DESIGNING OF FLARE SYSTEM

A PROJECT REPORT

Submitted by

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with specialization in

PROCESS DESIGN ENGINEERING

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CERTIFICATE

This is to certify that the thesis titled **DESIGNING OF FLARE SYSTEM** submitted by **MOHAMED ISMAIL (R670215011)**, to the University of Petroleum & Energy Studies, for the award of the degree of **MASTER OF TECHNOLOGY** in Chemical Engineering With Specialization in Process Design Engineering is a bonafide record of project work carried out by him/her/them under my/our supervision.

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Sincerely

Mohamed Ismail

NOMENCLATURE

Abbreviations

- PRD - Pressure Relief device
- PRV - Pressure Relief Valve
- KOD - Knock Out Drum
- CV - Control Valve
- MAWP - Maximum allowable working pressure
- LPG- Liquified Petroleum Gas
- API - American Petroleum Institute
- ASME - American Society of Mechanical Engineers
- OSBL -Outside Battery limit
- ISBL -Inside Battery limit
- HC - Hydrocarbon
- O_2- Oxygen
- N_2- Nitrogen
- Btu - British thermal unit
- scf - standard cubic feet
- scfm - standard cubic feet per minute
- kPa - kilo Pascal
- NA - Not Available

ABSTRACT

Process safety is a disciplined framework for managing the integrity of operating systems and processes handling hazardous substances. It is achieved by applying good design principles, engineering and operating & maintenance practices. Flare systems plays an important role in the safety of Oil and Gas installations by serving as outlets for emergency pressure relief in case of process upsets. Accurate design of the flare system plays a key role in containing possible process safety hazards on the oil and gas installation, petrochemical industries, especially oil and gas offshore platforms. This Project is focused on designing the a flare network for a particular plant and optimizing the design procedure of the flare system. This project will give the optimized procedure involved in detailed design engineering for selection and sizing of Pressure relieve devices (PRD) and designing of Flare headers, Knock out drum, water seal drum, Flare stack, Flare seal, Flare tip, Flame Front Generator and documentation of Flare system. The Various relief scenario is considered like fire case, CV failure case, reflux failure case, electricity failure case, etc. The simulation tool used in this project is Aspen Techs Flare system analyzer (known as FLARENET),which is a steady state simulation tool.

Keywords: Process safety, Flare system, Pressure relief device, Knockout Drum, MAWP, Water seal drum, Flare seal, Aspen Tech, Steady state, FLARENET, Hydrocarbon

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CHAPTER 1

INTRODUCTION

1.1 General background

Across the global oil & gas industry, considerable effort has been focused on the prevention of major incidents. For the oil & gas industry the emphasis of process safety is to prevent unplanned releases which could result in a major incident. Process safety is a disciplined framework for managing the integrity of operating systems and process handling hazardous substances(C.L. Beyler, 2002). It is achieved by applying good design principles, engineering and operating & maintenance practices. It deals with prevention and control of events that have potential to release hazardous materials and energy. Such incidents can result in toxic exposures, fires or explosions and could ultimately result in serious incidents including fatalities, injuries, property damage, lost production or environmental damage.
As a major safety requirement at oil and gas installations such as refineries and process facilities, a flare system is usually installed to relieve built up pressure that may occur during shut down, start up or due to process system failure, reducing other safety hazards associated with process emergencies. Accurate design of the flare system plays a key role in containing possible process safety hazards on the oil and gas installation, especially oil and gas offshore platforms(CCPS, 1993).

1.2 Flare System

The flare system is the single largest pipe network in an oil & gas processing plant. It serves as a relief system for depressurizing different process and production units in cases of shut down or unexpected cases of hazardous process emergencies, by collecting excess fluid through relief devices and a pipe network and disposing of it to the required outlet. The light hydrocarbons and other gases are released by combustion into the atmosphere while the heavier hydrocarbon, liquids are let out through drains and are often pumped back into the separation system. Flare knock out drums are installed in the units to remove liquids from the relieved vapors. These liquids are routed to the units closed blow down system. Vapors from the units flare knock out drum leave the units battery limit and join the OSBL flare header. OSBL flare headers from various units in a complex are combined and routed to centralized flare systems. Depending upon the quality and quantity of flare loads and the physical dimensions of the complex, one or more centralized flare systems may be installed.The centralized flare system typically consists of an OSBL Flare Knock Out Drum, a Water Seal Drum, a Flare Stack, a Flare Seal, a Flare Tip and a Flame Front Generator(CCPS, 1993).

1.3 Potential causes of overpressure

There are various causes for overpressure in a process unit. Some of the major causes are reproduced here below(CCPS, 1993 ; API, 2014):

1. **Blocked Outlet:**
   Inadvertent closure of a valve at the outlet of equipments, instruments or piping may subject them to overpressure. Example: Blocking the outlet of a reciprocating compressor or pump subjects the intervening equipments, instruments and piping to overpressure.

2. **Inadvertent Valve Opening:**

   Inadvertent opening of a valve at the inlet of equipments, instruments or piping may subject them to overpressure. Example: Opening of valves designed for the reduction in steam pressure may subject the downstream system to overpressure.

3. **Check Valve Malfunction:**

   When fluids from a system with a low design pressure combine with fluids from a system with a high design pressure, usually a check valve is specified in the line with the low design pressure. The design pressure of the low design pressure line from the point of joining up to the check valve is specified as the higher one. This protects the low design pressure system. However, during check valve failure, low design pressure system gets subjected to high pressure. Example: Discharge check valve failure of high discharge pressure pumps when the discharge pressure of the pump gets transmitted to its suction side, which may not have been designed for the same.

4. **Utility failure**

   Loss of utility to equipments or instruments can lead to overpressure. Some of the possible utility failures and the equipments that get affected due to the same are as follows:

   (a) **Electric Power:**

   Pumps of Cooling water, boiler feed water, quench or reflux; FANS of air cooled heat exchangers, cooling towers, combustion air or flue gas; COMPRESSORS for process vapours, instrument air, vacuum or refrigeration, etc.

   (b) **Cooling Water:**

   Condensers for process or utility service; Coolers for process fluids, lubricating oil or seal oil; Jackrets on rotating or reciprocating equipment.

(c) **Instrument Air:**

Control valves for various process control functions Possibility of sudden instrument air loss to units is mitigated by the installation of a instrument air surge vessel of sufficient size to allow either rectification of instrument air loss (e.g. restarting of spare instrument air compressor) or safe shut-down of complex.

(d) **Steam:**

Turbine for pumps, compressors, blowers, combustion air fans or electric generators; Reboilers; Equipments using direct injection of steam (e.g. Strippers); ejectors and eductors

(e) **Fuel:**

Boilers, Reboilers, Engine drivers for pumps or electric generators, Gas turbines.

(f) **Inert Gas (Nitrogen):**

Seals, Purge for instruments and equipment.

5. **Loss of Fans:**

Loss of fans due to electrical or mechanical failure can lead to loss of cooling in air cooled heat exchangers thus causing overpressure. Loss of fans of cooling towers can cause high cooling water temperature which could lead to loss of cooling in units, causing overpressure.

6. **Loss of heat in series fractionation systems:**

In systems where the bottoms product from a column becomes the feed of the subsequent column, loss of heat in an upstream column can lead to higher volatility material going as feed to the downstream column, thus increasing its vapour load. As the downstream condenser is not designed for this additional vapour load, overpressurization of this column can occur. Example: Vacuum column downstream of a Atmospheric column in a Crude Distillation Unit.

7. **Reflux Failure:**

   Loss of reflux in a column due to reflux pump or reflux control valve failure can cause overpressure.

8. **Abnormal Heat Input from Reboilers:**

   Failure of temperature control valve of reboilers can cause excessive vapour generation leading to over pressurisation of column.

9. **Heat Exchanger Tube Failure:**

   Failure (rupture) of tubes of heat exchangers handling two fluids of different design pressures can lead to over pressurisation of low pressure side.

10. **Plant Fire:**

    Fire outside vessels can heat their surface and lead to vaporization of fluid contained in them thus leading to over pressurisation.

11. **Process Changes / Chemical Reactions:**

    Certain processes can lead to uncontrolled chemical reactions called runaway reactions if the process conditions are altered as a result of improper process control. If these reactions are exothermic, they may lead to over pressurisation.

CHAPTER 2

LITERATURE SURVEY

Process safety is a disciplined framework for managing the integrity of operating systems and processes that handle hazardous substances(API 520, 2015). It relies on good design principles, engineering, operating and maintenance practices. In recent years, major incidents in both the upstream and downstream industries have highlighted the importance of having these robust processes and systems in place.

2.1 Flare systems contribution for overall Process safety

Flare and disposal system plays an important role to prevent major incidents and it is part of process safety design of a plant. As seen in figure the Swiss cheese model, hazards are prevented/contained by multiple protective barriers. Barriers may have weaknesses or holes. When holes align hazard energy is released, resulting in the potential for harm. Barriers may be physical engineered containment or behavioral controls dependent on people. Holes can be latent/incipient, or actively opened by people. Flare and disposal system is one of the major prevention barriers for the safety and integrity of the operating assets(API 520, 2015).

2.2 Flare stack

Flares are generally categorized in two ways:

- height of the flare tip (i.e., ground or elevated)
- method of enhancing mixing at the flare tip (i.e., steam-assisted, airassisted, pressure-assisted, or non-assisted).

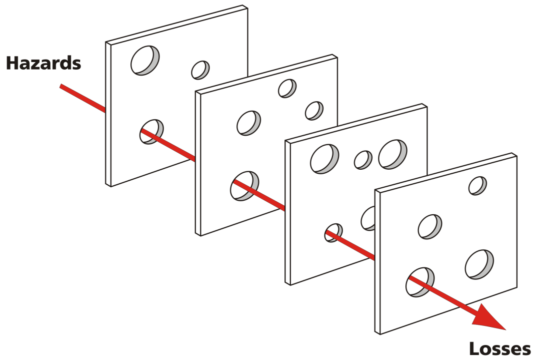


Figure 2.1: Swiss Cheese Model

Elevating the flare can prevent potentially dangerous conditions at ground level where the open flame (i.e., an ignition source) is located near a process unit. Further, the products of combustion can be dispersed above working areas to reduce the effects of noise, heat, smoke, and objectionable odors.
In most flares, combustion occurs by means of a diffusion flame. A diffusion flame is one in which air diffuses across the boundary of the fuel. This mixture on ignition establishes a stable flame zone around the gas core above the burner tip. Cracking can occur with the formation of small hot particles of carbon that give the flame its characteristic luminosity. If there is an oxygen deficiency and if the carbon particles are cooled to below their ignition temperature, smoking occurs. As in all combustion processes, an adequate air supply and good mixing are required to complete combustion and minimize smoke. The various flare designs differ primarily in their accomplishment of mixing(API 520, 2015).

2.2.1 Steam Assisted flares

Steam-assisted flares are single burner tips, elevated above ground level for safety reasons, that burn the vented gas in essentially a diffusion flame. To ensure an adequate air supply and good mixing, this type of flare system injects steam into the combustion zone to promote turbulence for mixing and to induce air into the flame.

2.2.2 Air Assisted flares

Some flares use forced air to provide the combustion air and the mixing required for smokeless operation. These flares are built with a spider shaped burner with many small gas orifices. The principal advantage of the air-assisted flares is that they can be used where steam is not available. This type of flare is not generally economical when the gas volume is large. The air assist flare is not usally used on large flares.

2.2.3 Enclosed Ground flares

The enclosed flare's burner heads are inside a shell that is internally insulated. This shell reduces noise, luminosity, and heat radiation and provides wind protection. A high nozzle pressure drop is usually adequate to provide the mixing necessary for smokeless operation and air or steam assist is not required. The height must be adequate for creating enough draft to supply sufficient air for smokeless combustion and for dispersion of the thermal plume. These flares are always at ground level. Enclosed flares generally have less capacity than open elevated flares and are used to combust continuous streams. Stable combustion can be obtained with lower Btu content stream gases (50 to 60 Btu/scf).

2.3 Flare system limitations in Oil and Gas industry

Flare, vent and blow down system are very critical systems in oil & gas plant. Initial system design for a typical topside facility is for maximum relief from the largest source for a particular relief scenario decided during design phase of the plant. As the time goes, subsequent modification projects, subsea tie-in to the existing topside facility makes flare system vulnerable. Some times each and individual project estimates the additional relief loads they will put into the existing flare system and compare with the available capacity in the flare system. In most of the cases the new sources are added to the existing flare system without any modification or upgrade of current system. Again, building a new flare system (which includes tail pipes, main header, KO drum and flare stack) requires

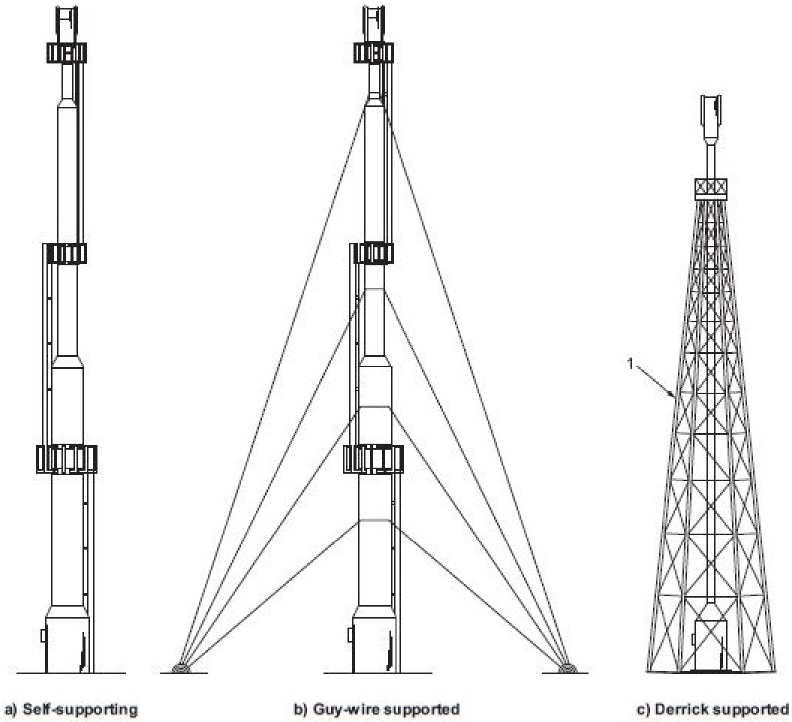


Figure 2.2: Flare stuctures

heavy investment and typically in an offshore installation where there is restriction on total allowed weight on top side equipment makes it no feasible.
Many cases it is gaseous/vapor phase fluid goes through the flare system. In certain cases the flow could be two phase with both liquid and gas phase present. Then it necessitates a detail study of the flow regimes, velocities of different phases, reaction forces and change in fluid property along the flare header(API 520, 2015).

