Exploration & Production

GENERAL SPECIFICATION

PROCESS

GS ECP 103

Process sizing criteria

02	09/03	Change of Group name and logo
01	09/02	Replaces GS EXP 303
00	11/01	First issue
Rev.	Date	Notes

Contents

1. Scope

This general specification gives minimum requirements for Process Sizing Criteria at either basic or detailed engineering phase.

This general specification is intended for use for facilities such as oil and gas production plants, gas plants, LNG plants and LNG import terminals.

2. Reference documents

The reference documents listed below form an integral part of this General Specification. Unless otherwise stipulated, the applicable version of these documents, including relevant appendices and supplements, is the latest revision published at the EFFECTIVE DATE of the CONTRACT.

Standards

Reference	Title
Not applicable	

Professional Documents

Reference	Title
API RP 14 E	Recommended Practice for Design and Installation of Offshore Production Platform Piping Systems
API RP 521	Guide for Pressure - Relieving and Depressuring Systems
API RP 520 Part II	Sizing, Selection and Installation of Pressure - Relieving Devices in Refineries

Regulations

Reference	Title
Not applicable	

Codes

Reference	Title
Not applicable	

Other documents

Reference	Title
Not applicable	

Total General Specifications

Reference	Title
GS SAF 228	Liquid Drainage
GS SAF 261	Pressure Protection & Relief, Emergency Shutdown & Depressurisation
GS SAF 262	Safety Rules For Hydrocarbon Disposal Systems
GS SAF 321	Fire Pump Stations and Fire Water Mains
GS PW 211	Design And Fabrication Of Pressure Vessels
GS ECP 101	Glycol Package Unit

3. General

3.1 Applicability

This specification is **not** retroactive. It shall apply to new installations and to major modifications or extensions of existing installations, both onshore and offshore. It is also applicable to Vendor's packages.

3.2 Terminology and definitions

Heat and material balance: It is the document giving for each process cases, the normal operating conditions, compositions, thermodynamic and physical properties for all streams.

Maximum operating conditions: There are the maximum pressure, temperature, flow rate, etc., in the equipment when the plant operates at unstable conditions corresponding to the high alarm set point.

Minimum operating conditions: There are the minimum pressure, temperature, flow rate, etc., in the equipment when the plant operates at unstable conditions corresponding to the low alarm set point.

Minimum Service Operating Temperature (mSOT): It is the minimum temperature obtained during normal operation, start-up, shutdown or process upset.

Normal operating conditions: There are the pressure, temperature, flow rate, etc., in the equipment when the plant operates at steady state conditions corresponding to the values of the heat and materiel balance.

Normal Service Operating Pressure (NSOP): It is the highest (for pressure above the atmospheric pressure) or the lowest (for pressure below the atmospheric pressure) normal operating pressure of all process cases when the plant operates at steady state conditions. If there is only one process case, NSOP = Normal Operating Pressure.

Normal Service Operating Temperature (NSOT): It is the highest normal operating temperature of all process cases when the plant operates at steady state conditions. If there is only one process case, NSOT = normal operating temperature.

P&ID (Piping and Instrumentation Diagram): It is the details drawing used for assistance of construction and operation of the processing plant including associated off sites and utility fluids.

P&ID's symbols are required for a good understanding and also to permit to give some symbolic representations on P&ID's.

Process: Discipline(s) in charge to study the facilities such as oil and gas production plants, gas plants, LNG plants, LNG import terminals, etc., but also all associated off sites and utility fluids.

Process data sheet: It is the data sheet for equipment, packages, control valves, PSV's, rupture disks, instruments, etc., containing the process information required for sizing and giving sometimes the sketch with main dimensions such as internal diameter, length, level elevation for equipment such as vessels, columns. It is issued by Process discipline.

The information given in the process data sheets are usually specified in the "mechanical data" sheet, issued for the purchasing of the said equipment. Refer to API, ASME, TEMA Codes or similar for such proposed Mechanical Data Sheet's.

PFD (Process Flow Diagram): It presents the succession of the operations in the fluid(s) processing to reach the required products specifications set in the objectives of the plant processing including also associated off sites and utility fluids.

Process rated flow: This terminology is mainly used for rotating equipment. It is the normal Process flow rate multiplied by the design factor and it shall be specified on the Process data sheet. It is generally the flow rate at the guaranteed point. The Process rated flow rate is the net Process flow because it does not take into consideration the losses (i.e. internal recycle, external consumption such as the air needed for air dryers regeneration for an air compressor package).

3.3 Abbreviations

LAH	Level Alarm High
LAHH	Level Alarm High High
ILAH	Interface Level Alarm High
ILAL	Interface Level Alarm Low
ILALL	Interface Level Alarm Low Low
INLL	Interface Normal Liquid Level
LG	Level Gauge
LAL	Level Alarm Low
LALL	Level Alarm Low Low
LSHH	Level Switch High High
LSLL	Level Switch Low Low
mSOT	minimum Service Operating Temperature
NLL	Normal Liquid Level
NPSH	Net Positive Suction Head
NSOP	Normal Service Operating Pressure
NSOT	Normal Service Operating Temperature
PAH	Pressure Alarm High
PSHH	Pressure Switch High High

PSV Pressure Safety Valve

SDV Shut Down Valve

4. Design pressures

4.1 General rules

The design pressure is the pressure for the mechanical design of a given piece of equipment.

The design pressure of a piece of equipment (excluding storage tanks, atmospheric tanks and pipelines) shall be taken as the following:

NSOP (bar g)	Minimum design pressure (bar g)
< 1	2 or 3.5 (*)minimum
1 – 10	NSOP + 1 bar (**)
> 10	NSOP + 10%

Notes: A margin of 10% is generally adequate in order to insure protection of the equipment with one conventional pressure relief valve (PSV), one Pressure Alarm High (PAH) and one Pressure Switch High High (PSHH).

It is reminded that the following tolerances are generally admitted for conventional instrumentation.

- PSV opening: ± 3%
- PSV closing: ± 5%
- PSV recommended leak test: 10% below set point
- Pressure transmitter (or pressure switch derived from a transmitter): ± 1%

In case of absolute necessity (for example in case of revamping or high pressure), the use of piloted operated PSV (tolerance of ± 1% on set point) could help to reduce the margin between the design pressure and the NSOP.

Unless otherwise specified, the design pressure specified by Process applies to the vapour phase at the top of the vessel.

The liquid density and the maximum liquid height shall be specified on the Process data sheet to allow the vessel designer to calculate the bottom thickness.

The design pressure shall also account for upset or transient conditions such as pressure surge, settle-out pressure at compressor suction, etc.

4.2 Pipelines

The Process Department usually provide only the MOP (Maximum Operating Pressure).

MOP is normally the design pressure of the last equipment upstream the pipeline plus the hydrostatic pressure due to the pipeline profile.

(*) 2.0 bar g for PSV discharging to atmosphere, 3.5 bar g for PSV discharging to flare networks.

(**) 2.0 or 3.5 bar g as minimum design pressure.

