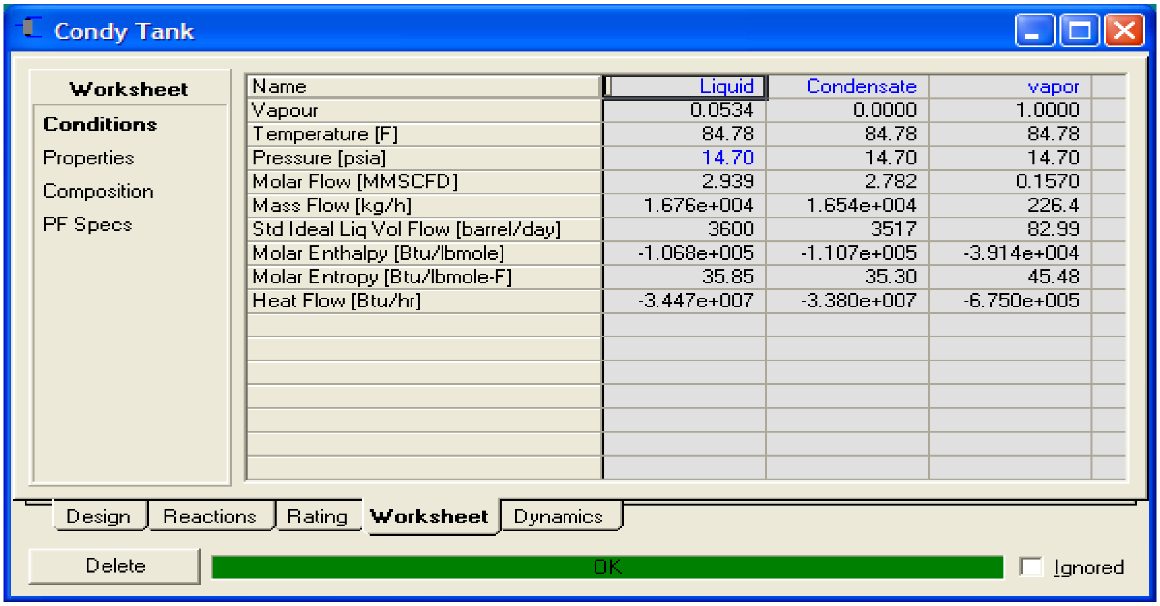


Figure-3.3: Condensate tank upstream pressure reducing control element

Figure 3.4 describes the condition of different parameters for inlet and outlet streams of condensate tank. Here input given for pressure, temperature and flow rate that specify remaining different parameters.


Condy Tank

Worksheet
Conditions
Properties
Composition
PF Specs

Name	Liquid	Condensate	vapor
Vapour	0.0534	0.0000	1.0000
Temperature [F]	84.78	84.78	84.78
Pressure [psia]	14.70	14.70	14.70
Molar Flow [MMSCFD]	2.939	2.782	0.1570
Mass Flow [kg/h]	1.676e+004	1.654e+004	226.4
Std Ideal Liq Vol Flow [barrel/day]	3600	3517	82.99
Molar Enthalpy [Btu/lbmole]	-1.068e+005	-1.107e+005	-3.914e+004
Molar Entropy [Btu/lbmole-F]	35.85	35.30	45.48
Heat Flow [Btu/hr]	-3.447e+007	-3.380e+007	-6.750e+005

Design | Reactions | Rating | Worksheet | Dynamics

Delete | OK | Ignored

Figure-3.4: Material stream worksheet that specify variables of different streams

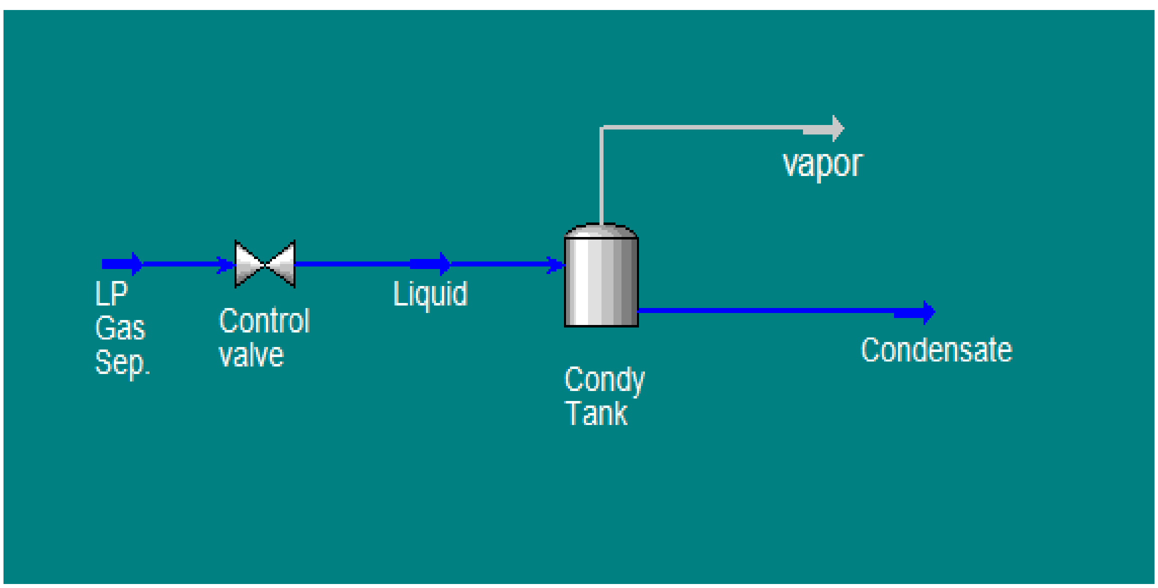
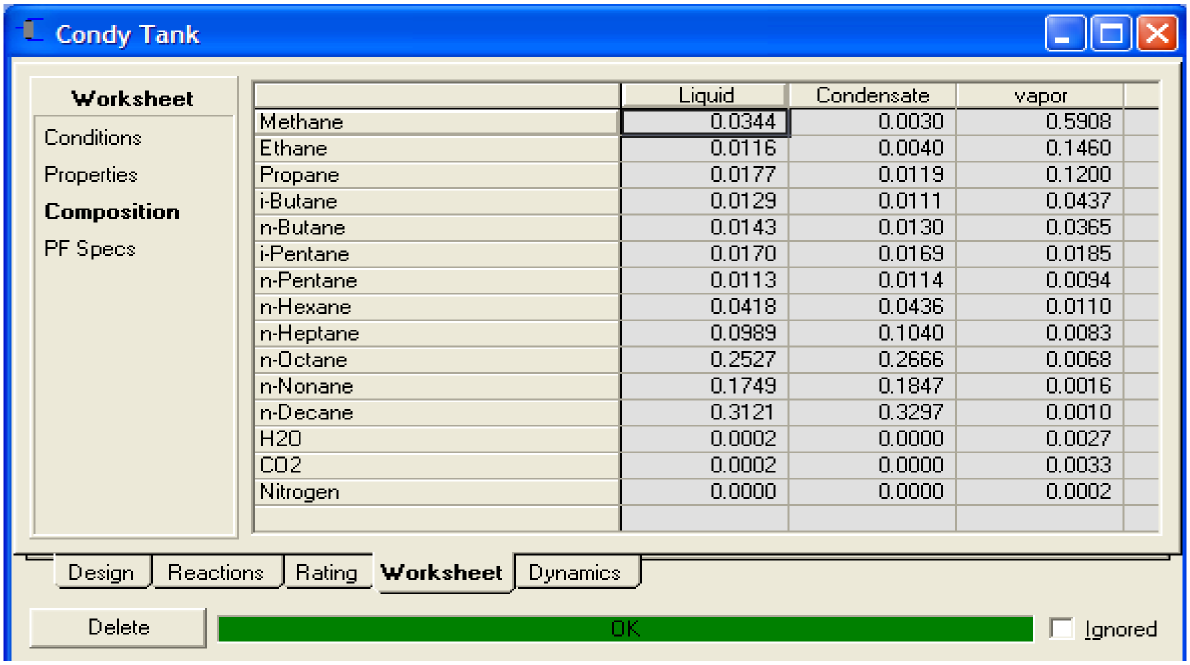


Figure-3.5: A snapshot of HYSYS environment

In figure 3.6, Condensate tank inlet liquid composition is given in mole fraction and HYSYS calculates outlet condensate and vapor composition. Inlet liquid (condensate) compositions are find out by HYSYS calculation of sales condensate analysis from Bibiyana gas plant laboratory.



	Liquid	Condensate	vapor
Methane	0.0344	0.0030	0.5908
Ethane	0.0116	0.0040	0.1460
Propane	0.0177	0.0119	0.1200
i-Butane	0.0129	0.0111	0.0437
n-Butane	0.0143	0.0130	0.0365
i-Pentane	0.0170	0.0169	0.0185
n-Pentane	0.0113	0.0114	0.0094
n-Hexane	0.0418	0.0436	0.0110
n-Heptane	0.0989	0.1040	0.0083
n-Octane	0.2527	0.2666	0.0068
n-Nonane	0.1749	0.1847	0.0016
n-Decane	0.3121	0.3297	0.0010
H2O	0.0002	0.0000	0.0027
CO2	0.0002	0.0000	0.0033
Nitrogen	0.0000	0.0000	0.0002

Figure-3.6: HYSYS exhibits tank inlet-outlet streams composition

Figure 3.7 exhibits vapor estimation with inlet and outlet streams different process parameters. HYSYS results detailed datasheet are provided in Appendix-C.

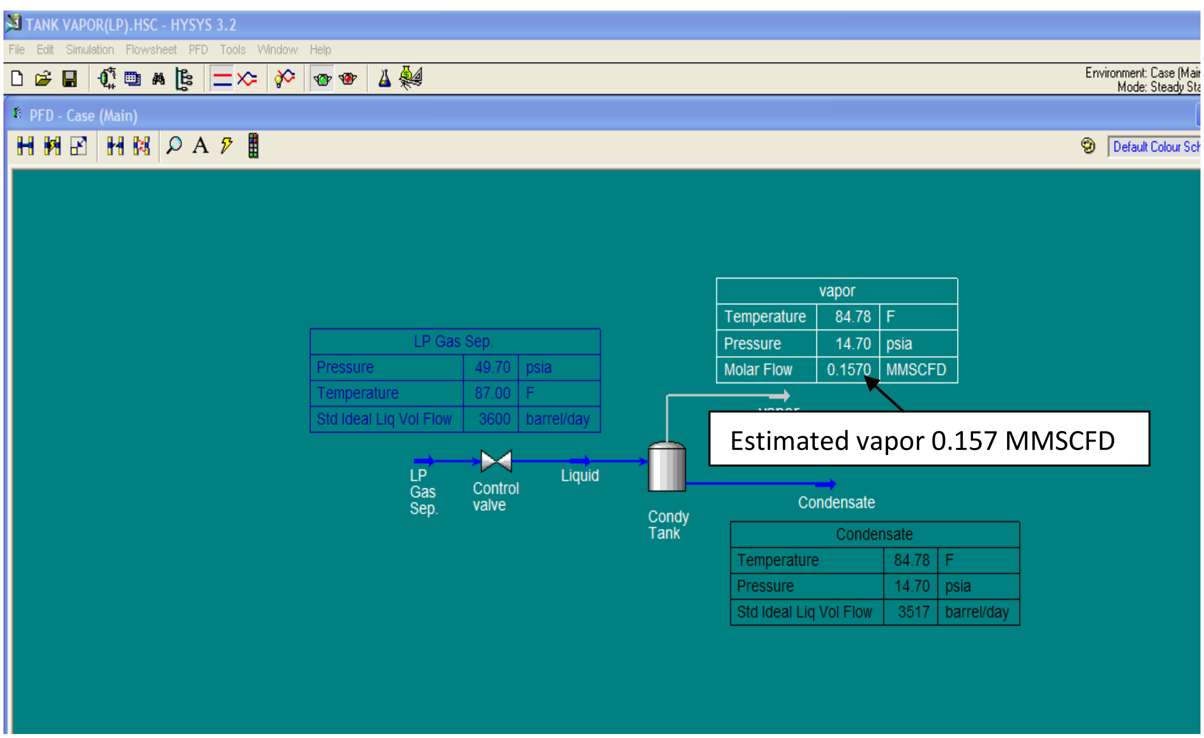


Figure-3.7: A snapshot of condensate tank vapor estimation

3.4 Tank Vapor Estimation Results

The outcome of condensate tank vapor calculation by three different methods is tabulated here

Table-3.3: Condensate tank Vapor calculation results

No.	Method	Estimated Vapor (MMSCFD)
3.1	Condensate Shrinkage Volume approach	0.215
3.2	API Gravity and separator pressure to determine Gas-Oil Ratio(GOR)	0.199
3.3	HYSYS Simulation approach	0.157
Average value		0.190

Estimated vapor from condensate storage tanks is 157 to 215 MSCFD range with an average value considering of 190 MSCFD. But vapor flow rate depends on condensate daily production. In Appendix-A, condensate daily production data has been shown. Condensate production varies with swing gas production. In table 3.3, estimated gas on HYSYS simulation method shows some inconsistent compare to other two methods. For HYSYS simulation, condensate upstream composition data were used from simulated result of bibiyana sales condensate analysis. Tank upstream condensate composition lab analysis was not being performed in location due to technical problem. Therefore condensate composition input data in HYSYS might be some differ from original value.

