

GENERAL SPECIFICATION

PROCESS

GS EP ECP 103

Process sizing criteria

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1. Scope

1.1 Purpose of the specification

This general specification gives minimum requirements for process sizing criteria to be used during pre-project and basic or detailed engineering studies.

1.2 Applicability

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

This specification is not retroactive. However 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 and to pipelines and subsea pipeline systems (refer to specific notes).

For oil sand froth production or subsea process equipment (separator, pump etc.), a specific design basis document shall be developed and submitted to Company for approval.

2. Reference documents

The reference documents listed below form an integral part of this General Specification.

External Documents

Unless otherwise stipulated, the applicable version of these documents, including relevant appendices and supplements, is the latest revision published at the effective date of this document.

Reference	Title
API STD 520 PT I	Sizing, Selection, and Installation of Pressure-relieving Devices in Refineries Part I - Sizing and Selection -
API RP 520 PT II	Sizing, Selection, and Installation of Pressure-Relieving Devices in Refineries Part II—Installation -
EI	Guidelines for the avoidance of vibration induced fatigue failure in process pipework (2008)
ISO 23251 / API STD 521	Petroleum, petrochemical and natural gas industries Pressure relieving and depressuring systems
TEMA R	Tubular exchanger manufacturers association

Total General Specifications

Unless otherwise stipulated, the applicable version of these documents, including relevant appendices and supplements, is the latest revision published in the applicable yearly collection.

Reference	Title
GS EP COR 004	Limiting flow velocities calculations for erosion corrosion purpose
GS EP ECP 101	Glycol package unit
GS EP ECP 105	Process documents to be prepared during engineering phase
GS EP INS 101	Instrumentation engineering, supply and construction general requirements
GS EP INS 120	Control Valves
GS EP MEC 028	Rotating Machine Package Specification - Positive Displacement Motopump according to API Standard 674
GS EP MEC 251	Supply of Axial and Centrifugal Compressors and Expander-compressors for Petroleum, Chemical and Gas Industry Services according to API Standard 617
GS EP MEC 271	Centrifugal Pumps for the Petroleum, Chemical and Gas Industry Services According to API Standard 610
GS EP PLR 100	Submarine pipeline systems
GS EP PVV 107	Piping Stress Analysis
GS EP PVV 112	Piping material classes
GS EP PVV 119	Design and fabrication of metallic piping supports - Standard for piping supports
GS EP PVV 211	Design and fabrication of pressure vessels according to ASME VIII
GS EP PVV 212	Design and fabrication of pressure vessels according to BSI PD 5500
GS EP SAF 228	Liquid drainage
GS EP SAF 253	Impacted area, restricted area and fire zones
GS EP SAF 260	Design of High Integrity Protection Systems (HIPS)
GS EP SAF 261	Emergency Shutdown and Emergency De-Pressurisation (ESD & EDP)
GS EP SAF 262	Pressure protection relief and hydrocarbon disposal systems
GS EP SAF 321	Fire pump stations and fire water mains
GS EP SAF 361	SIL assignment
GS EP TEC 260	HIPS Design, Implementation and Life cycle

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3. General

3.1 Terminology and definitions

There are five types of statement in this specification, the “shall”, “should”, “may”, “can” and “must” statements. They are to be understood as follows:

Shall	Is to be understood as mandatory. Deviating from a “shall” statement requires derogation approved by Company.
Should	Is to be understood as strongly recommended to comply with the requirements of the specification. Alternatives shall provide the same level of protection or performance and this shall be documented.
May	Is to be understood as permission.
Can	Is to be understood as a physical possibility.
Must	Expresses a regulatory obligation

For the purpose of this specification, the following definitions shall apply:

Heat and material balance: this is the document giving the normal operating conditions, compositions, thermodynamic and physical properties of all streams for each steady state process case.

Normal operating conditions: these are the range of possible internal pressure, temperature, flow rate and fluid composition in stable steady state conditions considering seasonal changes, production changes during facility lifetime etc. Normal operating conditions are required to maintain product specifications and system performance.

Normal Operating Pressure/Temperature (normOP, normOT): It is the highest (positive) or the lowest (negative) normal operating pressure/temperature amongst all steady state process cases.

Maximum/Minimum Operating Pressure/Temperature (MOP, MinOP, MOT, MinOT): these are the maximum/minimum acceptable internal pressure/temperature beyond which product specifications or system performance are not maintained. Maximum/minimum operating pressure/temperature are the upper/lower boundaries of the normal operating range and correspond to high/low alarm set points.

Note for pipelines, subsea pipeline systems and reception facilities: maximum/minimum operating pressure/temperature take into account transient modes such as start-up, shutdown, turndown conditions, slugging regime, pigging, packing, and normal preservation sequence (if applicable).

Minimum wall temperature: this is the lowest temperature the equipment wall can reach. It takes into account emergency depressurization final temperature, LPG or refrigerant fluid atmospheric boiling point etc.

Minimum design temperature: this is the internal lower temperature used for the system mechanical design and material selection (i.e. coincident with maximum design pressure).

