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1.0 INTRODUCTION

The purpose of this document is to provide process design criteria for NIGC's BID BOLAND GAS TREATING PLANT II Project, including the following:

- Hydraulic calculations, including piping systems
- Pumps and compressors
- Vessels
- Heat exchangers and air coolers
- Fired heaters
- Storage tanks
- Relieving systems, including safety relief valves
- Flare and blowdown system.

This document shall be used as the standard reference for process design work, including equipment in packages. This will ensure consistent quality process deliverables.

The standards used in this document were mainly derived from IPS (Iranian Petroleum Standards) and supplemented by SNC-Lavalin standards, where applicable.

The design criteria described here are intended to serve as a guideline. In case of discrepancy between this document and IPS documents, this document shall govern.

2.0 REFERENCES

IPS-E-PR-360	(Process Design of Liquid & Gas Transfer & Storage)
IPS-E-PR-440	(Process Design of Piping System)
IPS-E-PR-460	(Process Design of Flare & Blowdown Systems)
IPS-E-PR-450 Inclusive Safety	(Process Design of Pressure Relieving Systems Relief Valves)
IPS-E-PR-750	(Process Design of Compressors)
IPS-E-PR-771	(Process Requirements of Heat Exchanging Equipment)
IPS-E-PR-785	(Process Design of Air Cooled Heat Exchangers)

IPS-E-PR-830 (Process Design of Valves & Control Valves)

IPS-E-PR-850 (Process Equipment of Vessels, Reactors and Separators)

IPS-E-PR-880 (Engineering Standard for Process Design of Gas (Vapour)-Liquid Separators)

3.0 HYDRAULIC CALCULATIONS

This section provides some general rules for line sizing. These rules should cover most situations, but might not be suitable for all cases. For critical services or long headers, a study of the hydraulic system, to confirm total pressure drops and pressure balance, shall be carried out, regardless of whether lines meet the allowable pressure drop and velocity criteria provided in this document. The frictional pressure losses in valves and fittings is accounted for in calculating equivalent pipe length.

The design criteria below, expressed as recommended flow velocities and maximum pressure drops, are provided as general guidelines to avoid erosion, vibration and noise problems. These guidelines are overall experience factors and are not intended to be a truly quantitative analysis of all variables involved. These criteria apply only to carbon steel (CS) pipes. For more expensive materials such as stainless steel (SS) or alloy, economic considerations may overrule and lead to higher velocities or smaller pipe sizes than normally allowed for CS.

3.1 Line Size

The nominal pipe size (NPS) will be designated DN (Diameter Nominal), although in calculations the diameter is normally expressed in millimeters (mm). Table 3.1-shows DN and NPS equivalents.

Table 3.1-1 DN and NPS Equivalents

DN (mm)	NPS (inches)	DN (mm)	NPS (inches)
15	1/2	350	14
20	3/4	400	16
25	1	450	18
40	1 1/2	500	20
50	2	600	24
80	3	650	26
100	4	700	28

DN (mm)	NPS (inches)	DN (mm)	NPS (inches)
150	6	750	30
200	8	800	32
250	10	900	36
300	12	1000	40

3.2 Pump Lines

3.2.1 Pump Suction Lines

Table 3.2.1-1 shows recommended velocities and allowable pressure drop.

Table 3.2.1-1 Recommended Velocities and Delta Ps

	Max. Allowable Velocity m/s	Max. Allowable ΔP, Bar/100 m
Sub-Cooled Liquid	2.4	0.079
Boiling Liquid	1.2	0.045

Maximum normal flows shall be used for calculations. For viscous liquids, the velocity criteria could lead to excessive pressure drops; in those cases, pressure drop criteria shall govern.

Gas blanketed liquids shall be considered boiling liquids for line sizing, pump calculations, specifications, etc.

In cases where permanent strainers are provided, a pressure drop of 0.035 Bar shall be added.

3.2.2 Pump Discharge Lines

Table 3.2.2-1 shows recommended velocities and allowable pressure drop.

Table 3.2.2-1 Recommended Velocities and Allowable Pressure Drop

Pipe Diameter	Maximum Velocity m/s		Maximum Allowable Pressure Drop Bar/100m	
2" and less	1.	2.7	0.23	0.7
3" to 10"	1.5	3.7	0.23	0.7
12" or greater	2.2	4.3	0.23	0.7

3.3 Gravity Transfer Lines

Gravity transfer lines shall have a pressure drop within the range of 0.023 to 0.079 Bar/100 m, depending upon available head.

In cases where no controller is provided for liquid level in the liquid draw-off tray, flow velocity in the first 3 m of the vertical line shall be less than 0.762 m/s.

Where a boiling liquid enters a control valve, line sizes shall be based on the following.

With consideration given to static head and length of the line from liquid level to control valve, no vapourization shall occur at the inlet of the control valve. In this case, the following should be satisfied:

$$4.91 \times 10^{-5}\,\rho.H > \Delta P\ (\text{flow meter}) + \Delta P\ (\text{line friction}) \qquad \text{(Eq. 3.3-1)}$$

Where,

ρ: density, kg/m^3

H: static head, m

ΔP (flow meter): pressure drop in flow meter, Bar

ΔP (line friction): pressure drop in line, Bar

3.4 Limitations Owing to Erosion Preventive Measures

Velocity of fluid plays an important role in erosion and corrosion. Velocity often strongly influences the mechanism of corrosion. Mechanical wear of lines results at high velocities and particularly when the solution contains solids in suspension. The following limitations should be considered for amine solution:

	Carbon Steel	Stainless Steel
Liquid	3 m/s	9 m/s
Vapour-liquid	30 m/s	36 m/s

3.5 HC Vapour Lines

In the case of vapours, two approaches may be followed, depending on the magnitude of the compressibility effect. For vapours with low pressure drop or low velocity (pressure drop equal to or less than 10% of the upstream pressure, or the velocity equal to or less than 60 m/s) the following ranges for velocity and pressure drops may be used for vapour line sizing. Should pressure drop exceed 10% of the upstream absolute pressure, the calculation should be split into a number of segments with corresponding physical properties.

3.5.1 Maximum Allowable Pressure Drop

For lines not exceeding 100 meters in length, the following shall apply:

Pressure Level	Pressure Drop, Bar/100 m
0 to 3.5 Bar(g)	0.056 to 0.113*
3.5 to 10.5 Bar(g)	0.113 to 0.339*
10.5 to 35 Bar(g)	0.339 to 0.678*
Over 35 kPa(g)	(0.5% of pressure level)

* Interpolate linearly for intermediate pressure levels.

3.5.2 Allowable Velocities

Maximum for gas = 122/SQRT(ρ_v) m/s, where ρ_v is the gas density in kg/m^3.

3.5.3 Centrifugal Compressor Suctions

Recommended velocity and pressure drop for compressor suction lines are as follows:

Velocity m/s		Pressure Drop
Operating Range	Maximum Limit	
6 to 15	v = 122/SQRT(ρ_v)	0.12 Bar for reciprocating 0.7% inlet Bara for centrifugal

3.6 Two-Phase Process Lines (Vapour/Liquid Mixtures)

The primary concern in designing a pipe for two-phase flow is that various flow patterns can occur. Particular attention shall be paid to the flow regime to avoid slug flow, where possible. For preliminary sizing, line velocities shall be limited by the following:

Maximum mixture velocity for two-phase flow =122 / SQRT(ρ_{mix}), m/s

Minimum mixture velocity for two-phase flow = 3.0, m/s where (ρ_{mix}) is the density of the mixture, kg/m^3.

Velocity may be less than 3.0, m/s at the vessel inlet or for horizontal runs where separation of the two phases is desirable.

For critical two-phase process lines, detailed calculations shall be made for pressure drop and flow regime.

3.7 Pressure Safety Valve (PSV) Lines

Line sizing shall be according to API RP 520, Part I and Part II, and API RP 521.

During an emergency relief to flare, discharge line and header velocities shall not exceed 70% of sonic velocity

Where,

sonic velocity: 91.215 $(k*T/M)^{0.5}$

T: °K,

M: mol. wt.,

k: Cp/Cv.

Flare headers shall be self-draining toward the flare knockout drum, which shall be located near the flare. Flare piping slopes shall be provided as follows:

- Main Header 1:1200
- Unit Header 1:500
- Unit Laterals 1:250

Discharge lines from relief and depressuring valves shall preferably enter the headers at the top but shall, in any case, free the drain into the headers.

3.8 Control Valve Pressure Drop Allowance

Unless otherwise dictated by process, control valve pressure drop shall be calculated on the following basis.

3.8.1 Pumping Services (Discharge Side)

Control valves (including actuators) should be capable of shutoff, and operating with maximum upstream pressure and minimum downstream pressure. On a pump discharge system, the maximum upstream pressure is the pump shutoff pressure. The maximum downstream is normally atmospheric pressure.

At normal flow rate, the pressure drop allowed for the control valve shall be 1.03 Bar or 20% of the variable friction pressure drop (excluding the valve), whichever is greater.

At rated pump flow rate, the pressure drop allowed should be 0.70 Bar or a minimum of 15% of the variable system drop (excluding the control valve), whichever is greater.

For large variable system pressure drops (1.035 Bar or over), such as in heaters, furnaces, etc., 15% of the variable system pressure drop (excluding the control valve) shall be allowed for control valve pressure drop.

3.8.2 Other Liquid Services (Gravity Flow, Cascading from One Pressure Level to Another, etc.)

Pressure drop is a function of the system under consideration, but as a general rule, the pressure drop assigned to the control valve shall not be less than 10%, or greater than 85%, of the pressure differential available for frictional loss.

3.8.3 Gas Services

For compressor services, the engineer should be guided by the criteria in Section 3.5.3.

Normally, for other gas services, the pressure drop assigned to the gas control valve shall not be less than 10%, or greater than 85%, of the pressure differential available for frictional loss.

4.0 PUMP , FAN, AND COMPRESSOR SIZING

4.1 Over Capacity Factors

The following overcapacity factors shall be used.

Pumps:	Charge, product transfer services:	1.1
	Critical services (e.g., reflux pump):	1.2
	Chemical Injection:	1.2
	Metering:	1.3
	Reciprocating and Rotary:	1.15
Compressors:		1.1
Fan:		1.1

Fluids shall be considered bubble fluids (boiling liquids) if the operating pressure is less than 110% of the liquid vapour pressure at operating temperatures, or if fluids contain dissolved or entrained gas.

4.2 Calculation of Maximum Pump Suction Pressure and Shutoff Pressure

4.2.1 Pump Maximum Suction Pressure

Maximum pressure at suction of pumps is calculated as follows:

- Taking suction from tankage open to the atmosphere:

  MSP (max. suction pressure) = static head measured from high-high liquid level in the tank and zero frictional losses across piping.

- Taking suction from a pressure vessel:

  MSP = setting of relief valve at top and static head measured from highest liquid level (zero frictional line loss).

- For a system unprotected by a relief valve that has a pump upstream:

  MSP = shutoff pressure of an upstream pump.

4.2.2 Pump Shutoff Pressure for Centrifugal Pumps

Centrifugal pump shutoff pressure is the highest pressure of the following:

- Pump Shutoff Pressure = MSP plus (pump differential at maximum specified specific gravity), or

- Pump Shutoff Pressure = Normal suction plus 1.2 x (pump differential at maximum specified specific gravity), or

- If a blocked discharge causes the suction pressure to rise to its maximum (i.e., reflux pumps and pump-around pumps), then

  Max. Shutoff Pressure = Max. suction plus 1.2 x (pump differential at maximum specified specific gravity).

Pump shutoff pressure as calculated above is used to designate preliminary equipment design pressure; but shall be confirmed by the Mechanical Group on selection of final pump curves.

4.2.3 Pump Shutoff Pressure for Positive Displacement Pumps

Pump shutoff pressure for positive displacement pumps is normal suction pressure plus the stalling pressure of the pump.

4.3 Centrifugal Compressor Design Discharge Pressure

Centrifugal compressor design discharge pressure is the higher pressure of the following:

- Design Pressure = normal suction pressure + surge differential pressure, or

- Design Pressure = maximum suction pressure + normal differential pressure.

If the compressor curve is not available, surge differential pressure may be estimated as 1.2 x normal differential pressure. However, this needs to be checked after getting Vendor information.

Where the compressor is equipped with a variable speed driver, surge differential pressure shall be 125% of the highest design differential pressure, based on the 105% RPM (revolutions per minute) curve of performance characteristics of the compressor.