Chapter 4

DESIGNING THE VAPOR RECOVERY SYSTEM

4.1. Existing VRU Capacity Analysis

Before starting a modification, it is necessary to check out the existing mechanical Vapor Recovery Unit (VRU) present production rate and total production capacity as manufacturer recommended. Figure 4.1 exhibits the VRU monthly average production which is below 1 MMSCFD. VRU collects vapor from Flash gas separators, LP gas separators and LP gas boots. It consists of a 3 stage reciprocating Compressor. The VRU however, cannot recover vapor from the condensate storage tanks because of low pressure of vapor (nearly atmospheric).

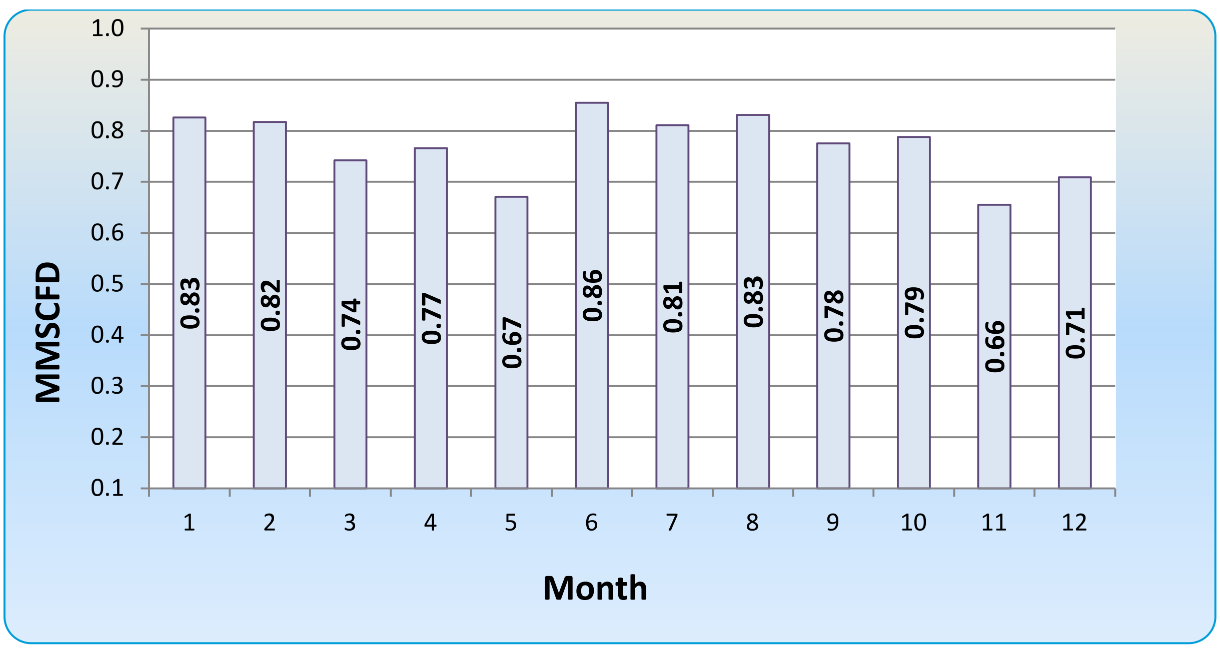


Figure 4.1: VRU Monthly average production [MMSCFD]

Table 4.1: Vapor Recovery Compressor capacity

Manufacturer	Dresser-Rand
Model	6 HOS 4
Cylinders	4 throw 3 cylinder, max 1200 rpm
1st Stage / 2nd Stage/ 3rd Stage	1st Stage / 2nd Stage/ 3rd Stage
Suction, Discharge Pressure	25,96/ 89,410/ 408,1217 psig
Suction, Discharge Temperature	64,199/ 72,259/ 115,271 °F
Flow Rate capacity	**0.16/ 1.67/ 1.67 MMSCFD**
Compressor Engine Driver	495 HP@ 800 rpm , 1200 rpm max
Driver Manufacturer, Model	Waukesha F3514 GSI ESM
Driver Cylinders, Displacement	6, 3520 in^3 total

The total design capacity of the existing VRU is 1.67 MMSCFD. According to VRU daily average production that was mentioned in Appendix A, it presently handles only average 0.811 MMSCFD [50% of capacity]. The VRU 1st stage, 2nd and 3rd stage capacity is 0.16, 1.67, 1.67 MMSCFD respectively. In this case Second and Third stage have almost 50% capacity left to use; so there is an opportunity to feed more gas. According to calculation, shown in Chapter-3, section 3.4, estimated gas from condensate tanks is on average 0.190 MMSCFD and peak rate is 0.215 MMSCFD. If maximum vapor recovery is considered then VRU daily production would reach 1.065 MMSCFD that is 64% of total rated capacity. In this case, configurations of VRU compressor will not be required to change. Rather, added gas will increase compressor efficiency. In this project, Ejector discharge gas will be feed into VRU second stage inlet at approximately 100 psig.

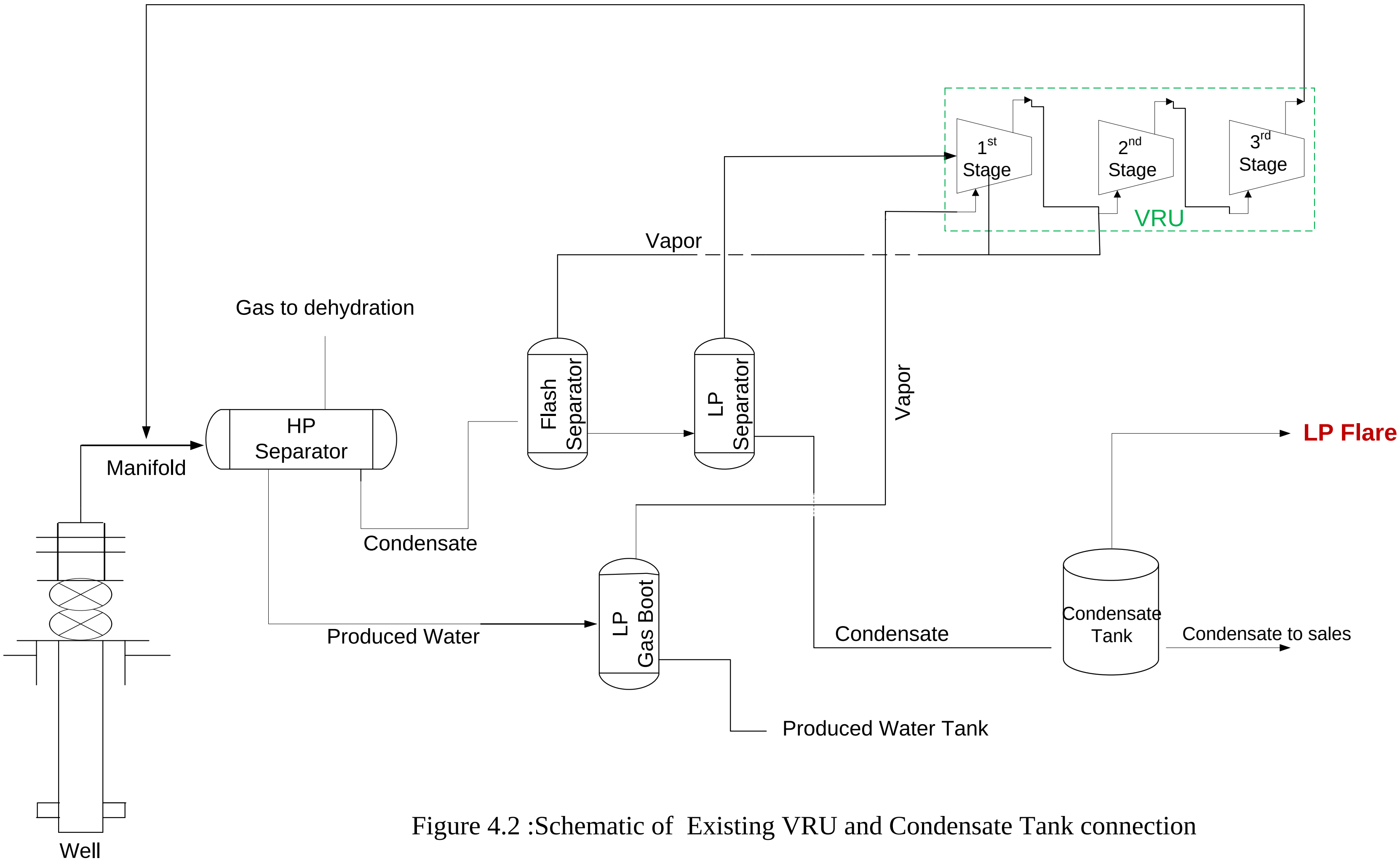


Figure 4.2 :Schematic of Existing VRU and Condensate Tank connection

4.2. Ejector Design Considerations

For the sake of simplicity in designing an ejector, the following assumptions were considered:

a) Flow is frictionless (viscosity=0)

b) Expansion in the nozzle and compression in the diffuser are isentropic, with *k* (ratio of specific heats, dimensionless) remaining constant

c) Motive fluid properties are the same as the suction fluid properties

d) Mixture occurs at constant pressure and is adiabatic

e) Total enthalpy of the final mixture equals the weighted average of the initial components

Information required for the gas ejector design included:

- Operating pressure of low pressure system (e.g. pressure range of source gas, the condensate tank vapor pressure in this case)
- Estimated amount of tank vapor that can be captured by ejector
- Source of high pressure gas (e.g. motive gas which comes from VRU third stage discharge)
- Operating pressure of high pressure gas system (e.g. motive gas pressure range, here VRU third stage gas pressure)
- Fluid properties of Suction and Motive gas (e.g. here assumed that Motive fluid properties are the same as the suction fluid properties)
- Present and future estimated volume of gas
- Amount of spare horsepower available to booster compressor for compressing intermediate pressure gas from venturi jet ejector
- Tank dimensions, tank operating pressure, location and dimensions of piping
- Volume of gas recovered by existing mechanical vapor recovery

4.3 Key Design Parameters

The main ejector design parameters are the nozzle section and the mixing tube length. The nozzle controls the drive energy and the mixing tube controls the compression rate.

With the new technology, the nozzle section can be controlled and a little bit modified by an external valve. The valve stem transmits an axial motion to the tapper pin positioned inside the nozzle, by changing the pin position it is possible to regulate the nozzle flow area.