Maximum design pressure: this is the internal upper pressure used along with the design temperature for the system mechanical design. This is also the reference pressure used to

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determine at which pressure the system will be tested prior to being put on line. In case the system is protected against the risk of overpressure by a PSV, the PSV set-point is the maximum design pressure. Design codes stipulate the allowable overpressure limits (i.e. transient excursions beyond the system design pressure).

Minimum design pressure: this is the internal lower pressure used along with the design temperature for the system mechanical design. In case the system is protected against the risk of collapse with a vacuum breaker, the vacuum breaker set point is the minimum design pressure.

Maximum design temperature: this is the internal upper temperature used for the system mechanical design (i.e. coincident with design pressure). No system shall be operated beyond its maximum design temperature. Fire conditions are not taken into account.

Process Flow Diagram (PFD): it presents the succession of operations in the fluid(s) processing to reach the required products specifications including associated off sites. Refer to [GS EP ECP 105](#).

Utility Flow Diagram (UFD): equivalent of PFD for utilities generation and distribution systems. Refer to [GS EP ECP 105](#).

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

P&ID's symbols are used for a good understanding and also in order to simplify the representations on P&ID's. Refer to [GS EP ECP 105](#).

Process data sheet: it is the data sheet for all process components (packages, control valves, PSV's, rupture disks, instruments, etc.) containing the information required for sizing. A sketch with main dimensions such as internal diameter, length, liquid levels for vessels and columns. It is issued by Process discipline. Refer to [GS EP ECP 105](#).

The information given in the process data sheets are also used in the "mechanical data" sheet, issued for the purchasing of the said equipment. Refer to API, ASME, TEMA codes for Mechanical Data Sheet templates.

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 (100% rotating speed). The Process rated flow rate is the net Process flow because it does not take into consideration losses (i.e. internal recycle, external consumption such as the air needed for air-dryers regeneration for an air compressor package etc.).

Standpipe: is an external extension of the pressure vessel, to which multiple level instruments can be connected, but no instrument is installed inside the standpipe itself. No isolation valves are installed between the vessel and the standpipe. Each instrument is provided with individual isolation valves. Not to be confused with level instrument sensor cage.

Sensor cage: is an individual chamber in which the level sensor is installed, part of a single level instrument (normally supplied by the instrument manufacturer). The sensor cage is installed either on the pressure vessel or on a standpipe, and has dedicated process isolation valves and vent/drain facilities, allowing the level sensor to be safely removed whilst the system it is connected to is on stream. Not to be confused with standpipe.

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Toxic fluid: refer to **GS EP SAF 253**.

3.2 Abbreviations

BDV	Blow Down Valve
BS&W	Basic Sediment and Water
CFU	Compact Flotation Unit
CI	Corrosion Inhibitor
EDP	Emergency Depressurization
EOR	Enhanced Oil Recovery
FC/FO	Fail Closed / Fail Open
FVF	Fluid Viscosity Factor
HIPS	High Integrity Protection System
ID	Internal Diameter
ILAH	Interface Level Alarm High
ILAL	Interface Level Alarm Low
ILSHH	Interface Level Safety High High
ILSLL	Interface Level Safety Low Low
INLL	Interface Normal Liquid Level
LAH	Level Alarm High
LAL	Level Alarm Low
LCV	Level Control Valve
LG	Level Gauge
LSHH	Level Safety High High
LSLL	Level Safety Low Low
LT	Level Transmitters
MinOP	Minimum Operating Pressure
MinOT	Minimum Operating Temperature
MOP	Maximum Operating Pressure
MOT	Maximum Operating Temperature
MSL	Mean Sea Level
NLL	Normal Liquid Level
normOP	normal Operating Pressure
normOT	normal Operating Temperature

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NPSH	Net Positive Suction Head
OD	Outside Diameter
P&ID	Piping & Instrumentation Diagram
PAH	Pressure Alarm High
PFD	Process Flow Diagram
PFP	Passive Fire Protection
PSHH	Pressure Safety High High
PSLL	Pressure Safety Low Low
PSV	Pressure Safety Valve
PWL	Power Level
RO	Restriction Orifice
RVP	Reid Vapour Pressure
SDV	Shut-Down Valve
SIL	Safety Integrity Level
SIS	Safety Instrumented System
TEG	Tri Ethylene Glycol
TSHH	Temperature Safety High High
TSLL	Temperature Safety Low Low
UFD	Utility Flow Diagram
WHSIP	Wellhead Shut-In Pressure

4. Mechanical design

The mechanical integrity of all process, pipeline and piping systems and their operability must be maintained under all possible coincident pressure and temperature conditions. As a general statement, all assumptions made during the early phase of the study shall be firmed up as and when the design develops.

All modes of operation with associated pressure and temperature conditions shall be described in an auditable document as well as the interface conditions at the system battery limits.

Several sets of design pressure and temperature (one per operating mode) shall be defined where necessary.

The thermal specifications for pipeline systems shall be issued during basic engineering study and revised in detailed engineering study if required.

For specific applications such as deepwater field developments, temperature and pressure profiles along the pipeline system may be required.

The risk of overpressure due to reverse flow conditions shall be duly accounted for.

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Additionally, the potential impacts of a connection on an existing system (the design conditions of which may be different from the new ones) shall be thoroughly reviewed.