Particular attention shall be paid to the transient conditions such as equilibrium pressure plus hydrostatic pressure, water hammer, etc. which may induce a highest MOP.

4.3 Particular cases

4.3.1 The design pressure of a piece of equipment (excluding storage tanks, atmospheric tanks and pipelines) subject to operate at pressure below atmospheric pressure shall also be taken as the following:

NSOP (bar a.)	Design Pressure (bar a.)
≤ 0.35	Full vacuum
0.35 – 0.6	NSOP – 0.1 bar
> 0.6	0.5

4.3.2 Equipment that could face vacuum under abnormal conditions such as:

- Vacuum conditions during start-up, shutdown and/or regeneration purges
- Normally operated full of liquid but that could be blocked in and cooled down
- Containing condensable vapour but that could be blocked in and cooled down
- Could undergo a vacuum condition through the loss of heat input or external cooling (e.g. deluge)

will be treated case by case. They will not be designed for full vacuum if protective devices are provided (vacuum breaker, pressurisation gas, low pressure switch, etc.).

4.3.3 Vapour pressure at design temperature should be considered as design pressure except when safety relief valves are provided.

4.3.4 For equipment in equilibrium with flare and without isolation on the flare side, the design pressure of the said equipment is at least the flare system design pressure.

4.3.5 At the discharge of reciprocating compressor (piping and equipment):

- Design Pressure = NSOP + 2 bar for NSOP ≤ 20 bar g
- Design Pressure = NSOP + 10% for NSOP > 20 bar g.

PSV's are required.

4.3.6 At the discharge of centrifugal compressor:

- Design Pressure = NSOP + 1 bar for NSOP ≤ 10 bar g
- Design Pressure = NSOP + 10% for NSOP > 10 bar g.

For centrifugal compressor driven by a variable speed driver, the design pressure may be increased when the compressor operating curves are known in order that the compressor can operate in its full operating range.

PSV's are required.

4.3.7 Consideration to be given to compressor arrangement to determine the **"settling pressure"** of the isolated system.

The settling pressure is the equilibrium pressure reached between the suction and discharge isolating valves of the compressor system when the compressor is stopped or shutdown. The design pressure of the equipment and piping at compressor suction shall be above this settling pressure as a general rule in order to prevent unnecessary PSV's lifting.

4.3.8 Generally no PSV's are provided at the discharge of centrifugal pumps and the design pressure shall be the discharge pressure of the pumps at no flow with the maximum suction pressure and the maximum specific gravity.

When the discharge pressure of the pumps at no flow is not available, this pressure can be estimated by the following formula:

$$P_d = Ps\max. + \frac{1.2 * head * d_{max}}{10.2}$$

P_d =	design pressure at pump discharge (bar g).
Psmax. =	design pressure of suction drum + static head at d $_{max}$ and at LAH
Head =	head of the pump at design point (m).
d_{max} =	maximum specific gravity of liquid pumped under normal operating conditions.

Note: when pump curves are known, this design pressure has to be checked with actual head of the pump at no flow condition with the maximum suction pressure and the maximum specific gravity.

4.3.9 At discharge of volumetric pumps:

- Design Pressure = NSOP + 2 bar for NSOP $\leq$ 10 bar g
- Design Pressure = NSOP + 20% for NSOP > 10 bar g.

PSV's are required.

4.3.10 For columns, the same design pressure will be selected for the top of a fractionation tower and its associated condenser, reflux drum and inter connecting piping.

Liquid density and maximum liquid height in the bottom will be specified on the process data sheet to allow the vessel designer to calculate the bottom thickness.

4.3.11 Atmospheric tanks shall be designed full of water (or full of product if the product specific gravity > 1) as a minimum. Depending on the type of tank, higher design pressures could be specified. To be treated case by case depending on tank type.

4.3.12 Hydraulic pressure due to the relative elevation between equipment and also the PSV's location shall be considered.

4.3.13 Downstream of wellheads, the design pressure should be at least the wellhead shut in pressure.

5. Design temperatures

5.1 General rules

The design temperature is the temperature used for the mechanical design of equipment.

The design temperatures shall be specified as follows, with due consideration for the special cases discussed at section 5.2:

- Max. design temperature = NSOT + 15°C (consideration to be given to ambient conditions such as solar radiation temperature)
- Min. Design temperature = mSOT - 5°C, or minimum ambient temperature.

5.2 Particular cases

5.2.1 Consideration for the minimum design temperature must take into account any de-pressurisation and re-pressurisation (depending of material selection) of the equipment/piping that may occur during emergency or shutdown situation or gas blow-by from one equipment to another equipment and to the possible consequence of change of material.

5.2.2 Consideration for design temperature definition should be given to cooling medium failure when coolers are used.

Downstream of an air cooler, the design temperature is determined considering that 20% of the duty is provided by natural draft.

Downstream of other coolers, the design temperature will be the upstream NSOT.

These considerations generally apply up to the next equipment only.

5.2.3 Consideration to be given to conditions and fluids used for cleaning (e.g. steam). In this case, both pressure and temperature conditions for this operation have to be specified.

5.2.4 Consideration to be given for upset and transient conditions such as start-up, shut-down, etc. In this case, both pressure and temperature conditions have to be specified.

5.2.5 The exceptional temperature generated by fire will not be considered to specify the design temperature selection.

5.2.6 The “cold” design temperature case for cryogenic equipment can be supplemented with an exceptional “hot” temperature case (the related design pressure shall also be indicated).

5.2.7 A specific design temperature will be given together with the specified vacuum design pressure.

5.2.8 For discontinuous/cycled processes, various pressure and temperature conditions sets will be specified for each phase of equipment operation.

One typical example is molecular sieve vessel design conditions specification. The set of design conditions of each phase of the operating cycle shall be specified, i.e. design conditions during the adsorption phase, during the regeneration phase, etc.

6. Design factors

6.1 Pumps

The minimum margin between the normal and rated flow for a pump will be as below:

• Reflux pump and boiler feed water pump	20%
• Other process pump	10%
• Utility pump	10%
• Loading pump (to a tanker)	0%
• Export pump from storage to pipeline (continuous operation)	15%

Notes:

- When a non automatically minimum flow protection has been installed, the permanent re-circulation flow if required must be added to the net process flow rate.
- Normal and rated flows will be identical in such instance as:
  - Intermittent service pumps
  - When the pump has been overrated to allow for a centrifugal type and if overrating is ≥ 10%
  - Re-circulation flow such as for product loading lines or through amine filtration system.
- For fire water pumps see specification GS SAF 321: "Fire Pump Stations and Fire Water Mains".

6.2 Compressors

If the flow is constant, no flow margin is required, but if the flow is coming from a production separator a flow margin of 10% is recommended in order to take into account the possible slugging regime at the inlet of these production separators.

The variations of gas compositions, molecular weight, Cp/Cv, etc., and the operating conditions (mainly suction pressure and temperature) shall be written on Process data sheets and shall be taken into account for the sizing of the said compressors.

6.3 Glycol package unit

See general specification GS ECP 101: "Glycol Package Unit".