The compression ratio of the ejector is defined as ratio of the ejector's outlet pressure P_{Out} to the inlet pressure of the $P_{Suction}$. The compression ratio depends on the ejector geometry and on the motive and suction pressure ratio.

The entrainment ratio of the ejector is defined as the amount of motive gas W_m required to entrain and compress a given amount Ws of suction gas.

The compression ratio and the entrainment ratio are key parameters in designing an ejector

Thus for prediction of the ejector performances the others input data have to be taken into account, those are the dimensions of the ejector, the flow rates, composition, pressure and temperature of the motive and suction fluid. [12]

$$\text{Compression ratio} = \frac{Discharge\ gas\ pressure}{Suction\ gas\ pressure}$$

$$= \frac{(90+14.7)\ PSIA}{(0.6+14.7)PSIA}$$

$$= 6.8$$

$$\text{Entrainment ratio} = \frac{Motive\ \ gas\ flowrate}{Suction\ gas\ flowrate}$$

$$= \frac{(440\sim590)MSCFD}{(157\sim215)\ MSCFD}$$

$$= 2.8$$

According to ejector manufacturer, COMM Engineering, USA, Motive Volume needs to be approximately 3 times than the Suction Volume to be picked up, depending on pressure. Hence, higher motive gas pressure is used (approx 750 psig), the higher the pressure of the motive gas, the more beneficial would the ejector be in generating the required level of boost for the suction gas, using a minimum amount of HP gas. A further benefit of using HP gas from sources such as outlet of a compressor is that this pressure is relatively high and remains unchanged with less fluctuation. In this project, entrainment ratio is estimated to be about 2.8 that would support the optimum ejector operations, since motive stream flows relatively at higher pressure. Compression ratio is to be 6.8 with low suction pressure. It is possible to achieve high compression ratio with a supersonic nozzle for Ejector. [13]

4.4 Supersonic Nozzle and Subsonic Diffuser

A brief discussion on the theory of flow through nozzle and diffuser is presented in this section. To achieve better performance, modern gas ejectors are normally operated in a supersonic condition at the exit of primary nozzle. Therefore, it is necessary to introduce the choking phenomena which occur at the throat of the primary nozzle. The basic idea of a gas ejector is to accelerate the motive flow to supersonic by a converging-diverging nozzle, primary flow exit at the suction chamber where secondary flow is induced by this high-velocity, depressurized flow. In most cases, there is also a diffuser installed at the exit of the mixing section to induce pressure recovery.

The conservation equations and ideal gas law for steady one dimensional (1-D) compressible flow in an arbitrary variable-area control volume as sketched in Fig.4.3. The definitions of termininologies can be found in the nomenclature section.

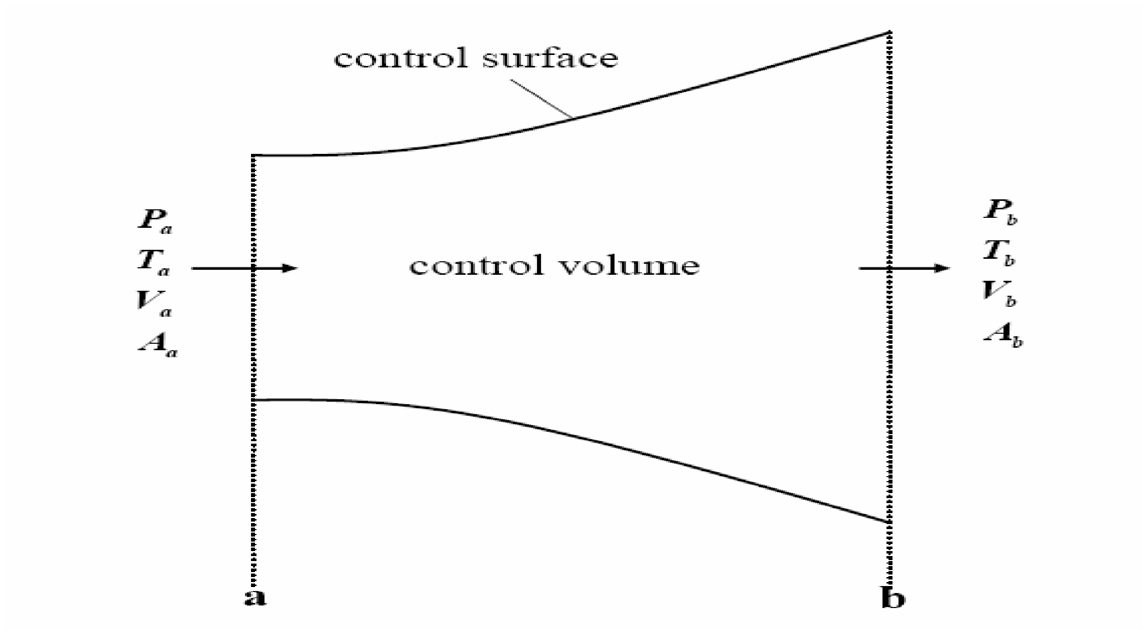


Figure 4.3: Control volume for 1-D flow through converging-diverging nozzle

Continuity equation

$$m = \rho_a V_a A_a = \rho_b V_b A_b \tag{4.1}$$

Momentum equation

$$P_a A_a + m_a V_a + \int_{Aa}^{Ab} P dA = P_b A_b + m_b V_b \tag{4.2}$$

Energy equation

$$h_a + \frac{V_a^2}{2} = h_b + \frac{V_b^2}{2} \tag{4.3}$$

Ideal gas law

$$\frac{P}{\rho} = RT \tag{4.4}$$

Where R is the gas constant with unit of $J\ (kg.K)$. R is related to its molecular weight

by following equation:

$$R=\frac{\bar{R}}{W} \tag{4.5}$$

In the above equation, $\bar{R}$ is the universal gas constant with unit of J (kmol.K) and W is

the molecular weight with unit of kg (kmol).

4.4.1 Mach Number

Mach number, M , is a very important dimensionless parameter for compressible flow, specially, for supersonic flow. Mach number is defined as the ratio of the fluid velocity to the local sonic speed.

$$M=\frac{\text{local fluid velocity}}{\text{local sonic speed}}=\frac{V}{C} \tag{4.6}$$

The local sound speed C in a medium with temperature T is given by:

$$C=\sqrt{\gamma R T} \tag{4.7}$$

4.4.2 Isentropic Expansion

Equation (4.8) is the process equation for isentropic flow of an ideal gas:

$$=\frac{P}{\rho^{\gamma}}=\text{Constant} \tag{4.8}$$

The flow basic equations: continuity, momentum, energy, second law, equation of state, as well as the above-mentioned process equation, the local pressure, temperature and density can be related with their corresponding values at stagnation condition by isentropic flow functions expressed in Equations (4.9) through (4.11). The parameters with subscript 0 refer to the stagnation properties. Stagnation properties are constant throughout a steady, isentropic flow field.

Pressure: $$\frac{P_0}{P} = \left(1+ \frac{\gamma-1}{2} M^2\right)^{\frac{\gamma}{\gamma-1}} \quad (4.9)$$

Temperature: $$\frac{T_0}{T} = 1+ \frac{\gamma-1}{2} M^2 \quad (4.10)$$

Density: $$\frac{\rho_0}{\rho} = \left(1+ \frac{\gamma-1}{2} M^2\right)^{\frac{\gamma}{\gamma-1}} \quad (4.11)$$

4.4.3 Choking Phenomena

To explain the choking phenomena, a convergent-divergent nozzle with its static pressure distribution along the flow direction are shown in Fig. 4.4 [14]. Flow through the Converging-diverging nozzle of Fig. 4.4 is induced by an adjustable lower downstream pressure; upstream supply is constant and stagnation conditions with $V_0 \cong 0$. Pe and P_b represent the static pressure at the nozzle exit plane and the back pressure, respectively. Fig. 4.4 illustrates graphically the effect of variations in back pressure P_b on the pressure distribution through the nozzle.

The flow rate is low when back pressure P_b is slightly less than P_0 ; curve i shows the pressure distribution in the nozzle for this case. If the flow rate is low enough, the flow will be subsonic and essentially incompressible (if $M < 0.3$) at all points on this curve. Under this condition, the nozzle will behave as a venturi, with flow accelerating in the converging portion until a point of maximum velocity and minimum pressure is reached at the throat, then decelerating in the diverging portion to the nozzle exit. When the back pressure is reduced further, the flow rate increases but, is still subsonic everywhere and the pressure distribution is shown as curve *ii* similar to curve *i* although the compressibility effects become important. As P_b continues to be reduced, the flow rate will continue to increase. [15]

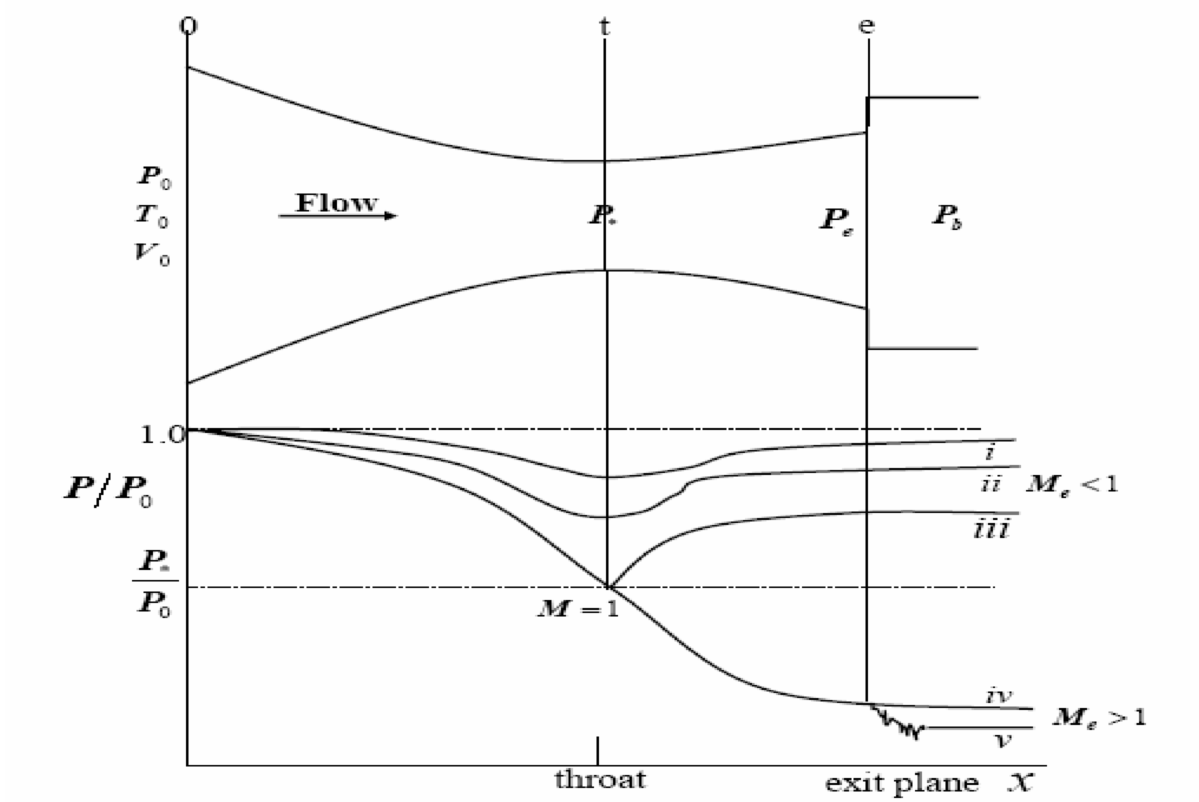


Figure 4.4: Pressure profile for isentropic flow in a converging-diverging nozzle [15]

If back pressure P_b is lowered far enough, ultimately the flow reaches $M = 1$at nozzle throat—the section of minimum flow area, as shown on curve iii of fig.4.4 and the nozzle is choked. When curve *iii* is reached, critical conditions are present at the throat and the mass flow rate attains the maximum possible for the given nozzle and stagnation conditions. The corresponding pressure is the critical back pressure, P_*. The definition of critical condition is the state at which the Mach number is unity. Substituting $M = 1$ into Equations (4.9) through (4.11) and considering the definition of Mach number, we have following relationships.

$$\frac{P_*}{P_0} = \left(\frac{2}{\gamma+1} \right)^{\frac{\gamma}{\gamma-1}} \tag{4.12}$$

$$\frac{T_*}{T_0} = \frac{2}{\gamma+1} \tag{4.13}$$

$$\frac{\rho_*}{\rho_0} = \left(\frac{2}{\gamma+1} \right)^{\frac{1}{\gamma-1}} \tag{4.14}$$

$$V_* = C_* = \sqrt{\frac{2\gamma RT_0}{\gamma+1}} \tag{4.15}$$

For gas, $\gamma = 1.29$, and $\frac{P_*}{P_0} = 0.547$, $\frac{T_*}{T_0} = 0.873$, $\frac{\rho_*}{\rho_0} = 0.626$. The mass flow rate at critical condition is calculated by

$$m = \rho_* V_* A_* \tag{4.16}$$

As shown in Fig. 4.4, when back pressure is reduced further, below P_* , such as conditions of *iv* and *v* , information about the throat downstream conditions cannot be transmitted upstream. Consequently, reductions in P_b below P_* has no effect on flow conditions in the converging nozzle portion; thus, neither the pressure distribution through the converging nozzle, throat exit pressure, nor mass flow rate is affected by lowering P_b below P_*. This is an important phenomenon which is used to design critical flow process for gas flow measurement.