As stipulated in **GS EP SAF 361**, a SIL assignment analysis shall be conducted for all instrumented safety functions (PSHH, TSLL, LSHH etc.).

Reference is made to **GS EP SAF 260** for protection by a SIS or HIPS.

For a system protected by a PSV, the system pressure will increase above the PSV set point to open the valve to its full lift and full capacity position. For a system protected by a SIS, the system pressure will increase above the SIS set point until the trip signal shuts down the source of pressure. The SIS set point and response time shall be selected so that it offers a protection equivalent to the one offered by a conventional overpressure protection system. The pressure excursions beyond the system design pressure due to PSV accumulation and SIS response time shall be within the overpressure limits established by the applicable design code. The same applies to all process parameters (temperature, flowrate, level etc.).

All pressures shall be given at the selected pressure reading (gauge or absolute).

4.1 Process equipment

The following requirements apply to process components and piping designed to **GS EP PVV 112**.

4.1.1 Design pressure

4.1.1.1 Maximum design pressure

The maximum design pressure shall be selected by adding a safety margin on normOP.

As a minimum, the safety margin between normOP and the maximum (upper) design pressure of a process equipment (excluding storage tanks, atmospheric tanks, pipelines and subsea systems, receiving facilities) shall be taken as follows:

Table 1 Design pressure of equipment with positive operating pressure

normOP (bar g)	Upper design pressure (bar g)
< 1	2 or 3.5 ^(*) minimum
1 – 10	normOP + 1 bar, 2 or 3.5 ^(*) minimum
> 10	normOP + 10%
> 150	normOP + 15 bar

The Process engineer shall define MOP which is the upper boundary of the normal operating range. The high pressure alarm set point PAH shall correspond to MOP.

It shall be checked that the maximum design pressure takes into account all possible excursions beyond MOP due to upset, blocked outlet, instrument failure etc. Refer to section 4.1.1.3.

PSV set point (if any) shall correspond to the maximum design pressure.

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

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The safety margin between normOP and the maximum design pressure may need to be increased above Table 1 minimum values in order to stagger successive layers of overpressure protection (PAH, PSHH, PSV and/or SIS). The interval between these set points shall be at minimum 0.5 bar.

Reference is made to [API STD 520 PT I](#) for typical pressure relief systems characteristics (tolerances etc.).

The maximum design pressure of all process equipment and piping shall take into account transient conditions such as surge pressures due to water hammer phenomena. The water hammer surge pressure shall be calculated as per the method described in section 11: the sum of this surge pressure and normOP shall be checked against the system maximum design pressure and appropriate surge protection shall be implemented if required.

All pressures shall be given for the vapour phase. The equipment elevation relative to grade (reference level) shall be indicated.

The liquid density and the maximum liquid height shall be specified on the Process data sheet.

4.1.1.2 Minimum design pressure

The Process engineer shall define MinOP which is the lower boundary of the normal operating range. The low pressure alarm set point PAL shall correspond to MinOP.

For equipment operating under vacuum, the minimum design pressure shall be selected by adding a safety margin on normOP.

The minimum design pressure of a process equipment (excluding storage tanks, atmospheric tanks, pipelines and subsea systems) operating at pressure below atmospheric pressure shall also be taken as follows:

Table 2 Design pressure of equipment operating under vacuum conditions

normOP (bar a.)	Lower design Pressure (bar a.)
≤ 0.35	Full vacuum
0.35 – 0.6	normOP – 0.1 bar
> 0.6	0.5

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 on a case by case basis. They may not be designed for full vacuum if protective devices are provided (vacuum breaker, pressurization gas, low pressure switch, etc.).

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4.1.1.3 Particular cases

4.1.1.3.1 For receiving facilities, all possible operating modes such as packing, pigging, shut-down and restart, slugging regime etc (and associated dynamic effects) shall be taken into account to determine the required safety margin between normOP and the maximum design pressure.

4.1.1.3.2 Where no PSV are provided, the maximum design pressure shall take into account the fluid vapour pressure at maximum design temperature.

4.1.1.3.3 The design pressure of a process equipment connected to flare shall be equal to or greater than the flare system design pressure.

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

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

PSV's shall be provided at compressor discharge.

4.1.1.3.5 At the discharge of centrifugal compressor:

- Design Pressure = normOP + 1 bar minimum for normOP $\leq$ 10 bar g
- Design Pressure = normOP + 10% minimum for normOP > 10 bar g

For centrifugal compressor driven by a variable speed driver, the design pressure should be reviewed when the compressor operating curves are issued so that the compressor can operate in its entire operating range.

PSV's are required at compressor discharge if not fully rated. For the sizing of PSV's, the following upset case shall be considered:

- Compressor discharge blocked outlet.
- PSHH condition in inlet scrubber.
- Driver at maximum capacity (e.g. turbine at maximum speed).

4.1.1.3.6 Consideration shall be given to centrifugal compressor arrangement to determine the "settle out pressure" of the isolated system.

The settle out pressure is the equilibrium pressure reached between the suction and discharge isolating valves of the compression train when the compressor is stopped. The design pressure of the equipment and piping at compressor suction shall be equal to or greater than the maximum settle out pressure in order to prevent unnecessary PSV's lifting. As specified in **GS EP MEC 251**, it shall be possible to restart the compression train from the settle out pressure without depressurization.