4.4 Miscellaneous

The pump data sheets should specify if minimum flow requirements apply to the process during start-up, shutdown, turndown, regeneration, etc.

Pump calculations for net positive suction head (NPSH) and power consumption shall be made at the design flow rate. The NPSH is calculated based on LLL (low-liquid-level) of the vessel or tank that could be reached during normal or other operation at full pumping rate.

A pressure drop of 0.035 Bar through any permanent strainer shall be based on 50% clogging. Static head requirements for the discharge pressure calculation shall be based on actual operating temperature where the static head occurs (not necessarily pumping temperature).

The required NPSH shall be no greater than 80% of the NPSH available, but at no time less than 0.5 m over the full operating range specified, or an NPSH demonstration test shall be required before leaving the Vendor's shop.

Air blowers serving fired heaters shall not be spared.

Single speed electric motors are the preferred drivers for primary and spare pumps, as well as for compressors. The exception is to use turbine drivers are to be used if the required power is greater than 1.5 MW. Air coolers are to be equipped with variable frequency drive motors.

Drivers for air blowers shall be designed for minimum ambient temperature operation.

5.0 MECHANICAL DESIGN

Heat exchangers are considered pressure vessels. The same mechanical design conditions are used for vessels and heat exchangers.

5.1 Mechanical Design Pressure

5.1.1 Operating Pressure

The operating pressure (OP) is the pressure applied during its normal operation.

Maximum operating pressure (MOP) is a temporary peak pressure that could occasionally be reached during an operating cycle (e.g., end of run), due to a foreseeable event that may decrease pressure drop (e.g., bypassing an upstream exchanger), closing of a valve downstream of the vessel, or temporary overpressure due to a temporary increase in temperature. In determining the maximum operating pressure, variations in pressure that could be expected to arise because of operating fluctuations, other than upsets, shall be considered. Liquid static head, pump and compressor characteristics, and pressure pulsation should be considered.

5.1.2 Internal Design Pressure

Vessels/heat exchangers shall normally be protected by pressure relief valves set according to code. Relieving pressure of a safety relief valve shall be set at the vessel design pressure. If the valve is not directly mounted on the vessel, then the PSV setting shall also include the effects of static head and line pressure drop.

The design pressure at the top of the vessel shall be set at the highest of the following:

- Upstream pump shut-off pressure
- Upstream compressor discharge design pressure
- Any upset conditions that can be identified
- According to the following:

Maximum Operating Pressure Bar(g)	Mechanical Design Pressure Bar(g)
< 1.5	3.5
0-35	MOP + 3.5
>35	MOP+ 10% of MOP

The mechanical design pressure for drums directly connected to the flare network (flare knockout drums) is 3.5 Bar(g) minimum.

Vessels/heat exchangers without pressure relieving devices shall be provided with an outlet that cannot be completely blocked off. The minimum outlet opening shall be sized so that the maximum pressure developed is not greater than the design pressure.

Design pressure for heat exchangers, low-pressure side (when technically feasible), shall be not less than 0.769 times the high-pressure side design pressure, to avoid the need to install a PSV for tube rupture. This higher value of design pressure should apply up to the first block valves.

5.1.3 External Pressure

Normal vacuum services shall include effects of the regular process, start-up, shutdown, condensation or other suction devices. Vessels not designed for full vacuum shall be suitable for partial vacuum. At minimum, vessels shall be designed for 0.5 Bar(g) external pressure using ASME Code.

Exchangers operating under a vacuum shall be designed for full vacuum.

5.2 Mechanical Design Temperature

5.2.1 Operating Temperature and Maximum Operating Temperature

Operating temperature (OT) is the temperature to which the inside of the vessel/heat exchangers is exposed during its normal operation.

The maximum operating temperature (MOT) is a temporary peak temperature reached systematically during an operating cycle (e.g., end of run) or occasionally due to a foreseeable event (e.g., bypassing an upstream exchanger).

In defining design temperature, consideration shall be given to start-up, shutdown, upset or any other condition that could result in a temperature lower than normal operating temperature. However, for all of the above conditions, the corresponding pressure shall also be considered.

Exchangers that will operate at temperatures 0°C and lower shall be designed for minimum anticipated operating temperature, including depressurization.

Design temperatures for multiple exchangers in series shall be selected according to maximum temperatures likely to occur on each exchanger in both clean and fouled condition. The design temperature indicated on the process data sheet is the temperature of the hottest exchanger.

For fixed tubesheet exchangers without expansion joints, the differential between the average shell metal temperature and the average metal temperature of any one tube pass shall not exceed 28°C. When temperature differentials exceed 28°C, an expansion joint shall be provided.

For two-pass shell exchangers, the differential between inlet and outlet temperatures of the shell side fluid shall not exceed 194°C.

The following table provides the basis for design temperature of vessels and heat exchangers.

Operating Temperature (OT)	**Design Temperature (DT) (Note 1 and 4)**

Less than -100°C	min. oper. Temp / 86 °C min. (Note 2)
Between –40°C and -100°C	-100°C / 86 °C min. (Note 2)
Between -30°C and -39°C	-45°C / 86 °C min. (Note 2)
Between -29°C and +60°C	(min. oper. Temp – 6) °C./ 86 °C min. (Note 2)
Between 60°C and 343°C	max. oper. temp. +25°C. (Note 3)
Above 343°C	To be specified according to selected material and process requirements

Notes:

1. Black temperature for Bid Boland Facility is 86 °C. Black temperature for Mahshahr Facility is 87 °C.
2. Design temperatures lower than 86 °C are allowed only for insulated equipment, for which a design temperature of 60°C shall be selected
3. Indicate maximum sun temperature for non-insulated exchangers. The insulated case should also be calculated.
4. For systems requiring a minimum design metal temperature specification, including specific depressuring Facilities, auto-refrigeration should be considered in the calculation.

6.0 VESSEL SIZING

6.1 Surge Volume for Separators

Surge volume to inlet separators shall be estimated based on pipeline hydraulic calculations.

6.2 Liquid Hold-up, Surge and Liquid Levels

6.2.1 Total Minimum Vessel Retention Time

The vessel retention time (hold-up + surge) between low liquid level (LLL) and high liquid level (HLL), which shall be applied in the absence of other overriding process considerations, is as follows:

- 4 minutes for product to storage (2 minutes hold up time + 2 minutes surge time)
- 5 minutes for feed to a furnace (3 minutes hold up time + 2 minutes surge time)
- 4 minutes for other applications (2 minutes hold up time + 2 minutes surge time).

6.2.2 Low-Low or Low Liquid Level

The following are minimum distances above the vessel bottom. If low-low liquid level is present, these values refer to low-low liquid level.

Horizontal Vessels	150 mm (6") above bottom
Vertical Vessels	200 mm (8") above lower tangent line

6.2.3 Between Low-Low Liquid Level (LLLL) and Low Liquid Level (LLL)

LLL shall be at least 200 mm (8") above LLLL.

6.2.4 Vapour Height Above High-High Liquid Level (HHLL) or High Liquid Level (HLL)

The following shall be the minimum distances from HLL. If HHLL is present, these values refer to HHLL.

Horizontal Drum:

- Without demister – 600 mm (24")
- With demister — 20% of the drum diameter, or 300 mm (12"), whichever is greater.

Vertical Drum:

- With demister — 610 mm (24") from the top of the feed inlet arrangement to the bottom of the demister
- Without demister — 1220 mm (48") from the top of the feed inlet arrangement to the top of the vessel.

6.2.5 Between High-High Liquid Level (HHLL) and High Liquid Level (HLL)

HHLL shall be at least 200 mm (8") above HLL.

6.2.6 Between High Liquid Level (HLL) and Low Liquid Level (LLL) for KO Drums

Compressor suction drums shall have minimum of 356 mm (14").

Other types of KO drums shall have minimum of 356 mm (14") or a volume equivalent to 15 m of the inlet line, whichever is greater.

6.2.7 Slugs

- When liquid slugs formation is expected, total liquid volume provided between NLL and HLL shall include the surge volume required plus the expected slug volume.
- In the absence of other information, slug volume should be taken as 2 to 5 seconds of flow with the normal feed velocity and 100% liquid filling of the feed pipe.

6.2.8 Foam Formation

If the feed to the drum produces foam, an additional 250 mm of liquid level between HLL and HHLL shall be added.

6.3 Vane Type Demister

The following equation shall be used to calculate critical velocity:

$$V_c = K * (\rho_L/\rho_V - 1)^{0.5} \qquad \text{(Eq. 5.5-1)}$$

Where,

V_c: critical velocity, m/s

ρ_L: liquid density, kg/m^3

ρ_V: vapour density, kg/m^3

K: empirical constant for separator sizing, m/s

Refer to the following table for values of K:

Separator Type	K (MAX)
Horizontal	0.12-0.15
Vertical	0.05-0.11
Spherical	0.05-0.11
Wet Steam	0.076
Most Vapours under vacuum	0.061
Slug Catcher	0.21

Separator Type	**K (MAX)**
Adjustment Factor (multiply K by the factor)	Factor
1. Pressure	
Atmospheric	1
10 Bar	0.9
20 Bar	0.85
40 Bar	0.8
80 Bar	0.75
2. For glycol and amine solutions,	0.6 - 0.8
3. Vertical Separators without wire demisters	0.5
4. Compressor suction scrubbers and expander inlet separators	0.7 - 0.8

Maximum setting velocity of the heavy liquid out of the light liquid shall be calculated as follows:

$$U_{hl} = k_s(\rho_h - \rho_l) / \mu_l$$
(Eq. 5.5-2)

$$k_s = 5.45 * 10^{-10} * D_{ph}^2$$
(Eq. 5.5-3)

Where,

U_{hl}: settling velocity of heavy liquid out of the light liquid, m/s

k_s: settling factor

μ_l: viscosity of light phase, cP

ρ_h: heavy phase density, kg/m^3

ρ_l: light phase density, kg/m^3

D_{ph}: heavy liquid droplet diameter, micron (10^{-6} m)

Table 5.5-1 shows maximum velocity of water out of oil fractions.

Table 5.5.1- Maximum Velocity of Water Out of Oil Fractions

Properties			Maximum water droplet velocity, m/s	
	S.G.	μ (cP)	400-500 micron	1000 micron
Gas-oil	0.85	5.0	0.003	0.02

6.4 Boot

Design considerations are as follows:

There shall be a minimum 5 minutes of residence time.

The height/diameter shall range from 2:1 to 5:1.

Diameter shall be less than 1/3 of the vessel's inside diameter.

Minimum diameter shall be 305 mm.

6.5 Three-Phase Vessel Weir Plates

Generally, the weir plate should be 150 mm (minimum) above the oil-water interface. It can vary in height, bottom inside shell wall to top of plate, from 300 mm to the midpoint of the vessel.

6.6 Feed and outlet nozzles sizing

Nozzle sizing shall be based on actual flow rates (excluding the appropriate design margins).

Sizes of inlet and outlet nozzles are normally the same as those of the connecting process. However, there are cases where the liquid outlet nozzle size might have to be larger than the line size to avoid vortex formation. In these cases, the transition piece connecting the oversized outlet nozzle with the smaller size process line should be thirty nozzle diameters long (30 d) but not more than 3 m.

6.6.1 Inlet Nozzle

Knockout Drum Without Mist Extractors:

The diameter of the nozzle, d_n, shall satisfy the following criteria:

Horizontal drum – $\rho mix * V^2 mix < 1000\ kg/(m*s^2)$ (Eq. 5.9-1)

Vertical drum – $\rho mix * V^2 mix < 1500\ kg/(m*s^2)$ (Eq. 5.9-2)

Where,

ρ_{mix}: liquid-vapour mixture density, kg/m^3

V_{mix}: liquid-vapour mixture velocity, m/s

Demister Separators:

The diameter of the nozzle, dn, shall satisfy the following criteria:

$$\rho mix * V^2 mix < 6000 \text{ kg/(m*s}^2) \quad \text{(Eq. 5.9-3)}$$

$$\rho v * V^2 g,\text{ in} < 3750 \text{ kg/(m*s}^2) \quad \text{(Eq. 5.9-4)}$$

Where,

ρv: vapour density, kg/m^3

Vg, in: vapour inlet velocity, m/s

The length of the vane type inlet nozzle should be about five times the feed nozzle diameter.