CHAPTER 3

RELIEF SYSTEM DESIGN

3.1 Flare load calculation

For each of the overpressure scenarios, individual relieving rates need to be determined in order to size the safety valves that need to be installed to protect against overpressure. Generally liquid and vapour relief loads are determined by estimating net energy inputs. Energy inputs can be as heat input leading to an increase in pressure or it can be a direct pressure input from a higher pressure source. In determining the relief loads, the probability of two unrelated failures occurring simultaneously and normally does not need to be considered. The effect of pressure and temperature need to be considered to calculate the relief load. This means that relieving loads need to be considered at the set pressure of the safety device and not the normal operating pressure.
The location of the PRD devices is based on the nature of the vessel, operating conditions of the vessel,etc. The failure case scenario is to be analysed for that particular vessel like fire case, CV failure case, etc. The flare load for each PRD devices is calculated based on the components inside the vessel, failure case scenario, geometry of the vessel, height and thickness of the vessel, design pressure and set pressure of the vessel. After determining the individual relieving rates of a unit for all possible relief scenarios, the same are tabulated in the form of a Unit Flare Load Summary. A tabulation of flare loads from all units connected to a flare system is then prepared. This is called an Overall Flare Load Summary. The highest load out of all the loads in a flare load summary is called the Controlling Load or Governing case and determines the size of the relevant flare system facilities. (API521, 2014 ; API 520, 2015)

3.1.1 Load calculation for fire case

If the entire plant or a small portion catches fire, the components in the vessels in that fire area absorbs heat and it will pressurize the vessel. The fluid in that vessel should be relieved to depressurize and protect the vessel. The relieving load from each vessel is calculated from wetted area of the vessel. The wetted area depends on the geometry and orientation of the vessel. If the vessel located above 7.6 m from the base, there is no need to fix the PRV to that vessel for fire case. The wetted area to be calculated upto 7.6 m from the base.

- Dish area of horizontal cylinder

$$= \frac{\pi D_i^2}{8}\left[\left(\frac{h}{D_i} - 0.5\right)B + 1 + \frac{1}{4\epsilon}\ln\frac{4\epsilon\left(\frac{h}{D_i} - 0.5\right) + B}{2 - \sqrt{3}}\right]$$

Where B

$$= \sqrt{1 + 12\left(\frac{h}{D_i} - 0.5\right)^2}$$

D_i = Internal Diameter

h = Height

- Wetted Area of the Horizontal Cylinder

$$= \left[D_i\left(\pi - \arccos\left(\frac{(h-r)}{r}\right)\right)\right]L$$

Where L = Length

- Dish area of Vertical Cylinder

$$= \frac{\pi D_i^2}{8}\left[2 + \frac{1}{4\epsilon}\ln\left(\frac{2\epsilon + 2}{2 - \sqrt{3}}\right)\right]$$

- Wetted Area of vertical Cylinder

  $= \pi D_i h$

- Heat input to the wetted surface ,

  $Q = C_1 F A_{ws}^{0.82}$

  Where $C_1 = 43,200$

  Q = Total heat absorption.

- Mass flow rate due to choked flow

  $$= C_d A \sqrt{k \rho_0 P_0 \left(\frac{2}{k+1}\right)^{\left(\frac{k+1}{k-1}\right)}}$$

  Where C_d = Discharge coefficient

  A = Cross sectional Area

  $k = \frac{c_p}{c_v}$

  c_p = specific heat of the gas at constant pressure

  c_v = specific heat of the gas at constant volume

  ρ_0= Density

  P_0 = Absolute Pressure

3.1.2 Load calculation for control valve failure case

If the control valve is failed to close, the fluid flow from the valve is maximum and it will pressurize the vessel next to it. The fluid from the control valve is to be relieved to depressurize and protect the vessel. The maximum flow of the control valve is calculated.

The choked flow is the relieving load for control valve failure case.
Mass flow rate due to choked flow

$$= C_d A \sqrt{k \rho_0 P_0 \left(\frac{2}{k+1}\right)^{\left(\frac{k+1}{k-1}\right)}}$$

3.1.3 Load calculation for blocked outlet case

If any valve is failed to open or blocked, the vessel before to the valve will be pressurized. The maximum flow from the valve or the pipe is the relieving load for blocked outlet case.

3.1.4 Load calculation for reflux failure case

If the condenser, heat exchanger or cooling fans in a column is failed to operate due to electricity failure or pump failure, the column will be pressurized. The reflux stream will be failed to cool the stream from the tray next to it. The vapour load from that tray is the relieving load for that column. Probably this load will be the governing case among the other failure case.

3.2 Inlet line sizing calculation of PRV

The pressure loss between the vessel and PRD devices is calculated to design the diameter and thickness of the pipe connecting vessel. The Governing case load for vessel is used to determine the pressure loss. The diameter of the connecting pipe is designed in such a way that pressure loss in the line is less than 3% of the set pressure. The Darcy equation is used to find the pressure losses in the line.
Darcy Equation,

$$\frac{\Delta P}{L} = f_D \frac{\rho}{2} \frac{v^2}{D}$$

Where ΔP = Pressure drop

f_d = Friction Factor

v = velocity

3.3 PRV selection and sizing

The Pressure relief valve is used to control the pressure in a particular system which can build up for process failure like fire, instrument or equipment failure. The relief valve is designed to open at a set pressure to protect the system. The pressure is relieved by allowing the pressurised fluid to flow from the system (API 520, 2015).

3.3.1 General

The process in the particular system is operating under a certain pressure called operating pressure. The design temperature of that system is determined by the design rules of the pressure design code. The design temperature is based on the minimum permissible thickness and characteristic of each component. The Maximum allowable working pressure(MAWP) is the maximum pressure that the weekest component of the system can handle at a designed temperature. This is based on the design codes, fabrication of the vessel, piping, etc. The MAWP doesn't remain constant throughout the life of the system. It will reduce due to corrosion, wear and fatique. The design Pressure is the maximum pressure that the system that can be exposed to. Design pressure should be less than or equal to MAWP. The set pressure is the inlet pressure at which that PRV set to open to protect the vessel. The set pressure should be equal to or less than MAWP. Accumulation is pressure above the MAWP of the vessel. Accumulation is expressed as a percentage of MAWP. Overpressure is pressure above the set pressure of the PRV. The overpressure is expressed as a percentage of set pressure. The relieving pressure is equal to the set pressure plus the overpressure. The temperature of the flowing fluid at relieving conditions can be higher or lower than the operating temperature.

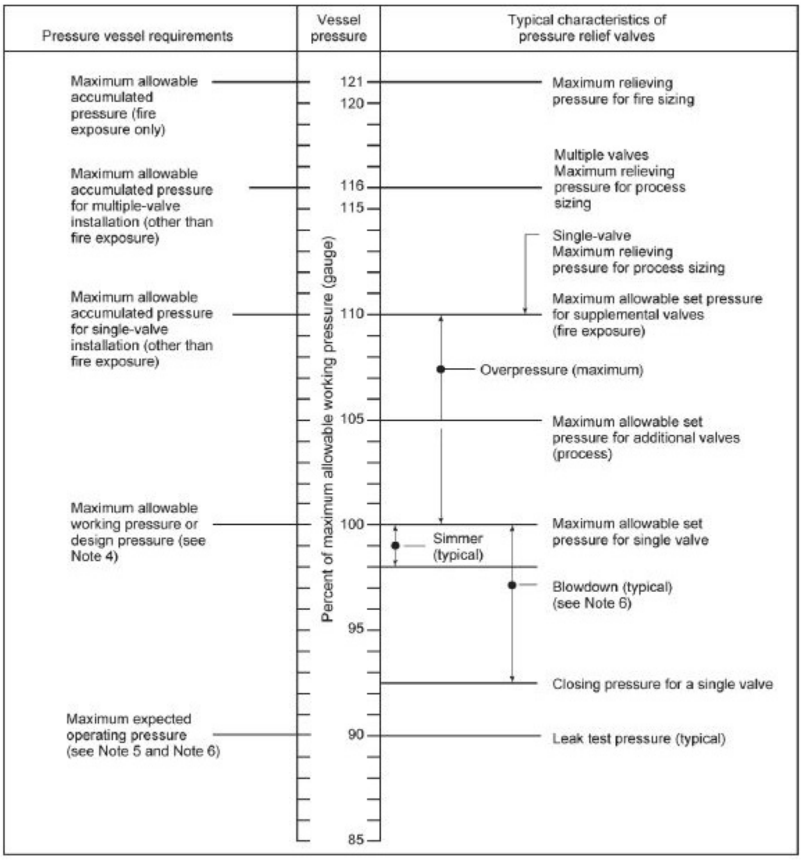


Figure 3.1: Pressure level relationship of PRV

3.3.2 Types of PRV

The three basic types of pressure relief valves are conventional, balanced and pilot operated.

3.3.2.1 Conventional PRV

The conventional valve is used in where the backpressure is less than 10% of the setpressure. The conventional PRV is a self actuated spring loaded PRV that is designed to open at a predetermined set pressure and protect the system from excess pressure by relieving the fluid from that system. The basic elements are an inlet nozzle connected to the system to be protected, a movable disk that controls flow through the nozzle, and a spring that controls the position of the disc(API 520, 2015). Under normal conditions the disc is seated on the nozzle preventing flow through the nozzle. When the inlet pressure is below the set pressure, the disc remains seated on the nozzle in the closed position. When the inlet pressure exceeds set pressure, the pressure force on the disc overcomes the spring force and the valve opens. When the inlet pressure is reduced to the closing pressure, the valve recloses.

3.3.2.2 Balanced PRV

A balanced PRV is a spring-loaded PRV that incorporates a bellows on the valve disc to minimize the effects of backpressure on the valve.For balanced PRV the allowable backpressure is 10 - 50% of the set pressure. When the backpressure is constant, the spring load in conventional valve can be reduced to compensate for the effect of backpressure on set pressure, and a balanced valve is not required. But the backpressure(superimposed + built up) is not constant always. The superimposed backpressure may be variable(API 520, 2015; Peter Smith *et al.*, 2003). In a balanced PRV, bellows is attached to the disk holder with an effective bellows area is equal to or greater than seating area of the disc. This isolates the seating area from the variable superimposed back pressure. Thus the variable back pressure would not affect the PRV opening pressure.

3.3.2.3 Pilot operated PRV

In pilot operated setup, main relief is combined with and controlled by a smaller self actuated pilot valve.This relief valve valve uses the process fluid (circulated through a pilot valve) to apply the closing force on the valve disc. The pilot valve is a small safety valve with a spring. The main valve does not have a spring but is controlled by the process fluid from pilot valve. The piston on the valve disc is designed to have a larger area on the top than on the bottom. Up to the set pressure, the top and bottom areas are exposed to the same inlet operating pressure(Adam Bader, *et al.*, 2011). Because of the larger area on the top of the piston, the net force holds the piston tightly against the main valve. As the

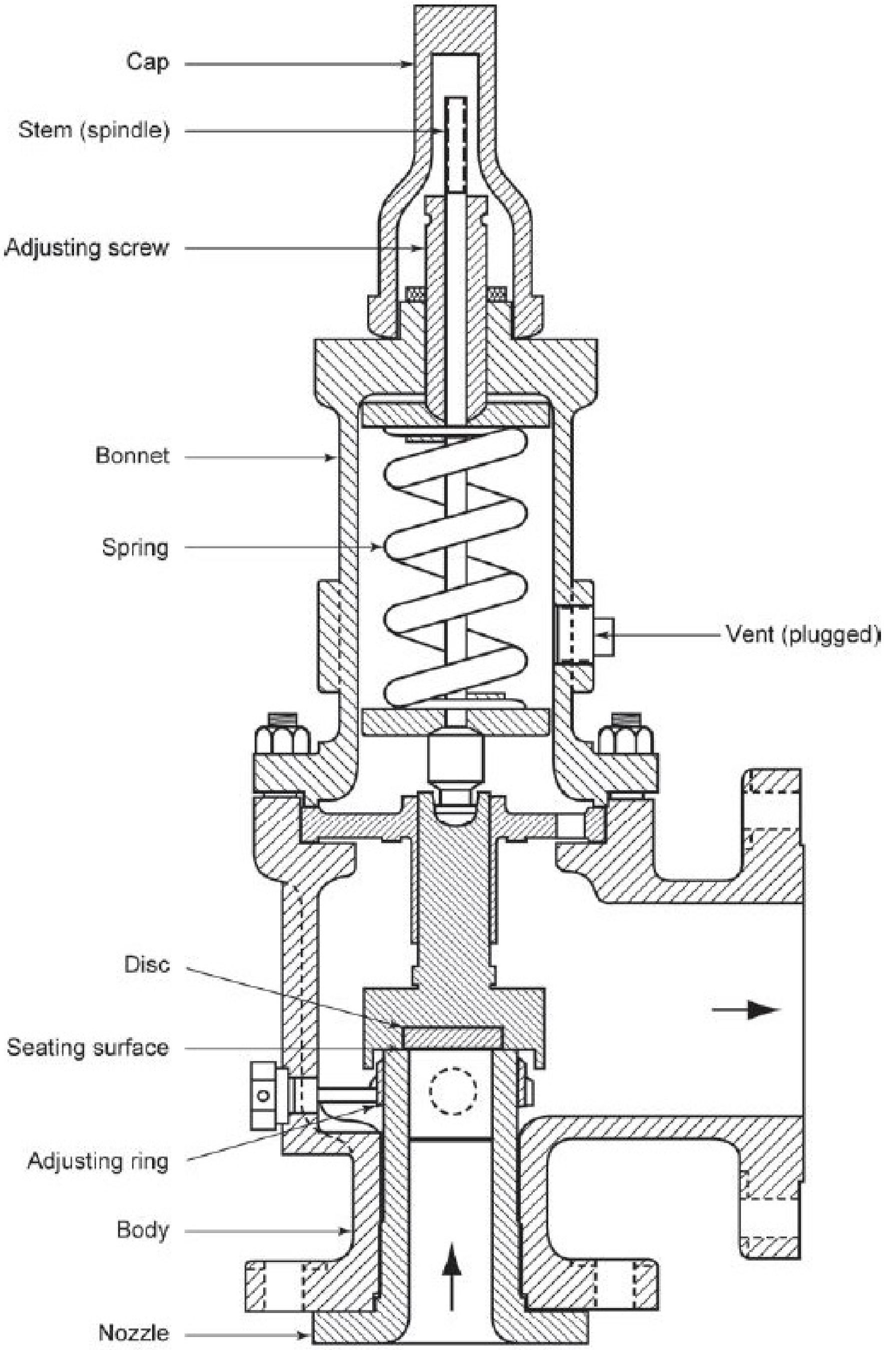


Figure 3.2: Conventional PRV

operating pressure increases, the net seating force increases and tends to make the valve tighter. At the set pressure the resulting net force is now upward causing the piston to lift and process flow is established through the main valve.This arrangement allows operation of pilot operated valves with a very narrow margin between set pressure of the relief valve and operating pressure of the protected equipment. The lift of the main valve piston is not

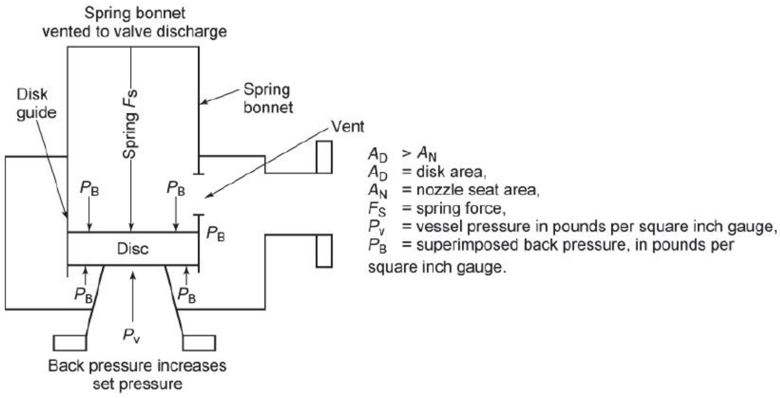


Figure 3.3: Effects of backpressure on convetional PRV

affected by built up backpressure. Allowable back pressure is typically more than 50% of the set pressure.