4.1.1.3.7 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 (maximum shut-off pressure).

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If the pump characteristic curve is not yet available, the maximum shut-off pressure shall be estimated with the following formula:

$$P_d = P_{smax} + \frac{1.2 * head * d_{max}}{10.2}$$

with:

- P_d = design pressure at pump discharge, in bar g
- P_{smax} = design pressure of suction drum + static head at d_{max} and at LAH, in barg
- head = head of the pump at design point, in m
- d_{max} = maximum specific gravity of liquid pumped under normal operating conditions (obtained for maximum BS&W),

Note: when the pump characteristic curve becomes available, the system design pressure shall be checked against the actual maximum shut-off pressure.

4.1.1.3.8 At discharge of volumetric pumps:

- Design Pressure = normOP + 2 bar minimum where normOP ≤ 10 bar g
- Design Pressure = normOP + 20% minimum where normOP > 10 bar g

PSV's shall be provided at pump discharge.

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

4.1.1.3.10 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 (treated on a case by case basis depending upon tank type).

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

4.1.2 Design temperature

The design temperature is the temperature used along with the design pressure for the mechanical design of equipment.

4.1.2.1 Maximum design temperature

The maximum design temperature shall be specified as follows, with due consideration for the special cases discussed in section 4.1.2.3.

Maximum (upper) design temperature = normOT + 15°C or maximum surface temperature due to solar radiation, whichever is greater.

The 15°C margin between normOT and the upper design temperature covers the following uncertainties:

- Process model accuracy.
- Reservoir fluid temperature.

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- Reservoir fluid thermal properties.
- Reservoir model accuracy.
- Well model accuracy etc.

The Process engineer shall define MOT which is the upper boundary of the normal operating range. The high temperature alarm set point TAH shall correspond to MOT.

The margin between normOT and the upper design temperature may be reduced where the system MOT can be determined accurately and/or when better quality information becomes available. In any case, it shall not be lower than 5°C.

The margin between normOT and the upper design temperature shall not be reduced where a SIS (TSHH) is implemented. The interval between the successive layers of protection shall be at minimum 5°C.

4.1.2.2 Minimum design temperature

The Process engineer shall define MinOT which is the lower boundary of the normal operating range. The low temperature alarm set point TAL shall correspond to MinOT.

For process equipment and associated piping (designed to **GS EP PVV 112**), MinOT excludes exceptional cases such as emergency depressurisation (EDP).

Minimum design temperature = MinOT - 5°C or minimum wall temperature or minimum ambient temperature, whichever is lower.

The minimum design temperature takes into account the minimum wall temperature (resulting for instance from EDP) unless it is demonstrated that there is no possibility to immediately repressurize the system after an emergency depressurisation event (i.e. a protection system is implemented to prevent coincident minimum wall temperature and maximum design pressure).

Refer to section 14 for depressurization calculation methodology and determination of minimum wall temperature.

Where a SIS (TSLL) is implemented, TSLL set point shall be at least 5°C lower than MinOT and 5°C above the minimum design temperature.

4.1.2.3 Particular cases

4.1.2.3.1 Equipment handling LPG or refrigerant fluid shall be designed for the fluid atmospheric boiling point.

4.1.2.3.2 The minimum design temperature shall take into account potential gas blow-by from a high pressure system into a low pressure system.

4.1.2.3.3 Consideration shall be given to cooling medium failure when coolers are used.

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

Downstream of other coolers, the maximum design temperature shall be the upstream MOT.

These considerations generally apply up to the following equipment only.

4.1.2.3.4 Steam out conditions (pressure and temperature) shall be specified where steaming is considered.

4.1.2.3.5 Due consideration shall be given to upset and transient conditions such as start-up, shut-down, etc. In this case, both pressure and temperature conditions shall be specified.

4.1.2.3.6 The exceptional temperature generated by fire shall not be considered to specify the maximum design temperature.

4.1.2.3.7 For cryogenic equipment, the lower design temperature case may not be the governing case for mechanical design; the upper design temperature case (along with the associated design pressure) may prevail.

4.1.2.3.8 Where a minimum design pressure is specified, the coincident design temperature shall be defined.

4.1.2.3.9 For discontinuous/cycled processes, a set of pressure and temperature conditions shall be specified for each mode of operation.

One typical example is molecular sieve vessel for which different design conditions shall be specified for the adsorption phase and the regeneration phase.

4.2 Pipelines and risers

Reference is made to **GS EP PLR 100**.

For pipelines and subsea pipeline systems, MOP, MinOP, MOT and MinOT shall take into account transient modes of operation including packing, pigging, shut-down and restart and normal preservation sequence (if applicable). Commissioning and installation activities are excluded (for example, onshore testing of a subsea pipeline system).

4.2.1 Design pressure

All pressures shall be given at the reference level (MSL). The fluid type and its properties (minimum and maximum densities) shall be specified.