6.6.2 Vapour Outlet Nozzle

$$\rho_v \cdot V^2_{g,\,out} \leq 3750 \text{ kg/(m*s}^2) \quad \text{(Eq. 5.9-5)}$$

Where,

$V_{g,\,out}$: vapour outlet velocity, m/s

6.6.3 Liquid Nozzle

The liquid outlet nozzle velocity, V_L, should be less than 1 m/s. The minimum nozzle diameter shall be 50 mm (2 in.).

6.7 Inlet Diverter

An inlet diverter should always be included, because it will break up the bulk of the inlet stream into smaller particles. There are various types available, all of which are used by manufacturers. These types are as follows:

- The dished end type inlet diverter directs the inlet fluid back into the vessel head. It is used in cases where the inlet nozzle is in the head of the vessel.
- A 90 degree elbowed pipe is used for gases and directs the fluid back against the vessel head.
- Currently, the most common types are vane, angle or pipe inlet diverters in a box type arrangement.

6.8 Vortex Breakers

Vortex breakers should always be installed on liquid outlets (two-phase separators), or on water and oil nozzles (three-phase separators) when the liquid goes to pump suction.

6.9 Miscellaneous

All vessels shall be provided with clean-out connections for water flush.

All necessary nozzles shall be provided for start-up, normal operations, emergency operations, safety, shutdown and maintenance of the vessel. These include, but are not limited to the following:

- Nozzle(s) for safety valve(s), mounted on the vessel. If a mist eliminator pad is in the vessel, nozzle(s) shall be located upstream of the pad.
- A vent at the highest possible point of the vessel, sized according to IPS-E-PR-200, Section 6.5.5.1.10
- A 2-in. NPS (minimum) drain, if the vessel is not naturally drained by a bottom process line
- Nozzles for utility connection, as required
- Instrument nozzles.

7.0 AIR COOLED HEAT EXCHANGERS

7.1 General Design Criteria

General design criteria are as follows:

- Horizontal type air coolers are preferred.
- For design maximum ambient air temperature, see the Design Basis Memorandum (DBM). For example, in Bid Boland, the design maximum ambient temp is 48 °C.
- If the fluid being handled is subject to hydrate formation over a range of atmospheric temperatures encountered, provision shall be made to control the extent of cooling at the lower ambient air temperatures.
- For Bid Boland and Mahshahr Facility, air cooling shall usually be the best choice.
- Over-design factor of 25% shall apply to refrigeration condensers.

7.2 Number of Fans

At least two fans shall be provided for each bay. Any deviation shall require prior Approval from Owner. Bid Boland facility shall use variable frequency drive motors. All fans shall have variable frequency drivers.

7.3 Fans in Various Duties

Where, for reasons of control, an air-cooled heat exchanger is provided with variable frequency drive, it shall not share its fans with air-cooled heat exchangers on other duties.

7.4 Chemical Cleaning Connections

If chemical cleaning maintenance is specified, connections shall be provided according to the following:

- Connections shall be installed only in nozzles DN 100 mm (NPS 4 in.) and larger. For smaller nozzles, connections shall be made in the attached piping.
- The minimum size connection shall be DN 50 mm (NPS 2 in.).
- Connections shall be installed horizontally.
- For bundles in series or series-parallel arrangement, only one chemical cleaning connection need be provided in the inlet nozzle and one in the outlet nozzle of each series group.

7.5 Air-Side Design

Environmental factors pertinent to design of the exchanger shall be supplied to the Vendor according to IPS-E-PR-785, Table 1. These factors shall be taken into account in the air-side design.

Air coolers shall be designed for summer and winter conditions, as specified in the job specifications.

Proper fouling resistance shall be applied to the inside surface of the tube.

All heat transfer surfaces and coefficients shall be based on total effective outside tube and fin surface.

Inside fouling and inside fluid film resistance shall be multiplied by the ratio of the total effective outside surface to the total effective inside surface for calculating heat transfer coefficients.

Pressure drops shall not exceed the maximum allowed values specified. These include total pressure drops across nozzles, headers and tubes.

Fouling factor on the air side of exchangers shall be 0.0004 h-m^2-C/kcal (0.35m^2.C/kW).

7.6 Winterization

- All air-cooled exchangers for which winterisation may be required should be forced draft units with top louvers. Winterization shall include control of ambient temperature under the plenum using louvers.

- Process outlet temperatures should be controlled, if required, variable frequency motor. All exposed headers should be steam traced and/or insulated.

- Slopping may be considered to facilitate complete drainage of the tube fluid during the shutdown period.

8.0 SHELL AND TUBE HEAT EXCHANGER

8.1 Design Factor

Exchangers should be designed for 110% of normal flow rates (shown in heat and material balance of the unit).

It is not good practice to over-design an exchanger, because potential control problems can result. Over-design factors greater than 30% should not be used.

8.2 Tubular Units

- Tubular units in general should have removable tube bundles and should be of the floating head type, with removable shell covers. Typical exceptions are the following:

  - Fixed tube sheet exchangers, such as refrigeration condensers and vacuum condensers. In this type of construction, differential expansion of the shell and tubes due to different operating metal temperatures might require the use of an expansion joint or a packed joint.

  - U-tube type for reboilers using steam in the tube. Removable shell covers are not required for this type.

- Fluids having a fouling factor above 0.0004 h-m²-C/kcal (0.00035m².C / W) should be routed on the shell side of U-tube exchangers and on the tube side for floating head type exchangers. In all cases, U-tubes should be horizontal.

- Fixed head exchangers should be used when the shell side fluid is non-fouling. Where the shell side fluid is fouling, U-tubes or floating head type bundles should be used, and when both sides are fouling, floating head type tube bundles should be used.

- Floating head type tube bundles are to be avoided for kettle type reboilers and chillers, unless agreed by Owner.

- See IPS-E-PR-785, Appendix B, Table B.1 for typical TEMA recommended fouling resistances for industrial fluids.

- The selection of TEMA Class R or TEMA Class C exchangers shall be governed by the following:

  - TEMA R is required under the following conditions:

    i. Tube side or shell side fouling factor is greater than 0.0004 h-m²-C/kcal (0.00035m².C / W), or

    ii. Shell side corrosion allowance is greater than 3.175 mm (1/8 inch).

    iii. Shell side corrosion rate is greater than 0.254 mm/y (10 mils per year).

  - TEMA C may be used when the exchanger is designed for chemical cleaning maintenance and if the fouling factor does not exceed 0.00035 m²-C / W on either tube or the shell side.

8.3 Horizontal and Vertical Exchangers

Heat exchangers should be of the horizontal type; however, for process requirements and where cleaning and other maintenance will be infrequent and space requirements make it more attractive, the vertical arrangement may be considered. This should be discussed with Owner. Centerline elevation of the top bundle of stacked exchangers shall be limited to 3.5 m, except for large exchangers, which shall be limited to two stacked shells.

When horizontal arrangements are preferred, stacking of exchangers should be considered, to conserve space in the structure.

8.4 Selection of Type

Fixed tube sheet heat exchangers should only be used in services under the following conditions:

- Differential expansion between the tube and the shell does not give rise to unacceptable stresses.

- Tube side cleaning, if required, can be done in situ.

- Shell side fluid is non-fouling, or

- Shell side fouling can be removed by chemical cleaning.

U-tube bundle heat exchangers shall only be used in services under the following conditions:

- Tube side fouling resistance is less than 0.0004 h-m²-C/kcal (0.00035m².C / W).
- Tube side fouling can be removed by chemical cleaning.

U-tube shall not be applied when tube side mechanical cleaning is required.

Floating head heat exchangers should be used in all other services.

8.5 Shell Selection

The single-pass shell, Type E (see IPS-E-PR-771, Appendix A, Figure A.1)), has the widest application and should be selected for general duties, except where significant advantage can be obtained by using one of the other shell types, as indicated below:

- Where shell side pressure drop is a restricting factor, the divided flow shell Type J, cross flow shell Type X, or double-split flow shell Type H should be considered.
- For horizontal shell side thermosiphon reboilers, split flow shell Type G or Type H should be selected.
- The kettle type (shell Type K) should be selected for boiling, where there is an almost 100% vapourization, or where a phase separation is required.

Use of the two-pass shell with the longitudinal baffle Type F should be avoided.

8.6 Front End and Rear End Selection

Front end bonnet Type B is generally used for heat exchangers where cleaning on the tube side will be infrequent. Rear end Type S should be used for floating head heat exchangers.

Rear end Type M should be applied for fixed tube sheet design.

When frequent tube side cleaning is anticipated, and the tube design pressure is low, the front end stationary head shall be Type A. However, for the corresponding rear end, Type L may be selected.

For high-pressure and/or very toxic service, where it is desirable to limit the number of external joints, stationary heads Type B, Type C or Type N should be selected for the front end, and the corresponding Type M or Type N for the rear end.

Outside packed floating head Type P and externally sealed floating tubesheet Type W rear ends are not acceptable.

8.7 Fluid Allocation

Fluid allocation shall be made under the following conditions:

- Dirty fluids shall be passed thorough the tubes because they can be easily cleaned, particularly if the tube bundle cannot be removed. The fluids shall be passed through the shell if the tubes cannot be cleaned (hairpin bundles) or if large amounts of coke or debris are present, which can be accumulated in the shell and removed by dumping the shell.

- High-pressure fluids, corrosive stock, and water are sent through the tubes, because the strength of small-diameter (and thin) tubes surpass that of shells, corrosion-resistant tubes are relatively inexpensive, and corrosion or water scale can be easily removed.

- For the same pressure drop, higher heat transfer coefficients shall be obtained on the tube side, rather than the shell side. Large-volume fluids (vapours) are passed through the shell because of adequate space, but small-volume fluids are also passed through the shell where cross baffles can be used to increase the transfer rates without producing an excessive pressure drop.

- Vapours that contain non-condensable gases are sent through the tubes so that the accumulation of non-condensables will be swept out.

- If pressure drop must be low, fluids are sent through the shell. The same applies to viscous or low transfer rate fluids, because the maximum transfer rates for a fixed pressure drop can be obtained by the use of cross baffles in the shell.

- In fin tube equipment, high-pressure, dirty, or corrosive stock is sent through the fin tube, because it is relatively inexpensive, can be easily cleaned, and has a higher strength than the outside tube.

- Condensing steam is normally passed through the tubes.

- If the temperature change of one fluid is very large (greater than about 145°C to 175°C), that fluid is usually passed through the shell, rather than the tubes, if more than one tube pass is made. This minimizes the construction problems caused by thermal expansion. In addition, to avoid thermal stress problems, fluids with a greater than 175°C temperature change cannot be passed through the shell side of a two-pass shell.

- If temperatures are high enough to require use of special alloys, placing the higher temperature fluid in the tubes will reduce overall cost. At moderate temperatures, placing the hotter fluid in the tubes will reduce shell surface temperatures, and the need for lagging to reduce heat loss, or it can be done for safety reasons.

- If one of the fluids is clean (fouling factor 0.00021 h-m^2-C/kcal (0.00017 m^2.°C / W) and is only mildly corrosive to the material selected, this fluid is passed through the tubes, and U-tube construction is used, where economical.

- Normally, a higher heat transfer coefficient will be obtained by allocating the more viscous material to the shell side, provided the flow is turbulent. The critical Reynolds number for turbulent flow in the shell is about 200. If turbulent flow cannot be achieved in the shell, it is better to place the fluid in the tubes, as the tube side heat transfer coefficient can be predicted with more certainty.

8.8 Nozzle Location

The following rules are suggested as a guide for locating heat exchanger nozzles:

- Streams being heated or vapourized should flow from bottom to top, whether on the tube side or the shell side.
- Streams being condensed should flow from top to bottom, whether on the tube side or the shell side.
- The direction of flow for streams being cooled should be dictated by piping economics.

8.9 Pressure Drop and Velocities

Maximum acceptable pressure drops indicated on the process data sheet shall be understood for fouled exchangers, and shall include pressure drops through inlet and outlet nozzles. In cases of alternate conditions, these shall apply to the worst operating condition.

High velocities will give high heat transfer coefficients but also a high-pressure drop. Velocity must be high enough to prevent suspended solids settling, but not so high as to cause erosion. High velocities will reduce fouling. Plastics inserts are sometimes used to reduce erosion at the tube inlet. Typical design velocities are provided below.

8.9.1 Liquid Velocities

Design velocities are as follows:

Tube side, process fluids: 1 to 2 m/s, maximum 4 m/s if needed to reduce fouling

Water: 1.5 to 2.5 m/s

Shell side: 0.3 to 1 m/s.