6.4 Other equipment

• Shell and tube heat exchangers and air coolers:	10% on heat exchange surface
• Fired heaters and furnaces:	10% on duty
• Boilers:	10% on steam flow rate
• 1st stage production separator:	10% on inlet gas flow rate and 0% on liquid inlet flow rate (attention shall be paid to the liquid slug, see section 7.2)
• Other drums:	10%

No over design is to be applied on other equipment unless specified on case by case.

For sizing factor to be applied for selection of equipment such as electrical motors, diesel engines, gas turbines, etc., refer to the relevant specifications issued by other disciplines.

7. Vessels sizing criteria

7.1 Vapour area

The following excludes flare/vent KO drums, desalters and electrostatic dehydrators.

When internals are installed, the common vapour internal shall be a wire mesh, but for some services vane pack can be used after approval from COMPANY.

The use of others vapour internals such as cyclones, etc., requires also COMPANY approval.

The basis of sizing of the vapour area is the critical velocity Vc, with:

$$Vc = 0.048 * \sqrt{\frac{\rho l - \rho g}{\rho g}}$$

with:

ρl = liquid density in kg/m^3

ρg = vapour density in kg/m^3

Vc = critical velocity in m/s

The maximum allowable gas velocity across the vapour area or across the vapour internals where installed is K*Vc.

K is a coefficient depending of the service, and the use or the absence of wire mesh.

The recommended K values are given hereafter for different services.

Service	Without wire mesh	With wire mesh
Production separator	1.7	2.2
Fuel gas drum	0.8	1.7
Compressor suction drum	0.8	1.7
Glycol or amine contactor inlet drum	0.8	1.7
Reflux drum	1.7	2.2
Steam drum	////////	1.3

If a vane pack internal is used, the recommended K value is 3.3. This has to be confirmed with the vane pack SUPPLIER.

For horizontal vessels without vapour internal (wire mesh, vane pack, etc.), the minimum distance between the top of the vessel and the LSHH is the largest of 300 mm or 0.2* internal diameter.

Vessels handling paraffinic oil shall not be equipped with vapour internals.

7.2 Surge volume for liquids

If the vessel is sized to receive a liquid slug, that liquid slug volume shall be taken between NLL and LAH.

7.2.1 Two-phase vessels

The minimum liquid surge times between LAL and LAH shall be specified as follows:

Services	Time (Minutes)
Unit Feed Surge Drum A. to heater B. to others	 5 3 without pump 5 with pump
Reflux Drum	5
Fractionation tower bottom	3 without pump 5 with pump
Steam flash drum (process units) Steam drum (utility generation) Steam deaerator	5 10 15
Atmospheric degassing drum “gas boot”	15 for storage tank with floating roof 10 for storage tank with fixed roof
Others Drums	2 without pump 3 with pump

The minimum distance between LAL and LAH shall be 300 mm.

The lower level instrument connections on liquid outlet piping are forbidden except for vessels dry in normal operation.

The LALL should be generally be located at least 300 mm above the tangent line for vertical vessel and at least 150 mm above the bottom for horizontal vessel, but alternative arrangements are acceptable for clean services (no risk of plugging of instrument nozzle) where the lower instrument nozzle may be located on the bottom.

When applicable, the surge time below the LALL has to be compatible with the time required to close a SDV or to stop a pump.

7.2.2 Three-phase separators

For three-phase separators, the retention time for the two liquid phases shall be considered.

The effective retention volume to be considered for the design of a vessel is the portion of the vessel in which the two liquid phases remain in contact with one another. As far as the two liquid phases separation is concerned, once either substance leaves the primary liquid section, although it may remain in the vessel in a separate compartment, it cannot be considered as a part of the retention volume.

The retention volume of the highest density liquid is taken between the bottom and INLL.

The retention volume of the lightest density liquid is taken between the INLL and the NLL.

The retention time for each phases depends on the effluents characteristics, the operating temperature, etc., and is fixed case by case.

The minimum distance between ILAL and ILAH and between LAL and LAH shall be 200 mm.

If the liquid is pumped, the surge time (volume between ILAL and ILAH for the highest density liquid and volume between LAL and LAH for the lightest density liquid) shall be at least the time required for two phase vessels (section 7.2.1) which could increase the retention time of the considered liquid.

The ILALL and the LALL should be generally be located at least 300 mm above the tangent line for vertical vessel and at least 150 mm above the bottom for horizontal vessel, but alternative arrangements are acceptable for clean services (no risk of plugging of instrument nozzle) where the lower instrument nozzle may be located on the bottom.

When applicable, the surge time below the ILALL and the surge time below the LALL have to be compatible with the time required to close the corresponding SDV or to stop a pump.

7.3 Levels positions

The upper and lower LG connections shall be located so that it is possible to read all alarms and shutdown levels.

For vertical vessels, if the operating levels are located in the bottom head, the non linear evolution of level with draw-off shall be considered.

- NLL — at 50% of the controllable range, except in separators having an over capacity assuming a slug catcher function
- LAL — minimum at 10% of the controllable range
- LAH — maximum at 90% of the controllable range
- LALL/LSLL — at typically 1 to 2 minutes with 200 mm minimum between LAL and LSLL for vertical drum and 100 mm for horizontal drum
- LAHH/LSHH — at typically 1 to 2 minutes with 150 mm minimum between LAH and LSHH for vertical drum and 100 mm for horizontal drum

Connections for level instruments for LSLL/LSHH will be independent from the other instrument connections and not connected on stand pipe (that is also valid for other functions such as pressure, temperature, flow rate, etc.).

The above levels positions are also applied for interfaces levels.

Stand Pipes:

- Level controllers and level gauges can be normally grouped on standpipes if 3 or more instruments are required.
- Minimum size of standpipe connections is 3".
- Standpipes are installed without isolating valves.

8. Diameter of vessels

As a general rule, inside diameter will be specified on process data sheets (in mm)

If the required internal diameter for a vessel is lower than 800 mm, a note shall be added specifying that a piping element is acceptable.

For vessels with internal diameter less than 1000 mm I.D., flanged heads may be accepted.

9. Heat exchangers

9.1 Fouling factors for shell and tube heat exchangers and air coolers

The following gives some typical fouling factors for process and utility fluids.

9.1.1 Process fluids

	m^2 °C/W
Heavy oil	0.00050
Oil	0.00040
Heavy gas oil	0.00035
Light gas oil	0.00030
Gasoline	0.00020
LPG (liquid)	0.00020
Natural gas	0.00015
Regeneration gas (dryers)	0.00017
Amine solution	0.00040
Glycol	0.00040
Refrigerant (propane or mixed refrigerant)	0.00010
Oily water	0.00030

9.1.2 Utility fluids

	m^2 °C/W
Sea cooling water	0.00030
River cooling water	0.00040
Fresh (desalinated) cooling water in closed loop	0.00020
Well water	0.00040
Atmospheric air for air coolers	0.00035
Fuel gas	0.00017
Hot oil	0.00020
Super heated steam	0.00010
Saturated steam / steam condensate	0.00017
Boiler feed water	0.00017
Instrument air, nitrogen	0.00017

9.2 Fouling factors for plate type heat exchangers

For Plate Frame Heat Exchangers, an overall fouling factor of 0.00005 m².°C/W shall be taken for all fluids.