3.3.3 Selection of PRV

The selection of PSV among Conventional, Balanced and pilot operated PRV is depends on the pressure drop in inlet line of PRV, backpressure, set pressure, number of relief valves, failure scenario, nature of fluid, inlet and discharge piping, tail pipe sizing, space available, cost feasibility and backpressure in flare header while process failure.
The conventional valves requires the backpressure is less than 10% of the set pressure. The balanced valves allows the use of smaller tail pipes and relief headers because of the larger pressure drops allowed, as a result of higher allowable backpressure (10-50%). The balanced valve is more expensive than conventional valves; however, the total cost of the use of balanced valves plus the smaller header system may be lower(Adam Bader, *et al.*, 2011). The bellows should be checked periodically for leakage because A leaking bellow does not provide backpressure compensation. If the super imposed backpressure is constant, the spring load in conventional valve can be reduced to compensate for the effect of backpressure on set pressure, and a balanced valve is not required. The superimposed backpressure may be constant or variable. If it is variable, the balanced valve is preferable. When the pressure drop in inlet line of PRV exceeds 3% of set pressure, the pilot operated valve should be used.
The conventional valve is used in non corrosive services and the balanced valve is used

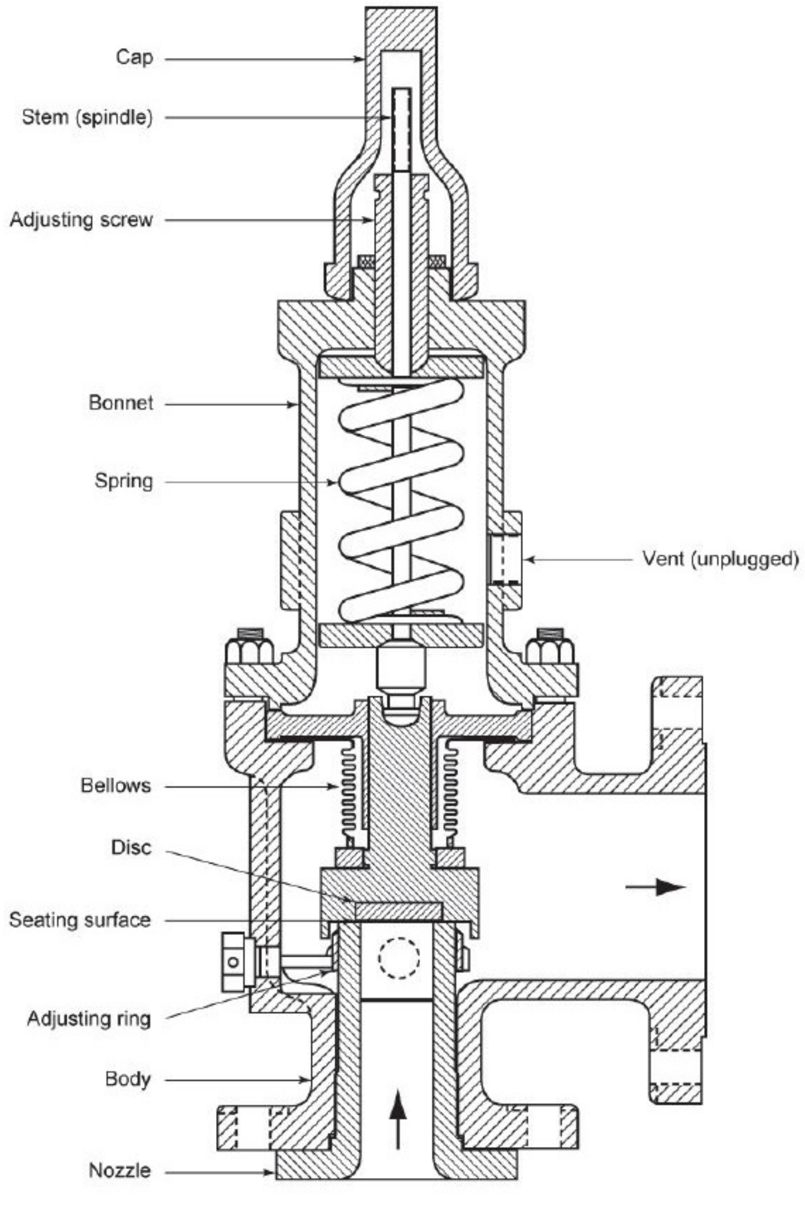


Figure 3.4: Balanced PRV

in corrosive services because the bellows isolates the disk from corrosive fluid. The pilot operated valves should be considered for clean services within their temperature limitations. The pilot operated valves cannot be used in multiphase fluids. For multiphase fluids, balanced valve is preferrable.

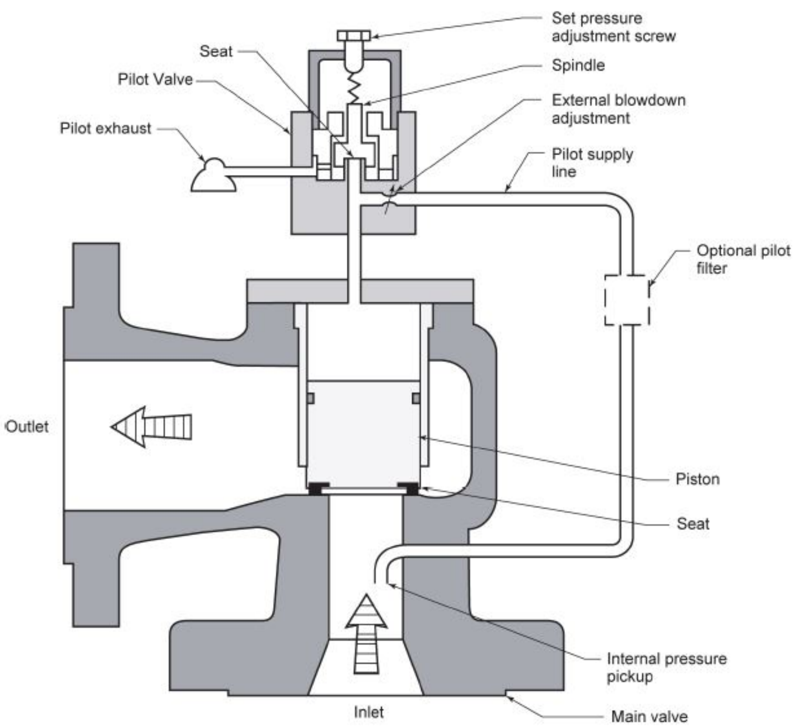


Figure 3.5: Pilot operated PRV

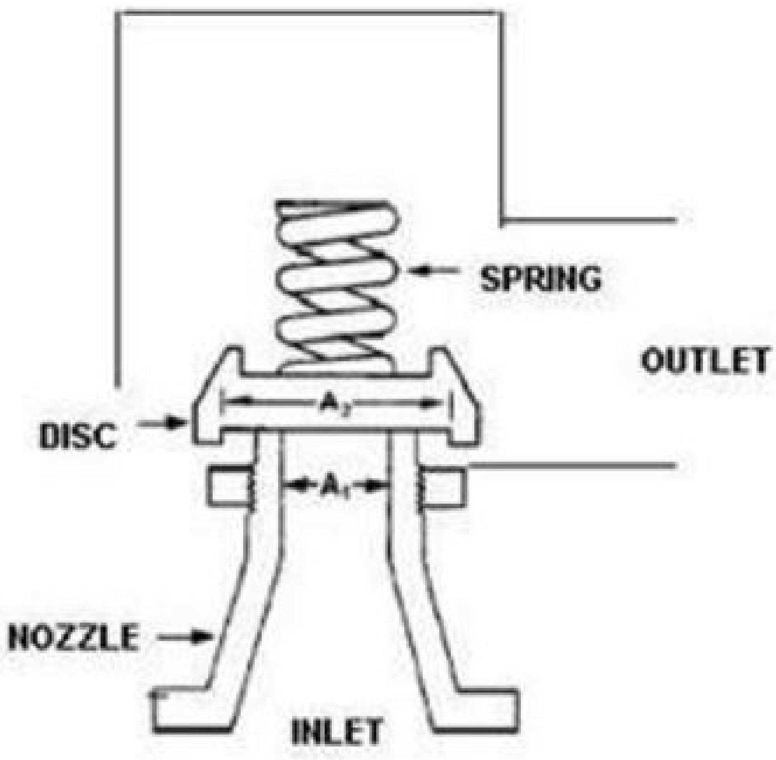


Figure 3.6: PRV opening

3.3.4 Orifice area calculation

The Orifice area needs to be calculated in order to have proper amount flow of the process fluid. The disc valve in the PRV held in the closed position by the spring. When the system pressure reaches the desired opening pressure. The pressure force of the process fluid pass through the inlet and then it is acting over Area A_1 equals the force of the spring, and the disc will lift and allow fluid to flow out through the outlet(Adam Bader, *et al.*, 2011). When pressure in the system returns to a safe level, the valve will return to the closed position. The orifice area has been calculated from the set of formulae given in the Appendix A. Some certain area has been standardized in API 526 (Flange Steel Pressure Relief Valves) and designated into certain alphabetic as shown on Table. The actual orifice area should be equal to or greater than calculated orifice area.

$$\mathrm{A} = \frac{W}{CK_dP_1K_bK_c}\sqrt{\frac{TZ}{M}}$$

or

$$\mathrm{A} = \frac{2.676V\sqrt{TZM}}{CK_dP_1K_bK_c}$$

or

$$\mathrm{A} = \frac{14.41V\sqrt{TZG_v}}{CK_dP_1K_bK_c}$$

Where

A = Required discharge area in mm^2

W = Flow in kg/hr

$$\mathrm{C} = 0.03948\sqrt{k\left(\frac{2}{k+1}\right)^{\frac{(k+1)}{(k-1)}}}$$

K_d = Effective coefficient of discharge

P_1 = Upstreamm relieving pressure in kPa

K_b = Capacity correction factor due to backpressure

K_c = Combination correction factor

T = Relieving temperature in K

Z = Compressibility factor

M = Molecular Weight

V = Flow in Nm^3 /min

G_v = Specific gravity

Table 3.1: Standardized Orifice Areas and Letter Designations

Designation	Orifice Area (in)
D	0.11
E	0.96
F	0.307
G	0.503
H	0.785
J	1.287
K	1.838
L	2.853
M	3.6
N	4.34
P	6.38
Q	11.05
R	16
T	26

3.3.5 Rated flow calculation

The orifice area has been calculated based on the actual relieving load. If the actual orifice area is greater than calculated orifice area, the rated relieving load has to be calculated(Adam Bader, *et al.*, 2011). The formulae to calculate the rated relieving rate is given in Appendix A. The tail pipe and flare header to be sized based on the governing case of rated flow.