The diverging section accelerates the flow to supersonic speed from $M=1$ at the throat. Accelerating the flow in the diverging section results in a pressure decrease. This condition is illustrated by curve *iv* shown in Fig. 4.4. If the back pressure is set at P_{iv}, flow will be isentropic through the nozzle, and supersonic at the nozzle exit. Nozzles operating at $P_b = P_{iv}$ are said to operate at design conditions. Lowering the back pressure below condition *iv* , say to condition *v* , has no effect on flow in the nozzle. The flow is isentropic from the plenum chamber to the nozzle exit (same as condition *iv*) and then undergoes a three-dimensional irreversible expansion to the lower back pressure. A nozzle operating under these conditions is said to be under expanded, since additional expansion takes place outside the nozzle.

4.4.4 Flow Acceleration and Deceleration

Equation (4.17) and Equation (4.18) are convenient differential forms of the momentum and continuity equations, respectively, for isentropic flow.

$$\frac{dp}{\rho} + d\left(\frac{v^2}{2}\right) = 0 \tag{4.17}$$

$$\rho VA = Constant \tag{4.18}$$

Starting from the above two equations, a relationship between the flow area change and the velocity change can be derived as given by Equation (4.19).

$$\frac{dA}{A} = - \frac{dV}{V}\left(1\text{-}M^2\right) \tag{4.19}$$

From the above equation, it is clear that for $M < 1$ an area change causes a velocity change of opposite sign and a pressure change of the same sign; *for* $M >1$ an area change causes a velocity change of the same sign and a pressure change of opposite sign. Fig. 4.5 summarizes these relationships. For subsonic flows $(M <1)$, flow acceleration in a nozzle requires a passage of diminishing cross section; area must decrease to cause a velocity increase. A subsonic diffuser requires that the passage area increase to cause a velocity decrease. In supersonic flows $(M >1)$, the effects of area change are the opposite. A supersonic nozzle must be built with an area increase in the flow direction. A supersonic diffuser must be a converging channel. Where $M =1$

and the above mentioned choking phenomena occurs, the channel area is at its minimum. At that point, it is called the throat.

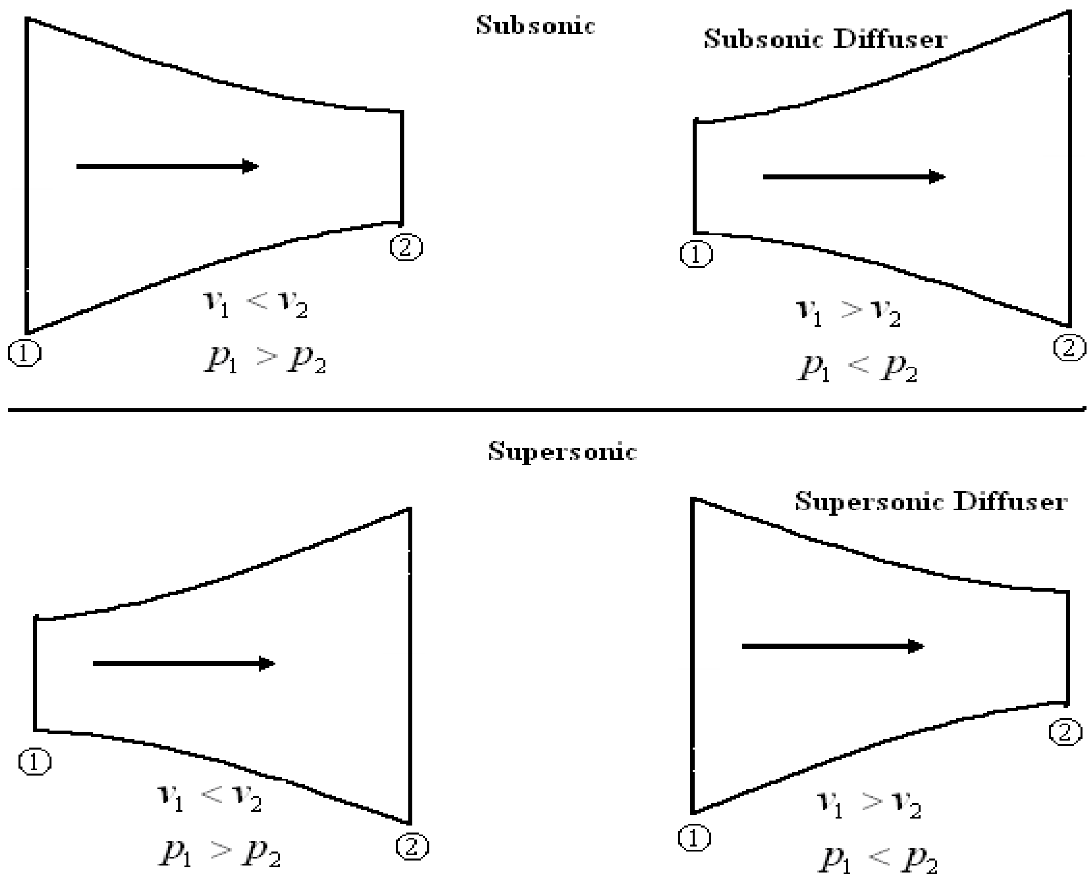


Figure 4.5: Nozzle and diffuser shapes for subsonic flow and supersonic flow

To accelerate flow from rest to supersonic speed requires first a subsonic converging nozzle. Under proper conditions, the flow will be choked at the throat, where the area is a minimum. Further acceleration is possible if a supersonic divergent nozzle segment is added downstream of the throat. This is the reason for the primary nozzle of modern gas ejectors to be designed in the convergent-divergent shape, as shown in Fig. 4.4.

To decelerate flow from supersonic to subsonic speed requires first a supersonic converging diffuser. The flow will be at $M=1$ at the throat, theoretically. In practice, supersonic flow cannot be decelerated to exactly $M=1$ at a throat because sonic flow near a throat is unstable due to a rising, reverse, pressure gradient. Further isentropic deceleration could take place in a diverging subsonic diffuser section. Flow at the ejector mixing section exit is subsonic for this study, and a diverging subsonic diffuser is used. [16]

Table 4.2: Ejector design boundary conditions

Parameters		**Units**	**Values**
Suction Gas	Flow Rate	MSCFD	155-215
	Pressure	PSIG	0.4-0.6
	Temperature	^{0}F	80-85
	C_p/C_v		1.283
Motive Gas	Flow Rate	MSCFD	440-590
	Pressure	PSIG	750
	Temperature	^{0}F	80-95
	C_p/C_v		1.629
Discharge Gas	Pressure	PSIG	90-100

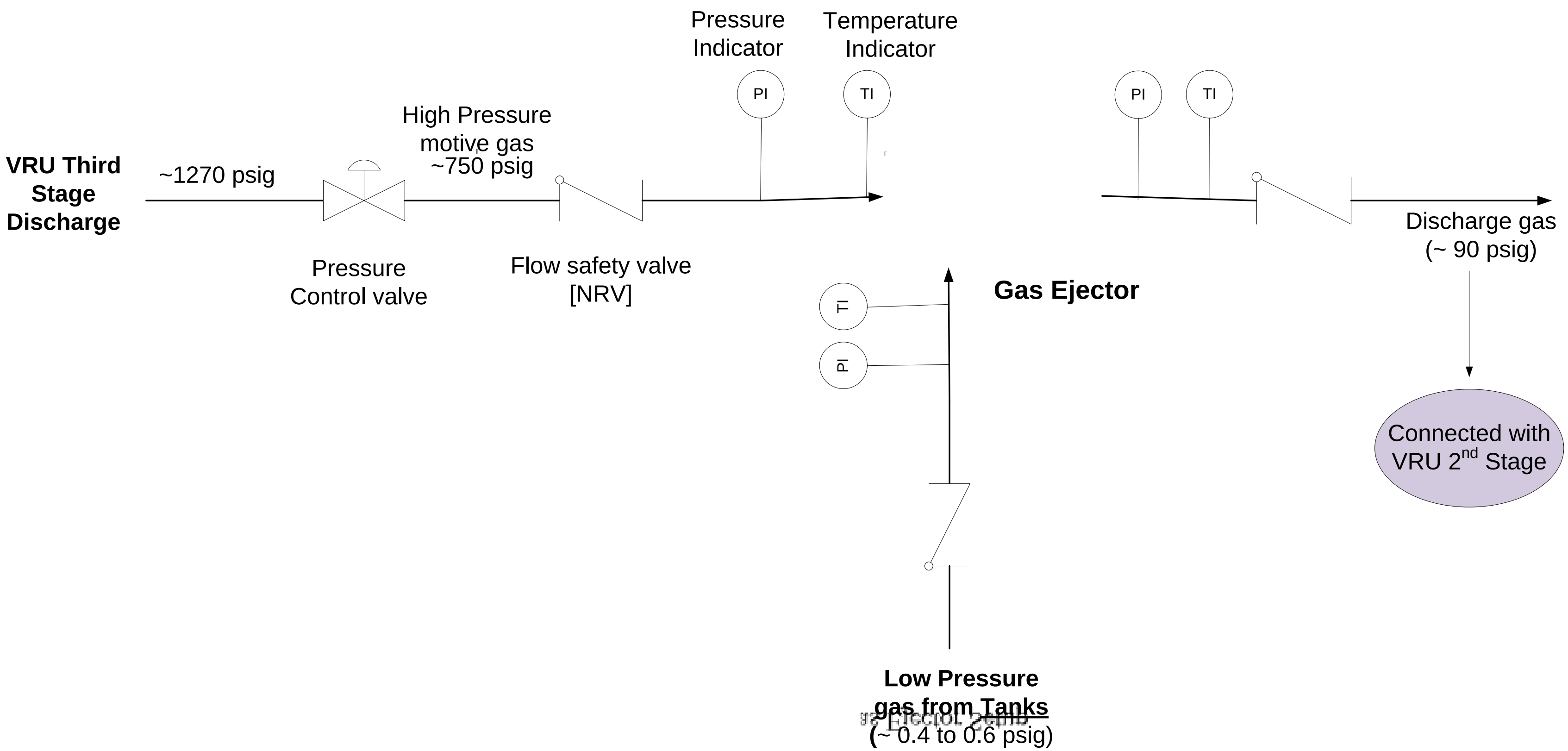


Figure-4.6:Schematic of Gas Ejector Setup

4.5 The Effect of Liquid in Gas Phase on Ejector Performance

Using a gas ejector to transfer two-phase flow with a very high portion of liquid is impractical, however it is feasible if the liquid portion (condensate or water) is very low and the mixture behaves like a gas. If, the amount of the liquid increases to more than 1% or 2% by volume, its effect on the performance of ejector starts to become decreasing. It is worth mentioning that in excess of 1% or 2% by volume, separation of the gas and liquid phases is required for the HP flow and in many cases, also for the LP flow. [17]. In this project, motive gas comes out from VRU third stage after passing 1st, 2nd and 3rd stage scrubber, so, gas is almost dry but pressure drop down by pressure regulator from 1270 psig to 750 psig creates liquid knockout .To estimate amount of liquid drop out, HYSYS simulation was used for justifying whether liquid scrubber was required or not. The result shows only 1.38 barrels [0.56% of total volume] liquid knock out, with respect to the total volume of 245 barrels. The mixture behaves like a gas, therefore, liquid would not create any problem for ejector performance. The suction gas is collected from the storage tanks. It is at low pressure and is almost dry. There would not be any liquid drop out from this stream.