For Plate Fin heat Exchangers, no fouling factor shall be applied but an extra surface of 15% shall be added on calculated area.

9.3 Temperature approach for heat exchangers

The temperature approach shall be optimised for heat exchangers but it shall not be smaller than:

- 3°C for kettles
- 5°C for others shell and tubes exchangers
- 10°C for air coolers
- Case by case for plate type heat exchangers.

9.4 Specific requirement for heat exchangers

The Code "TEMA R" will be generally used for all shell and tubes and hair pin type heat exchangers.

Fixed tube sheet exchangers are acceptable for non fouling service on the shell side. In this case, the Designer shall define all exceptional operating conditions (start-up, shutdown, etc.) to assess the necessity to provide an expansion bellow on the shell.

For air coolers, 50% of the fans can be equipped with variable/dual speed motors (instead of auto-variable blade pitch control) when process control is required.

10. Equipment nozzles

10.1 Nozzles and manholes

See general specification GS PW 211 "Design And Fabrication Of Pressure Vessels".

The number of manholes depends on internals. Internals shall be removable through the manhole. In any case for vessel having a length higher than 9 m, two manholes are required.

10.2 Vent, drain and overflow connections

The drain nozzles of the vessel should be connected:

- To the outlet line at low point for vertical vessel
- And directly on the capacity for horizontal vessel and for vertical vessel with outlet line entering inside vessel.

For horizontal vessels having a length greater than 6 m, additional drain connections are required with a maximum distance of 3 m between each drain connection.

For vessels equipped with internals (baffle), a drain connection is required on each compartment.

Overflow connections: For vessels equipped with overflow connections, the overflow nozzle and line size will be one size greater than the inlet/outlet nozzle (whichever is greater).

Vent and drain connections for vessels shall be sized as follows:

Volume or diameter of vessel	**Minimum**	**Minimum**
	Vent diameter	**Drain diameter**
(m3 or mm)		
V ≤ 15 or D ≤ 2500	2"	2"
15 < V ≤ 75 or 2500 < D ≤ 4500	2"	3"
75 < V ≤ 220 or 4500 < D ≤ 6000	3"	4"
220 < V ≤ 420 or D > 6000	4"	4"
V > 420	6"	4"

Vents and drains for packages and rotating equipment shall be discussed with Vendors for size, back pressures, temperatures, products, etc.

10.3 Utility connections

Utility connections (2” minimum) shall be sized as follows:

- Drums and heat exchangers (when applicable): 2"
- For large vertical drums, two 2” connections will be provided for diameter ≥ 4.5 m
- For horizontal vessel with a length > 6 m and operating in toxic service, two 2” connections will be provided

- If vessel is equipped with internals (baffle), one 2” connection will be provided on each compartment
- Columns: as follows with regard to the column diameter, D (m)
    - $D \leq 4$:   2"
    - $4 < D \leq 5.5$: 3"
    - $D > 5.5$: 4"

Note: Utility connections, when specifically required, are not necessarily located on vessels (advantage will be taken of available connections such as drains for steamout service) but should remain operational when the vessel is isolated.

11. Elevation of equipment

As a conventional rule for a vessel containing a product at its boiling temperature, a minimum elevation of 3000 mm will be specified between the bottom tangent line for vertical vessel or the bottom line for the horizontal vessel and the grade when a recovery bottom centrifugal pump is provided.

That elevation will be confirmed when NPSH requirements are finalised with piping/layout and rotating equipment disciplines.

If there is no other process requirement regarding the elevation, a note on P&ID will be indicated “minimum for piping”.

12. Piping

12.1 General design and hydraulics

12.1.1 Line velocity and friction loss for liquid line and gas line

The size of each line shall be estimated based on the mass flow rate and in accordance with the velocity range criteria. It shall be verified then in accordance with the friction loss range criteria given in section 12.3 “Line sizing criteria”.

12.1.2 Minimum piping sizes

Except for instrument piping, connections to equipment or piping in which minimum flow velocity requirements govern, the minimum size shall be:

- ¾” for pipe when located above ground
- 2” for process line on pipe rack
- 2” for utility line on main pipe rack
- 2” for pipe on pipe sleeper
- 2” for underground steel pipe
- 2” for underground non metallic piping.

12.1.3 Pump suction line

Pump suction lines shall be sized to provide a net positive suction head available ($NPSH_A$) 1 meter above than the net positive suction head required ($NPSH_R$) by the pump selected.

Pump suction lines shall not be smaller than pump suction nozzle. Therefore, eccentric reducer at pump suction is acceptable provided that the available calculated $NPSH_A$ (with pressure loss in the reducer) remains acceptable regarding the required $NPSH_R$.

Pump suction valves shall be in the same diameter as the line.

The static head used in calculating the available $NPSH_A$ shall be taken from the LSLL or the tangent line in the suction vessel to one of the following:

- The centreline of a horizontal pump
- The suction impeller on a vertical centrifugal pump.

For the value of NPSH specified on process data sheet, the referenced elevation shall be indicated (e.g. grade, pump centreline, etc.).

The design of suction lines from storage tanks shall be based on a NPSH taken from the lowest specified liquid level in the tank at which rated pump capacity is required.

When sizing suction lines for reciprocating pumps, acceleration head shall be considered.

For fire water pump see specification GS SAF 321: "Fire Pump Stations and Fire Water Mains".

12.1.4 Isolation

For isolation of equipment, control valves, PSV's, etc., or section of piping removable or open in operation, double block and bleed isolation is required on all fluids having one of the following conditions:

- Operating pressure above 70 bar g. (refer to specification GS SAF 261)
- Operating pressure above 35 bar g. only for LNG
- H_2S partial pressure > 1 bar (refer to specification GS SAF 261)
- Fluid very corrosive and abrasive.

De-pressurising, drainage or sample manual globe valves which may induced freezing during operation shall be located at least 600 mm downstream of a block valve.

All equipment will be generally isolated by spectacle blinds, or spacer and ring blinds (when isolation for maintenance) installed at minimum distance of equipment and easily accessible except for PSV's for which a spool piece is required.

Permanent spectacle blind and spacer and rings will be installed depending on diameter and rating as follows:

Rating ANSI	No positive isolation device	Permanent spectacle blinds	Spacers or ring blinds
150 RF	$\varnothing$ < 8"	8" ≤ $\varnothing$ < 14"	14" ≤ $\varnothing$
300 RF	$\varnothing$ < 6"	6" ≤ $\varnothing$ < 10"	10" ≤ $\varnothing$
600 RF and RJ	$\varnothing$ < 4"	4" ≤ $\varnothing$ < 8"	8" ≤ $\varnothing$
900 + RJ	$\varnothing$ < 2"	2" ≤ $\varnothing$ < 6"	6" ≤ $\varnothing$

12.1.5 Double block and bleed

Double block and bleed will be used for condition stated on section 12.1.4. This applies to equipment and on-line instrument (PSV's, control valves, etc.).