Mass flow rate due to choked flow

$$= C_d A \sqrt{k \rho_0 P_0 \left(\frac{2}{k+1} \right)^{\left(\frac{k+1}{k-1} \right)}}$$

	SPRING-LOADED PRESSURE RELIEF VALVE SPECIFICATION SHEET		Sheet No. Page of
			Requisition No.
			Job No.
			Date
			Revised
			By
	GENERAL		**BASIS OF SELECTION**
1.	Item Number: *Example 1*	5.	Code: ASME VIII [X] Stamp Required: Yes [X] No [ ]
2.	Tag Number:		Other [ ] Specify:
3.	Service, Line, or Equipment Number:	6.	Comply with API 526: Yes [X] No [ ]
4.	Number Required:	7.	Fire [ ] Other [X] Specify:
		8.	Rupture Disk: Yes [ ] No [X]
	VALVE DESIGN		**MATERIALS**
9.	Design Type: XXXX	17.	Body:
	Conventional [X] Bellows [ ] Balanced Piston [ ]	18.	Bonnet:
10.	Nozzle Type: Full [X] Semi [ ]	19.	Seat (Nozzle): Disk:
	Other [ ] Specify:	20.	Resilient Seat:
11.	Bonnet Type: Open [ ] Closed [X]	21.	Guide:
12.	Seat Type: Metal-to-Metal [X] Resilient [ ]	22.	Adjusting Ring(s):
13.	Seat Tightness: API 527 [X]	23.	Spring: Washer:
	Other [ ] Specify:	24.	Bellows: *N/A*
		25.	Balanced Piston: *N/A*
	CONNECTIONS	26.	Comply with NACE: Yes [ ] No [X]
14.	Inlet Size: *4* Rating: *150* Facing: *RF*	27.	Internal Gasket Materials:
15.	Outlet Size: *6* Rating: *150* Facing: *RF*	28.	Other [ ] Specify:
16.	Other [ ] Specify:		
	SERVICE CONDITIONS		**ACCESSORIES**
34.	Fluid and State: *Hydrocarbon Vapor*	29.	Cap: Screwed [ ] Bolted [X]
35.	Required Capacity per Valve and Units: *53,500 lb/hr*	30.	Lifting Lever: Plain [ ] Packed [ ] None [X]
36.	Mass Flux and Basis:	31.	Test Gag: Yes [ ] No [X]
37.	Molecular Weight or Specific Gravity: 51	32.	Bug Screen: Yes [ ] No [X]
38.	Viscosity at Flowing Temperature and Units: ---	33.	Other [ ] Specify:
39.	Operating Pressure and Units: *50 psig*		
40.	Set Pressure and Units: *75 psig*		
41.	Blowdown: Standard [X] Other [ ]		
42.	Latent Heat of Vaporization and Units:		**SIZING AND SELECTION**
43.	Operating Temperature and Units: *100 °F*		
44.	Relieving Temperature and Units: *167 °F*	51.	Calculated Orifice Area (in square inches): *5.73*
45.	Built-up Back Pressure and Units:	52.	Selected Effective Orifice Area (in square inches): *6.38*
46.	Superimposed Back Pressure and Units: *0 psig*	53.	Orifice Designation (letter): *P*
47.	Cold Differential Test Pressure and Units:	54.	Manufacturer: *
48.	Allowable Overpressure in Percent or Units: *10*	55.	Model Number: *
49.	Compressibility Factor, Z: *0.90*	56.	Vendor Calculations Required: Yes [X] No [ ]
50.	Ratio of Specific Heats: *1.11*		

Figure 3.7: Sample of PRV specification sheet

3.4 Tail pipe sizing

Tail pipe is the outlet line of the PRV. Maximum load for each PRV is the governing case of that PRV. The outlet line is sized accorinding to that governing case load. The allowable mach number in the tail pipe is usually fixed to size the tail pipe. It depends upon the MOC and support of that pipe. Usually it is 0.7 in tail pipe. The pressure drop in the tail pipe also calculated(Peter Smith *et al.*, 2003). The tail pipe should join the flare header at the angle 45 degrees to minimise the pressure drop. There are various method to calculate the pressure drop.

3.5 Flare header sizing

Under segregation, usually two flare headers are provided to collect the flare releases from the various safety valves. These are generally called LP and HP flare headers. All safety valves with a low value for the maximum allowable back pressure are connected to the LP flare header, whereas safety valves which can take a higher back pressure are connected to the HP flare header. As flare headers are long uninsulated pipes carrying condensable vapors. The condensate needs to be separated from vapors before reaching the flare. For this purpose, flare headers are built with no pockets and continuously sloping (typical slope- 1:500) towards the knock out drum. Typically in a straight segment, expansion loops are installed at a center to center distance of 80-100 m for the impact of thermal stress(Scandpower, 2004; Dr.A. Alizadeh *et al.*, 2007). The allowable mach number in the flare header is based on the flare support. It is usually 0.5 in header. The flare header is sized according to governing case scenario. The governing case scenario is maximum load of each governing case. For example if 5526 kg/hr load for fire case and 15625 kg/hr load for reflux failure case, the governing case scenario is reflux failure case. The flare header is need not to be sized for double jeoperdy or two failure scenario at a time. The pressure drop in the flare header is calculated by various methods in the literature.

3.6 Pressure drop calculation

The pressure drop is need to be calculated from tail pipe to flare tip. The pressure in the entire flare network should be greater than atmosphere(API 521, 2014). The flare fluid will move towards the flare tip because of the positive pressure in the network. The pressure drop is calculated from one of the following methods.

3.6.1 Isothermal Pressure drop

$$\left(\frac{G}{a}^2\right) ln\left(\frac{P_1}{P_2}\right) + \frac{M(P_2^2 - P_1^2)}{2RT} + 2f\left(\frac{L}{\Phi}\right)\left(\frac{G}{a}\right)^2 = 0$$

Where

G = Mass flow

a = Cross sectional area of the pipe

P_1 = Upstream Pressure

P_2 = Downstream Pressure

R = Universal gas constant

f = Moody friction factor

ϕ = Internal diameter

L = Equivalent length

T = Temperature

M = Molecular weight

3.6.2 Adiabatic Pressure drop

$$4f\left(\frac{L}{\phi}\right) = \{\frac{\gamma - 1}{2\gamma} + \frac{P_1}{V_1}\left(\frac{a}{G}\right)^2\}\{1 - \left(\frac{V_1}{V_2}\right)^2\} - \frac{\gamma + 1}{\gamma} ln\frac{V_2}{V_1}$$

Where

$\gamma =$ Ratio of specific heats

3.7 Backpressure and Mach number calculation

The allowable backpressure in outlet of PRV is based on the type of PRV. The actual backpressure should not exceed the allowable backpressure. The actual backpressure in the tail pipe and header for particular failure scenario is calculated by

$$\frac{fl}{d} = \frac{1}{Ma_1^2}[1 - (\tfrac{P_2}{P_1})^2] - ln(\frac{P_1}{P_2})^2$$

and

$$\frac{fl}{d} = \frac{1}{Ma_2^2}[\frac{P_1}{P_2}][1 - (\tfrac{P_2}{P_1})^2] - ln(\frac{P_1}{P_2})^2$$

Where

$f =$ Moody friction factor

$l =$ Equivalent length

$d =$ Pipe inside diameter

$Ma_1 =$ Mach number at pipe inlet

$Ma_2 =$ Mach number at pipe outlet

$P_1 =$ Upstream pressure in kPa

$P_2 =$ Downstream pressure in kPa

$$Ma_2 = 3.23 \times 10^{-5} \left(\frac{G}{P_2 d^2}\right) \left(\frac{ZT}{M}\right)^{0.5}$$

From the combination of Mach number and ratio of upstream and downstream pressure, the diameter of the pipe is calculated after some iterations.

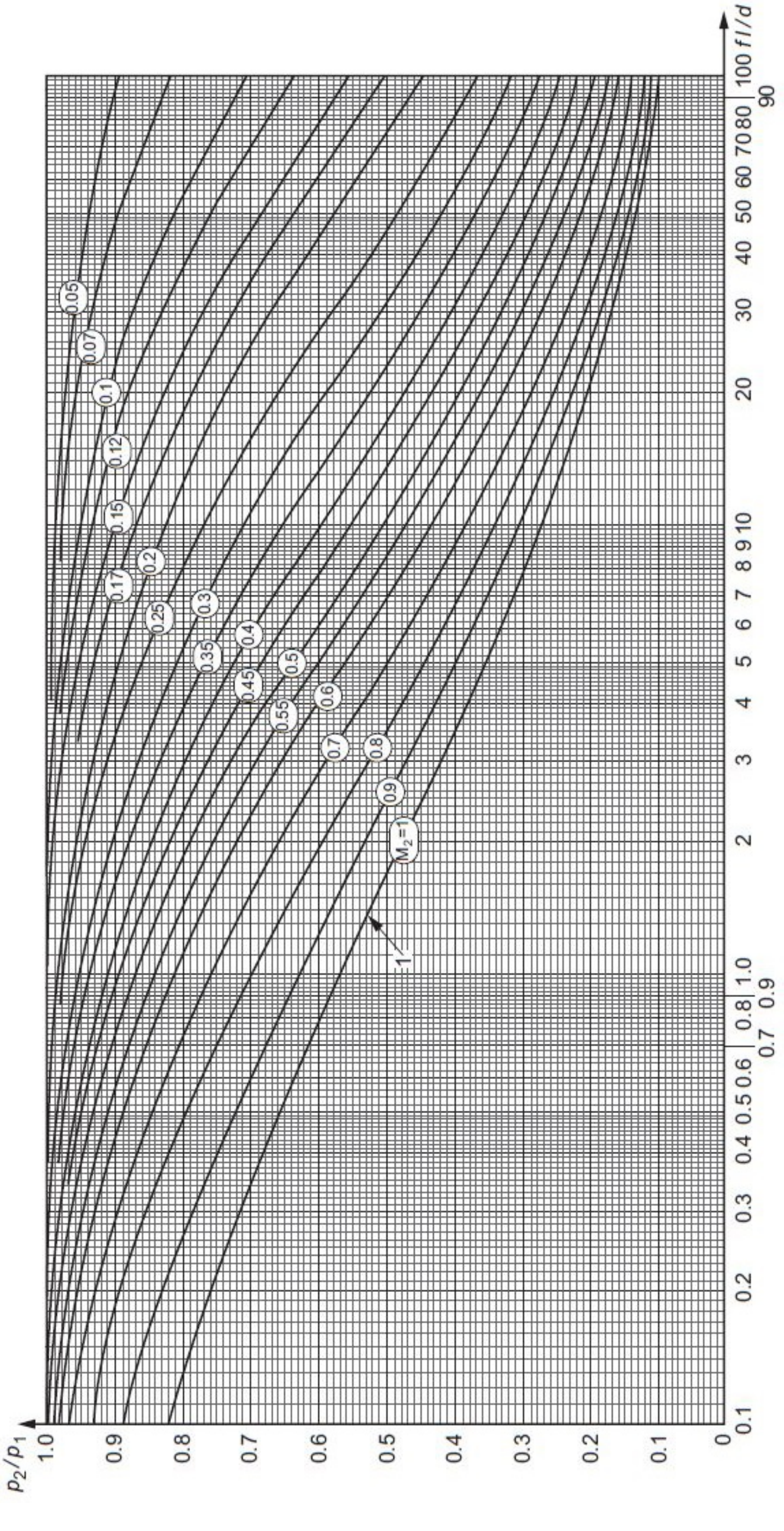


Figure 3.8: Isothermal flow chart

CHAPTER 4

KNOCKOUT DRUM DESIGN

Flare systems generally require a flare knockout drum to separate liquid from gas in the flare system and to hold the maximum amount of liquid that can be relieved during an emergency situation.

4.1 Types of Knockout drum

The cost of drum design can influence the choice between a horizontal and a vertical drum. If a large liquid storage capacity is desired and the vapor flow is high, a horizontal drum is often more economical. Also, the pressure drop across horizontal drums is generally the lowest of all the designs. Vertical knockout drums are typically used if the liquid load is low or limited plot space is available. They are well suited for incorporating into the base of the flare stack.
The various types are

- Horizontal drum with the vapor entering at the top end the vessel and exiting at the another end of the vessel with no internal baffling.
- Vertical drum with the vapor inlet nozzle entering the vessel radially and exiting at top of the vessel in vertical axis.
- Vertical vessel with a tangential nozzle.
- Horizontal drum with the vapor entering at each end exiting at center of the drum.
- Horizontal drum with the vapor entering at the center and exiting at each end.
- Combination of vertical drum and horizontal drum.

The typical horizontal KOD is shown in figure 4.1

Key

1 - vapor and liquid pressure relief valve releases

2 - level instrument

3 - minimum vapor space for dropout velocity

4 - liquid holdup from pressure relief valves and other emergency releases

5 - slop and drain liquid

6 - to flare

7 - pump out

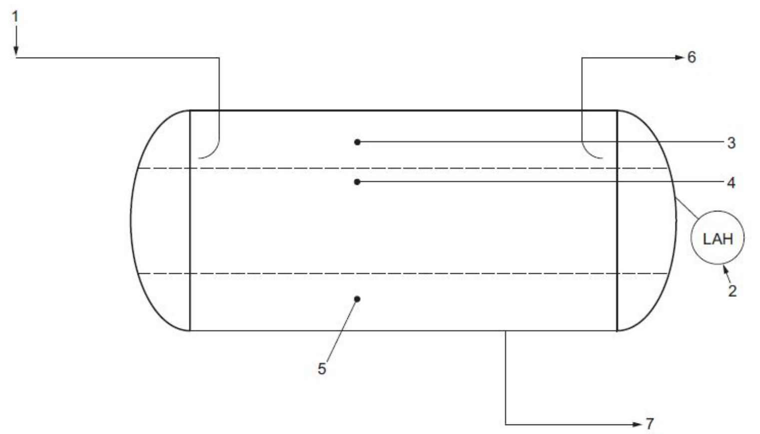


Figure 4.1: Flare Horizontal Knockout Drum

4.2 Droplet size criteria

The function of the knockout drum is to provide residence time for liquid discharges and to limit the size of droplets directed to the flare burner. Large liquid droplets and liquid loading can cause smoke.The phenomenon is called as burning rain, occurs when a liquid hydrocarbon droplet does not burn completely within the flare flame envelope and the rate of burning is lower than the rate of settling of the liquid droplet. Liquid droplets 300μm and larger may drop out of the gas stream at less than 2 m/s. If liquids are not drained from the system, flare flows with gas velocities exceeding about 3 m/s or 4 m/s can entrain liquid droplets up to 1000μm in size. Liquid droplets exceeding 1000 μm can readily lead to burning rain regardless of flare type. Burning rain can occur at smaller droplet sizes for some flare types(Daniel A.Crowl *et al.*, 2013).