4.2.1.1 Maximum design pressure

As some standards allow transient excursions beyond the system design pressure during surge incidents, the Process engineer in charge shall provide the following information:

- MOP.
- The list of all potential pressurization sources including, but not limited to WHSIP, compressors, dead oil pumps, water injection pumps, production pumps, chemical injection pumps.
- The maximum design pressures associated to these pressure sources (with no consideration of potential surge pressure).
- The potential pressure increase due to surge.
- PSV accumulated pressure or the overpressure reached before completion of the shutdown sequence. In case of a full pressure rated system, there is no relieving device and the ultimate pressure corresponds to the maximum design pressure with surge.

Exceptional scenarios such as the pressurization of a production pipeline or jumper by chemical injection (methanol, inhibitors etc.) are not expected to occur during normal operation cycles,

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and therefore are not included in MOP. They shall be taken into account for determining the system maximum design pressure.

4.2.1.2 Minimum design pressure

MinOP will be used to evaluate the risk of collapse due to external forces and shall at least take into account the following scenarios: depressurization, cooling of a liquid full system, atmospheric conditions in a water reinjection system etc.

4.2.2 Design temperature

4.2.2.1 Maximum design temperature

Maximum (upper) design temperature = MOT + 15°C or maximum surface temperature due to solar radiation, whichever is greater.

This margin covers the same uncertainties as for a process system. Refer to section 4.1.2.1

The margin between MOT and the upper design temperature may be reduced where the system MOT can be determined accurately and/or when better quality information becomes available. In any case, it shall not be lower than 5°C.

The margin between MOT and the upper design temperature shall not be reduced where a SIS (TSHH) is implemented. The interval between the successive layers of protection shall be at minimum 5°C.

4.2.2.2 Minimum design temperature

MinOT shall take into account the following operating conditions: high rate depressurization, well opening, riser start-up sequence with cold preservation fluid and gas-lift injection (Joule-Thomson effect). However, MinOT does not take into account exceptional cases such as hydrate removal operation.

Minimum design temperature = MinOT - 5°C or minimum ambient temperature, whichever is lower.

Where a SIS (TSLL) is implemented, TSLL set point shall be at least 5°C lower than MinOT and 5°C above the minimum design temperature.

5. Design methods

5.1 Pumps

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

- 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%

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Notes:

- When a RO is used as minimum flow protection, the permanent re-circulation flow shall be added to the net process flow rate
- Normal and rated flows should be identical for the following services:
 - Intermittent service pumps.
 - When the pump has been overrated to allow for a centrifugal type and if overrating is $\geq 10\%$.
 - Re-circulation flow, such as product loading and amine filtration.
- For fire water pumps see specification [GS EP SAF 321](#)

5.2 Compressors

If the flowrate is constant, the compressor process rated capacity shall be the maximum expected flowrate in normal operating conditions with no additional margin. If the flow is coming from a production separator, a flow margin of 10% should be added in order to account for the possible slugging regime at the inlet of the production separator.

All possible operating conditions (including but not limited to variations of gas compositions, molecular weight, Cp/Cv, suction/discharge pressure, suction temperature, flowrate) shall be considered when sizing the compressor. These operating conditions shall be reported on the compressor Process data sheet.

The compressor associated equipment (suction drums, after-coolers etc.) shall be sized adequately to allow operation of the compressor at its maximum capacity (i.e. at compressor driver maximum power output). Where more than one gas compression trains operate in parallel, in case of shutdown of one train, the throughput through the train(s) remaining in service should not be limited by the size of the compressor associated equipment or the capacity of an inter-stage amine or TEG unit.

5.3 Gas Dehydration Unit (TEG)

Reference is made to [GS EP ECP 101](#).

5.4 Other process equipment

- Shell and tube heat exchangers and air coolers: 10% on heat exchange surface and additional margin for fouling
- Plate heat exchangers: 15% on heat exchange surface and no additional margin for fouling
- Fired heaters and furnaces: 10% on duty
- Boilers: 10% on steam flow rate
- First stage production separator: 10% on nominal gas inlet flow rate and 10% of nominal liquid inlet flow rate (attention shall be paid to liquid slugs, refer to section 6.2)

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- Other drums: 10% of inlet flow rates
- Pipelines transporting multiphase fluids 10% on overall pressure drop

No over design should be applied on other equipment unless project-specific documentation specifies these margins.

For the sizing factor to be applied on electrical motors, diesel engines, gas turbines, etc., refer to the relevant specifications issued by other disciplines.

5.5 Storage / RVP

When the crude oil must meet a RVP specification, the API crude method shall be applied. In addition it shall always be checked that for all normal operating situations no degassing occurs within neither the degassing boot nor the storage tank. Indeed, the use of a degassing boot shall be limited to upset conditions in the upstream process facilities. Degassing in a storage tank is not accepted with the only exceptions of either tank filling operation or fluctuation of oil vapor pressure with atmospheric environment changes (pressure and/or temperature).

In the case of light crude oil or condensate (with specific gravity lower than or equal to 0.85), the most conservative method between API Naphta method and API crude method shall be used.

6. Vessel sizing criteria

6.1 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 packs and cyclones can be used after written approval from Company Process engineer in charge.