The valves which are not used when the line or equipment is under pressure (vessel vents for example) are not doubled, but will be provided with blind flange.

The bleed valve is a single ¾” or 1” with a blind flange.

12.1.6 Control valve

The pressure drop across a control valve should be a minimum of 20% of the system dynamic pressure loss at normal flow rate or 0.7 bar whichever is greater. If a lowest minimum pressure drop is required, a derogation is possible for specific types of control valve; COMPANY instrument specialists shall be consulted.

Non continuous control valves do not require by-passes. Control valves in continuous service and less than or equal to 6” should have block and by-pass valves manifold if operation can be achieved in manual operation.

Above 6", the control valve can be provided with a hand wheel and no by-pass, subject to COMPANY approval.

For main utility control valves (fuel gas and air), the by-pass should be replaced by a spare control valve.

In case of misoperation, the gas blow-by situation shall consider the flow rate through the control valve full open and its by-pass also fully open when it is installed. If that flow rate becomes the sizing case for the flare, the manual by-pass could be deleted to avoid an increase of the flare size or alternatively a mechanical interlock between the upstream control valve manual block valve and upstream by-pass valve manual block valve should be installed.

Sizing of bleed devices of control valve manifolds:

- Upstream control valve line size ≥ 6":
  - For control valve open by fluid failure (FO), one maintenance bleed valve (¾” block valve with blind flange) will be installed downstream the control valve.
  - For control valve closed by fluid failure (FC), two maintenance bleed valves will be installed, one upstream and one downstream the control valve.
- Upstream control valve line size < 6":
  - One maintenance bleed valve will be installed downstream the control valve whatever the control valve position (FO or FC) by fluid failure.

12.2 Insulation and heat tracing

Thermal insulation for hot or cold services is required when applicable for:

- Heat or cold conservation of equipment and piping
- Personnel protection of equipment for operating temperatures above 70°C. A physical barrier with warning signs attached to hot surface is preferred to insulation

- To avoid external water condensation or ice
- Steam, hot water or electrical tracing.

In all cases, insulation shall be designed in order to limit CUI (Corrosion Under Insulation) and to allow access to instruments and pipe fittings without destroying the shield.

Fire resistant insulation may be specified in order to reduce the relief load to the flare system in case of fire. However the normal engineering practice is to take no credit for heat insulation.

12.3 Line sizing criteria

This section shall not be applied to the flare lines (see § 13.5 hereafter) and fire water lines (see GS SAF 321: "Fire Pump Stations and Fire Water Mains").

12.3.1.1 Line sizing criteria for gases and steam

Vapor and steam lines (Flare Lines Excluded)	Maximum ρV^2 (1) $(kg \cdot m \cdot s^{-2})$	Maximum Velocity (m/s)	ΔP bar/km	
			Normal	Maxi.
Vapour lines				
Continuous operation:				
• $P \leq 20$ bar g.	6,000			
• $20 < P \leq 50$ bar g.	7,500			
• $50 < P \leq 80$ bar g.	10,000 (2)			
• $80 < P \leq 120$ bar g.	15,000		ΔP must be Considered compatible with the corresponding service	
• $P > 120$ bar g.	20,000			
Discontinuous operations e.g.:				
• Compressor anti-surge:				
- $P \leq 50$ bar g.	10,000 (2)			
- $50 < P \leq 80$ bar g.	15,000			
- $P > 80$ bar g.	25,000			
• Compressor suction	To be compatible with above		0.2	0.7
• Compressor discharge			0.45	1.15
Steam lines				
• - $P \leq 10$ bar g.				
- Short line $L \leq 200$ m			0.5	1.0
- Long line $L > 200$ m			0.15	0.25
• $10 < P \leq 30$ bar g.				
- Short line $L \leq 200$ m		42	1.2	2.3
- Long line $L > 200$ m		42	0.25	1.0
• $P > 30$ bar g.				
- Short line $L \leq 200$ m		30	1.2	2.3
- Long line $L > 200$ m		30	0.35	1.0

For gas and two phases flow lines, the ρV^2 is calculated for all operating conditions but the criteria selection is based on the design pressure. However, the other criteria (pressure drop, velocity for the film inhibitor integrity, erosion velocity, etc…) shall be based on all operating conditions;

(1) ρ = gas density in kg/m^3, V = gas velocity in m/s

(2) Value could be increased but not exceeding 15,000 in case of debottlenecking

12.3.2 Line sizing criteria for liquids

Liquid line type	ΔP bar/km		Maximum velocity m/s			
	Normal	Maxi	∅ ≤ 2"	3" ≤ ∅ ≤ 6"	8" ≤ ∅ ≤ 18"	∅ ≥ 20"
Pumps suction						
• Liquid at bubble point or with dissolved gas	0.6	0.9	0.6	0.9	1.2	1.5
• Non boiling liquid	2.3	3.5	0.9	1.2	1.5	1.8
Unit lines						
• Liquid at bubble point or with dissolved gas	0.6	1.0	0.6	1.0	1.4	1.8
• Non boiling liquid	2.3	3.5	0.9	1.2	1.8	2.4
Pump discharge						
• Discharge pressure ≤ 50 barg	3.5	4.5	Velocity 1.5 to 4.5 m/s [1]			6.0 [1]
• Discharge pressure > 50 barg	7.0	9.0	Velocity 1.5 to 4.5 m/s [1]			6.0 [1]
Column outlet	0.6	0.9	0.6	0.9	0.9	0.9
Gravity flow	0.25	0.45	0.6 (normal velocity)			
Water (carbon steel) lines [2]						
• Cooling water and service Water						
- Large feeders between	1.5		Velocity 1.5 to 3.0 m/s			
- Pumps and units						
- Unit lines: long lines		1.5	1.5	2.5	3.0	3.0
Short lines		3.5	1.5	2.5	3.0	3.0
- Boiler feed:						
- Pressure ≤ 50 barg	3.5	4.5	Velocity 1.5 to 4.5 m/s			6.0
- Pressure > 50 barg	7.0	9.0	Velocity 1.5 to 4.5 m/s			6.0
• Seawater lines			Velocity 2.5 to 3.5 m/s (2 m/smin.)			
Steam condensate return			Velocity 1 to 1.5 m/s			

[1] 3.0 m/s maxi (2 m/s average) at storage tank inlet or in loading.

[2] Special considerations can be applied for copper-nickel or glass reinforced plastic piping upon COMPANY approval.

12.3.3 Line sizing criteria for two phase flow

The criteria stated for gas lines can be used in two phase line, using the fluid apparent density ρ_m expressed as:

$$\rho = \rho_m = \frac{W}{\frac{Wl}{\rho l} + \frac{Wv}{\rho v}} \text{ in kg/m}^3$$

W = Wl + Wv = total flow rate in kg/h

ρl = liquid density in kg/m^3

Pv = vapour density in kg/m^3

Wl = liquid flow rate in kg/hρv = vapour density in kg/m^3

Wv = vapour flow rate in kg/h

and the apparent fluid velocity Vm expressed as:

$$V = Vm = \frac{4 * W}{3600 * \rho_m * \pi * D^2} \text{ in m/s}$$

D = internal diameter of the line in meter.