4.3 Designing KOD

Sizing a knockout drum is generally a trial-and-error process. The first step is to determine the drum size. The KOD sizing procedure for simple horizontal configuration is described below

1. The governing load has been calculated for the failure case scenarios.
2. Calculated the fluid equilibrium. Usually 10% of the total load is liquid.
3. The temperature of the governing stream is calculated.
4. The pressure of the KOD is calculated by back calculation from flare tip. The pressure at the flare tip is considered as slightly greater than atmoshpheric pressure. Then the pressure drop is calculated for flare stack, liquid seal drum, molecular seal drum, fittings and joints and other components. Then the pressure at the oulet and inlet of KOD is calculated.
5. The viscosity, Density, Specific gravity and Molecular weight of the fluid is calculated.
6. The dead volume of the KOD is considered based on requirements.
7. The volumetric flow rate of vapour is calculated.
8. The particle diameter is calculated based on requirements. Usually it is assumed as $300\mu m$.
9. The $C(Re)^2$ is calculated from the set of equations.
   In SI units,

$$C(Re)^2 = \frac{0.13 \times 10^8 \rho_v (pd)^3 (\rho_l - \rho_v)}{\mu^2}$$

In USC units,

$$C(Re)^2 = \frac{0.95 \times 10^8 \rho_v (pd)^3 (\rho_l - \rho_v)}{\mu^2}$$

Where C = Drag Coefficient

Re = Reynolds number

pd = Particle diameter

$\mu =$ Viscosity

$\rho_v =$ Density of gas

$\rho_l =$ Density of liquid

10. The Drag Co-efficient, C is determined from the drag co-efficient graph.
11. The residence time for the fluid from low liquid level to High liquid level in KOD is 30 minutes. The liquid holdup for 30 minutes is determined.
12. The Cross sectional area for liquid and vapour flow is determined by assuming the length and diameter of the drum.
13. Check L/D ratio. The L/D ratio should be less than 6 and greater than 5.
14. The vertical height of the liquid level in horizontal drum is calculated based on the geometry of the vessel.
15. The dropout velocity of the stream is calculated using this equation,

$$u_c = 1.15 \sqrt{\frac{gD(\rho_1 - \rho_v)}{\rho_v C}}$$

16. The liquid dropout time is calculated using this equation,
    In SI units,

$\theta = (\frac{h_v}{100})(\frac{1}{100})$

In USC units,

$$\theta = (\frac{h_v}{12})(\frac{1}{100})$$

17. The vapour velocity is determined from volumetric flow rate.

    In SI units,

    $$u_v = (\frac{7.34}{N})(\frac{1}{A_v})$$

    In USC units,

    $$u_v = (\frac{260}{N})(\frac{1}{A_v})$$

    Where N = Number of Passes

18. The Minimum required length L_{min} is determined.

    $$L_{min} = u_v \theta N$$

The assumed length should be greater than or equal to the minimum required length. The snapshot of this calculation can be seen in the figure. The model calculation for KOD Sizing is in Appendix. In vertical vessel, the vapour velocity is equal to dropout velocity. The required cross sectional area is determined by dividing volumetric flow rate by dropout velocity.

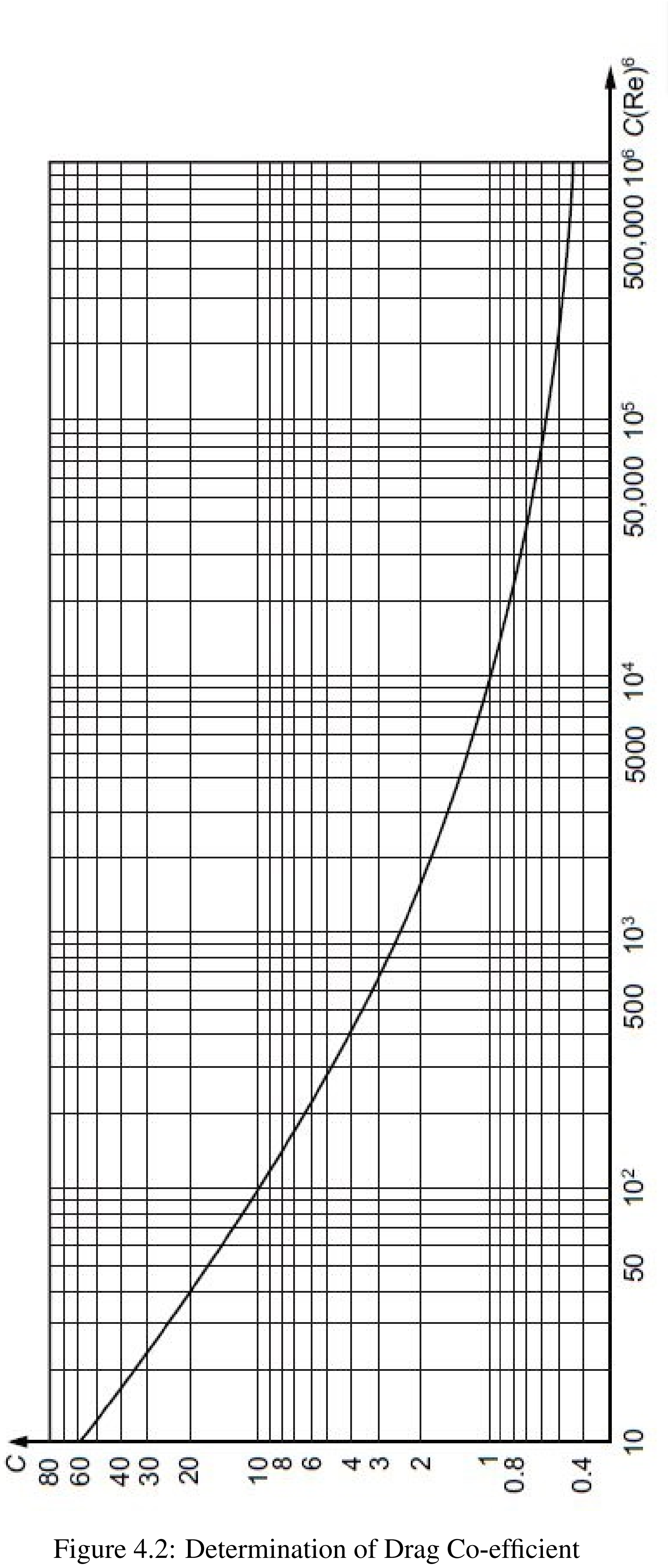
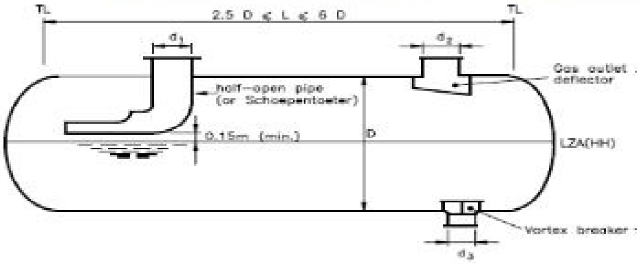


Figure 4.2: Determination of Drag Co-efficient

Horizontal Knockout Drum Sizing

Project		**Unit**	
Type		**Tag No**	



Input			
INPUT	**Units**	**Horizontal Drums**	**Guidelines**
Gas Flow			
Vapour flow rate	kg/hr	36769	
Design Margin, %		20	For Refineries - 10% to 50%, Different margins can be used for gas & liquid respectively based on the requirement.
Density	kg/m³	8.718	
$\kappa = C_p/C_v$		1.09	Value of C_p/C_v shall be between 1.0 to 1.4. If value k is not available, input
Molecular weight of gas		56.2	
Temperature of gas	K	422.35	
Liquid Flow			
Liquid Flow Rate	kg/hr	3676.9	
Design Margin, %		20	For Refineries - 10% to 50%, Different margins can be used for gas & liquid respectively based on the requirement.
Density	kg/m³	799.6	
viscosity	cP	1.4	
Nozzle Sizes			
Feed Inlet	in	20	
Gas Outlet	in	20	
Liquid Outlet	in	12	
Vessel Dimensions			
Enter minimum diameter, D	m	4.2	Initial values to be assumed such that 2.5 < L/D < 6.
Enter length of the vessel, L	m	16	
Value of L/D		3.81	L/D Criteria Met.
Type of vessel head		Semi-Elliptical Head	The most common type of head is the semi-elliptical head.
Type of feed inlet device		Half-Open Pipe	
Residence Times			
Control time for LLLL to LLL	sec	300	
Control time for LLL to HLL	sec	1800	
Control time for HLL to HHLL	sec	300	
OUTPUT	**Units**	**Horizontal Drums**	**Shell Standards**
Calculated length of the vessel	m	17.20	Vessel length assumed initially is correct. Proceed with assumed length value.
Check for gas handling criteria	Gas handling criteria met		The separator must be large enough to handle the gas flow rate under the most severe process conditions. Moreover to limit liquid carry-over, the vessel dia should satisfy the gas handling criterion.
Sizing of feed inlet nozzle	Nozzle Size is Correct		Sizing of the nozzles is based on maximum flow rates, including the appropriate design margin. Sizes of the nozzle may vary depending on the type of feed inlet
Sizing of gas outlet nozzle	Nozzle Size is Correct		When the gas outlet velocity is very high, it may lead to a high pressure drop. In that case, gas outlet shall be sized such that the pressure drop requirements are satisfied.
Sizing of liquid outlet nozzle	Nozzle Size is Correct		The liquid outlet nozzle dia shall be chosen such that the liquid velocity does not exceed 1 m/sec.
Pressure drop across inlet and vapor outlet	Pa	323.2201784	

Figure 4.3: Snapshot of Horizontal KOD Calculation

CHAPTER 5

LIQUID SEAL DRUM

5.1 Purpose of liquid seal drum

The purpose of the liquid seal drum in flare system is to prevent flashback originating from the flare tip from propagating back through the flare system and to maintain positive pressure to ensure no air leakage into the flare system. The iquid seal also used to prevent air ingress into flare system during sudden temperature changes or condensation of flare gas.

To prevent air entry, it is necessary that the seal dip-leg height and the density and amount of seal liquid within the drum be sufficient to prevent the seal from being broken as a result of the vacuum formed in the flare header.The physical dip-leg height is measured from the top opening of the seal head or end piece to the bottom of the horizontal section of the flare header piping immediately upstream of the inlet leg(Daniel A.Crowl *et al.*, 2013).The seal drum should be designed to provide the volume of liquid to fill the vertical seal leg up to the specified vacuum. Experience has shown that a minimum dip-leg height of 3 m (10 ft) above the liquid level is effective in minimizing the ingress of air into the flare header from flare stacks for typical refining applications.

5.2 Liquid seal selection

Liquid seals typically use water as the seal medium, however, other fluids are possible. Fluid selection requires consideration of freeze protection in cold climates, hydrocarbon/water separation, implications of carryover, compatibility with the relief stream, cost, availability, and disposal(Daniel A.Crowl *et al.*, 2013). In facilities that have cryogenic products released into the flare header, consideration should be given to the effect of the

cold material on the seal medium. Water seals are not recommended if there is a risk of obstructing the flare system due to an ice plug. Alternate sealing fluids such as a glycol/water mixture or other means to prevent freezing can be required.
Consideration should be given to providing a continuous flow of seal fluid (typical for water seals), which allows for the continuous skimming of hydrocarbons as well as maintaining liquid level. Proper liquid seal drum operation is dependent upon maintaining the design liquid level in the seal. Routine surveillance and hydrocarbon skimming, if applicable, are required to ensure proper seal operation. Seal drums that overflow to open sewer should be evaluated as to whether condensed flammable and toxic vapors can be discharged and the need to provide suitable containment and mitigation systems.

5.3 Liquid drum sizing

Sizing depends on whether the drum is horizontal or vertical. The minimum diameter of a vertical seal drum is determined by the total seal liquid volume.

$$D = d\sqrt{(\frac{H}{h} + 1)}$$

For a horizontal drum this criteria typically does not control the drum diameter because the length of the drum can easily be adjusted to get sufficient seal liquid volume. The volume of seal fluid in a horizontal seal drum shall be adequate to fill the dip-leg(Daniel A.Crowl *et al.*, 2013). Optimization of the L and w dimensions can be done subsequently.

$$L.w = \frac{\pi}{4}d^2\frac{H}{h}$$

$$h = \frac{102p}{\rho}$$

The area for the gas flow above the liquid level should be at least three times the inlet pipe cross-sectional area to prevent surges of gas flow to the flare. The diameter of a vertical drum based upon avoiding pulsing is determined by providing an area ratio 1:3.
$D^2 - d^2 = 3d^2$

$D = 2d$

The typical horizontal liquid seal drum is shown in figure 5.1.

Key

1 - to flare

2 - flare header

3 - try cocks used to check for hydrocarbons

4 - vent

5 - to sewer

6 - water supply

7 - submerged weir welded on end of flare line

8 - water level

9 - baffle

10 - drain

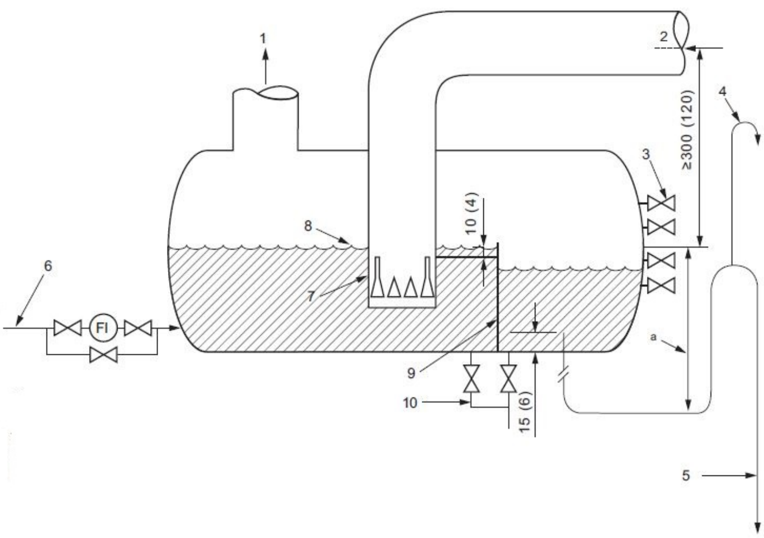


Figure 5.1: Horizontal Liquid Seal Drum

CHAPTER 6

FLARE STACK

6.1 Thermal Radiation

Flare system design and plant equipment layout should minimize the need for operator attendance and equipment installed in locations of high radiant heat intensity. The design of towers or other elevated structures exposed to flare radiation should consider radiation effects on the ability to safely egress.If personnel exposure to radiant heat exceeds the set of guidelines, then shielding or other protection should be considered. It is often most effective to accomplish this by locating ladders and platforms on a side away from the flare. Personnel are commonly protected from high thermal radiation intensity by restricting access to any area where the thermal radiation can exceed $6.31 kW/m^2$. The boundary of a restricted access area can be marked with signage warning of the potential thermal radiation exposure hazard(Daniel A.Crowl *et al.*, 2013). There are practical differences between laboratory tests and full-scale field exposure. Heat radiation is frequently the controlling factor in the spacing of equipment such as elevated and ground flares. The table presents recommended design total radiation levels for personnel at grade or on adjacent platforms. The extent and use of personal protective equipment can be considered as a practical way of extending the times of exposure beyond those listed. Each company may select the radiation level to which personnel can be exposed, either for a short duration or continuously. The wind and ambient temperature can influence the amount of radiation a person can withstand.