8.9.2 Vapour Velocities

For vapours, the velocity used will depend on the operating pressure and fluid density, as well as allowable pressure drop.

8.9.3 Pressure Drop

Maximum acceptable pressure drops indicated on the process data sheet shall be for fouled exchangers, and shall include pressure drops through inlet and outlet nozzles. In cases of alternate conditions, these shall apply to the worst operating condition.

8.10 Bypasses and Block Valves

Bypasses and block valves shall be used on exchangers for one or more of the following reasons.

Process Control:

When it is necessary to control heat transfer in an exchanger, use one of the following:

- A single bypass and two valves, or
- A bypass and a three-way valve (or two butterfly valves), which splits the flow between bypass and exchanger.

A three-way valve is necessary when exchanger pressure drop is so small that insufficient fluid diversion will take place through a simple bypass in the wide open position.

Leakage:

Where leaking of one side of a heat exchanger will result in intolerable contamination of the other fluid (e.g., in an overhead vapour-feed exchanger), blocks and bypasses may be installed to permit isolating the leaking unit from the system. In addition, welded tube to tube sheet joints or double tube sheets should be considered.

9.0 RELIEVING SYSTEMS (INCLUDING SAFETY/RELIEF VALVES)

9.1 Design Requirements

9.1.1 General

Safety relief valves shall be provided to protect all equipment subject to overpressure, and under certain other conditions specified here. Safety devices shall be provided when the overpressure exceeds the design pressure of the equipment.

The pressure/vacuum relief requirement and relief load capacity for atmospheric and low-pressure storage tanks shall be evaluated according to API Standard 2000, unless otherwise specified in IPS-E-PR-450.

For applicable design codes and standards, see IPS-G-ME-250, unless otherwise specified in this document.

9.1.2 Provisions of Pressure Safety Relief Valves

Pressure safety relief valves shall be provided for all cases specified below.

9.1.2.1 Vessels

Safety relief valves shall be provided in the following cases:

- When designed according to ASME, Section VIII, Unfired Pressure Vessel Code, and the overpressure exceeds the design pressure
- When designed according to ASME, Section I, Power Boiler Code, and the overpressure exceeds the maximum allowable working pressure as defined in that code.

9.1.2.2 Pumps

Safety relief valves shall be provided in the following cases:

- On discharge of positive displacement pumps
- On discharge of centrifugal pumps to protect downstream equipment from overpressure based on pump shutoff pressure
- On pump suction lines from a "bottled in" system, where overpressure can be imposed on suction piping by backflow through the pump or through a control valve bypassing from pump discharge to suction.

9.1.2.3 Compressors

Safety relief valves shall be provided in the following cases:

- At each stage of reciprocating compressors to protect interstage, intercooler, compressor frame or cylinder.
- On suction lines where overpressure occurs on suction lines or frame, or overloads the electric motor driver before interstage or discharge safety valves open.

9.1.2.4 Fired Process Heaters

Safety relief valves shall be provided to prevent overpressure due to heat input resulting from blocking the lines downstream of the heater, where check valves or other valves upstream of the heater are closed by the same action blocking the upstream line(s), except for the condition covered in the next paragraph. The safety valve may be located anywhere between the upstream and downstream blocking valves.

A safety valve shall not be required to protect the heater if the only block valve(s) between a fired heater and a tower are operable manually and intended to be used to prevent backflow from the tower to the heater in case of heater tube failure.

9.1.2.5 Piping and Connected Equipment

Safety relief valves shall be provided in the following cases:

- To protect piping, heat exchangers and other equipment served by the piping against overpressure in the following locations:
  - o At the fuel inlet to gas engine drivers
  - o Downstream or upstream of all control valves when the piping or equipment could be subjected to overpressure, assuming that the control valve will fail in the open or closed position
- On pedestal or gland water systems to pumps and turbines where overpressure is caused by water pump shut-off pressure. One safety valve may be provided on the supply line to a group of pumps, turbine glands or pedestals in lieu of a safety valve on the line to each piece of equipment.

9.1.2.6 Cold Side Blockage and Tube Failure in Exchangers

Safety relief valves shall be provided in the following cases:

- To prevent overpressure due to heat input resulting from blocking the line(s) downstream of the cold side of the exchanger, where check valves or other valves upstream of the exchanger are closed by the same action blocking the upstream line(s). A safety valve is required where heating the cold fluid within the exchanger, when the hot side inlet temperature will raise the pressure of the fluid contained between the upstream and downstream blockages to more than 1.3 times the design pressure of any item of equipment (excluding piping) in the contained system.

9.1.2.7 Pressure Safety Relief Valves Not Required

Pressure safety relief valves shall not be provided on the following systems:

- Interconnected vessels (excluding those falling under the requirements of ASME, Section I, Power Boiler Code,), if they meet the following conditions:
  - o The vessel that is the source of pressure is equipped with a safety valve sized to protect all of the interconnected vessels and the interconnected piping (including heat exchanger equipment).
  - o At least one interconnecting piping system between the protected vessel and any other vessel must be free of the following:
    - – Any equipment that may fail or stop in a closed position
    - – Any block valves, control or check valves
    - – Any orifices or similar restrictions to flow.
- A lower pressure piping system such as a pump-out is routed to an offsite slop or emergency tank with overshot connections (connections entering the top of the tank without block valves).

- Depressuring systems are routed to a flare, if all valves at the unit limit, cooling boxes, or downstream depressuring valves are locked open.

9.1.3 Provisions of Thermal Expansion Safety Valves

Thermal expansion safety valves shall be provided for all cases, as described below.Heat Exchangers

Thermal expansion safety valves shall be provided on the cold side of heat exchange equipment, including compressor jackets operating full of liquid and subject to being blocked off, where heat may be applied internally or externally and other forms of pressure relief valves are not provided.

9.1.3.1 Piping

Thermal expansion safety valves shall be provided for the following:

- Lines containing cold solvent or refrigerant. For blocked-in sections of pipe that contain cold liquid and can develop excessive pressure under ambient temperature.
- Exposed liquid-filled lines. For blocked-in sections of pipe that may be subjected to excessive pressure from sun rays when the exposed length of line is 30 m or greater.

9.1.4 Provisions of Vacuum Safety Valves

Pressure vessels in the following services shall be provided with a vacuum breaking device if they cannot withstand full vacuum:

- Vessels operating under partial vacuum
- Vessels operating full of liquid and having a vapour pressure less than atmospheric pressure at the operating temperature
- Vessels on suction to compressors.

9.1.5 Spare Safety Valves

The necessity for spare safety valves shall be based upon the following considerations:

- On process units, spare safety valves shall be required where the required time interval between safety valve inspection periods is less than the time interval between designated inspection and test periods of the unit.
- When a spare safety valve is required, the size of the safety valves shall be such as to provide the least number of safety valves for a total capacity equal or greater than the calculated required relieving capacity.
- Spare thermal expansion safety valves are not required.

9.1.6 Provisions of Fire Relief Valves

Fire relief valves shall be considered for liquid-containing equipment (e.g., vessels, columns and settlers) if all of the following circumstances apply:

- The equipment is located where a sizable fire could occur.
- The equipment can be blocked while the unit remains in operation.

9.2 Selection of Type

9.2.1 Conventional (Unbalanced) Safety Relief Valves

The conventional safety relief valve shall be used where the service is as follows:

- Clean and non-corrosive
- Corrosive, with provision of corrosion resistant materials.

Balanced bellows valves shall be considered where the service is as follows:

- Corrosive and corrosive materials may damage the guide and disk, or guide and spindle, or spring and bonnet.
- The variable back pressure is greater than 10% where 10% accumulation is allowed, or greater than 20% where 20% accumulation is allowed under fire conditions.
- The differential pressure is less than 10% when the valve is relieving, compared to the normal differential pressure across the protected equipment. Also, where the service is corrosive, with provision for corrosion resistant materials.

9.2.2 Balanced Bellows Safety Relief Valves

Use the balanced bellows safety relief valve where the service is as follows:

- The relieving pressure is to be independent of the back pressure.
- The bellows is used to prevent corrosive products from damaging the guiding surfaces, spring or associated pieces.
- Back pressure as specified in 11.2.1 above exists and either liquid or vapor relief is to be handled.

Bellows balanced valves shall not be used for services involving materials with a pour point at or above the lowest ambient temperature (e.g., materials containing wax) or where coking can be expected. Balanced valves with a piston only shall be used in such cases.

9.3 Set Pressure

9.3.1 General

Unless otherwise specified in this section, safety relief valves shall be set to relieve initially at the design pressure of the equipment, within the limitations of the allowable blowdown and accumulation specified in this section.

Unless otherwise specified in this section, the stated set pressure is the initial relieving pressure.

In general, set pressures (SP) and maximum relief pressures (MRP) of safety relief valves, expressed in relation to the design pressure of the protected equipment, and in gauge pressure, shall not exceed the values provided in Table 11.3.1-1.

Note: For the pressure setting of the safety relief valve instead of the design pressure (DP), the maximum allowable working pressure (MAWP) is used.

Table 11.3.1-1 – Set pressure and Maximum Relieving Pressure

	Set pressure		Maximum relieving pressure	
	Non-fire conditions	Fire conditions	Non- fire conditions	Fire conditions
Single valve	100% of DP	100% of DP	110% of DP	121% of DP
Multiple valves	One valve 100% of DP the others at 105% of DP**	110% of DP*	110% of DP	121% of DP

* Relief valves for fire protection may be set only at 110% of DP if they are installed in addition to adequate relief protection of process equipment against non-fire situations.

** For set pressures below 10 Bar, staggering set pressure becomes impractical because of the difference between the set pressure tolerance of 3% (according to ASME VIII UG 134) and the value of 5% of the DP becomes too small.

The above shall also apply to safety relief valves discharging liquid and flashing liquid.

9.3.2 Pressure Safety or Relief Valve Set Pressure

Maximum pressure accumulations above set pressure (initial relieving pressure) of safety or relief valves shall be limited, as shown, for the following services and equipment:

10% for:

- Compressors

- Pumps
- Turbine cases
- Piping
- Heat exchanger units
- Vessels and fired heaters other than those governed by ASME Power Boiler Code, Section I.

21% covers fire conditions for:

- Air receivers
- Piping
- Heat exchanger units
- Vessels other than those governed by ASME Power Boiler Code, Section I.
- The following allowable blowdown shall be specified for pressure safety relief valves:
- 3% blowdown for all steam service
- 4% blowdown for all vapour and gas service other than steam.

9.3.3 Equipment and Piping Safety Relief Valve Set Pressure

9.3.3.1 Vessels

Set pressure for vessels is as follows:

- ASME, Section VIII,Unfired Pressure Code Vessels: Set at design pressure.
- ASME, Section VIII, Unfired Pressure Code Vessels (special conditions only): For special cases when the overpressure exceeds the design pressure, set pressure shall be the MAWP as defined in the relevant code.
- ASME, Section I, Power Boiler Code: Set at MAWP, as defined in ASME, Section I.

9.3.3.2 Pumps and Compressors

Under blocked discharge, the horsepower rating (in kilowatts) of the electric motor driver may be exceeded. To prevent this, the motor shall be rated at a set pressure plus the overpressure that does not exceed the pressure developed at the maximum safe horsepower rating (in kilowatts) as recommended by the pump or compressor manufacturer.

9.3.3.3 Compressor and Pump Couplings

To protect against overpressure on discharge, set pressure shall be at the manufacturer's recommendation.

9.3.3.4 Turbines

When turbines exhaust into a separate steam system, the turbine case safety valve shall be set at a minimum of 70 kPa(g) above the normal operating pressure of the system into which the turbines exhaust. Set pressure shall not exceed turbine case design pressure.

For safety valves to protect gas turbine cases, set pressure shall be at the turbine manufacturer's recommendation.

Set pressure for cold side blockage and tube failure in exchangers shall be 1.3 times the lowest design pressure for any item of equipment (excluding piping) in the system contained between the blockages.

9.3.3.5 Piping and Connected Equipment

For safety valves required to protect piping and connected equipment (including fired heater tubes), set pressure shall be at the lower of the following:

- Design pressure of equipment or fired heater tubes
- Short-time pressure of the piping as determined by the classification and limits of short-time conditions defined in the relevant ASME/ANSI B.31.3.

For safety valves required to protect piping only (no connected equipment involved), set pressure shall be at the pressure specified in the second point, above.