The basis for sizing the vapour area is the Souders and Brown coefficient K:

$$v_g = K \sqrt{\frac{\rho_l - \rho_g}{\rho_g}}$$

with:

ρ_l = liquid density in kg/m³

ρ_g = vapour density in kg/m³

V_g = gas velocity in m/s

K is a coefficient depending on the service as well as on the presence or absence of wire mesh. The following Souders and Brown coefficient K values shall be adopted:

Configuration	Without mist extractor	With wire mesh	With vanes
Horizontal	0.08	0.10	0.16
Vertical	0.04	0.08	0.16

Steam drums shall be provided with wire mesh and sized for a K value of 0.06.

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The mist eliminator performance shall be guaranteed by the vendor for the project-specific operating conditions. For vane packs, the gas velocity shall not exceed the following value:

$$V_{\max} = F_{\text{vane}} * 2.46 * \frac{(g \cdot \sigma \cdot (\rho_{\text{liq}} - \rho_{\text{gas}}))^{0.25}}{\sqrt{\rho_{\text{gas}}}}$$

where:

$V_{\max}$ is the maximum acceptable gas velocity (m/s) in the vane pack,

ρ_{liq} is the actual liquid density (kg/m³) in the vane pack,

ρ_{gas} is the actual gas density (kg/m³) in the vane pack,

σ is the actual surface tension (N/m) in the vane pack,

F_{vane} is a dimensionless correction factor to be provided by the vane pack supplier. This parameter aims at accounting for the area inside the vane which is used for liquid drain and not for gas flow. As a first estimate, the value of 0.74 can be retained.

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 the fifth of the internal diameter.

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

6.2 Surge volume for liquids

6.2.1 Two-phase vessels

The following surge times (LAL-LAH) shall be met at design capacity:

Table 3 Two-phase vessel surge volume

Services	Time (Minutes)
Production separators (no slug-catcher service)	
A. to heater	5
B. to others	without pump: 3 with pump: 5 of the pump rated flow rate
Reflux Drum	5
Fractionation tower bottom	without pump: 3 with pump: 5 of the pump rated flow rate
Steam / condensate flash drum (process & utilities units)	5
Steam drum (steam generation)	10
Steam deaerator	15
Atmospheric degassing drum "gas boot"	15 for storage tank with floating roof 10 for storage tank with fixed roof
Others Drums (scrubbers, etc.)	without pump: 2 with pump: 3 of the pump rated flow rate
CFU	To be provided by the supplier

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

When applicable, the retention time below LSLL should be compatible with the time required to close the outlet SDV or to stop the pump.

6.2.2 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 phase 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 higher density liquid is taken between the bottom and INLL.

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

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

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

If the liquid is pumped, the retention time (based on the volume between ILAL and ILAH for the higher density liquid and volume between LAL and LAH for the lighter density liquid) shall be at least the time required for two phase vessels (section 6.2.1) which could increase the retention time of the considered liquid.

When applicable, the retention time below the ILSLL and the retention time below LSLL should be compatible with the time required to close the corresponding SDV or to stop the corresponding pump.

6.2.3 Production separator ensuring slug catcher service

If the vessel is a production separator designed to receive liquid slugs (slug catcher service), the determination of the type and volume of the sizing liquid slug should normally be based on hydrodynamic study of the upstream pipeline, flowline or network.

The liquid slug volume shall be accommodated in the vessel as follows:

- For continuous intermittent slugs (i.e. cyclic hydrodynamic or severe slugs), between LAL and LAH.
- For transient slugs (i.e. induced by variations in the operating conditions of the upstream system, such as well or flowline start-up/restart, flow ramp-up, flowline pigging), between NLL and LAH.
The NLL shall be set to provide, between this level and LSLL, a minimum retention time, at normal operating flowrate, of:
 - 1.5 minutes without pump.
 - 2.5 minutes with pump.

In addition, if the vessel is designed as a three-phase separator with separate oil/condensate and water retentions accommodation capability, the above sizing specification shall apply to both liquid compartments.

6.3 Level measurement

Intrusive (top mounted) level instruments are not permitted for pressure vessels: level instruments shall be mounted externally and accessible. The sensor cage shall be provided with individual isolation valves. Reference is made to **GS EP INS 101** for level transmitter type selection.

The upper and lower Level Gauge (LG) and Level Transmitters (LT) connections used for control function shall be located so that it is possible to read all alarms and shutdown levels. The connections for safety functions LSLL and LSHH shall be separate from the connections for control.

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

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For both horizontal and vertical vessels, the following criteria shall be applied:

- LAL minimum at 10% of the measurement range.
- LAH maximum at 90% of the measurement range.
- NLL at 50% of LAH-LAL range, except in separators having an over capacity ensuring a slug catcher function.
- LSLL typically at 1 to 2 minutes of the draw-off capacity with 200 mm minimum between LAL and LSLL for vertical drum and 100 mm for horizontal drum.
- LSHH typically at 1 to 2 minutes of the draw-off capacity with 150 mm minimum between LAH and LSHH for vertical drum and 100 mm for horizontal drum. In the case the separator serves as slug-catcher, 2 minutes of the draw-off capacity shall be the minimum.

The above level positions also apply for interfaces levels.

Connections of lower (respectively upper) level instrument tapping on liquid outlet (respectively gas outlet) piping are forbidden. LSLL or ILSLL should 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. A connection on the bottom of a horizontal vessel is acceptable for clean services (no risk of plugging of instrument nozzle).