6.2 Flare Height and Flare tip diameter

The flare tip is a proprietary device. The flare tip design is based on the net heating value, maximum velocity of flare gas, composition of flare gas, steam rate, auxillary fuel

Permissible Design Level (K) Btu/h·ft^2			Conditions
wo Solar	Solar	Total	
2,700	300	3,000	Maximum radiant heat intensity at any location where urgent emergency action by personnel is required.
1,700	300	2,000	Maximum radiant heat intensity in areas where emergency actions lasting up to 30 s can be required.
1,200	300	1,500	Maximum radiant heat intensity in areas where emergency actions lasting 2 min to 3 min can be required.
200	300	500	Maximum radiant heat intensity at any location where personnel with appropriate clothing a can be continuously exposed.

Figure 6.1: Recommended Design Thermal Radiation

rate and ignition systems(Adam Bader, *et al.*, 2011). The flare tip is a complex design. The flare tip diameter is calculated from maximum allowable velocity and total flow rate including auxillary fuel rate.

Table 6.1: Maximum allowable velocity

Net heating value of Flare gas(Btu/scf)	**Maximum velocity V$_{max}$(ft/s)**
300	60
300 - 1000	$\log_{10} \mathrm{V}_{max} = \dfrac{(B_v + 1214)}{852}$
$>$1000	400

Minimum flare tip Diameter, $\mathrm{D}_t = 1.95\sqrt{\dfrac{Q_{tot}}{V_{max}}}$

The flare tip diameter is rounded upto the next commercially available size. The minumum diameter is 1 inch. Larger sizes are available in 2 inch increments from 2 to 24 inches and in 6-inch increments above 24 inches. The maximum size commercially available is 90 inches.

The height of a flare is based on the ground level limitations of thermal radiation intensity, luminosity, noise, height of surrounding structures, and the dispersion of the exhaust gases. In addition, consideration must also be given for plume dispersion. Industrial flares are normally sized for a maximum heat intensity of 1500-2000 Btu/hr-ft^2 when flaring at

their maximum design rates. The intensity of solar radiation is in the range of 250-330 Btu/hr-ft ^{2}. The flare height also depends on wind effect and safely dispersion of flare gas. However the wind effect is reduced by flare support structure like self supported, derrick supported and guy supported flare. The safely dispersion is acheived by plume dispersion model. If the flare stack height increases, the thermal radiation intensity and the noise decreases in the ground level. Usually upto 90 m , there should be no hydrocarbon processing unit and anyother flammable material. This 90 m may vary depends on the height of the flare stack.
Heat transfer from flare tip is propagated through three mechanisms: conduction, convection, and radiation. Thermal radiation may be either absorbed, reflected, or transmitted. Since the atmosphere is not a perfect vacuum, a fraction of the heat radiated is not transmitted due to atmospheric absorption due to humidity(Adam Bader, *et al.*, 2011). In general, the fraction of heat radiated increases as the stack diameter increases. The heat release from the flare tip is calculated from the flare gas flow rate and net heating value. From the heat release and fraction of radiation, the thermal radiation intensity in ground level is calculated.

$$\text{Heat release} = Q_{tot}\ B_v$$

6.3 Flare Diameter Calculation

The Flare diameter is determined from Mach number in the flare stack. The recommended Mach number for continuous flare is 0.2 and for emergency flare is 0.5. The maximum mach number is 0.7. The Flare stack sizing procedure for continuous flare is described below,

1. The governing load has been calculated for the failure case scenarios.

Table 6.2: Fraction of heat radiation

Gas	Flare dip diameter(in)	Fraction of heat radiated
Hydrogen	>1	0.1
	1.6	0.11
	3.3	1.6
	8	1.5
	16	1.7
Butane	>1	0.29
	1.6	0.29
	3.3	0.29
	8	0.28
	16	0.3
Methane	>1	0.16
	1.6	0.16
	3.3	0.15
Natural gas	8	0.19
	16	0.23

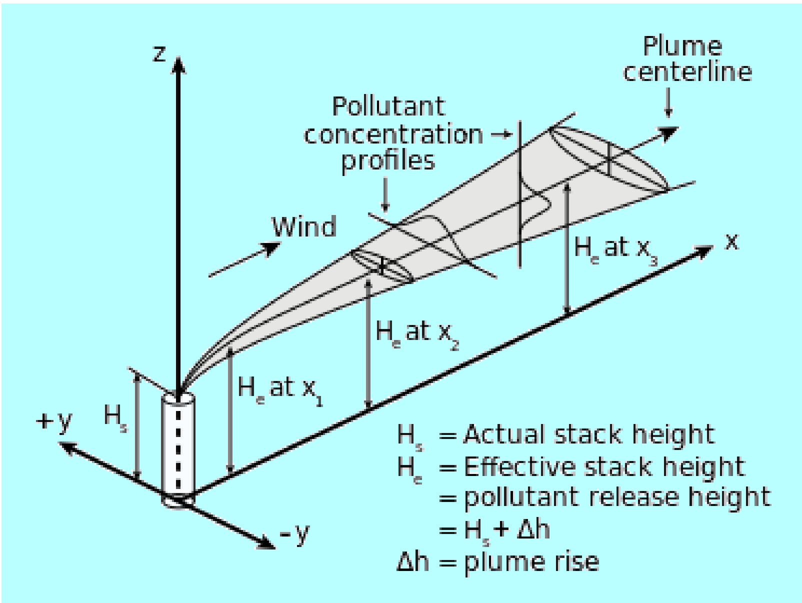


Figure 6.2: Plume dispersion model

2. The Average molecular weight, Temperature, Heat of Combustion and specific heat ratio of that stream is calculated.

3. The Mach number is assumed as 0.2 for continuous flare system.

4. The Maximum Wind velocity is observed.

5. The Maximum allowable radiation at the ground level is fixed as $2000 Btu/hr.ft^2$

($6.31 kW/m^2$) for continuous flare system.

6. The Ambient air temperature and relative humidity is calculated.
7. The inside diameter is calculated from the mach number,

$$Ma = 3.23 \times 10^{-5} [\frac{q_m}{p_2 d^2}][\frac{ZT}{M}]^{0.5}$$

8. Calculated the heat liberated from the heat load.
9. From the heat liberated, the flame length is calculated by the graph.
10. The vapour volumetric flowrate and the flare tip velocity is calculated.

$$u_j = \frac{q}{\frac{\pi d^2}{4}}$$

11. The ratio of the wind velocity and flare tip velocity is calculated.

$$\frac{u_\infty}{u_j}$$

12. From the ratio, the flame perimeters $\Sigma\frac{\Delta x}{L}$, $\Sigma\frac{\Delta y}{L}$ are calculated by the graph.
13. The distance from the flame center to the grade level boundary, D is then calculated by the equation.

$$D = \sqrt{\frac{\tau F Q}{4\pi K}}$$

Where D = minimum distance from the epicenter of the flame to the object being considered.

τ = fraction of the radiated heat transmitted through the atmosphere.

F = fraction of heat radiated.

Q = Heat released, kW.

K = Heat Intensity,kW/m^2.

14. Then distance from the flare and object being considered in the ground is calculated from the flare perimeters.
    The height of the flare stack and flame is calculated.

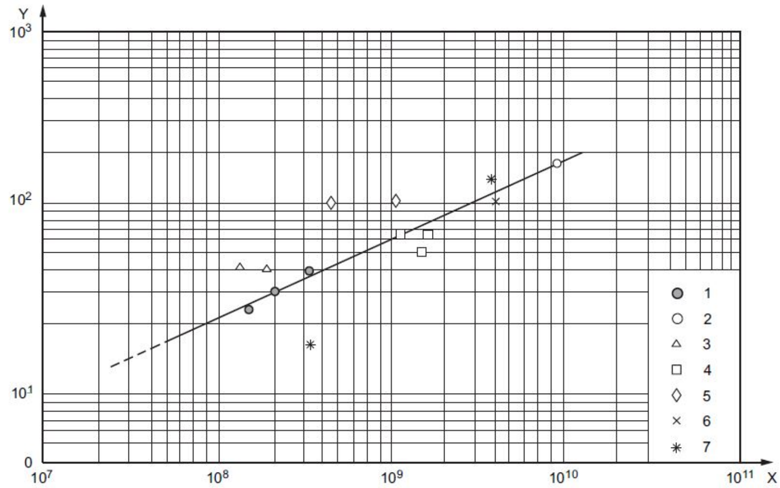


Figure 6.3: Flame Length vs Heat Release

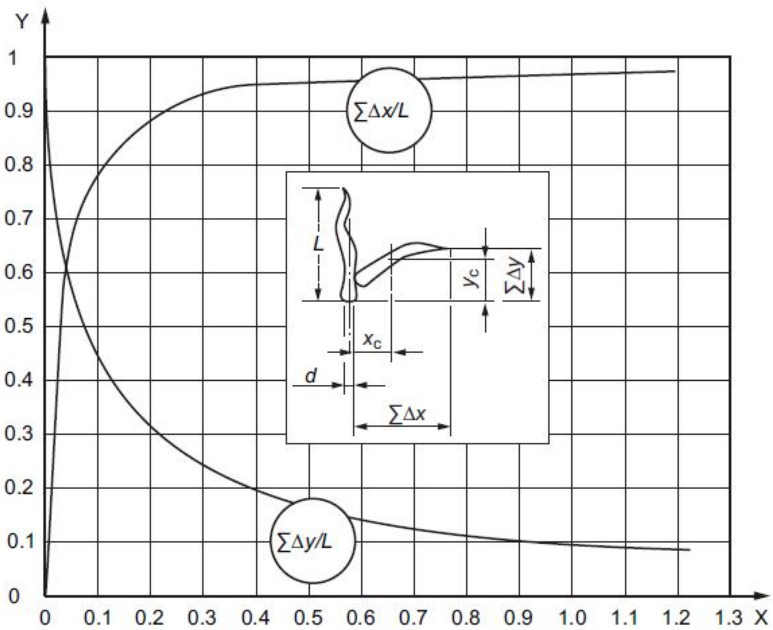


Figure 6.4: Flame Distortion Due to Wind Velocity

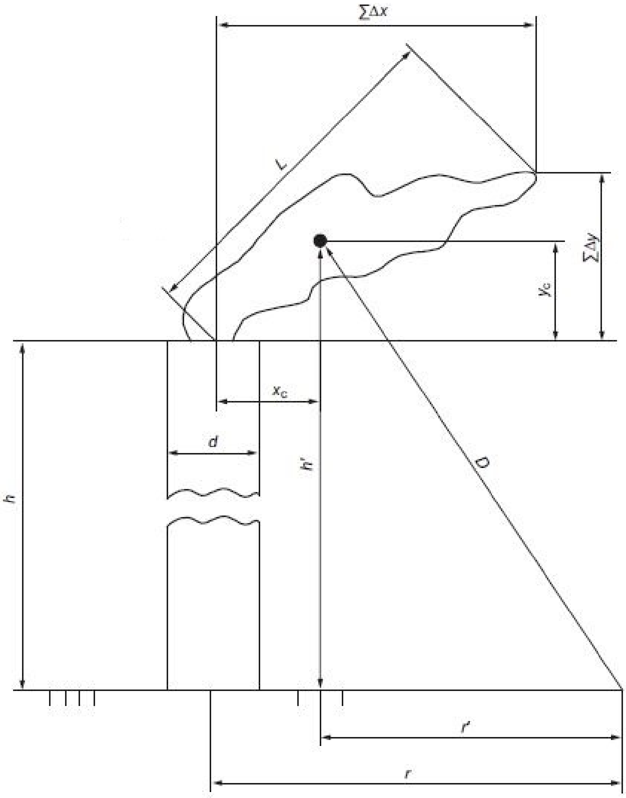
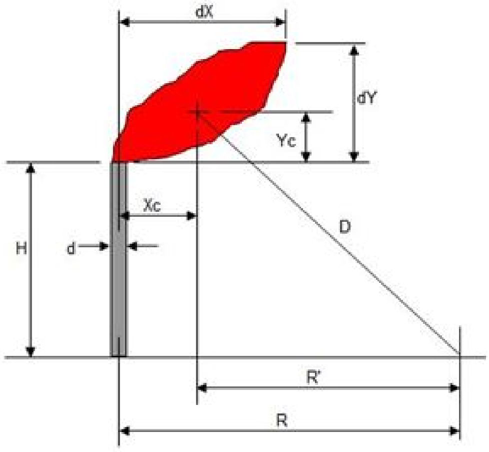


Figure 6.5: Sizing a Flare Stack

Flare Stack Sizing

Project		**Unit**	
Type		**Tag No**	



INPUT DATA		
Material Flowing In Stack	**HC VAPOUR**	
Flow Rate, kg/hr	**45,360**	
Average Molecular Weight	**46.1**	
Heat Of Combustion, kJ/kg	**50,000**	
Specific Heat Ratio, k Cp/Cv	**1.1**	
Design Mach Number For Flare	**0.2**	
Design Wind Velocity m/s	**8.94**	
Fraction Of Heat Radiated	**0.3**	
Max Allowable Radiation, K, at Dist R , BTU/hr/sf	**2,000**	
Distance, R, To Max Allowable Radiation ,Ft	**150**	
Relative Humidity	**50.0%**	
CALCULATION OF FLARE DIAMETER		
Specified Mach Number	0.2	
Diameter Squared	0.219	
Flare Diameter, m	0.468	
CALCULATION FOR FLAME LENGTH		
Heat Liberated, Btu/hr	2.15E+09	
Flame Length, m	50.0	
CALCULATION OF FLAME DISTORTION CAUSED BY WIND VELOCITY		
Wind Speed,m/s	8.94	
Actual Exit Gas Velocity, m/s	55	
Ratio Wind Velocity / Exit Velocity	0.162	
SUM Delta X / Length	0.85	
SUM Delta X , m	42.50	
SUM Delta Y / Length	0.3600	
SUM Delta Y, m	18.0	
CALCULATE REQUIRED FLARE STACK HEIGHT		
Specified Max Allow Radiation, BTU/hr/sf	2,000	
D	48.90	m
R'	24.40	m
H'	42.30	m
H	33.30	m

Figure 6.6: Snapshot of Stack Sizing

CHAPTER 7

ASPEN FLARENET MODEL

7.1 Introduction

The design of flare and vent system piping is an important part of the overall system design for any chemical process. FLARENET has been designed to facilitate the design and rating of flare and vent system piping throughout the entire design process. The program interface uses a flow diagram for direct visualisation of the piping network. This is supported by detailed tables of all pertinent data and calculated results. FLARENET can model the piping system topologies most commonly found in flare systems. Multiple relief scenarios such as ”Plantwide Power Failure”, ”Plantwide Cooling Water Failure” and ”Localised Fire” cases, as well as the individual relief valve loads can be maintained within a single file model of the flare system. The following calculations can be done simply from a consistent data set(Adam Bader, *et al.*, 2011).