9.3.4 Multiple Safety Valves

Safety valve settings for multiple safety valves shall conform to the following:

- For safety valves protecting vessels or other equipment falling under the jurisdiction of ASME, Section I, Power Boiler Code, the first valve shall be set at or below the MAWP. The other safety valve(s) may be set at a pressure not to exceed MAWP by 3%. The range of settings between the first and other safety valves shall not be more than 10% of the setting of the highest set valve.
- For safety valves protecting vessels falling under the jurisdiction of ASME, Section VIII, Unfired Pressure Vessel Code, the first valve may be set at or below the design pressure. The other safety valve(s) may be set at pressures that do not exceed design pressure by more than 5%. For fire conditions set pressure of all valves shall not exceed 110% of design pressure.

9.3.5 Thermal Expansion Valve Set Pressure

For valves discharging to atmosphere, the following shall apply:

- Heat exchangers and equipment excluding piping: Set pressure shall be at .design pressure.
- Piping only: Set pressure shall be at the maximum allowable short-time pressure, permitted under the applicable class of short-time conditions, defined in the relevant ASME/ANSI Code B.31.3, for the anticipated blocked-in temperature.
- Where piping and other equipment are involved in a common system is to be protected : Set pressure shall be at design pressure of governing equipment or at the pressure defined in the point above, whichever is lower.

Valves discharging to a closed system:

- In a consecutive system of piping (including other equipment) containing block valves, check valves, pumps or other closed devices, a series of temperature safety valves might be desired so that each valve relieves to the section of pipe protected by the next valve, with the last safety valve in the series relieving pressure to the destination of the low pressure end of the system. In such a consecutive system, the set pressure shall be determined as follows:
  - Determine the limiting system allowable pressure (LSAP), for example, design pressure of equipment in the system or short-time pressure of the piping, whichever is smaller.
  - The set pressure of each safety valve must be checked to ensure the sum of the set pressures of all safety valves in series in the system does not exceed the LSAP defined in the point above. If the sum of the set pressures exceeds the LSAP, each safety valve set pressure shall be reduced to bring the sum within prescribed limits.

9.3.6 Vacuum Relief Valve Set Pressure

Vacuum relief valves are set as close to atmospheric pressure as possible while not being affected by normal upsets. This allows as much accumulation as possible while keeping valve size to a minimum.

Vacuum relief valves are allowed to accumulate to the maximum external design pressure.

9.4 Sizing

9.4.1 Pressure Safety Relief Valves

9.4.1.1 General

Safety relief valve capacity formulas used for calculation of orifice area shall be according to API RP 520, Part I, Section 4, Procedures for Sizing.

Standard effective orifice areas and corresponding letter designations shall be according to API Standard 526.

Thermal expansion valves shall be DN 25 × DN 25 (1 inch × 1 inch) with a flange orifice area of 38.7×10^{-6} or 71.0×10^{-6} m^2,as required.

All safety relief valves shall have flanged inlet and outlet connections of 300 # RF and 150 # RF, respectively, unless the service requires a higher rating or a different type of facing.

When estimating the normal fluid mass inflow to the system at blocked outlet conditions, credit shall be given that, the vessel under emergency is at relieving conditions (i.e., at PSV set pressure plus accumulation). When evaluating relieving requirements, it is assumed that any automatic control valve that is not the cause of upset will remain in the normal position. Credit may, therefore, be taken for the normal capacity of these valves, corrected for relieving conditions and limited to the flow rates that downstream equipment can safely handle.

A load summary table showing the following information shall be provided for each safety relief valve (a typical arrangement of a pressure safety relief valve load summary table is shown in IPS-E-PR-450, Appendix A):

- Item number
- P&ID number
- Protected equipment
- Size and type
- Set pressure kPa(g)
- Discharge to ....
- Emergency failure. All applicable emergency causes shall be recorded. For each applicable emergency cause, the following information shall be provided:
    - MW for vapour and kg/m^3 at flowing conditions for liquid
    - kg/h for vapour and m^3/h at flowing conditions for liquid
    - °C
    - V or L

  - Area (only for fire case).

- Remarks.

9.4.1.2 Relief Load Requirements

Double jeopardy shall not be used and safety valve size shall be based on a single involuntary occurrence. Sizing shall not be based on double or multiple occurrences that are completely independent of one another, except where one of the occurrences happens frequently. Where multiple occurrences are dependent on one another and are likely, this shall be taken into account.

For each safety relief valve, the maximum relief condition shall be determined for each applicable cause, with the following modifications:

- For new designs, no corrections shall be made for the difference between operating and relieving pressure, except and subject to Onwer's Approval where there is an appreciable difference between maximum operating pressure and maximum allowable working pressure (e.g., a vacuum column or a near atmospheric column with a MAWP of 6 to 8 Bar(g)).

- For new designs, the cooling effect of an air cooling heat exchanger in case of fan failure due to natural drought shall not be taken into account.

- In case of lean oil failure to an absorber, the quantity to be released shall be taken as the total incoming vapour plus the quantity generated under normal conditions.

In addition, the quantities to be discharged in the event of each emergency case, as specified in this section, shall be determined for each valve. The conditions resulting in a maximum relief quantity in all possible emergency cases shall be evaluated. The size of the relief valve and of its outlet pipe up to the relief header shall be determined based on the largest of the relief quantities.

The following emergency cases should be evaluated for each system. Each system shall be considered equipment or a group of equipment that can be isolated by control valves or other valves, but is without other isolation between.

Utilities Failures:

- Power failure

- Cooling water failure (main system)

- Cooling water failure (within process unit limits)

- Cooling water and power failure (simultaneous failure)

- Instrument air failure

- Steam failure

- Fuel oil or fuel gas failure
- Air fan coolers failure.

Impact of utility failures shall be evaluated according to API RP 521, Section 2, Clause 2.3.5 and the criteria below. Utilities failures can be plant-wide or local.

Power Failure:

Power failure effects on unit operation and safety valve size shall be considered and shall include the following:

- Complete unit-wide loss of low and medium volt supply while high volt supply power source, steam drivers, gas engine drivers, etc. are still operable.
- Complete unit-wide loss of all types of electricity while steam drivers (if any are used), gas engine drivers (if any are used), etc. are still operable.
- Consideration shall be given to any automatic shutdown of drivers caused by power failure affecting controls.

Instrument Air Failure:

The relief load for instrument air failure shall be calculated based on the consequences, as follows:

- Blocking the vapour or liquid outlet, when a fail close type control valve is applied on the outlet
- Cutting out heat input to the system (e.g., when a fail close type control valve on steam to the reboiler is used)
- Others.

Fuel Oil and Fuel Gas Failure:

The impact of fuel oil or fuel gas failure on the unit and associated equipment shall be investigated in light of the equipment affected. For example, loss of fuel oil or gas may lead to loss of heater duty.

Air Fan Coolers Failure:

It shall be assumed that air fan condensers and coolers will fail if air fans fail.

Operating Failures:

Potential operating contingencies shall be considered for evaluating overpressure in the system. Probable cases are as follows:

- Blocked outlet

- Control valve failure
- Reflux failure
- Side stream failure/pumparound failure
- Loss of fractionator or column or liquid gas separator drum condensing duty
- Simultaneous failure of reflux and condensing duty.

Blocked outlet:

The protected system shall be considered, that is, tower, overhead condensing system and accumulator, provided there is no control valve or other restrictions between tower and accumulator.

No credit shall be allowed for operation of any control valve in overhead, bottom or feed system.

Only those sources of fluid having sufficient pressure to open the safety valve shall be considered:

- Where the design pressure of the vessel is greater than stalling pressure of the positive displacement pump, the safety valve on the pump shall provide protection from this source of pressure.
- Where discharge pressure of a centrifugal pump is above design pressure of the vessel, flow quantity is the pumped quantity at the accumulated relieving pressure.

Liquid Flow:

- Flow quantity shall be the quantity available at the accumulated pressure.
- Liquid flow quantity and relative density (specific gravity) shall be determined at the temperature existing at the accumulated pressure.
- Circulating streams pumped from a vessel and back into the same vessel shall be excluded when determining flow quantity.
- Reflux flow quantities shall be excluded, except when these flows are furnished from tankage outside the unit or where there is a control valve or other restriction in the overhead line before the reflux drum.

Vapour Flow:

- Only the vapour input shall be considered if the vessel has a high level alarm and at least 10 minutes liquid residence time after the alarm.

- Large vessels that have over 30 minutes liquid residence time above high liquid level need not have a high level alarm, and only vapour input is considered

- Where only vapour input is considered, flow quantity shall be the vapour generated at the controlled temperature at a pressure equal to the safety valve accumulated relieving pressure.

Control Valve Failure:

Control valve failure should be considered for both open and closed positions. Special attention shall be given to the inadvertent opening of a control valve on a high pressure source such as steam, which may lead to overpressure in the system.

Reflux Failure:

In towers where the source of heat is in the charge to the tower, special attention shall be given to the location of the safety valve and to the charge flow rate at relieving conditions:

- A flash curve shall be calculated to determine vapour quantity at relieving conditions using heat input, if feed is pressured into the tower.

- The available feed rate should be investigated, using pump characteristics at relieving conditions.

- Stripping steam, if used in the tower, should be added to the hydrocarbon vapour if the stripping steam pressure is above the safety valve relieving pressure.

For towers where reboilers, heaters or feed preheaters are used, the heat input is calculated as follows:

- Heat Input Via Fired Heaters:

  The heater outlet shall be on temperature control, with the fuel gas or fuel oil rate adjusted by the outlet temperature.

  Simultaneous failure of the temperature reflux shall not be considered.

  Special consideration shall be given to a multi-service heater where the entire heat input to a stream is in the convection section.

- Heat Input Via Exchangers:

  The charge rate shall be the normal design rate unless the rate is limited by charge pump flow characteristics or flow from a higher pressure vessel at relieving pressure.

The heating medium fluid throughput shall be the design rate, even if on temperature control, unless the relieving conditions affect the normal design rate.

After determining new hot and cold fluid rates, heat exchange shall be calculated using inlet and outlet temperatures at relieving conditions. The heat transfer coefficient used may be the design heat transfer coefficient, unless it is different at relieving conditions.

- Total Heat Input:

  Total heat input shall be the input from fired heaters plus exchangers plus heat inherently in the feed or sidecut streams.

  Sensible heat can normally be deducted from the total heat input where the feed enters the vessel below its bubble point at relieving conditions, that is, sensible heat from feed inlet temperature to its bubble point at relieving conditions.

  Sensible heat represented by net tower bottoms and/or sidecut streams at their relieving temperatures may also be deducted from total heat input.

  Credit for condensing duty shall not be taken unless the accumulator drum has a high liquid level alarm and at least 20 minutes residence time at the relieving condensing rate after the alarm is actuated and there is no buildup of vapour into the accumulator to prevent flow through the condensers at relieving conditions.

Side Stream Reflux Pumparound Failure:

Relief load for side stream reflux and pumparound failure shall be calculated according to API RP 521, Section 3, Clauses 3.6.7 and 3.6.8.

Mechanical Failures:

- External fire
- Heat exchanger tube failure
- Hydraulic expansion
- Other failures.

Failure of electrical or mechanical equipment such as fans, pumps, compressors, etc. shall be considered emergencies. Such mechanical failures can be evaluated as part of operating failures or as partial utility failures.

External fire:

External fire shall be evaluated according to API RP 520, Appendix D, unless otherwise specified in this section and in the following criteria.

General:

The surface area effective in generating vapour when exposed to fire is limited to areas wetted by internal liquid contents.

No credit shall be taken for insulation (i.e., F=1.0).

For establishing fire areas, each unit shall be divided into probable fire areas of approximately 460 m².

Air coolers shall be assumed under fire, even if installed above 8 m. Wetted surface area of the air fin cooler shall be according to API RP 521, Paragraph 3.15.4.1, depending on services.

Vertical Distance Covered by Fire:

Total wetted surface included within a height of 8 m above grade shall be used.

Where the vessel is on a platform or flat roof, the roof or platform shall be considered grade level if it can sustain a fire.

For spheres or spheroids, the vertical distance covered by fire is the elevation of the maximum horizontal diameter, or a height of 8 m, whichever is greater.

Area covered by Fire:

For horizontal accumulators, the fire covers the area in the two heads and the horizontal wetted surface within the height limitation above grade. The horizontal wetted surface depends on the internal arrangement of the vessel and the maximum liquid level covered by the level control instrument.