In addition, the flow regime may have to be checked using:

- Flow pattern for small diameter lines and low operating pressures (ex: reboiler lines)
- Computer programs for high pressure large diameter lines (ex: production headers).

12.3.4 Corrosion/erosion criteria

12.3.4.1 Corrosion

For corrosion resistant material (Stainless Steel (SS), special alloys, etc.), there is no limitation of flowing velocity up to 100 m/s.

For non corrosion resistant material such as carbon steel (CS), low temperature carbon steel (LTCS), etc., in corrosive fluid service, a corrosion allowance for the design service life and corrosion inhibitor injection is required. The flowing velocity is limited by the inhibitor film integrity.

12.3.4.2 Erosion

For Duplex, SS or alloy material, the flowing velocity shall be limited to:

- 100 m/s for single phase vapour lines and multiphase lines in stratified flow regimes (65 m/s for 13% Cr material)
- 20 m/s in single phase liquid lines and multiphase lines in annular, bubble or hydrodynamic slug flow regime
- 70 m/s in multiphase lines in mist flow regimes.

For Carbon Steel material:

- In case of continuous injection of corrosion inhibitor, the inhibitor film ensure a lubricating effect which allows for higher fluid velocity for a given corrosion/erosion rate. The corrosion inhibitor erosion velocity limit will be calculated taking into account the inhibitor film wall shear stress.
- In case of uninhibited fluid, the API RP 14 E recommendation should apply: the flowing velocity must be maintained below the erosional limit:

$$Ve = \frac{C}{\sqrt{\rho_m}}$$

With: Ve = erosional velocity in ft/s (m/s in SI units)

ρ_m = gas / liquid mixture density at flowing conditions in lb/ft^3 (kg/m^3 in SI units)

C = empirical constant equal to 150 to 170 (183 to 207.4 in SI units). C value up to 200 (244 in SI units) can be considered on peak flow rate only in case of absence of abrasive (solid) particles such as sand. When solid and/or corrosive contaminants are present C value shall not be higher than 100 (122 in SI units).

13. Flare and cold vent systems

13.1 Type of flare tip

For flares and cold vents, the tip can be normally conventional or sonic depending on the required back pressure and noise limitation.

When possible a sonic tip shall be preferred.

Flare tip design based on the use of Coauda effect or involving moving parts is prohibited, unless sufficient technology improvement from suppliers is found acceptable by COMPANY. In general tip with variables slots are forbidden.

The installation of flares to run in parallel is forbidden.

The flares shall be smokeless.

13.2 Flaring flow rates

The analysis of the causes of relief is required and an occurrence flaring loads balance including each individual relieving rate for each possible cause shall be performed.

The de-pressurisation rate shall follow the criteria of specification GS SAF 261: “Pressure Protection & Relief, Emergency Shutdown & Depressurisation”.

For de-pressurisation calculations at least, two cases shall be considered:

- Without fire:
  - For calculation, the following conditions shall be applied, except when COMPANY has agreed different criteria:
  - The initial pressure will be the system design pressure /safety relief valve set pressure, except for the compressor systems for which the settle out pressure will be considered.

- The initial temperature: mSOT. For equipment which could be isolated under pressure, the initial temperature will be the minimum ambient temperature when the mSOT is higher than the minimum ambient temperature, the initial pressure will be calculated due to the temperature change considering that at the isolation time the pressure and the temperature were respectively the design pressure and the mSOT.
- The initial level of liquid in drum: high or low operating level = LAH or LAL (both calculations shall be performed and the worst case considered).

• With fire: For calculation, the following conditions shall be applied, except when COMPANY has agreed different criteria :
  - Heat input, see specification GS SAF 261
  - The initial pressure will be the system design pressure/safety relief valve set pressure, except for the compressor systems for which the settle out pressure will be considered.
  - The initial temperature: mSOT. For equipment which could be isolated under pressure, the initial temperature will be the minimum ambient temperature when the mSOT is higher than the minimum ambient temperature, the initial pressure will be calculated due to the temperature change considering that at the isolation time the pressure and the temperature were respectively the design pressure and the mSOT.
  - The initial level of liquid in drum: high or low operating level = LAH or LAL (both calculations shall be performed and the worst case considered).

After selection of the blowdown restriction orifice size, the calculations shall be reworked for the non sizing case.

The de-pressurisation calculations shall not be stopped at the final required pressure but shall be performed to down the atmospheric pressure to be sure that the lowest temperature is obtained.

13.3 Radiation levels criteria

The radiation levels criteria shall follow specification GS SAF 262: “Safety Rules For Hydrocarbon Disposal Systems”. The minimum relative humidity stated on the basis of design shall be applied.

13.4 Emissivity coefficient

When the radiation calculations are performed by a flare Vendor it is mandatory that the Vendor provide the emissivity coefficient used for radiation calculations. It is necessary to check carefully the emissivity coefficient used because actual experience has shown that some Vendors take a too low value for this emissivity coefficient. Many times, the emissivity coefficient used by Vendors does not take into account the liquid carry over, they consider an ideal gas/liquid separation. The droplets size for the flare drum sizing and the expected liquid carry over shall be clearly indicated in the flare tip process data sheet.

Recommended emissivity coefficient

- For pipe flare:
  - Natural gas molecular weight of 18: 0.21
  - Natural gas molecular weight of 21: 0.23
  - Ethane: 0.25
  - Propane: 0.30
  - See also API RP 521.
- For sonic flare:

  The emissivity coefficient = 0.15 with liquid carry over not exceeding 5% weight (that is the general case). For a higher liquid carry over, the emissivity coefficient will be increased. For a gas without carry over, the minimum emissivity coefficient will be 0.13.

13.5 Flare and cold vent lines sizing criteria

13.5.1 Lines upstream relieving devices

- PSV's:

  For the line sizing, the maximum capacity of the PSV (recalculated with the selected orifice) shall be considered even if this figure exceeds the actual maximum flow rate due to process limitations.

  - ΔP between the protected equipment and the PSV < 3% of PSV set pressure (API RP 520 Part II)
  - Of line $\geq \varnothing$ PSV inlet
  - $\rho V^2 \leq 25{,}000$ kg/m/s^2 for $\varnothing$ of line $\leq$ 2"
  - $\rho V^2 \leq 30{,}000$ kg/m/s^2 for $P \leq 50$ bar g
  - $\rho V^2 \leq 50{,}000$ kg/m/s^2 for $P > 50$ bar g.
- Depressurisation device (BDV's + FO)
  - Minimum line size 2"
  - ρV^2 criteria are the same as for PSV's.

13.5.2 Lines downstream relieving devices, flare and cold vent headers and sub-headers

- Minimum line size 2".
- Back pressure to be compatible with the installed relieving device
- Velocity and ρV^2:

  Monophasic (GAS)

  - Intermittent flow:
    - . Lines downstream relieving devices and sub-headers: 0.7 Mach maximum and $\rho V^2 < 150{,}000$ kg/m/s^2 considering the maximum capacity of the relieving devices even if this figure exceeds the actual maximum flow rate due to process limitation and the relevant occurrence.