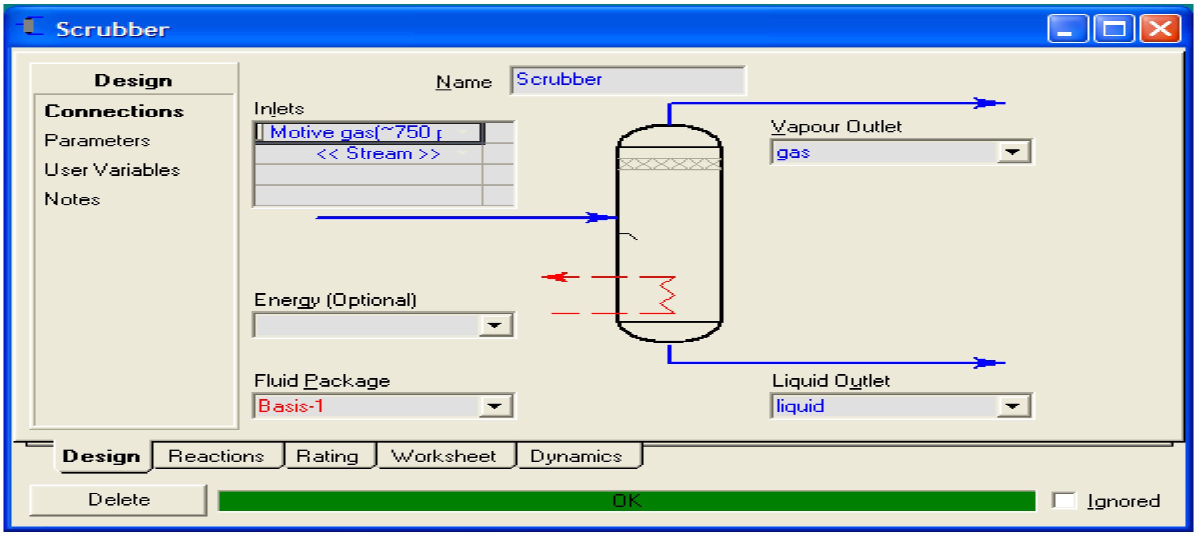
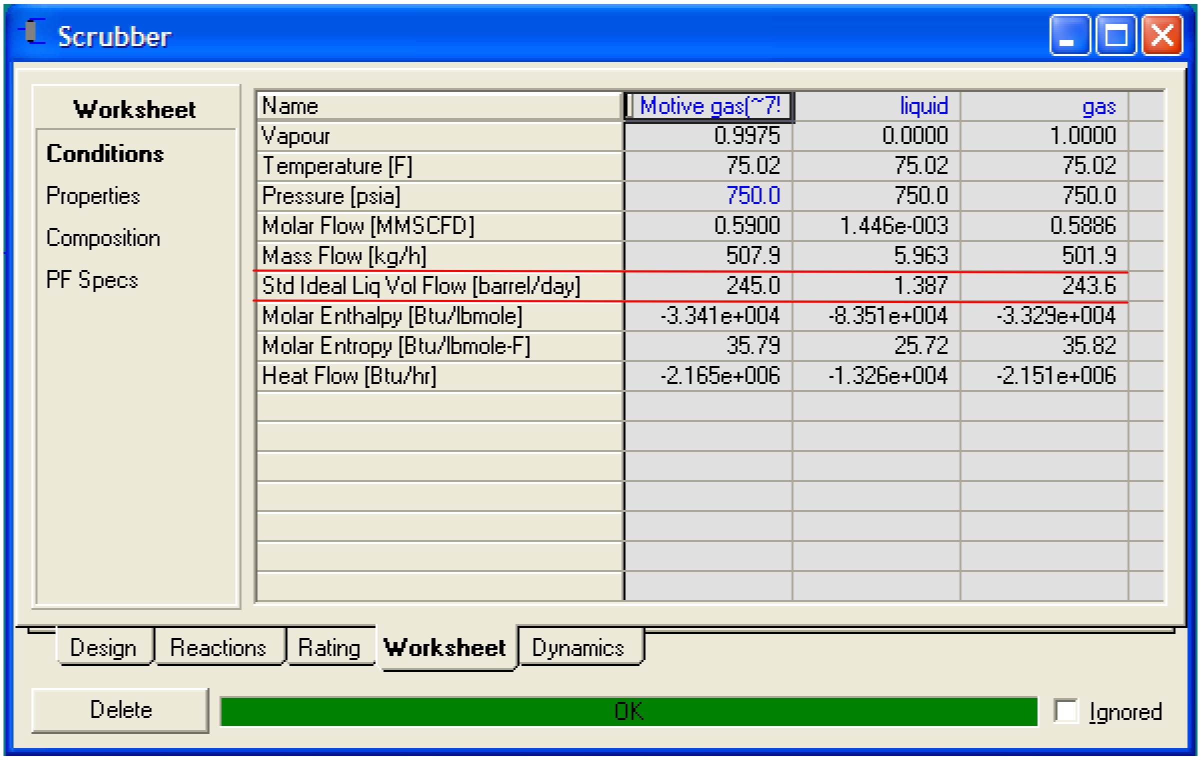


Figure 4.7: Liquid knockout scrubber inlet and outlet streams



Name	Motive gas(~7!	liquid	gas
Vapour	0.9975	0.0000	1.0000
Temperature [F]	75.02	75.02	75.02
Pressure [psia]	750.0	750.0	750.0
Molar Flow [MMSCFD]	0.5900	1.446e-003	0.5886
Mass Flow [kg/h]	507.9	5.963	501.9
Std Ideal Liq Vol Flow [barrel/day]	245.0	1.387	243.6
Molar Enthalpy [Btu/lbmole]	-3.341e+004	-8.351e+004	-3.329e+004
Molar Entropy [Btu/lbmole-F]	35.79	25.72	35.82
Heat Flow [Btu/hr]	-2.165e+006	-1.326e+004	-2.151e+006