On production separators, there shall be no vertical connections for instrument tappings on the vessel bottom. Vertical connections for instrument tappings are permitted on the vessel top.

The use of standpipes should be avoided. If a standpipe is definitely required, it shall be designed in accordance with **GS EP PVV 211** and **GS EP PVV 212**. No isolation valves shall be installed between a standpipe and the vessel it is attached to.

Refer to **GS EP TEC 260** for requirements regarding HIPS sensors.

The definition of alarm and safety levels shall be reviewed with piping and instrumentation specialists at an early stage of basic engineering in order to ensure feasibility. Dedicated level sketches combining process levels and vessel/tank dimensions shall be developed.

7. Diameter of vessels

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

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

For vessels with internal diameter lower than 1,000 mm, flanged heads may be used.

8. Heat exchangers

8.1 Fouling factors

The following fouling factors shall be used for the design of shell and tubes heat exchangers and air coolers.

8.1.1 Process fluids

Table 4 Process fluid fouling factors for shell and tube exchangers

	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

8.1.2 Utility fluids

Table 5 Utility fluid fouling factors for shell and tube exchangers

	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

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8.2 Plate type heat exchangers

No fouling factor shall be applied for plate and frame heat exchangers and plate fin heat exchangers but an extra surface of 15% shall be added to the calculated area.

When installed on critical services (such as the main cooling loop of the plant), n+1 plate exchangers shall be provided i.e. one exchanger can be isolated whilst the plant is operated at full capacity. The frame dimensions shall be sufficient to accommodate at least 10% more plates than originally required.

8.3 Temperature approach for heat exchangers

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

- 3°C for kettles.
- 5°C for others shell and tubes exchangers.
- 10°C for air coolers.
- For plate type heat exchangers, the temperature approach may be optimized on a case by case basis.

It shall be checked that in all operating and upset conditions the thermal stress due to the temperature difference between the tubes and the shell does not exceed the code-allowed limits.

8.4 Specific requirement for heat exchangers

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

For shell and tubes heat exchangers, where a tube rupture will not cause overpressure of the low pressure side beyond its hydrotest pressure (with due reference to the exchanger design code), then no overpressure protection is required; otherwise, bursting disks shall be installed.

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 expansion bellows 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) where process control is required.

9. Equipment nozzles

9.1 Nozzles and manholes

See general specification [GS EP PVV 211](#).

The number of manholes depends on the type of vessel internals. Internals shall be removable through the manhole. In any case for vessels longer than 9 m, at least two manholes are required for sufficient air ventilation during internal inspection.

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9.2 Vent, drain and overflow connections

The following shall apply, depending on the vessel position:

Vertical vessels:

- The drain nozzles of the vessel should be connected onto the outlet line at low point
- The drain nozzles of the vessel should be connected directly onto the capacity if the outlet line protrudes inside the vessel
- For vessels equipped with internals (baffle), a drain connection is required on each compartment

Horizontal vessels:

- The drain nozzles of the vessel should be connected directly onto the capacity
- For horizontal vessels longer than 6 m, additional drain connections are required with a maximum distance of 3 m in between any two drain connections
- 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:

Table 6 Vessel vent and drain connection size requirements

Volume or internal diameter of vessel (m ³ or mm)	Minimum Vent diameter	Minimum Drain diameter
$V \leq 15$ or $D \leq 2500$	2"	2"
$15 < V \leq 75$ or $2500 < D \leq 4500$	2"	3"
$75 < V \leq 220$ or $4500 < D \leq 6000$	3"	4"
$220 < V \leq 420$ 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.

9.3 Utility connections

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

- Drums and heat exchangers (where applicable): 2".
- For large vertical drums, two 2" connections will be provided for diameter above 4.5 m.

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- For horizontal vessel with a length above 6 m and operating in toxic service, two 2" connections will be provided.
- If the 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)
 - $D \leq 4$ m: 2"
 - $4 \text{ m} < D \leq 5.5$ m: 3"
 - $D > 5.5$ m: 4"

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

10. Elevation of equipment

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

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

If there is no other process requirement regarding the elevation, a note on P&ID will indicate "minimum for piping".

11. Piping

11.1 General design and hydraulics

11.1.1 Minimum piping sizes

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

- $\frac{3}{4}$ " for all lines located above ground except for the following:
 - 2" for process line on pipe rack
 - 2" for utility line on main pipe rack
 - 2" for pipe on pipe sleeper
- 2" for all lines located underground, whatever the material they are made of, i.e. steel pipe or non metallic piping.

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11.1.2 Pump suction line

Pump suction lines shall be sized so as the net positive suction head available (NPSH_A) at the rated point exceeds by 20 % at least the net positive suction head required (NPSH_R) as per **GS EP MEC 271**. It shall also be checked that all over the operating range of the pump the difference between (NPSH_A) and (NPSH_R) is not lower than 0.5 m.

Pump suction line shall not be smaller than pump suction nozzle. Where the suction line and the pump inlet nozzle are not the same size, an eccentric reducer shall be used in order to meet NPSH requirement.

Pump suction valves shall be of the same diameter as the line and full bore.