. Headers: 0.7 Mach maximum and ρV^2 < 150,000 kg/m/s^2 considering the maximum flow rate due to process limitations and for the relevant occurrence, however a velocity of 0.8 Mach could be accepted for a long straight line without elbows and connections (e.g. stack, line on bridge).

. For a ρV^2 > 100,000 kg/m/s^2 vibration and line support studies are required.

- Continuous flow:

Velocity < 0.35 Mach and $\rho V^2 \leq$ 50,000 kg/m/s^2

. Multiphase (2 phase flow at the inlet of relieving device):

- Velocity $\leq$ 0.25 Mach and $\rho_m V_m^2 \leq$ 50,000 kg/m/s^2.
- For ρ_m and V_m definitions see section 12.3.3.

The sizing shall be done for the line downstream each device with the built-up back pressure for the corresponding occurrence and not with the maximum built-up pressure for the maximum flow rate to the flare or cold vent. The same shall be applied for the headers and sub-headers.

13.6 Flare drum sizing

For flare drum and cold vent drum, the sizing criteria shall follow API RP 521 method with the following droplets size in microns:

- Vertical flare or cold vent on platform: 150
- Inclined boom offshore> 45° from horizontal: 150
- $\leq$ 45° from horizontal: 400
- Remote flare or cold vent offshore: 600
- Vertical flare or cold vent onshore: 600

13.7 Purge gas

The purge gas is provided to avoid:

- Air ingress into the flare or cold vent stack in order to avoid to form an explosive mixture in the stack or header
- For a flare, the risk of burn back which induces a quickest deterioration of the tip.

Since it is recognised that flare SUPPLIERS tend to underestimate the amount of purge gas, flare SUPPLIERS shall guarantee the purge gas flow rate to avoid air ingress and the burn back. In any case, the purge gas flow rate used for the design of the purge gas system shall not be lower than the value given by the following equation:

- Without gas seal: Purge gas flow = $24{,}000 * D^3 * MW^{-0.565}$
- With gas seal: Purge gas flow = $12{,}000 * D^3 * MW^{-0.565}$

Where:

- Purge gas flow in Sm^3/h
- D is the tip internal diameter (note 1) in m
- MW is the purge gas molecular weight in kg/kmole.

Note 1: For sonic flare, the tip internal diameter is taken as the equivalent diameter corresponding to the exit gas area.

When fuel gas is used for purge gas, the source of purge gas shall be common to the fuel source to the pilots in order to avoid a loss of purging while pilots remain in service.

The heaviest available gas should be preferably used as the normal source of purge gas in order to minimise the vacuum pressure in the flare header for an elevated flare or cold vent. However, the heavy gas tends to give some smoke.

Sometimes, nitrogen could be used as purge gas. In these situations, pilots able to run even in inert gas atmosphere shall be installed (to be studied case by case with the flare tip supplier).

13.8 Installation

13.8.1 Headers and sub-headers

Only one low point is permitted on a flare system, it is the flare drum. A continuous slope (at least 2 mm per m) from the relieving devices to the flare drum and from the flare tip to the flare drum is required.

The same principle is applied for the cold vent.

Check valves on flare headers/sub-headers are prohibited. Block valves shall be avoided. If block valves are required for maintenance purpose (e.g. relieving devices) or revamping, these block valves should be full bore ball valve and locked in open position. The installation of these valves shall be such that the line shall not be blocked in case of mechanical failure of these valves in case of ball valve is replaced by a gate valve for example.

Flow orifice or flowmeter, which can be blocked by foreign matters, shall not be installed. For flare flow metering, the types available are:

- Averaging pitot (annubar)
- Thermal
- Ultrasonic.

All connections shall be done with no low point and preferably on the top of the header and sub-header.

The connections shall be done at 45° angle except for small diameters up to and comprising 4" nominal diameter.

13.8.2 Relief valve arrangement

Refer to:

- Drawing 1: PSV's installation with one upstream isolating valve
- Drawing 2: PSV's installation with two upstream isolating valves.

Spare relief valve: A spare relief valve is always installed except:

- If the protected equipment can be isolated and de-pressurised/drained without production loss (e.g. test separator, pig trap, etc.).

- If the installed relief valve is only for fire case, in this case the upstream and downstream block valves should be installed to avoid the purging and inerting of the protected equipment and the flare lines for the relief valve maintenance and the spare relief valve is stored in the workshop (note on P&ID to be added).

13.8.3 De-pressurisation valve arrangement

Refer to drawing 3.

13.8.4 Drainage arrangement

See general specification GS SAF 228 “Liquid Drainage”.

14. Softwares

The parties in charge of the Process design shall include in their proposal a list of the Process softwares they intend to use, with their capability if they are not listed hereafter.

This list shall be approved by COMPANY.

The following softwares are accepted by COMPANY for process related calculations:

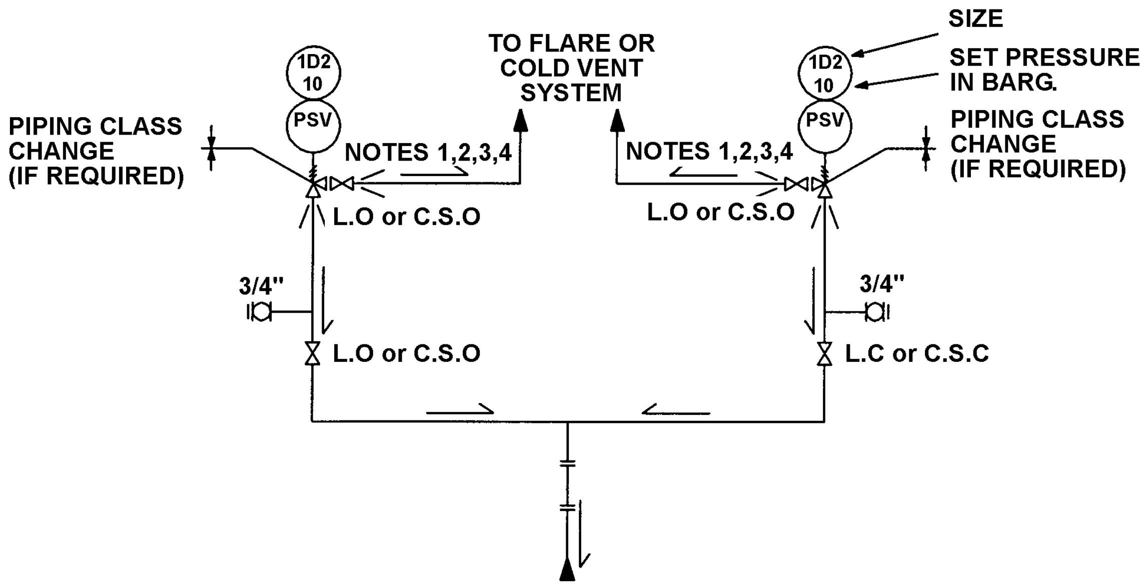
- Plant thermodynamic simulations:
  - PRO II, PROVISION
  - HYSIM, HYSYS [upon written COMPANY approval].
- Multiphase pipelines calculations:
  - Steady state:
    . PIPEPHASE (option TACITE or OLGA compulsory)
    . OLGA [(1)]
    . PIPESIM (option OLGA compulsory).
  - Transient conditions:
    . OLGA [(1)]
- Heat exchangers thermal calculations:
  - Shell and tube:
    . HTRI
  - Air coolers
    . HTRI
    . HTFS.