Figure 4.8: Scrubbing different streams process parameters

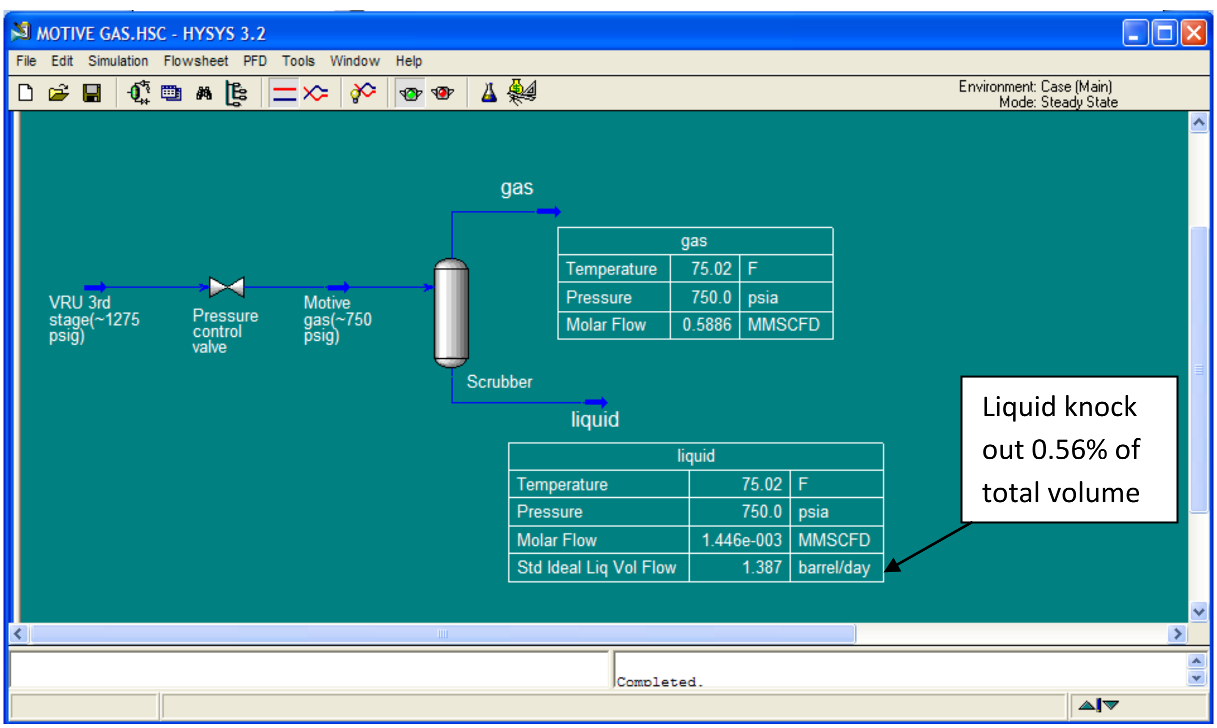


Figure 4.9: Liquid knockout in scrubber due to pressure reduction

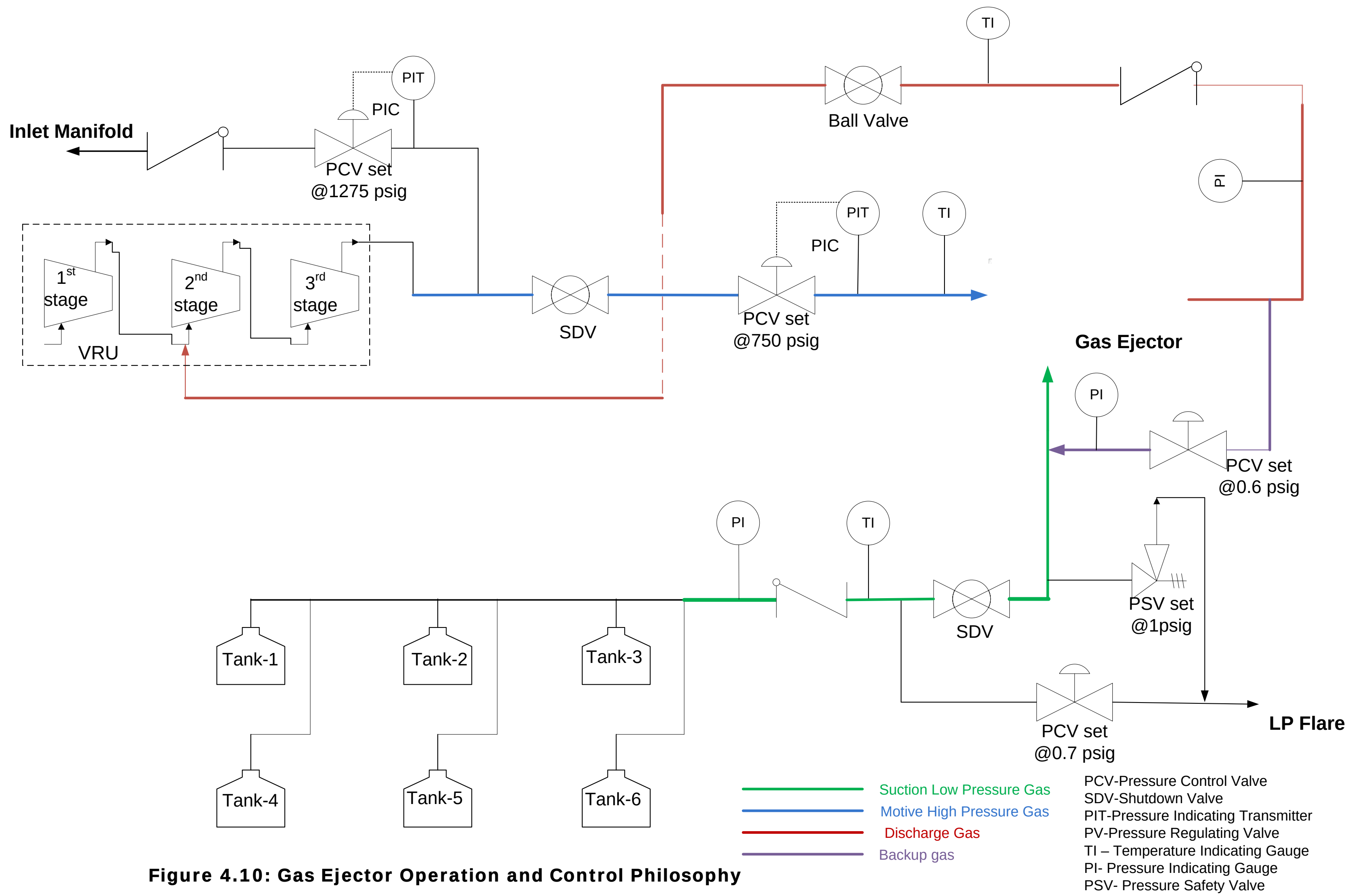


Figure 4.10: Gas Ejector Operation and Control Philosophy

4.6. Gas Ejector Operations and Control Philosophy

Gas ejector operations and control philosophy here discussed in reference to figure 4.10. The source gas will come from the condensate tanks, total 6 condensate tanks with a common header gas feed into the ejector suction side of operating pressure range 0.4 to 0.6 psig. The settings for the pressure relief valves for the tanks are PVSV (pressure vacuum safety Valve) set at 0.8 psig and PSV (pressure relief valve) set at 0.9 psig. The pressure and vacuum setting for the tank pressure relief devices are not being changed from the existing settings in place. Tank positive pressure maintaining will not be interrupted by installing the ejector. Tank existing process parameters will remain the same as before, only vapors that is now flaring, will be fed into ejector as suction gas. The pressure range for ejector as suction gas is 0.4 to 0.6 psig and flow rate range is 157 to 215 MSCFD with an average value of estimated gas is 190 MSCFD. But flow rate depends on condensate daily production. Condensate production varies with swing gas production.

The high pressure gas will come from third stage of existing VRU, first priority to use as motive gas for ejector and the remaining gas will be fed into inlet manifold (upstream of high pressure separator) as currently the total production for VRU goes to inlet manifold. VRU discharge pressure is about 1265-1270 psig, whereas motive pressure for ejector would be about 750 psig (pressure step down by pressure regulator) and gas flow rate is controlled depending on suction gas. Motive flow rate range would be approximately 440 to 590 MSCFD, considering flow rate is on average 530 MSCFD. Incorporating VRU discharge gas some portion as a motive source for ejector, this will not change or have any effect on VRU normal operations as well as VRU safety system at present.

The discharge of ejector will be fed into VRU second stage; presently VRU second stage handles only average 811 MSCFD against its total capacity of 1670 MSCFD. Considering maximum recovery from tanks, second stage will handle totally 1065 MSCFD (64% of design rated capacity). Ejector discharge pressure would be about 90 psig and flow rate is summation of suction and motive flow. Total enthalpy of the final mixture equals the weighted average of the initial components.

The Gas Ejector is being designed to handle the maximum vapors from condensate tanks, but in order to maintain the constant pressure on the low pressure side (suction portion of ejector) a simple control system is used to recycle make-up gas from the discharge of the Ejector, if required.

Regarding the control and safety considerations, the ejector system piping used to control the motive gas pressure and flow is accomplished by a pressure control valve upstream of the motive side of the ejector. This valve regulates the pressure to the design motive pressure, so that there is consistency and reduces the chances of any surges reaching the ejector. On both the suction and discharge lines, there are facilitated check valves inline to prevent any back flow. Backflow into the tanks would cause the tanks to bent or burst, as they are not capable of holding high pressures for welded tanks. The discharge check valve is to prevent any pressure surges in the

VRU 2nd stage inlet from back flowing into the ejector system. To protect the piping of the ejector system from pressure high, a pressure safety valve or pressure relief valve (PSV) is located on the suction piping set to 1 psig.

VRU will shut as existing safety considerations (i.e. third stage discharge pressure high high at 1325 psig and other process parameters high high or low low logic activation at present), when VRU goes in shut in condition then ejector also will go to out of operations and trapped gas in ejector system release to flare . Mentioning that, VRU can independently run while ejector system is out of operation and in both cases gas plant operations are unaffected.

All piping connections have pressure and temperature indicators to show in real-time these parameters to an operator of the system.

Gas Ejector installation will not eliminate the LP flare system rather than it will cut off tank flare. Then LP flare will burn only purge and pilot gas necessary for safe operation of the plant. Purge and pilot gas volumes that go to the flares are vital to the safe operation of the flares and the overall Plant. This purge and pilot gas ensures reliable stand-by ignition of the flare and prevents generation of dangerous gas/air mixtures in the flare systems.

Chapter 5

ENVIRONMENTAL AND ECONOMICAL BENEFITS

Gas Flaring is a high temperature oxidation process used to burn combustible components, mostly hydrocarbons. In combustion, gaseous hydrocarbons react with atmospheric oxygen to form carbon dioxide (CO_2) and water. During a combustion reaction, several intermediate products are formed, and eventually most of them are converted to CO_2 and water. Some quantities of stable intermediate products such as carbon monoxide, hydrogen, and hydrocarbons will escape as emissions. According to General Electric (GE), Natural gas flaring emits 400 million metric tons of CO_2 annually, the same as 77 million automobiles. Gas flaring doesn't produce useful energy, electricity or heat. This means that worldwide, billions of cubic meters of natural gas are wasted every year, typically as a by-product of oil extraction. [18]

Flaring of gas is a problem with two key dimensions-

1. Environmental concerns
2. Economical concerns

5.1 Environmental Concerns

The Environmental problems caused by flaring are mainly global, but to some extent also regional and local. Global environmental impact is due to the burning of associated or solution gas, which produces carbon dioxide (CO_2) and methane (CH_4). Of these two, methane is actually more harmful than carbon dioxide. It is also more prevalent in flares that burn at lower efficiency. These emissions increase the concentration of greenhouse gases (GHG) in the atmosphere, which in turn contributes to global warming. Flaring may furthermore contribute to local and regional environmental problems, such as acid rain with attendant impact on agriculture, forests and other physical infrastructure· Heat, soot from smoky flares and emissions of noxious gases (NOx) may also have adverse environmental impacts, in particular on agriculture, and may also contribute to health problems.

The associated gas from condensate tanks, presently is not being utilized, is being flared through Low Pressure (LP) flares which contributes to global warming. Ejector installation is to reduce flaring is an important measure to curb emissions of greenhouse gases.

5.1.1 Calculate CO_2 Gas Flaring Emissions

The method for estimating emissions from natural gas flaring is based on the volume of natural gas is being flared. To calculate CO_2 emissions, the volume of flaring gas is converted into Btu, and then multiplied by an emissions coefficient *of 14.92 million metric tons of* CO_2 *per quadrillion BTU.* [19]

Gas Flaring = 14.92 million metric tons of CO_2 emissions per 10^{15} Btu of Natural gas

= 14.92 metric tons of CO_2 emissions/10^9 BTU of Natural gas ……………(5.1)

According to chapter-3 of Table 3.3, Condensate tank vapor calculations, estimated flaring gas from condensate storage tanks was 157 to 215 MSCFD range with an average value of 190 MSCFD.

For simplify the calculation, it was considered average amount of flaring gas of 190 MSCFD. Gross heating value of flaring gas was considered 1075 BTU/SCF (i.e. as like sales gas BTU of Bibiyana gas plant, though tank vapors contain higher heat content).

Amount of heat content of flared gas,

$$= (190*1000*1075) \text{ BTU/day}$$

$$= 204.25X10^6 \text{ BTU/day}$$

$$=74.55X10^9 \text{ BTU/year} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (5.2)$$

Converting heat content of gas to CO_2 emissions due to flaring, according to above equation (5.1),

Gas Flaring = 14.92 metric tons of CO_2 emissions/10^9 BTU of Natural gas

10^9 Btu heat content of gas is equivalent to 14.92 metric tons carbon emissions

According to equation (5.2),

Total CO_2 emissions = 1112 metric tons per year for tank vapor flaring

Installing ejector,

Net Heat Energy Saving	$74.55X10^9$ BTU per year
Flare Co_2 eliminated in gas plant locality	1112 metric tons per year

By capturing Tank hydrocarbon vapors, gas ejector reduces flaring and CO_2 emissions in gas plant locality. This ensures a cleaner environment (cleaner air) for on-site personnel who get benefit from a safer working environment.

5.2. Economical Concerns

- Volume of Gas collected: Gas ejector will recover gas from condensate tanks is on average 190 MSCFD or, 0.190 MMSCFD and maximum rate is 215 MSCFD. Considering the average value of three different methods for vapor estimation is 190 MSCFD, therefore, gas ejector will save the flaring gas of 5.7 MMSCF per month and 68 MMSCF per year

- Gross revenue per year = (Q x P x 365)

  Q = Rate of vapor recovery (MSCF per day)

  P = Price of natural gas

If price of natural gas is assumed $3/MSCF(considering minimal gas prices and BTU content just like sales gas heat content as gross heating value 1075 BTU/Scf, though, tank vapors is enriched with higher heat content)

Then Gross Revenue will be $ **208,050** per year

The main benefits of gas ejector in most applications are the increase in production and recovery from the field or preventing wastage or flaring of low pressure gas, using an available source of energy.

The capital spend for the system includes a single gas ejector and the associated connecting piping, controlling valves, instruments etc. The total capital spent can be compared against the increase in revenue from recovery gas.

Ejector systems are made by several manufacturers. Equipment costs are determined largely by the volume handling capacity of the unit, line pressure, the number of tanks, and the size and type of ejector.

Companies, who have installed Ejector in field, report that installation costs can add as much as 50 to 60 percent to the initial capital equipment and piping cost. Installation costs can vary greatly depending on location (remote sites will likely result in higher installation costs) and the number of tanks. [20]

Operating and maintenance costs are very low because an ejector has no moving parts. Ejector based vapor recovery system are almost maintenance free operation and zero operating cost.

A simple way to assess the economic aspects whether recovery is profitable or not, that is to work out Cost Analysis and the Payback period achieve from the increased revenue to recover the capital spent.

5.2.1 Cost Benefit Analysis

According to Ejector Manufacturer, COMM Engineering, the gas ejector that will serve to extract 190 MSCFD on average and maintaining motive and discharge flow boundary conditions as applicable for this project, cost of that Ejector is approximately $8,500. Other equipments such as control valves, instruments and connecting piping cost and also total installation cost are mentioned below.