- Design of an entire new flare system for a single relief scenario.
- Design of an entire new flare system for all relief scenarios.
- Debottlenecking design of an entire flare system for a single and multiple relief scenario.
- Rating of an entire flare system for a single and multiple relief scenario.

FLARENET has the option to calculate the pressure profiles using a range of single and two-phase pressure drop calculation methods. These methods may be used globally throughout the model or specified at a local level. Robust multiphase thermodynamic models back up the physical property predictions used by the pressure drop models.

7.2 Data Requirements

Before starting to build Flarenet Model, the data are defined to determine our system. The basic data is inlet pressure, inlet temperature, allowable backpressure, relieving pressure, type of orifice, allowable mach number, failure case scenarios, length and elevation of pipe, type of fittings, total mass flow, composition of the stream, methods to calculate pressure loss, properties of stream, properties of pipe and parameters of KOD and flarestack. The data input of PSV, Control valve, pipe, KOD and flare stack is shown in figure 7.1 7.2 7.3.

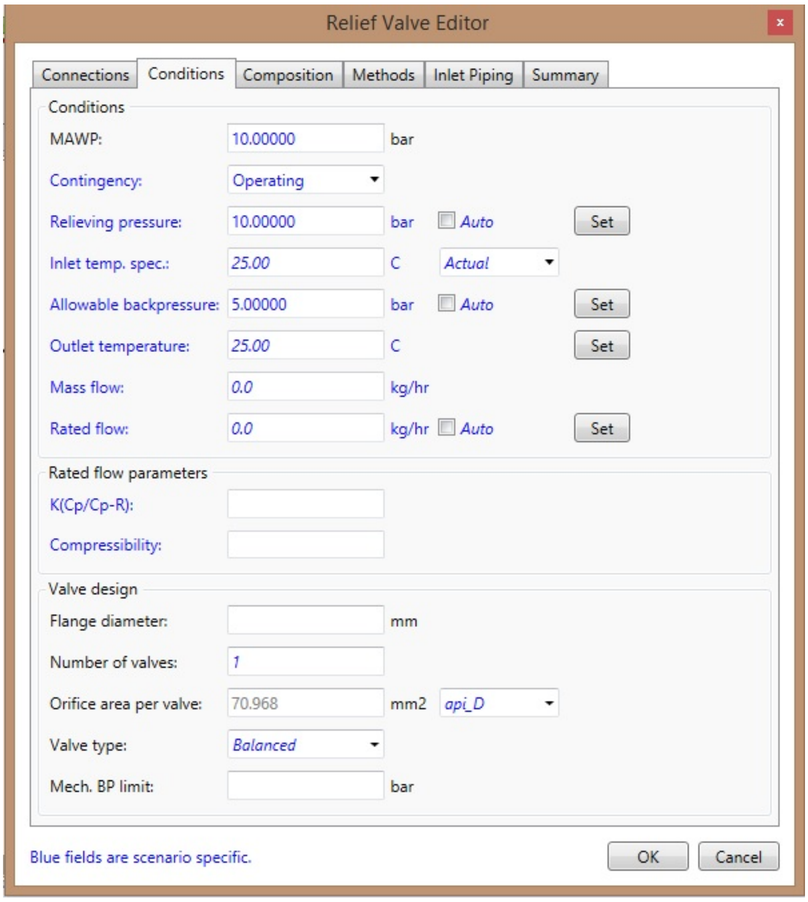


Figure 7.1: Relief Valve Editor

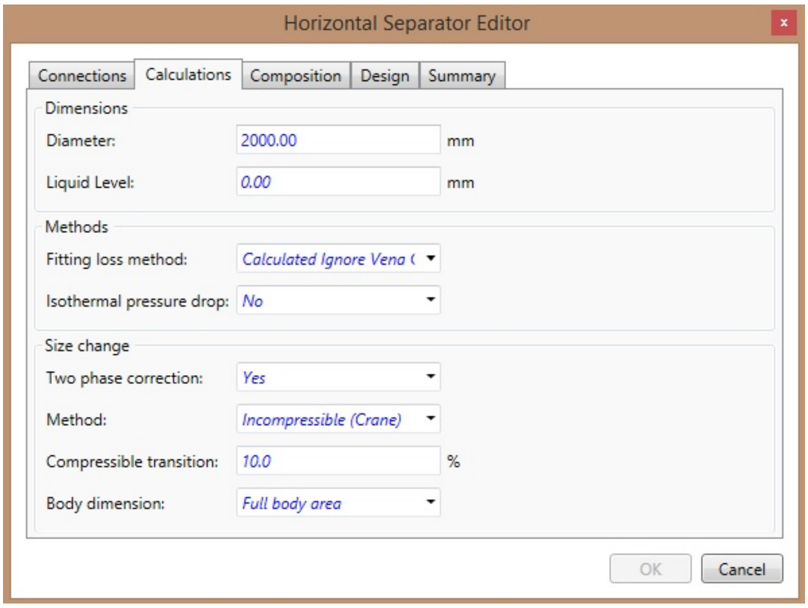


Figure 7.2: Knockout Drum Editor

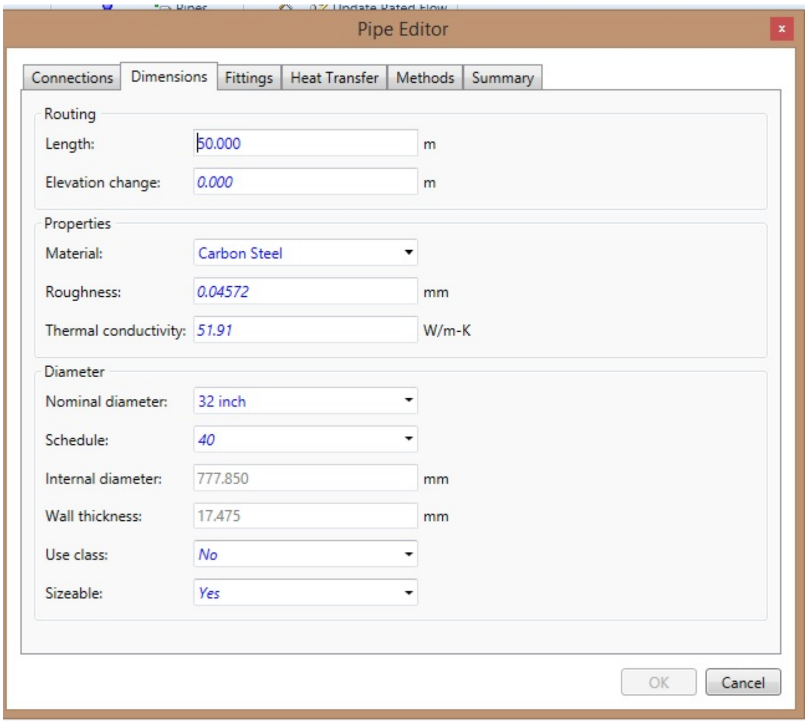


Figure 7.3: Pipe Editor

7.3 Viewing Data and Results

The results can be viewed in the form of tables, graphs and charts. Data can be interpreted from the graphs. Typical graphs are shown in figure 7.4, 7.5, 7.6.

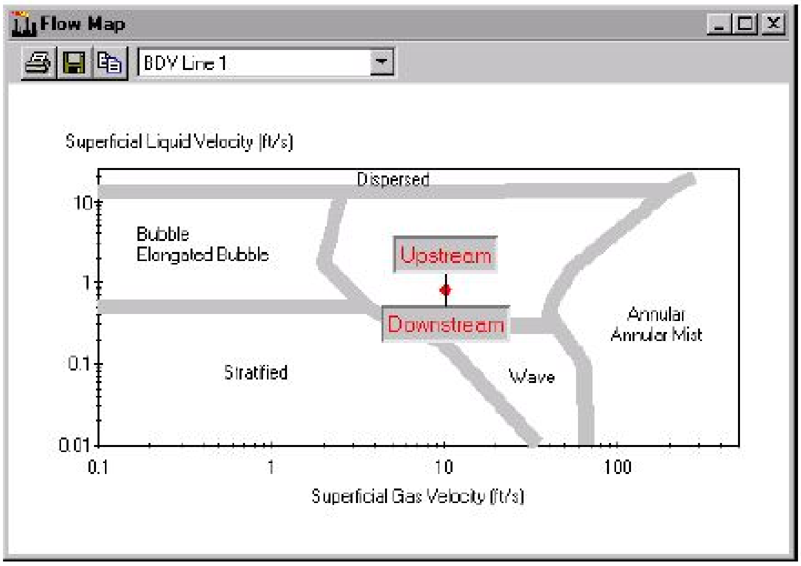


Figure 7.4: Flow Map

Scenario Summary — Source 1

Scenario	Default Scenario	Source 1 Only	Source 2 Only
Ignored	No	No	Yes
Mass Flow (kg/hr)	100000.0	100000.0	0.0
Rated Flow (kg/hr)			
Upstream Pressure (bar abs)	10.00000	10.00000	10.00000
Upstream Temperature (C)	15.00	15.00	15.00
Downstream Static Pressure (bar a	1.60888	1.14464	
Allowable Backpressure (bar abs)	5.00000	5.00000	5.00000
Downstream Temperature (C)	15.00	15.00	
Mol. Wt.	20.000	20.000	20.000
Composition Basis	Mol. Wt.	Mol. Wt.	Mol. Wt.
Methane	0.717894	0.717894	0.717894
Ethane	0.282106	0.282106	0.282106
Propane	0.000000	0.000000	0.000000

Figure 7.5: Scenario Summary

Physical Properties

Name	Phase	Upsteam Density kg/m3	Upstream Enthalpy kJ/kgmole	Upstream Entropy kJ/kgmole/K	Upstream Fraction
BDV 1 Discharge Line	F	6.330	9286.98	162.941	1.00000
HP Flare Stack Riser	F	4.227	11947.42	177.714	1.00000
HP Header 1	F	27.264	2216.46	176.185	1.00000
HP Header 11	F	3.583	12721.58	175.747	1.00000
HP Header 2	F	43.417	-23175.32	91.674	1.00000
HP Header 3	F	17.541	3335.39	193.133	1.00000
HP Header 4	F	9.739	6618.17	177.993	1.00000
HP Header 5	F	9.414	6616.50	178.219	1.00000
HP Header 6	F	9.074	6614.85	178.465	1.00000
HP Header 7	F	5.330	11843.46	176.109	1.00000
HP Header 8	F	4.997	11843.46	176.636	1.00000

Figure 7.6: Physical Properties

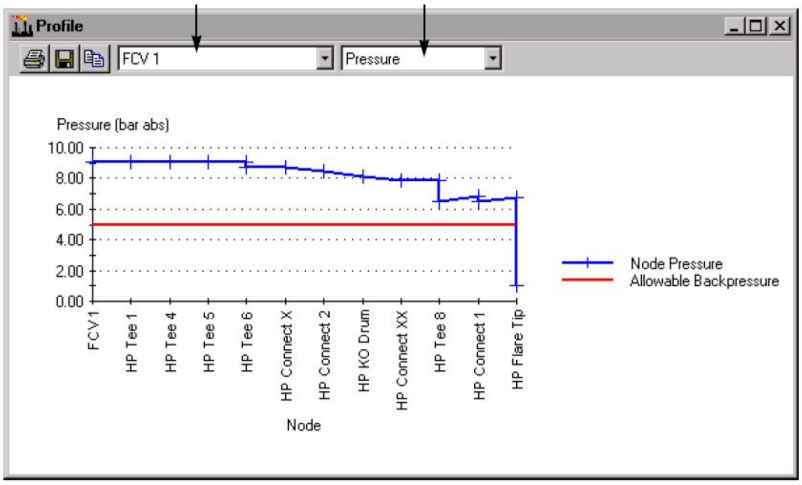


Figure 7.7: Pressure Profile

7.4 Flarenet Model

Based on the data availabe, the flarenet model has been built. Analysed the results in the form of tables and graphs. The snapshot of that model is shown in figure 7.8.

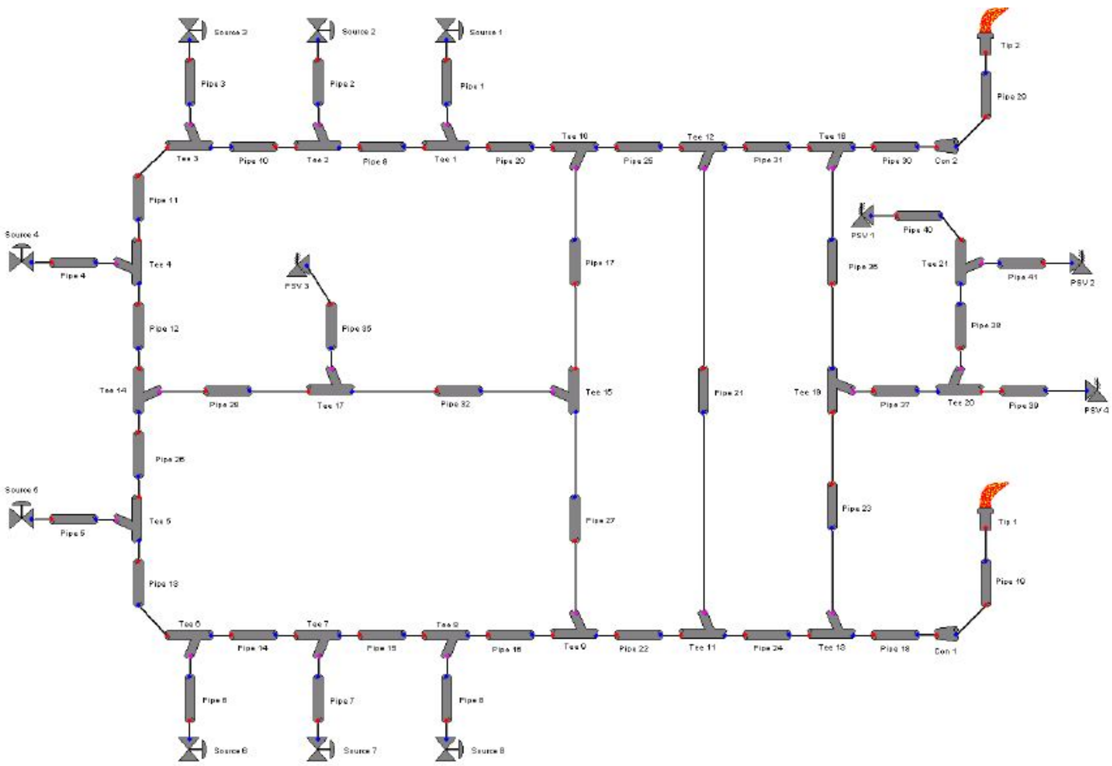


Figure 7.8: Flarenet Model

CHAPTER 8

DISPOSAL TO FLARE

8.1 Auxillary fuel requirement

The flares are provided with auxillary fuel to combust hydrocarbon vapours when the flare gas stream falls below the flammability range to sustain a stable flame. The amount of fuel is calculated based on the flare gas stream net heating value(Adam Bader, *et al.*, 2011) Typically the natural gas has a net heating value of 1000 Btu/scf. Automatic control of the auxillary fuel is recommended.

$$\text{Required fuel, F} = G\frac{300 - B_v}{B_f - 300}$$

Where

B_v and B_f are the heating value of flare gas and fuel gas respectively in Btu/scf.