[(1)] Options "slug tracking" and "water module" to be used at COMPANY request.

- De-pressurisation calculations:
  - "BLOWDOWN" software from Imperial College (UK)
  - "LNGDYN" software from Technip (France).
- Flare radiation levels for preliminary calculations, final one being performed by SUPPLIERS:
  - FLARESIM (mixed method with 25 points minimum or API method with 25 points are recommended, wind chill effect to be ignored).
- Surge protection on liquid systems and water hammer calculations :
  - "PIPENET" software from Sunrise (UK)
  - "FLOWMASTER" software from Flowmaster (France)

Drawings

Appendix 1 PSV's installation with one upstream isolating valve



LEGEND

BLOCK VALVE, GATE VALVE OR FULL BORE BALL VALVE

BALL VALVE

REDUCER

SLOPE

SPOOL PIECE

L.C	LOCKED CLOSED
L.O	LOCKED OPEN
C.S.O	CAR SEAL OPEN
C.S.C	CAR SEAL CLOSED

Notes:

1- Distance between PSV and reducer is minimum.

2- Downstream block valve is installed flange to flange with PSV; if the built-up back pressure is too high, this block valve is installed downstream the reducer.

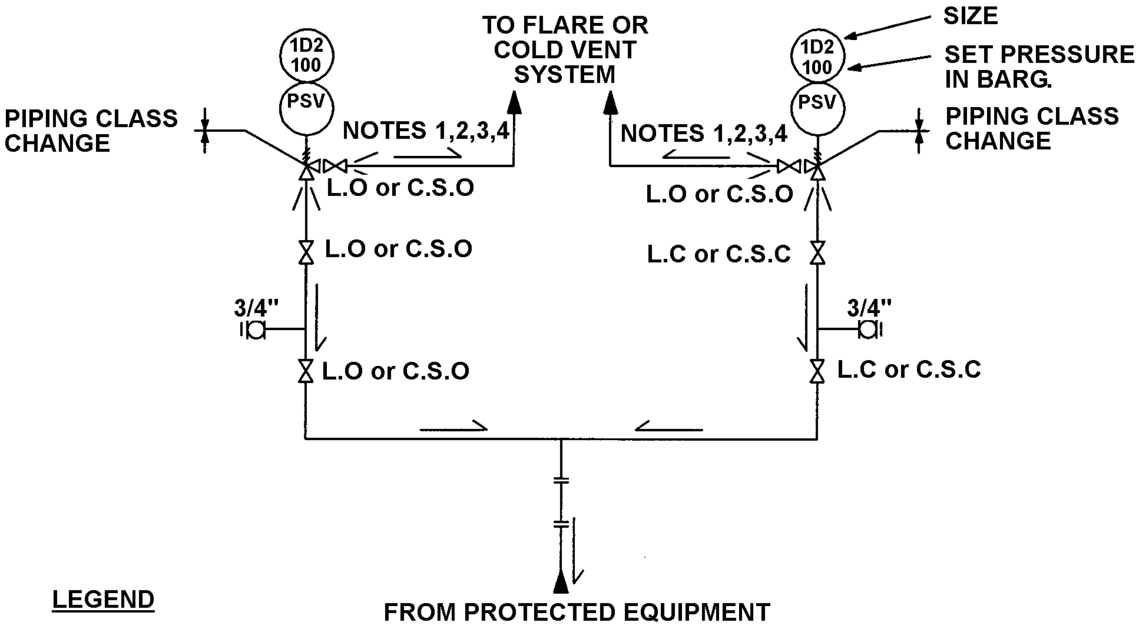
3- Installation of elbow between PSV and reducer is forbidden.

4- Piping class change could be required downstream PSV, depending of relief pressure and PSV size.

Appendix 2 PSV's installation with two upstream isolating valve

The PSV's installation with two upstream isolating valves is required on all fluids having one of the following conditions:

- Operating pressure above 70 barg
- Operating pressure above 35 barg, only for LNG
- H_2S partial pressure > 1 bar
- Fluid very corrosive and abrasive

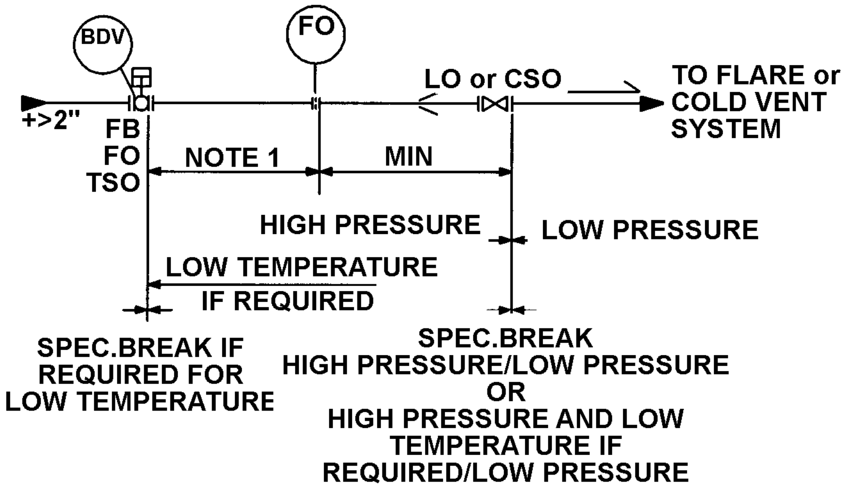


LEGEND

⋈	BLOCK VALVE, GATE VALVE OR FULL BORE BALL VALVE
	BALL VALVE
	REDUCER
	SLOPE
	SPOOL PIECE
L.C	LOCKED CLOSED
L.O	LOCKED OPEN
C.S.O	CAR SEAL OPEN
C.S.C	CAR SEAL CLOSED

Notes:
1- Distance between PSV and reducer is minimum.
2- Downstream block valve is installed flange to flange with PSV; if the built-up back pressure is too high, this block valve is installed downstream the reducer.
3- Installation of elbow between PSV and reducer is forbidden.
4- Piping class change could be required downstream PSV, depending of relief pressure and PSV size.Depressurisation valve arrangement

Appendix 3 Depressurisation valve arrangement



LEGEND

BLOCK VALVE, GATE VALVE OR FULL BORE BALL VALVE

BALL VALVE

FO RESTRICTION ORIFICE

REDUCER

SLOPE

F.O	FAIL OPEN
L.O	LOCKED OPEN
C.S.O	CAR SEAL OPEN
T.S.O	TIGHT SHUT-OFF
F.B	FULL BORE

Note: 1 - 600 mm minimum or specify BDV in piping class for high pressure and low temperature if required