For vertical accumulators, the fire covers the area in the bottom head and the vertical wetted surface up to the height limitation. The maximum liquid level covered by the level instrument governs the vertical wetted surface.

In settlers operating full of liquid, total surface area up to the height limitation shall be used as the wetted surface.

In charge drums or surge drums, the total surface area up to the height limitation shall be used as the wetted surface, because it is assumed these vessels can be operated full.

For vessels in pits, the fire is assumed to cover the entire surface area.

For fractionators, strippers and absorbers, the area of the bottom head exposed to fire and the vertical area within height limitation shall be considered as wetted surface.

The area covered by a fire is considered to be a circle 24 m in diameter around a given tower. Presence of firewalls to contain liquid shall also be considered.

Storage vessels are assumed to contain their maximum working volume.

Heat Input Due to Fire:

Heat input shall be calculated according to API RP 520 and the criteria set forth in this section.

Quantity to be Released from Liquid Storage:

Heat input due to fire is assumed to vapourize the stored products at relieving pressure.

Relieving temperature shall correspond to the boiling point of the material, corrected to the absolute relieving pressure.

Where material is vapourizing from a constant boiling mixture, the temperature shall be the boiling point of the mixture at absolute relieving pressure.

Where water, caustic or amine solution or other liquids exist in the presence of another liquid, the temperature shall correspond to the temperature at which the partial pressure of the different components add up to the relieving pressure.

Heat Exchanger Tube Failure:

Heat exchanger tube failure shall be evaluated according to API RP 521, Section 2, Clause 2.3.13, and according to Section 3, Clause 3.18 for relief load estimation.

Flow through the ruptured tube shall be determined according to CRANE Technical Paper No. 410, Section 3.4.

Hydraulic Expansion:

Hydraulic expansion shall be considered according to API RP 521, Section 3, Clause 3.14. Requirements for the thermal relief valves for process piping shall be evaluated according to pipe length and diameter.

9.4.1.3 Equipment Safety Relief Valve Capacity

Reciprocating Compressors:

Compressor safety valves shall be sized for the maximum capacity of each stage, at its recommended speed.

Positive Displacement Pumps:

Safety valves on positive displacement pumps shall be sized for the maximum pumping capacity at its recommended speed.

Where the pump and equipment pumped into are designed for stalling pressure of the pump under normal conditions, the safety valve shall be sized for 25% of the pump-rated capacity under continuous operating conditions.

Sizing for Protection of Electric Motors on Positive Displacement Pumps:

Safety valves for protecting electric motors shall be sized for the quantity that can be pumped at the relief valve setting without exceeding the horsepower rating (in kilowatts) recommended by the motor manufacturer.

Particular consideration shall be given to effects of viscosity on horsepower (in kilowatts).

Sizing for Protection of Gas Turbines:

Safety valves to protect the casing of gas turbines or expansion joints on turbines shall be sized for the quantity of effluent gases developed by the turbine at the safety valve relieving pressure. The quantity shall be checked with the equipment manufacturer. Consideration shall be given to pressure drop in the effluent gas header before the safety valve.

Sizing for Protection of Shell and Tube Heat Exchangers:

Safety valves shall be sized to protect shell and tube equipment for the pump capacity at relieving pressure.

Safety valves shall be sized for wide-open capacity of the control valve at its normal inlet pressure, and the control valve outlet pressure shall be equal to the relieving pressure of the safety valve.

Cold side blockage safety valves shall be sized according to assumed heat input equal to the design exchanger duty. The temperature and composition of the relieving fluid, and the required safety valve capacity, will depend upon the location of the safety valve relative to the exchanger. For example, a safety valve located on the exchanger will relieve heated vapours, whereas a safety valve in the piping 15 m upstream of the exchanger will relieve cold fluid. Each situation shall be individually examined.

Fired Heaters:

Safety valves on fired heaters (if required) shall be sized so that no less than 50% of normal heat input shall be used, due to residual heat input from the hot firebox. However, as cracking will often occur, each situation will require careful evaluation of the individual system conditions to determine the required relieving rate.

Vessels/Towers:

Safety valves for towers and other types of pressure vessels shall be checked by each of the following methods, and sized for the most severe condition:

- Vessel blocked outlet
- Failure of condensing duty
- Failure of reflux

- Simultaneous failure of reflux and condensing duty
- Fire
- Electrical or electrical plus cooling water failure (this case may be covered in the above points).

9.4.2 Vacuum Relief Valve Sizing

Vessels operating full of liquid shall be sized to provide protection against loss of liquid input, depending on rate of withdrawal.

Protection against loss of stripping steam depends on whether the vessel is bare or insulated, and on the maximum possible condensation that can occur in a rain or snow storm.

Where a column or vessel during start-up, abnormal operating conditions, heating medium failure, etc. can be subject to vacuum, it should normally be designed to withstand vacuum conditions. Only when this is very costly or impractical may ways for introducing fuel gas or inert gas at a sufficient rate be considered. Approval of Owner shall be obtained in this case, but steaming-out shall not be considered.

9.4.3 Thermal Expansion Valve Sizing

Thermal expansion valves are not sized. They are generally selected according to the minimum size connection permitted on piping for the particular service. For cases where expansion relief is required around check valves, block valves or pumps, sizes down to DN 15 (½ in.) may be acceptable, provided that applicable piping material service specification permits screwed connections for small sizes, and Approval by Owner is obtained.

9.4.4 Emergency Vapour Depressuring Systems

Vessels operating at less than 7 Bar(g) will use a PV as the depressuring valve.

Sizing of depressuring systems shall be based on the assumption that during the fire, all input and output streams to and from the system have been stopped and all internal heat sources have been shut off.

The pressure of the equipment shall be reduced to 7 Bar(g) or 50% of the MAWP, whichever is lower, within 15 minutes, while vapour is being generated at a rate corresponding to the following:

- Vapour generated from liquid by heat input from the fire, plus
- Density change of the vapour in the equipment during pressure reduction, plus
- Liquid flashing during pressure reduction.

Vessels and columns need not be provided with emergency vapour depressuring facilities if, as a result of a fire, the pressure in the equipment will not rise to 700 kPa(g) or 50% of the MAWP within 30 minutes.

Subject to corrections made necessary by the final plant design, the following assumptions may be made for estimating purposes:

- The liquid inventory of fractionating columns can be estimated as the normal column bottom and draw-off tray capacity plus normal tray liquid hold-up (total tray liquid hold-up shall be based on the pressure drop over the column).
- The liquid inventory of accumulators, flash drums, knockout vessels and the like may be based on normal operating levels.
- For shell and tube heat exchangers, it may be assumed that one-third of the total shell volume is occupied by the tube bundle.

Depressuring valves shall be located near the vessel or vapour lines to be protected.

Valves shall be spring-loaded, be pneumatic diaphragm operated, have tight shut-off, and spring to open.

The minimum size of depressuring valves shall be DN 25 (1 in.).

Block valves shall be provided to isolate the depressuring valve.

Air for depressuring valves shall be supplied from the instrument air system. They shall not be provided with a handwheel.

9.5 Arrangement of Safety Relief Valves

9.5.1 General

Safety devices shall be arranged so that their proper functioning is not hampered by the nature of the vessel contents. If necessary, the use of protecting devices such as rupture discs, swan-neck seals or purge arrangements is permissible, but these shall not interfere with the proper functioning of the safety device.

If a rupture disc is used in combination with a safety relief valve, a pressure gauge preceded by a block valve shall be provided between rupture disc and safety relief valve.

Pressure drop in the connection between equipment and safety relief valve should not exceed 3% of the valve set pressure.

Safety relief valves discharging to atmosphere shall in general be located at maximum practical elevation to economize on discharge piping, considering ease of maintenance.

Safety relief valves connected to a closed relief system shall be located slightly above the relief header, if possible. For example, valves protecting tall columns shall be put in a suitable position on the overhead vapour line.

If valves must be below the header, outlet lines leading to the header shall be steam traced from the safety relief valve to their highest point. However, Approval by Owner shall be obtained for lowering the valve in such a way. Branches should be connected to the headers so that the latter can not drain back into them, even with the header full of liquid.

9.5.2 Location on Vessels

Where a safety valve is provided for a vessel, the connection for the safety valve shall be provided on the vessel and not on the vapour line or discharge line from the vessel, except as follows:

When access to the safety valve or discharge piping arrangement can be improved by locating the safety valve on vessel piping, such a location is permissible, provided the maximum pressure drop permitted between the vessel and safety valve inlet does not exceed the allowable value according to the previous paragraph, and provided the safety valve can be properly supported considering reaction forces.

Safety valve nozzles shall be vertical when placed in the top head of vessels, except as follows:

In cases where vessel diameter is small enough to prohibit vertical nozzles, the safety valve nozzle should be attached at a 45 degree angle from the top head, with a 45° elbow so that the flange face is horizontal and above the nozzle.

9.5.3 Location of Safety Valve Nozzles to Minimize Turbulence

Where safety valves are installed to protect piping and are located downstream of control valves, pressure reducing stations, orifice plates, flow nozzles or pipe fittings such as elbows, which can cause turbulence, the safety valve shall be a sufficient distance downstream of these devices to avoid improper operation of the safety valves due to turbulence.

The minimum number of straight-pipe diameters that must be provided between the source of turbulence and the safety valve shall be according to API RP 520, Part II.

9.5.4 Location of Safety Valve Nozzles to Minimize Pulsation

Where there are pressure fluctuations at the pressure source (discharge of reciprocating pumps or compressors) that peak close to the set pressure of the safety valve, it is beneficial to locate the safety valves farther from the pressure source, in a more stable pressure region.

9.5.5 Inlet Piping of Safety Relief Valves

Piping from the vessel to the safety valve inlet shall be, at minimum, the same size as the safety valve inlet connection. It shall have a pressure drop, including that in any block valve or fittings, of 3% or less of the set pressure.

Where this requirement can not be met, ways must be found to reduce the pressure drop by rounding the entrance connection, using block valves (when required) with full area ports, or by enlarging the inlet piping.

The inlet piping or nozzles to safety valves on vessels shall be kept as short and direct as possible, preferably a single nozzle for each valve. Where more than one safety valve is involved, avoid mounting them on a common header tee type nozzle to prevent turbulence and excessive pressure drop.

9.5.6 Discharge Piping of Safety Relief Valves

9.5.6.1 General

In the design of outlet piping, the effect of superimposed or build-up back pressure on the particular type of valve and its service shall be considered.

In addition to any specific requirements set forth in this document, determination of discharging pressure and safety valves to atmosphere or to a closed system shall be governed by API RP 520, Part I and IPS-E-PR-460.

9.5.6.2 Permissible Pressure Drop in Closed Flare System

The permissible pressure drop from the outlet of a safety valve to a flare system shall be as given for the standard safety valve or balanced bellows safety valve.

The pressure drop permitted shall be limited to a great extent by the lowest set safety valve in the system, because this determines the pressure permitted at the point where this safety valve ties in to the flare header.

The pressure drop in any emergency cooler and the pressure required at the unit flare knockout drum must be subtracted from the permitted pressure drop.

A minimum of 35 kPa(g) shall be used for the pressure in the unit flare knockout drum. The type of safety valve used on the lowest set safety valve shall permit this back pressure to exist.

9.5.6.3 Closed header design considerations

A single maximum discharge occurrence for one safety valve or manual depressuring valve shall be used to determine header size.

The header shall be sized according to maximum relief load, as outlined in Section 9.1.

Consideration shall be given to using balanced bellows safety valves where pressures available are limiting.

Unless further restricted by job specifications, hot streams that flow into the closed above-ground headers shall be limited to 232°C maximum downstream of the safety valve. An emergency cooler shall be provided to cool to 232°C all relief streams that exceed this maximum, except for the following:

- A relief stream where a temperature higher than 232°C will be reached under fire conditions, and where the safety valve size is not set by fire conditions, need not have an emergency cooler unless it is required for other reasons. A cooler is required for such streams when the safety valve size is set by fire conditions.
- Under the conditions permitted in the point above, the piping and flare appurtenances shall be designed for the temperatures involved.

9.5.6.4 Vapour Safety Valve Piping

Sizing:

Sizing of discharge piping for vapour safety valves depends on whether the valve is relieving to atmosphere, or to a closed relieving system, and whether a standard or bellows type safety valve is used.

Permissable Pressure Drop:

Conventional (Standard) Safety Valve:

The permissible pressure drop in discharge piping from the safety valve in vapour service is 10% of the set pressure under normal relief where 10% accumulation is used, and 20% of the set pressure under emergency fire condition where 20% accumulation is used.