Table 5.1: Equipment and Installation cost

No.	Description	Number required	Cost($)
1	Pressure control valve, High Pressure[PCV set @1275 psig, size 2” and PCV set @750 psig, size 3”]	2	4,000
2	Pressure control valve, Low Pressure[PCV set @0.6 psig, size 2” and PCV set @0.7 psig, size 6”]	2	2,600
3	Shut down valve, High Pressure [SDV pressure rating 1350 psig, size 3”]	1	5,000
4	Shut down valve, Low Pressure [SDV pressure atmospheric, size 3”]	1	4,000
5	Ball Valve[VB size 4”, pressure rating 100 psig]	1	800
6	Pressure safety valve[PSV set @1 psig and size 2”]	1	600
7	Check valve[VC size 2” and pressure rating 1350 psig]	1	1,200
8	Check valve[VC size 4” and pressure rating 100 psig]	1	1,200
9	Check valve[VC size 3” and pressure atmospheric]	1	1,000
10	Gas Ejector[capacity described in Table 4.3:Ejector design boundary conditions]	1	8,500
11	Piping[scheduled Carbon steel] and Instrumentation[pressure transmitter, pressure & Temperature gauge]		40,000
12	Installation cost[consider 50% cost of equipment & piping]		36,000
13	Radiography Test (X-ray) for welding joint		6,000
14	Hydro test		1,000
15	Miscellaneous		8,000
Total Cost			**$119,900**

Under Cost Analysis, equipment cost and capitals spend for installations are covered. Equipment prices are taken from Manufacturer equipment price list and here approximate value is used. The Payback Period (PBP) is the time required for an initial investment to be covered, neglecting the time value of money. In this project total investment will be $119,900 which would returns $208,050 per year.

Therefore, Simple Payback Period is 7 Months

Economic value of vapor gas recovered is $208,050 per year (considering minimal gas prices $3/MSCF, if gas prices higher, obviously value will be increased). Payback period covers in 7 months. This has shown clearly the economic attractiveness of this concept.

Industry experience regarding gas ejector performance and economical return shows very impressive. The TotalFinaEIf E&P USA, Inc., EI Ebanito Facility has been collecting approximately 175,000 SCF per day and payback period is only 0.4 years. Oued Zar Plant in Tunisia, Gemini field in Mississippi, offshore Louisiana etc. got generous revenues from associated gas collecting by ejector and payback period did not exceed six months. Recent field examples experienced by Agip, Marahthon, BP(UK), Britannia, Philips petroleum, Shell and others has shown that payback period is just few weeks. [21]

Chapter 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 Summary of Results

Gas ejector based design will capture tank hydrocarbon vapors thus natural gas will be saved which would have been otherwise flared. Environmental and Economical benefits show its justification in this application. Proper implementation of the design should result into the following:

- ***Volume of gas collected:*** 157 to 215 MSCFD range with an average value considering of 190 MSCFD. This should result into a monthly saving of 5.7 MMSCF and 68 MMSCF Per year

- ***Potential value of gas recovered:*** $208,050 per year, based on $3/MSCF(considering minimal gas prices and BTU content just like sales gas heat content as gross heating value of 1075 BTU/SCF, though, tank vapors is enriched with higher heat content)

- ***Cost of installed unit:*** $119,900 (approx.)
- ***Payback Period:*** 7 Months
- ***Net Heat Energy Saving***: $74.55X10^9$ BTU per year, considering heat content 1075 BTU/SCF
- ***Flare Co_2 eliminated in gas plant locality:*** 1112 metric tons per year. This ensures a cleaner environment (cleaner air) for on-site personnel in gas plant locality.

6.2 Conclusions

A simple design based on gas ejector is proposed to recover low pressure vapor from the condensate storage tanks. This design is easily implementable, with minimal intrusion of the existing system. Gas Ejector is a viable technology for recovering condensate tank vapors. Ejector based vapor recovery is called the EVRU (Environmental Vapor Recovery unit) or vapor recovery of natural gas using non mechanical technology. The short pay-back period and the significant amount of natural gas vapor recovery make the project for economical benefits. Environmental benefits are like flare reduction and reduce the CO_2 flaring gas emissions which ensure the cleaner environment (cleaner air) in gas plant locality.

6.3 Recommendations

- For tank vapor calculations, different methods (shrinkage volume, gas-oil ratio, HYSYS simulation approach) were used but that result shows some inconsistent and estimated gas varies ± 15%. The best approach for facility design is to use Recording Manometer or Ultrasonic Meter over several cycles that provide realistic data. In this project, this system was not availed due to unavailable option for install.

- Since the venturi gas ejector is a pressure increasing device, it can be used to recompress low pressure gas from a variety of sources for injection into the system (e.g. fuel gas)

- In this study, ejector design analysis is not covered in details, so, there is an opportunity to investigate supersonic performance of gas ejector.

- Ejector based technology might be applicable for other gas fields in Bangladesh to promote flare reduction as well as conserving energy.

REFERENCES

[1]. Bibiyana Gas Field, Chevron Bangladesh Ltd., “Condensate Stabilization and Storage Manual”

[2]. Bibiyana Gas Field, Chevron Bangladesh Ltd., “Mechanical Vapor Recovery Unit Manual”

[3]. Bibiyana Gas Field, Chevron Bangladesh Ltd.,“South Pad Flare System Manual”

[4]. U.S. Environmental Protection Agency. 1996. *Methane Emissions from the U.S. Petroleum Industry* (Draft Document). DCN: 96-298-130-61-01.

[5]. U.S. Department of Commerce. 1993. Control of Volatile Organic Compound Emissions from Volatile Organic Liquid Storage in Floating and Fixed Roof Tanks. Available through NTIS. Springfield, VA PB94-128519.

[6]. Power, Robert B. (1993). Steam Jet Ejectors for the Process Industries (First Edition). McGraw-Hill. ISBN 0-07-050618-3.

[7]. Mark, A., Stoner, J. (2003), Vapor recovery of natural gas using non mechanical technology, paper SPE-80599, TX, USA, 10-12 March 2003

[8]. Transvac Ejector Technology,U.K. URL: www.transvac.co.uk

[9] P. Andreussi, A. Ansiati, S. Senna, S. Sodini, P. Guaita “ Application of MJPs in Oil & Gas Fields” SPE annual technical conference and exhibition held in Houston, Texas, U.S.A., 26-29 September 2004

[10]. Campbell, J. M. (2004), Gas Conditioning and Processing, Vol. 1, Campbell Petroleum Series, OK, USA.

[11]. Aspen HYSYS, version 3.2 (build 5029); Hyprotech Ltd.

[12]. R. Gicquel “Modelisation dun ejector dans Thermoptim” Centre for Energy and Processes of Ecole des Mines de paris-April 2007

[13]. COMM Engineering. http://www.commengineering.com/vapor-recovery.html

[14]. Fox, R. W., McDonald, A. T. and Pritchard, P. J., (2003), Introduction to Fluid Mechanics, 6th Edition, John Wiley & Sons, Inc., Hoboken, NJ.

[15]. Dutton, J. C. and Carroll B. F., (1986), “Optimal Supersonic Ejector Designs,” Journal of Fluids Engineering, 108, pp. 415-420

[16]. Bartosiewicz, Y., Aidoun, Z., Desevaux, P. and Mercadier, Y., (2005), “Numerical and Experimental Investigations on Supersonic Ejectors,” International Journal of Heat and Fluid Flow, 26,pp. 56-70.

[17] Sarshar, M.M. and BEG, Dr. N.A., 2001. The applications and performance of a novel compact separator, 2nd GCC-EU Advanced Oil and Gas Technology Conference, Abu Dhabi, UAE, May 21-22, 2001.

[18]. Environmental Vapor Recovery Unit (EVRU) COMM Engineering, “Environmental Technology Verification Report”, USA.SRI/USEPA-GHG-VR-19, September 2002.

[19]. Gervet, B. (2007), Gas flaring emission contributes to global warming, Renewable Energy Research Group, Lulea University of Technology, March 2007, SE-97187 Lulea, Sweden.

[20]. “Installing vapor recovery units on crude oil storage tanks” from Natural Gas Star Partners, Natural gas EPA pollution Preventer, EPA, Natural gas, USA.

[21]. Ashton, K., Green, A.J. and Read, A., “Gas Production improvements using ejectors”. SPE paper 26684, offshore Europe, Aberdeen, 7-10 September 1993.

APPENDIX - A

Table A-1 : Condensate Daily Production data in one month

Date of Month	Condensate Production (BPD)
1	3810.05
2	3604.15
3	3416.38
4	3457.08
5	3732.17
6	3753.80
7	3395.49
8	3494.40
9	3680.90
10	3614.18
11	3373.75
12	3403.40
13	3636.58
14	3711.65
15	3432.55
16	3403.14
17	3363.65
18	3641.48
19	3642.99
20	3505.71
21	3482.62
22	3359.89
23	3798.63
24	3303.86
25	3555.72
26	3396.08
27	3429.04
28	3298.26
29	3301.53
30	3415.70
31	3405.46
Average Condensate Production 3510 BPD	

Table A-2: VRU Daily Production data in one month

Date of Month	VRU Production(MMSCFD)
1	0.861
2	0.854
3	0.870
4	0.854
5	0.829
6	0.863
7	0.842
8	0.731
9	0.863
10	0.821
11	0.844
12	0.840
13	0.837
14	0.743
15	0.524
16	0.468
17	0.637
18	0.776
19	0.860
20	0.860
21	0.854
22	0.827
23	0.822
24	0.862
25	0.859
26	0.870
27	0.871
28	0.848
29	0.833
30	0.877
31	0.841
Average VRU Production	**0.811 MMSCFD**

APPENDIX - B

Ejector Datasheet and Specification Checklist

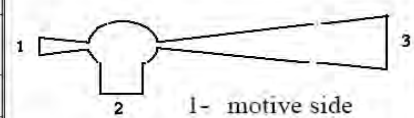
KARA TAHVIEH Co.
Design and manufacture of vacuum equipment

No.296 – 13th ave. – Toos(1) industrial town – Mashhad – IRAN
Tel: 00985115413211
Fax: 00985115413212
www.karajet.com
info@karajet.com

Enquiry form for ejectors

We thank you for contacting us and submitting an enquiry for quotation.
You are asked to fill up this form and return it by either uploading through our web, e-mail or fax. You can also send us your enquiry by mail or even call us for more details about our products.

Company	
address	
Contact name	
Phone	
E-mail	



1- motive side
2- suction side
3- discharge side

Field of application/ comments	
Receive a brochure from our company ☐	

Motive side	Nature of fluid	☐ Air		
		☐ Steam		
		☐ Gas	Name/ composition	
			MW(g/mol)	
			C_p/C_v	
		☐ Liquid	Name/ composition	
			Density	
			Viscosity	
	Pressure (bar(a),kPa)			
	Temperature(°C)			
	Flow rate(Kg/h)			
Suction side	Nature of fluid	☐ Air		
		☐ Steam		
		☐ Gas	Name/ composition	
			MW(g/mol)	
			C_p/C_v	
		☐ Liquid	Name/ composition	
			Density	
			Viscosity	
	Pressure (bar(a),kPa)			
	Temperature(°C)			
	Flow rate(Kg/h)			
Discharge side	Flow rate(Kg/h)			
	Pressure (bar(a),kPa)			
Construction material	Motive nozzle			
	Diffuser/shell			
	Inlet nozzle/flanges			
requisition	Quantity of identical equipment required			
	Attached documents/requisitions			

Date Sent to COMM

COMM ENGINEERING U.S.A.
P. O. BOX 53463 Lafayette, LA 70505-3463
PHONE (337)-237-4373 FAX (337)-234-1805

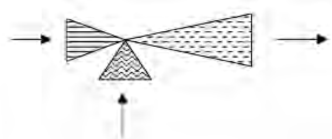
EVRU™ Based Vapor Recovery

Company			
Project Location		ON / OFF	Shore
Primary Contact			
Phone #		FAX #	
City		State	
Zip Code		Country	
Email			

Source of Vent / Flare Gas

Well / Tank (s) Name / # / Location	

Motive Gas / Available High Pressure Gas		
Gas		
Molecular Wt.		
Temperature		°F
Pressure		psig
Avg. Specific Heat		BTU/lb °F
Volume Available*		MScfD



Discharge Gas		
Pressure		psig
Avaliable Compression		
☐ Comp 1st Stage ______ MScfD		
☐ Comp 2nd Stage ______ MScfD		
☐ Comp 3rd Stage ______ MScfD		
☐ Other Pipeline ______ MScfD		

*: Motive Volume needs to be approximatly 3 times more than the Suction Volume to be picked up.