8.2 Purge gas requirement

The purge gas is used to maintain the positive pressure and to prevent the backfire, flashback and flame instability. Purge gas is inert gas usually N_2 . If there is a possibility of air formation in the leakage of flare manifold, the purge gas will prevent the formation of an explosive mixture in the flare system(Adam Bader, *et al.*, 2011). The purge gas requirement is determined by the design of flare seals, which are usually proprietary devices.

Purge flow rate, $F_{pu} = 190.8D^{3.46}\frac{1}{y}ln(\frac{20.9}{O_2})\Sigma C_i^{0.65}K_i$

Where D = Flare stack diameter.

y = Column width

O_2 = Oxygen volume fraction.

C_i = Volume fraction of Component i .

K_i = Constant for Component i.

This equation can be simplified to

$F_{pu} = 31.25D^{3.46}K$

Table 8.1: Constants for common used flare gas

Gas	**K**
Hydrogen	5.783
Helium	5.078
Methane	2.328
Nitrogen	1.067
Ethane	-1.067
Propane	-2.651
CO_2	-2.651
C_{4+}	-6.586

8.3 Steam requirement

The steam is used for the smokeless burning of flare gas. The steam requirement depends on the flare gas flow rate, composition of flare gas, steam velocity and flare tip diameter. Typically 0.01 to 0.6 kg of steam per kg of flare load is required(Adam Bader, *et al.*, 2011). This ratio is usually estimated from the molecular weight of the gas, the carbon to hydrogen ratio of the gas and whether the gas is saturated or unsaturated. For example, olefins such as propylene, require higher steam ratios than would paraffin hydrocarbons to burn smokelessly. The smokeless flare is a proprietary device. The manufacturer should be consulted about the minumum necassary steam rate for smokeless burning. Usually the amount of steam required is 0.4 kg per kg of flare load.

Required Steam, S $= 0.4$G

Where

G = Flare gas flow rate.

Table 8.2: Suggested Injection Steam Rates

Gases Being Flared	**Approximate Steam Rate** kg steam per kg gas
Ethane	0.10 to 0.15
Propane	0.25 to 0.30
Butane	0.30 to 0.35
Pentane	0.40 to 0.45
Ethylene	0.40 to 0.50
Propylene	0.50 to 0.60
Butene	0.60 to 0.70
Propadiene	0.70 to 0.80
Butadiene	0.90 to 1.00
Pentadiene	1.10 to 1.20
Acetylene	0.50 to 0.60
Benzene	0.80 to 0.90
Toluene	0.85 to 0.95
Xylene	0.90 to 1.0

CHAPTER 9

CONCLUSIONS

9.1 Conclusion

The flare network design is very complicated. Very few EPC companies in India doing the flare network design because most of the designing systems in flare network is proprietary one. This project optimized the flare network design procedure. The flare network in the old refining and petroleum industries requires huge amount of auxillary fuel and purge gas. It eliminates heavy radiation to the atmosphere and to the ground level. The carbon emission from the flare tip is heavy. The chunk cost for the piping and support of the flare system also heavy. Apart from this the thermal design of Flare tip and ignition systems also included in the designing of flare system. The supporting structure of the flare system should be designed before procurement. The expenditure for the entire flare system should be estimated. After the successful designing and procurement, the commissioning should be done in that particular plant.

Appendix A

A.1 Model calculation to determine Flare load

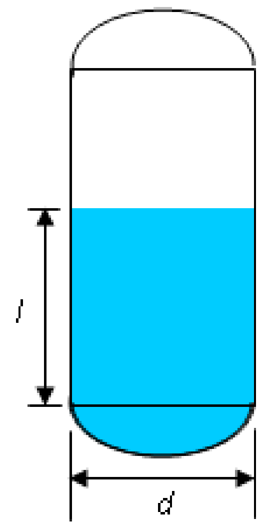


Figure 9.1: Vertical Cylinder

Unit No. = xx;

Location No. = xx

Item No. = xx

PSV No. = xx

Failure case = Fire case

Height of the vessel from the ground = 2 m;

Length of the vessel = 5 m;

Diameter, D_i = 1.1 m;

Normal Operating liquid level, l = 3.5 m;

ϵ = 0.866;

Dish Area = 1.623 m^2;

Wetted Area of Cylinder = 12.089 m^2;

Total Wetted area of cylinder = 13.7126 m^2;

Heat input = 369.76 kj/s;

Latent Heat = 1695.35 kj/kg; (Calculated from Individual latent heat and weight fraction of its components)

Total Load = 785. 17 kg/hr

A.2 Model calculation to determine line loss

Unit No. $=$ xx;

Location No. $=$ xx

Item No. $=$ xx

PSV No. $=$ xx

Governing case $=$ Fire case

Diameter $=$ 4 " ;(In this diameter, the line loss is less than 3% of set pressure);

Equivalent length $=$ 200 m ; (From Plot plan)

Density $=$ 0.9874 $\frac{kg}{m^3}$

Total Load $=$ 4555.497 $\frac{kg}{hr}$;

Flow rate $= 4613.424\frac{m^3}{hr}$;

Velocity $= 158.148\frac{m}{s}$;

Viscosity $=$ 0.018 cP;

Friction Factor $=$ 0.019;

Line loss $= 0.1196\frac{kg}{cm^2}$;

$\%$ of set pressure = 0.85$\%$;

A.3 Model Calculation for Flare KOD

Flow Rate = 25.2 kg/s;

Liquid density = 496.6 kg/m^3;

Vapour Density = 2.9 kg/m^3 ;

Temperature = $149C$;

Viscosity = 0.01 mPa.s ;

Vapour flowrate = 3.9 kg/s;

Liquid flowrate = 21.3 kg/s ;

Volumetric Vapour flowrate =$\frac{21.3}{2.9} = 7.34m^3/s$;

$$C(Re)^2 = \frac{0.13 \times 10^8 \times 2.9 \times (0.0003)^3 \times (496.6 - 2.9)}{(0.01)^2} = 5025;$$

From graph, C=1.3;

A.4 Model calculation for Flare Diameter

The Mach number is determined from the equation,

$$Ma = 3.23 \times 10^{-5}[\frac{q_m}{p_2 d^2}][\frac{ZT}{M}]^{0.5}$$

$$0.2 = 3.23 \times 10^{-5} (\frac{45,360}{101.3d^2}) \sqrt{\frac{1 \times 422}{46.1}}$$

$d^2 = 0.219$

d=0.468 m ;

A.5 Model Calculation for Flame Length

$$Q = (45,360kg/h) \times (5 \times 10^4 kj/kg) \times (1h/3600s) = 6.3 \times 10^5 kW$$

From graph , Flame Length is 50 m.

A.6 Model Calculation for flame parameters

$$u_j = \frac{q}{\frac{\pi d^2}{4}} ;$$

$$u_j = \frac{9.46}{\frac{\pi \times 0.468^2}{4}} = 55m/s;$$

$$\frac{u_\infty}{u_j} = \frac{8.94}{55} = 0.162;$$

From the graph $\sum \frac{\triangle x}{L} = 0.85$

$$\sum \frac{\triangle y}{L} = 0.36$$

$$\sum \triangle x = 0.85 \times 50 = 42.5m;$$

$$\sum \triangle y = 0.36 \times 50 = 18m;$$

A.7 Model calculation for Flare stack height

$$D = \sqrt{\frac{\tau F Q}{4\pi K}}$$

$$D = \sqrt{\frac{1.0 \times 0.3 \times 6.3 \times 10^5}{4\pi \times 6.3}} = 48.9m$$

$$h' = h + (0.5 \sum \triangle y)$$

$$r' = r - (0.5 \sum \triangle x)$$

$$r' = 45.7 - (0.5 \times 42.5) = 24.4m$$

$$D^2 = r'^2 + h'^2$$

$$48.9^2 = 24.4^2 + h'^2$$

$$h' = 42.3m$$

$$h = 2.3 - (0.5 \times 18) = 33.3m$$

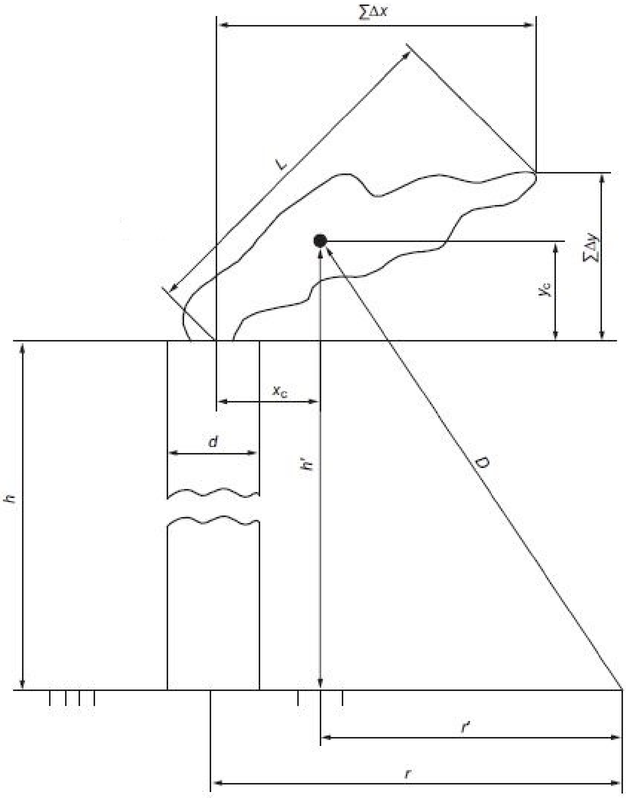


Figure 9.2: Flare Stack Parameters

Appendix B

B.1 Drag coefficient values for interpretation

Table 9.1: Drag Coefficient values

C(Re)2	C	C(Re)2	C
10.0	59.00	6,000.0	1.20
20.0	33.00	7,000.0	1.15
30.0	24.00	8,000.0	1.10
40.0	19.00	9,000.0	1.05
50.0	16.00	10,000.0	1.00
60.0	14.00	20,000.0	0.84
70.0	12.00	30,000.0	0.75
80.0	11.00	40,000.0	0.70
90.0	10.00	50,000.0	0.66
100.0	9.50	60,000.0	0.62
200.0	6.00	70,000.0	0.60
300.0	4.70	80,000.0	0.59
400.0	4.00	90,000.0	0.57
500.0	3.50	100,000.0	0.55
600.0	3.20	200,000.0	0.50
700.0	3.00	300,000.0	0.47
800.0	2.80	400,000.0	0.47
900.0	2.70	500,000.0	0.46
1,000.0	2.50	600,000.0	0.46
2,000.0	1.90	700,000.0	0.45
3,000.0	1.60	800,000.0	0.45
4,000.0	1.40	900,000.0	0.45
5,000.0	1.30	1,000,000.0	0.45

B.2 Properties of commonly used flare gas

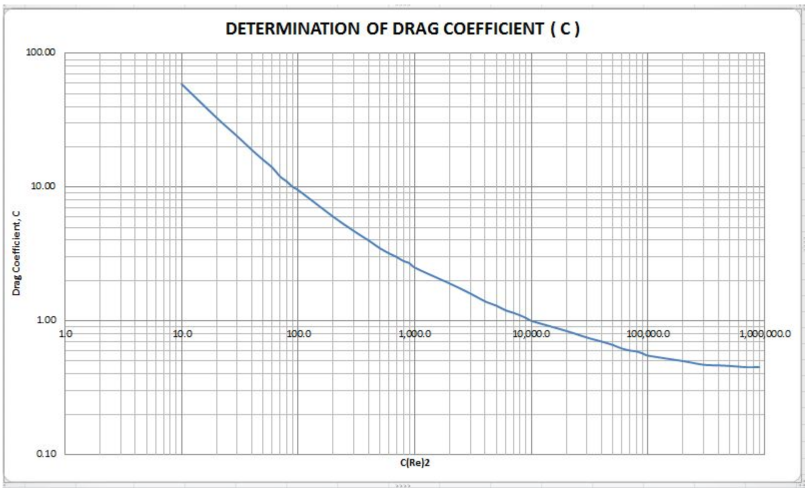


Figure 9.3: Determination of Drag Coefficient (C)

Table 9.2: Properties of commonly used flare gas

Gas	Mol.Wt.	Density kg/m^3	Sp. Gravity
Acetylene	26	1.0921	0.9
Air	29	1.205	1
Ammonia	17.031	0.717	0.59
Argon	39.948	1.661	1.38
Benzene	78.11	3.486	2.6961
Butane	58.1	2.489	2.0061
Butylene	56.11	2.504	1.94
Carbon dioxide	44.01	1.842	1.5189
Carbon monoxide	28.01	1.165	0.9667
Chlorine	70.906	2.994	2.486
Ethane	30.07	1.264	1.0378
Ethylene	28.03	1.26	0.9683
Helium	4.02	0.1664	0.138
Hydrogen	2.016	0.0899	0.0696
Hydrogen Chloride	36.5	1.528	1.268
Hydrogen Sulfide	34.076	1.434	1.1763
Methane	16.043	0.668	0.5537
Neon	20.179	0.8999	0.697
Nitric oxide	30	1.249	1.037
Nitrogen	28.02	1.165	0.9669
Oxygen	32	1.3313	1.1044
Ozone	48	2.14	1.66
Propane	44.09	1.882	1.5219
Propene	42.1	1.748	1.4523
Sulfur Dioxide	64.06	2.279	2.264
Steam	18.016	0.804	0.6218

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