Balanced Bellows Safety Valve:

The permissible pressure drop in discharge piping for balanced bellows safety valves in vapour service is 25% to 45% of valve set pressure. For these particular valves, the manufacturer's curve shall be used.

9.5.6.5 Liquid Relief Valve Piping

Because the pressure drop in the outlet piping affects the set pressure of a conventional safety relief valve and the capacity of both a conventional and a bellows relief valve, these impacts shall be evaluated when determining the size of outlet piping.

Most liquid relief valves have short discharge lines, for example, relieving into suction of a pump, where the discharge line relieves back to tankage or a tower. A considerable pressure drop can develop, and its effect must be considered.

Discharge piping from safety valves in liquid service shall be no smaller than the outlet connection on the safety valve and shall be increased where pressure drop significantly affects set pressure or capacity of the relief valve.

9.5.7 Block Valves

9.5.7.1 General

When spare safety valves are required, block valves shall be provided, as follows, to permit removal of all valves being spared and the spare safety valve:

- Block valves shall be provided only on the inlet of the safety valve where the safety valve discharges independently to the atmosphere.
- Block valves shall be provided on the inlet and outlet of a safety valve that discharges into a system operating under pressure or into a common discharge header.
- Block valves, according to the applicable piping material service specification, shall be provided in the pressure sensing piping to the pilot valves and in the inlet piping to the controlled relief valves, when spare safety valves are required.

Although it is not permitted, except as noted above, provision for block valves and bypass lines may be requested in the job specification for all safety relief valves providing protection of vessel(s), in order to make possible removal of safety relief valves for maintenance purposes while the unit or system is under normal operation. In such cases, safety features concerning the equipment or system shall be taken into account. The inlet and outlet block valves shall be locked open in any case.

9.5.7.2 Block Valves on Inlet of Safety Valves

Block valves conforming to applicable inlet piping material service specifications shall be used.

Block valves for the inlets of safety valves shall be, at minimum, the same size as the safety valve inlet. They shall be larger where necessary to reduce the pressure drop in the inlet piping to that essential for proper operation of the safety valve.

9.5.7.3 Block Valves on Outlet of Safety Valves

Block valves conforming to the piping material service specification for the service involved at the discharge shall be used.

Block valves for the outlet of safety valves shall be, at minimum, the same size as the outlet of the safety valve. They shall be larger where required to reduce the pressure drop in the outlet piping to ensure proper operation at the capacity required.

9.5.7.4 Locking Methods

Suitable devices shall be provided on all block valves at the inlet or the outlet of safety valves, which will allow locking of the valves by authorized persons in an open or closed position. The job specification will specify whether the valves are suitable for a padlock and chain or a car seal, depending on plant preference.

9.5.8 Discharge Piping of Temperature Safety Valves

9.5.8.1 Water service

Temperature safety valves in water service may discharge to grade or to a sewer.

9.5.8.2 Hydrocarbon Service

Temperature safety valves in hydrocarbon service on process units shall pipe to the nearest safe location, such as drain or sewer hub, with the end of the discharge pipe visible.

Temperature safety valves located in National Electrical Code, Division 2 electrical classification areas shall have discharge piped to an open drain or sewer, with the end visible outside the Division 2 area.

Temperature safety valves in storage or product loading areas shall not discharge to atmosphere but shall have discharge piped, using a spring loaded popped type check valve around block valves, check valves, pumps, etc. back to storage.

9.5.8.3 Chemical Service

The procedure for handling discharge from temperature safety valves in chemical service shall be specified in the job specification.

10.0 FLARE AND BLOWDOWN SYSTEM

10.1 Selection of Blowdown Systems

10.1.1 General

Although various systems for the disposal of voluntary or involuntary vapour or liquid are discussed below, actual selection of a disposal system shall be conducted according to expected frequency, duration of operation, required capacity and fluid properties.

10.1.2 Blowdown System for Vapour Relief Stream

Systems for the disposal of voluntary and involuntary vapour discharges are as follows:

- To atmosphere
- To lower pressure process vessel or system
- To closed pressure relief system and flare

- To acid gas flare.

10.1.2.1 Vapour Discharge to Atmosphere

Vapour relief streams shall be vented directly to atmosphere if all of the following conditions are satisfied (for a complete discussion on the subject see API RP 521):

- Such disposal is not in conflict with the present regulations concerning pollution and noise.
- Vapour is effectively non-toxic and non-corrosive.
- Vapour is lighter than air or of any molecular mass that is non-flammable, non-hazardous and non-condensable.
- There is no risk of condensation of flammable or corrosive materials.
- There is no chance of simultaneous release of liquid, except water.
- Relief of flammable hydrocarbons directly to the atmosphere should be restricted to cases where they will be diluted with air to below the lower flammable limit. This should occur well before they come in contact with any source of ignition.
- The above condition can most easily be met if the vapours to be released have a density less than that of air.

Exceptions:

- Vapour from depressuring valves shall be discharged to a closed pressure relief system.
- Vapour that contains 1% H_2S or more by volume shall be discharged to a closed pressure relief system.

10.1.2.2 Vapour Discharge to Lower Pressure Process Vessel or System

Individual safety relief valves may discharge to a lower pressure process system or vessel capable of handling the discharge. Although this is rarely used, it is effective for discharges that contain materials that must be recovered.

10.1.2.3 Vapour Discharge to Closed Pressure Relief System and Flare

In all cases where the atmospheric discharge of vapour to a lower pressure system is not permissible or practicable, vapour shall be collected in a closed pressure relief system that terminates in a flare system. Where the concentration of H_2S is such that condensation of acid gas is probable, provision for a separate line, heat traced, shall be considered. In all cases, the installation of a closed pressure relief system shall result in minimal release of combustion products.

In process plants where H_2S free and H_2S containing streams are to be flared, consideration should be given to the installation of a separate header and flare stack assembly for the H_2S containing streams. The following provisions should be considered for the acid flare assembly:

- Fuel gas can be automatically injected downstream of the H_2S pot in order to make the combustion stable.
- Steam injection for a smokeless operation shall not be considered for an H_2S flare tip.
- A common pilot igniter shall be used to ignite all flare stacks, including the acid flare.
- The H_2S flare header and subheaders shall be heat traced in order to prevent condensation of acid gas.

10.1.3 Blowdown System for Liquid Relief Stream

Systems for the disposal of voluntary and involuntary liquid discharges are as follows:

- To onsite liquid blowdown drum
- To lower pressure process vessel or system
- To oily water sewers, only if the material will not cause hazardous conditions
- To pump suction, if pump will not overheat or can withstand the expected temperature rise
- To burning pit
- To vapourizer.

Thermal expansion relief valves may discharge small quantities of volatile liquid or vapour into the atmosphere, provided the valve outlet is in a safe location.

10.1.3.1 Liquid Discharge to Onsite Liquid Blowdown Drum

The liquid shall be discharged to an onsite liquid blowdown drum capable of retaining the liquid discharged at the required liquid relief rate for a period of 20 minutes. This drum shall have a vapour discharge line to the closed pressure relief system.

10.1.3.2 Liquid Discharge to Lower Pressure Process Vessel or System

The liquid shall be discharged to a lower pressure process vessel or system of handling the required liquid relief rate and any flashed vapour.

10.1.3.3 Liquid Discharge to Oily Water Sewer

Liquid discharge to an oily water sewer shall be non-volatile and non-toxic. The required liquid relief rate shall be within the oil removal capability of the oily water treating system.

10.1.3.4 Liquid Discharge to Pump Suction

Required liquid relief shall discharge to an upstream liquid reservoir from which the pump takes suction. The liquid relief may discharge directly to the pump suction line if sufficient cooling is provided to prevent a temperature rise of the liquid recycled through the pump when the safety relief valve opens or when a constant displacement pump is used.

10.1.3.5 Liquid Discharge to Burning Pit

Liquid relief or voluntary liquid blowdown that does not need to be returned to the process, or discharged to an oily water sewer, shall be discharged to a burning pit, if environmentally acceptable.

10.1.3.6 Liquid Discharge to Vapourizer

The liquid shall be discharged to a vapourizer capable of vapourizing a liquid relief of no more than 5,000 kg/h.

10.2 Design of Disposal System Components

Depending on the process plant under consideration, a disposal system could consist of a combination of the following items: piping, knockout drum, quench drum, seal drum, flare stack, ignition system, flare tip, and burning pit.

10.2.1 Piping

10.2.1.1 General

In general, the design of disposal piping should conform to ANSI/ASME B31.3. Installation details should conform to API RP 520, Part II.

10.2.1.2 Inlet Piping

The design of inlet piping should be according to API RP 521, Section 5.4.1.

10.2.1.3 Vapour Relief Header

The size of the vapour relief header should be according to API RP521, Section 5.4.1.2; in conjunction with IPS-E-PR-460, Tables 1 & 2 and Appendix A; and API RP 520, Part I, Section 7 and Appendix C.

10.2.1.4 Liquid Blowdown Header

To reduce relief header loads and prevent surges due to two-phase gas/liquid flow as much as possible, it is advisable to direct all disposable liquids into a separate blowdown network.

Once maximum load and back pressure in each segment have been established, standard pipe sizing procedures shall be used (see IPS-E-PR-440).

Back pressure shall be determined as discussed in the section on discharge piping. The following shall also be taken into consideration:

- Flashing of liquid at the safety relief valve discharge or along the network due to pressure drop and/or warm-up to ambient temperatures should be analyzed.
- Solids formation due to auto refrigeration and presence of high melting point liquids should be determined.
- If flashing and auto refrigeration is possible, a temperature profile along the network should be established so that proper piping material selection and construction practices are undertaken.
- The network should be self-draining and should not include pockets.
- The network should be continuously purged by natural gas, controlled through an orifice.
- High liquid velocities should be monitored within the network (see IPS-E-PR-440).

10.2.1.5 Drainage

Disposal system piping should be self-draining toward the discharge end. Pocketing of discharge lines should be avoided. Where pressure relief valves handle viscous materials, or materials that can solidify as they cool to ambient temperature, the discharge line should be heat traced. A small drain pot or drip leg may be necessary at low points in lines that can not be sloped continuously to the knockout or blowdown drum. The use of traps or other devices with operating mechanisms should be avoided.

10.2.1.6 Details

Safety relief valve connections to the header normally have laterals from individual relieving devices enter a header from above.

Safety relief valve connections, when installed below the relief header, have laterals leading from individual valves located at an elevation above the header drain. Locating a safety valve below the header elevation in closed systems should be avoided. Laterals from individual valves that must be located below the header should be arranged to rise continuously to the top of the header entry point. However, ways should be provided to prevent liquid accumulation on the discharge side of these valves.

The following should be taken into consideration:

- For the branch header that must be connected to the main header from a lower level than the main header, such as sleeper flare piping, a drain pot must be installed.

- If a safety relief valve must be installed below the flare header, the outlet line leading to the flare header shall be heat traced from the safety relief valve to their highest point. This arrangement of safety relief valves shall be reviewed, because such an arrangement is not permitted for safety relief valves that discharge a medium that can leave a residue.

Heat tracing can be omitted if the safety relief valve in question handles only products that vapourize completely, or do not condense at all, at the lowest ambient temperature.

Purge Point of Gas for Dry Seal

A continuous fuel gas purge shall be installed at the end of the main header and any major subheader. The fuel gas purge shall be controlled by a restriction orifice.

Purge gas volume shall be determined so that a positive pressure is maintained and air ingress is prevented.

Insulation of Flare Line

Normally, insulation of the flare line (including the outlet line of the safety relief valve) is not required, except for Personnel protection. To avoid hydrate formation, ice accumulation, etc. within the flare line, however, insulation or heat tracing shall be considered.

Location of Safety Relief Valve

More than one piece of equipment may be protected by a common safety relief valve, provided they are connected by a line of sufficient size and no block valve exists on the connecting lines.

Valves on Inlet/Outlet Line of Safety Relief Valve

All safety relief valves shall have block valves on inlet and outlet to facilitate maintenance. Block valves must be locked open. Safety valves discharging to the atmosphere shall not have block valves on the outlet. A bypass line with a valve shall be provided for each safety valve, except thermal expansion valves.