Suction Gas From Well / Tank (s) / Equipment		
Gas		
Molecular Wt.		
Temperature		°F
Pressure		psig
Flow Rate		lb/hr
Avg. Specific Heat		BTU/lb °F
Volume to Move		MScfD

Construction Requirements

	Connection Type	Rating	Material	Weld
Motive	NPT / FF / RF	#	CS / SS	WN / SW
Suction	NPT / FF / RF	#	CS / SS	WN / SW
Discharge	NPT / FF / RF	#	CS / SS	WN / SW

Time to Install

- ☐ 0-2 Month Time Frame
- ☐ 3-6 Month Time Frame
- ☐ 7-12 Month Time Frame
- ☐ 12+ Month Time Frame
- ☐ Projected Install Date mm/dd/yyyy:______________

Service. Check all that apply

- ☐ Hydrocarbon
- ☐ H2S (Hydrogen Sulfide Gas)
- ☐ Aromatics
- ☐ Liquid/Gas Mixture
- ☐ Down hole Application

NPT = National Pipe Thread
FF = Flat Face Flange
RF = Raised Face Flange
CS = Carbon Steel

SS = Stainless Steel
WN = Weld Neck
SW = Socket Weld
MScfD = Thousand Standard Cubic Feet per Day

Additional Comments:

APPENDIX - C

HYSYS Result Datasheet [11]

HYPROTECH — TEAM LND, Calgary, Alberta, CANADA	Case Name:	F:\PMRE\THESIS\VR PROJECT\MAY'11\HYSYS PRACTICE\TANK VAPOR(
	Unit Set:	NewUser2
	Date/Time:	Fri Jul 10 03:24:42 2009

Valve: Control valve

PROPERTIES

Inlet Properties : LP Gas Sep.

		Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction		0.0270	0.0270	0.9730		
Temperature:	(F)	87.00	87.00	87.00		
Pressure:	(psia)	49.70	49.70	49.70		
Molar Flow	(MMSCFD)	2.939	7.936e-002	2.859		
Mass Flow	(kg/h)	1.676e+004	85.29	1.668e+004		
Std Ideal Liq Vol Flow	(barrel/day)	3600	36.54	3563		
Molar Enthalpy	(Btu/lbmole)	-1.068e+005	-3.529e+004	-1.088e+005		
Mass Enthalpy	(Btu/lb)	-932.8	-1635	-929.2		
Molar Entropy	(Btu/lbmole-F)	35.72	42.84	35.52		
Mass Entropy	(Btu/lb-F)	0.3118	1.985	0.3033		
Heat Flow	(Btu/hr)	-3.447e+007	-3.075e+005	-3.417e+007		
Molar Density	(lbmole/ft3)	0.1738	8.581e-003	0.3731		
Mass Density	(lb/ft3)	19.90	0.1852	43.70		
Std Ideal Liq Mass Density	(lb/ft3)	43.88	22.00	44.11		
Liq Mass Density @Std Cond	(lb/ft3)	44.33	---	44.47		
Molar Heat Capacity	(Btu/lbmole-F)	58.40	10.61	59.72		
Mass Heat Capacity	(Btu/lb-F)	0.5099	0.4915	0.5100		
Thermal Conductivity	(Btu/hr-ft-F)	---	1.772e-002	7.068e-002		
Viscosity	(cP)	---	1.127e-002	0.5346		
Surface Tension	(dyne/cm)	20.01	---	20.01		
Molecular Weight		114.5	21.58	117.1		
Z Factor		---	0.9872	2.271e-002		

Outlet Properties : Liquid					
	Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction	0.0534	0.0534	0.9466		
Temperature: (F)	84.78	84.78	84.78		
Pressure: (psia)	14.70	14.70	14.70		
Molar Flow (MMSCFD)	2.939	0.1570	2.782		
Mass Flow (kg/h)	1.676e+004	226.4	1.654e+004		
Std Ideal Liq Vol Flow (barrel/day)	3600	82.99	3517		
Molar Enthalpy (Btu/lbmole)	-1.068e+005	-3.914e+004	-1.107e+005		
Mass Enthalpy (Btu/lb)	-932.8	-1352	-927.0		
Molar Entropy (Btu/lbmole-F)	35.85	45.48	35.30		
Mass Entropy (Btu/lb-F)	0.3130	1.571	0.2957		
Heat Flow (Btu/hr)	-3.447e+007	-6.750e+005	-3.380e+007		
Molar Density (lbmole/ft3)	4.225e-002	2.533e-003	0.3677		
Mass Density (lb/ft3)	4.839	7.332e-002	43.89		
Std Ideal Liq Mass Density (lb/ft3)	43.88	25.71	44.31		
Liq Mass Density @Std Cond (lb/ft3)	44.33	---	44.59		
Molar Heat Capacity (Btu/lbmole-F)	58.04	13.10	60.58		
Mass Heat Capacity (Btu/lb-F)	0.5068	0.4526	0.5075		
Thermal Conductivity (Btu/hr-ft-F)	---	1.518e-002	7.167e-002		
Viscosity (cP)	---	1.058e-002	0.5576		
Surface Tension (dyne/cm)	20.56	---	20.56		
Molecular Weight	114.5	28.95	119.4		
Z Factor	---	0.9932	6.843e-003		

HYPROTECH LIFECYCLE INNOVATION	TEAM LND Calgary, Alberta CANADA	Case Name:	F:\PMRE\THESIS\VR PROJECT\MAY'11\HYSYS PRACTICE\TANK VAPOR(L
		Unit Set:	NewUser2
		Date/Time:	Fri Jul 10 03:45:30 2009

Tank: Condy Tank

Inlet Stream			
	Liquid		
Vapour Fraction	0.0534		
Temperature (F)	84.78		
Pressure (psia)	14.70		
Molar Flow (MMSCFD)	2.939		
MassFlow (kg/h)	1.676e+004		
Std Ideal Liq Vol Flow (barrel/day)	3600		
HeatFlow (Btu/hr)	-3.447e+007		

Outlet Stream				
	Condensate	vapor		
Vapour Fraction	0.0000	1.0000		
Temperature (F)	84.78	84.78		
Pressure (psia)	14.70	14.70		
Molar Flow (MMSCFD)	2.782	0.1570		
MassFlow (kg/h)	1.654e+004	226.4		
Std Ideal Liq Vol Flow (barrel/day)	3517	82.99		
HeatFlow (Btu/hr)	-3.380e+007	-6.750e+005		

PROPERTIES					
Liquid					
	Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction	0.0534	0.0534	0.9466		
Temperature: (F)	84.78	84.78	84.78		
Pressure: (psia)	14.70	14.70	14.70		
Molar Flow (MMSCFD)	2.939	0.1570	2.782		
Mass Flow (kg/h)	1.676e+004	226.4	1.654e+004		
Std Ideal Liq Vol Flow (barrel/day)	3600	82.99	3517		
Molar Enthalpy (Btu/lbmole)	-1.068e+005	-3.914e+004	-1.107e+005		
Mass Enthalpy (Btu/lb)	-932.8	-1352	-927.0		
Molar Entropy (Btu/lbmole-F)	35.85	45.48	35.30		
Mass Entropy (Btu/lb-F)	0.3130	1.571	0.2957		
Heat Flow (Btu/hr)	-3.447e+007	-6.750e+005	-3.380e+007		
Molar Density (lbmole/ft3)	4.225e-002	2.533e-003	0.3677		
Mass Density (lb/ft3)	4.839	7.332e-002	43.89		
Std Ideal Liq Mass Density (lb/ft3)	43.88	25.71	44.31		
Liq Mass Density @Std Cond (lb/ft3)	44.33	---	44.59		
Molar Heat Capacity (Btu/lbmole-F)	58.04	13.10	60.58		
Mass Heat Capacity (Btu/lb-F)	0.5068	0.4526	0.5075		
Thermal Conductivity (Btu/hr-ft-F)	---	1.518e-002	7.167e-002		
Viscosity (cP)	---	1.058e-002	0.5576		
Surface Tension (dyne/cm)	20.56	---	20.56		
Molecular Weight	114.5	28.95	119.4		
Z Factor	---	0.9932	6.843e-003		

HYPROTECH LIFECYCLE INNOVATION	TEAM LND Calgary, Alberta CANADA	Case Name:	F:\PMRE\THESIS\VR PROJECT\MAY'11\HYSYS PRACTICE\TANK VAPOR(LP)
		Unit Set:	NewUser2
		Date/Time:	Fri Jul 10 03:45:30 2009

Tank: Condy Tank (continued)

vapor					
	Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction	1.0000	1.0000	0.0000		
Temperature: (F)	84.78	84.78	84.78		
Pressure: (psia)	14.70	14.70	14.70		
Molar Flow (MMSCFD)	0.1570	0.1570	0.0000		
Mass Flow (kg/h)	226.4	226.4	0.0000		
Std Ideal Liq Vol Flow (barrel/day)	82.99	82.99	0.0000		
Molar Enthalpy (Btu/lbmole)	-3.914e+004	-3.914e+004	-1.107e+005		
Mass Enthalpy (Btu/lb)	-1352	-1352	-927.0		
Molar Entropy (Btu/lbmole-F)	45.48	45.48	35.30		
Mass Entropy (Btu/lb-F)	1.571	1.571	0.2957		
Heat Flow (Btu/hr)	-6.750e+005	-6.750e+005	0.0000		
Molar Density (lbmole/ft3)	2.533e-003	2.533e-003	0.3677		
Mass Density (lb/ft3)	7.332e-002	7.332e-002	43.89		
Std Ideal Liq Mass Density (lb/ft3)	25.71	25.71	44.31		
Liq Mass Density @Std Cond (lb/ft3)	---	---	44.59		
Molar Heat Capacity (Btu/lbmole-F)	13.10	13.10	60.58		
Mass Heat Capacity (Btu/lb-F)	0.4526	0.4526	0.5075		
Thermal Conductivity (Btu/hr-ft-F)	1.518e-002	1.518e-002	7.167e-002		
Viscosity (cP)	1.058e-002	1.058e-002	0.5576		
Surface Tension (dyne/cm)	---	---	20.56		
Molecular Weight	28.95	28.95	119.4		
Z Factor	0.9932	0.9932	6.843e-003		

Condensate					
	Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction	0.0000	0.0000	1.0000		
Temperature: (F)	84.78	84.78	84.78		
Pressure: (psia)	14.70	14.70	14.70		
Molar Flow (MMSCFD)	2.782	0.0000	2.782		
Mass Flow (kg/h)	1.654e+004	0.0000	1.654e+004		
Std Ideal Liq Vol Flow (barrel/day)	3517	0.0000	3517		
Molar Enthalpy (Btu/lbmole)	-1.107e+005	-3.914e+004	-1.107e+005		
Mass Enthalpy (Btu/lb)	-927.0	-1352	-927.0		
Molar Entropy (Btu/lbmole-F)	35.30	45.48	35.30		
Mass Entropy (Btu/lb-F)	0.2957	1.571	0.2957		
Heat Flow (Btu/hr)	-3.380e+007	0.0000	-3.380e+007		
Molar Density (lbmole/ft3)	0.3677	2.533e-003	0.3677		
Mass Density (lb/ft3)	43.89	7.332e-002	43.89		
Std Ideal Liq Mass Density (lb/ft3)	44.31	25.71	44.31		
Liq Mass Density @Std Cond (lb/ft3)	44.59	---	44.59		
Molar Heat Capacity (Btu/lbmole-F)	60.58	13.10	60.58		
Mass Heat Capacity (Btu/lb-F)	0.5075	0.4526	0.5075		
Thermal Conductivity (Btu/hr-ft-F)	7.167e-002	1.518e-002	7.167e-002		
Viscosity (cP)	0.5576	1.058e-002	0.5576		
Surface Tension (dyne/cm)	20.56	---	20.56		
Molecular Weight	119.4	28.95	119.4		
Z Factor	6.843e-003	0.9932	6.843e-003		