Provision for Installation of Drain Holes

Where individual valves are vented to the atmosphere, an adequate drain hole (DN 15 is suitable) should be provided at the low point, to ensure that no liquid collects downstream of the valve. The vapour flow that occurs through this hole during venting is not significant, but each case should be checked to see if the drain connection should be piped to a safe location. Vapours escaping from the drain hole must not be allowed to impinge on the vessel shell, because accidental ignition of such vent streams can seriously weaken the shell.

Angle Entry Into the Relief Header

The use of angle entry at 45 degrees (0.79 radian) or 30 degrees (0.52 radian) to the header axis for laterals is much more common in relieving systems than in most process piping systems.

Installation of Valves and Blinds in Relief Headers

Ways (valve and blind) must be provided to isolate each unit from the flare system, for reasons of safety and maintenance.

Slope of Flare Header

A slope of 1 m in 500 m is suggested for the flare header.

10.2.2 Sizing a Knockout Drum

See IPS-E-PR-460, Appendix B, Figures. 2 and 3 and Table 3.

10.2.3 Quench Drum

10.2.3.1 General

A quench drum is provided to prevent liquid hydrocarbon condensation in the flare system, to reduce flare capacity requirements, or to prevent discharge of condensable hydrocarbons to the atmosphere. In some cases, it reduces the maximum temperature of flare gases and, therefore, minimizes thermal expansion problems in the mechanical design of flare headers. The quench drum has a direct contact water spray arrangement that condenses entering heavy hydrocarbon vapours. Condensed hydrocarbons and effluent water are discharged through a seal to the sewer or through the pumpout to slop tankage, whereas uncondensed hydrocarbon vapours are vented to flare or to the atmosphere.

10.2.3.2 Details

The quench drum shall have design pressure capable of withstanding the maximum back pressure. Minimum design pressure is 3.5 Bar(g).

Water requirements are normally based on reducing gas and liquid outlet temperatures to about 50°C. Optimum temperature is based on temperature and composition of entering streams, and the extent to which condensation of effluent vapours downstream of the drum can be tolerated.

It is generally assumed that no more than 40% to 50% of the liquid feed will be vapourized. The water supply should be taken from a reliable water system. If a recirculating cooling water system is used, then the circulating pumps and cooling water basin must have adequate capacity to supply the maximum quench drum requirements for 20 minutes.

The seal height in the liquid effluent line (assuming 100% water) is sized for 175% of the maximum operating pressure, or 3 m, whichever is greater.

10.2.4 Sizing a Seal Drum

Sizing of a seal drum and design details should be according to API RP 521, Sections 5.4.2.2 and 5.4.2.4; and API RP 2001, Section 3.14.3.

For treating sour water discharge from seal drums, see IPS-E-PR-725.

10.2.5 Flares

10.2.5.1 General

Flare systems provide for safe disposal of gaseous plant wastes. Depending on local environmental constraints, these systems can be used for the following:

- Extensive venting during start-up or shutdown
- Venting of excess process plant gas
- Handling emergency releases from safety valves, blowdown, and depressuring systems.

  Designs vary considerably, depending upon the type of connected equipment and the complexity of the overall system.

  A flare system generally consists of an elevated stack, ways to maintain burning conditions at the top of the stack, and ways to prevent flashback within the system.

10.2.5.2 Sizing

Sizing of flares depends on required stack diameter and required stack height. Other factors governing size are wind effect and dispersion.

Since the flare tip is open to the atmosphere, high gas velocities are expected at this point. Very high tip velocities cause a phenomenon known as blowoff, where the flame front is lifted and could eventually turn into a blowout. Very low velocities could damage the flare tip, due to high heat intensities and smoking. In this case, ingress of air in the system and creation of a flammable mixture is possible. Therefore, the correct flare diameter is important for operation of the system.

The location and height of flare stacks should be based upon the heat release potential of the flare, the possibility of Personnel exposure during flaring, and exposure of surrounding plant equipment. There are exposure limitations that fix the distance between flame and object. If there are limitations on the distance, then stack height can be calculated. An optimum trade-off between height and distance should be applied.

Because it tilts the frame, wind changes flame distance and heat intensity. Its velocity should be considered, therefore, in determining stack height.

If the flare is blown out (extinguished), or if there are environmental hazards associated with flare output, the possibility of a hazardous situation downwind should be analyzed.

Diameter:

Flare stack diameter is generally based on velocity, although pressure drop should be considered. Depending on the volume ratio of maximum conceivable flare flow to anticipated average flare flow, the probable timing, frequency, and duration of those flows, and the design criteria established for the project to stabilize flare burning, it might be desirable to permit a velocity of up to 0.7 Mach for a peak, short-term, infrequent flow, with 0.2 Mach maintained for normal and, possibly, more frequent conditions. Smokeless flares should be sized for the conditions under which they are to operate smokelessly.

Height:

Flare stack height is generally based on the radiant heat intensity generated by the flame. See IPS-E-PR-460, Section 7.5.2, p. 19 for the applicable formula and methodology in calculating height.

10.2.5.3 Design Details

Flashback Protection:

The most common method of preventing propagation of flame into the flare system due to entry of air is installation of a seal drum. Flame arresters are occasionally used for flashback protection; however, they are subject to plugging and their application is limited.

Ignition:

To ensure ignition of flare gases, a continuous pilot with remote ignition is recommended for all flares. The commonly used type of igniter is the flame-front propagation type, which uses a spark from a remote location to ignite a flammable mixture.

- Pilot igniter controls should be located at a safe distance from the base of elevated flares and at least 30 m from ground flares.
- A low-pressure alarm for pilot gas is recommended so that the operator in the control room is aware of pilot blowout.
- Reliable pilot operation under all wind and weather conditions is essential. Flaring operations are, for the most part, intermittent and non-scheduled. The flare must be immediately available for emergency duty, to prevent any possibility of a hazardous or environmentally offensive discharge to the atmosphere. Windshields and flame retention devices may be used to ensure continuous piloting under the most adverse conditions.

Fuel System:

Fuel gas supply to the pilots and igniters must be reliable. Since normal plant fuel sources can be upset or lost, it is desirable to provide a backup system connected to the most reliable alternative fuel source, with provision for automatic cut-in on low pressure. Use of waste gas with low energy content or with unusual burning characteristics should be avoided. Parallel instrumentation for pressure reduction is frequently justifiable. The flare fuel system should be carefully checked to ensure that hydrates do not present a problem. Because of small lines' long exposed runs, large vertical rises up the stack, and pressure reductions, use of a liquid knockout pot or scrubber after the last pressure reduction is frequently warranted. If at all feasible in terms of distance, relative location, and cost, it is good practice to install a low-pressure alarm on the fuel supply, after the last regulator or control valve, so that operators will be warned of any loss of fuel to the pilots.

Location:

The location of flares in the vicinity of tall equipment should be examined. Flames or hot combustion products can be carried by the wind, which could cause problems and create hazards to Personnel working on these elevated structures at the time of a flare release. As discussed in the section on sizing, flare height and distance are dependent on radiation intensity. When either the height or the distance from the plant of a flare is fixed, the other can be determined. Usually there are constraints on the distance, so stack height is calculated. If there are no constraints on the distance, and flare height is to be determined, the following guideline is recommended:

For stack heights less than 23 m, a distance of 91 m; and for stack heights greater than 23 m, a distance of 61 m from the plant (see GPSA, Section. 2.1).

Consideration should be given to installation of flow measuring equipment on the flare system. Specifically, sub-headers handling relief loads from individual units shall be provided with proper flow elements.

10.2.6 Burning Pits

Burning pit flares can handle flammable liquids or gases, or mixtures of the two. A typical design is shown in IPS-E-PR-460, Figure 8. In this figure, a circular pit is illustrated, but any convenient shape may be used.

The burning pit is simply a shallow earth or concrete-surfaced pool area enclosed by a dike wall, with a liquid/vapour inlet pipe through the wall, and a pilot and igniter. Although the design basis flow is adequate for handling emergency releases, a more conservative approach is recommended for continuous flaring services, incorporating up to twice the calculated pit area.

10.2.6.1 Burning Pit Flare Sizing

The burning pit area size shall provide sufficient surface to vapourize and burn liquid at a rate equal to the maximum incoming liquid rate. The calculation procedure is as follows:

- Determine the linear regression rate of the liquid surface (i.e., the rate at which the liquid level would fall as a result of vapourization by radiant heat from the burning vapour above it, assuming no addition of incoming liquid).
- Determine the pit area necessary to vapourize and burn liquid at a rate equal to the liquid input rate.
- Design the dike wall height above the water level to provide hollow capacity for the largest liquid release resulting from a single contingency during 30 minutes, plus 460 mm free board. The liquid rate is based on the actual flashed liquid entering the pit, assuming no burning or further vapourization in the pit. The height of the dike wall above the water level should not, however, be less than 1.20 m.

10.2.6.2 Spacing of Burning Pit Flares

Spacing is based upon radiant heat consideration at maximum heat release, using a simplified calculation procedure. The equation has been described in the section on flare sizing. In this equation, absorption of radiation by surrounding air is neglected.

The fraction of heat radiated, as shown in IPS-E-PR-460, Table 4 refers to light gases, but in burning pits, combustion of liquid is under consideration. Therefore, good engineering judgement should be exercised in evaluating its effect when determining the distance.

The center of the flame is assumed to be 1.5 pool diameters from the center of the pool, in the direction of the point where radiant heat density is being considered. This assumption is used to allow for flame deflection by wind.

Although permissible radiant heat densities are provided in IPS-E-PR-460, Table 5, its value at the property line must not exceed 1.60 kW/m^2.

In addition, the following minimum spacing applies to burning pits:

- 150 m from property lines, roadways, or any process or storage facilities
- 60 m from any source of ignitable hydrocarbons (such as separators) or floating roof tanks.

Valves in the inlet, seal water and pilot gas lines should be located according to permissible radiant heat densities for Personnel. Piping to the burning pit should be suitably protected against flame impingement (e.g., by installation below grade).

In designing the burning pit, all safety precautions for Personnel and equipment should be observed.

11.0 EQUIPMENT AND PIPING INSULATION

TABLE 3 – COLD SERVICE INSULATION
Minimum Insulation Thickness (Millimetres)

Pipe size DN	PIPE SIZE NPS	Amb. TO 10°C	9°C TO 0°C	-1°C TO -10°C	-11°C TO -20°C	-21°C TO -30°C	-31°C TO -40°C	-41°C TO -50°C	-51°C TO -75°C	-76°C TO -100°C	-101°C TO -125°C
15	½	25	25	25	38	38	38	50	50	64	64
20	¾	25	25	38	38	38	50	50	64	64	75
25	1	25	25	38	38	38	50	50	64	75	90
40	1 ½	25	25	38	38	50	50	50	64	75	90
50	2	25	38	38	50	64	64	75	75	100	115
80	3	25	38	38	50	64	75	90	100	115	125
100	4	25	38	38	50	75	75	90	100	115	125
150	6	38	50	50	64	75	90	90	115	125	140
200	8	38	50	50	64	75	90	100	125	140	150
250	10	38	50	64	64	90	90	100	125	140	165
300	12	38	50	64	75	90	100	100	125	150	165
350	14	38	50	64	75	90	100	115	140	150	180
400	16	38	50	64	75	90	100	115	140	165	180
450	18	38	50	64	75	90	100	115	140	165	180
500	20	38	50	64	75	90	100	115	140	165	190
600	24	38	50	64	75	100	100	115	150	165	190
750	30	38	50	64	75	100	115	125	150	165	205
900	36	38	50	64	75	100	115	125	150	180	205
1050 & UP	42 & UP	38	50	64	75	100	115	125	165	190	205

NOTES:

1. The above cold insulation thicknesses are based on 80% humitdity, ambient are 32°C, zero km/hr velocity and a minimum surface temperature of 29°C.
2. Use double layer on all lines operating below -50°C and 100 mm or greater thickness.

12.0 EQUIPMENT SPARING PHILOSOPHY

The following is the general philosophy for sparing of rotating equipment, unless otherwise specified in the design documents.

° Critical Pumps:	2X50% or 2X100%
° Product Pipeline Pumps:	3X60%
° Non-critical or Intermittent Pumps:	1X100%
° Critical Compressors:	3X50% or 2X100%
° Non-critical compressors:	2X50% or 2X100%
° Residue Gas Compressors:	6X20%

In general, the selection of the number of pumps or compressors for different services has taken into account:

- The number of process trains
- The criticality of the service
- The reliability of the equipment (eg centrifugal versus reciprocating compressors)
- Life cycle cost of the equipment (this includes installation plus operating/maintenance costs).