The static head used in calculating the available NPSH_A shall be taken from the LSL of the vessel upstream of the pump to one of the following:

- The centerline for a horizontal pump.
- The suction impeller for a vertical centrifugal pump.

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

The design of suction lines from storage tanks shall be based on the 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, the acceleration head shall be considered. Reference is made to **GS EP MEC 028**.

11.1.3 Isolation

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

- Operating pressure above 70 bar g.
- Operating pressure above 35 bar g. only for LNG.
- H₂S partial pressure > 1 bar.
- Toxic fluid, very corrosive or abrasive.

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

All equipment will generally be isolated by spectacle blinds, or spacer and ring blinds (when isolated for maintenance) installed as close as possible to the 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:

Table 7 Equipment and piping positive isolation requirements

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

11.1.4 Double block and bleed arrangement

Double block and bleed valves shall be used for condition stated on section 11.1.3. 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 standard bleed arrangement shall be a single $\frac{3}{4}"$ or 1" ball valve with a blind flange. The bleed can be either atmospheric or routed to a closed system (flare or drain) depending on the quantity, rate and type of material being released.

11.1.5 Control valve

The pressure drop across a control valve shall be at least 20% of the system dynamic pressure at normal flow rate or 0.7 bar whichever is greatest. If a lower minimum pressure drop is required, Company's instrumentation specialists shall be consulted.

Reference is made to **GS EP INS 120**.

Non-continuous service 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 may be provided with a hand-wheel and no by-pass. Company approval shall be requested for implementation of a by-pass.

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

In case of mis-operation, the gas blow-by situation shall consider the flow rate through the control valve fully open and its by-pass, where it is installed, fully open too. 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.

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Sizing of bleed devices of control valve manifolds:

- Upstream control valve line size $\geq 6"$:
 - For control valve open by fluid failure (FO), one maintenance bleed valve ($\frac{3}{4}"$ block valve with blind flange) shall be installed downstream of the control valve
 - For control valve closed by fluid failure (FC), two maintenance bleed valves shall be installed, one upstream and one downstream of the control valve
- Upstream control valve line size $< 6"$:
 - One maintenance bleed valve shall be installed downstream of the control valve whatever the control valve position (FO or FC) on fluid failure

11.2 Material selection

Piping material selection shall take into account the potential for corrosion (internal and external) and erosion problems associated with the fluids carried (composition, free water, carbon dioxide, hydrogen sulphide, oxygen, amount and characteristics of solids, etc.). Company's corrosion specialists shall always be consulted to make or validate this selection.

Piping wall thickness results essentially from:

- The design pressure and temperature (refer to section 4.2 of the present specification) selected by the Process engineer.
- The material selected (yield strength), the corrosion allowance selected and the applicable design code (under the responsibility of Corrosion engineers and Mechanical/Piping engineers).

Reference is made to **GS EP PVV 112**.

11.3 Insulation and heat tracing

Thermal insulation for hot or cold services is required where 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 provide access to instruments and pipe fittings without damaging the shield.

PFPP may be specified in order to reduce the relief load to the flare system in case of fire. However no credit shall be given to heat insulation for relief load reduction.

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11.4 Line sizing criteria

This section applies to general steel piping between pieces of process equipment within a plant section (PVV piping).

It does not apply to:

- Relief piping (upstream or downstream pressure relieving elements: refer to section 13.5).
- Fire water lines, whatever the material (refer to [GS EP SAF 321](#)).
- Non-metallic materials or material with a Young's modulus substantially different from that of steel (typical range for steel: 150 – 250 GPa).
- Pipelines (PLR piping including offshore risers, sea lines, buried pipelines etc.).
- Vessel nozzles and piping immediately upstream (where velocity should be reduced before entering a scrubber for example).

Turbulent energy in process systems puts a strain on piping. This specification is concerned with permanent dynamic strain or wear (flow induced vibrations, pressure drop/turbulence, erosion/wear etc.) regarded as governing for line sizing. Static or transient dynamic strain is addressed by Piping Stress Analysis (Refer to [GS EP PVV 107](#)).

When sizing a new process line, i.e. selecting diameter and thickness of piping, or checking an existing plant for new conditions, the following process-related criteria shall be considered and the selected size shall comply with the most severe criteria:

- Avoidance of erosion-corrosion (velocity and API C factor criteria)
- Pressure drop limitation and prevention of deposition
- Avoidance of vibration induced fatigue failure (pv^2 criteria and risk of pressure surge)

Multiphase streams are considered as homogeneous mixtures (negligible slippage between the gas and liquid phases) and the velocity and density can be accurately derived from the following formulae:

$$\rho_m = \frac{W_g + W_o + W_w}{\frac{W_g}{\rho_g} + \frac{W_o}{\rho_o} + \frac{W_w}{\rho_w}} \quad \text{and} \quad V_m = \frac{W_g + W_o + W_w}{\frac{\rho_m}{\pi ID^2} \cdot 4}$$

with:

W_o , W_w and W_g are respectively the oil, water and gas mass flow rates (kg/s)

ρ_o , ρ_w and ρ_g are the oil, water and gas densities (kg/m³)

ID is the internal line diameter (m)

ρ_m is the mixture density (kg/m³)

V_m is the mixture velocity (m/s)

For the application of the following line sizing criteria, the boundary between gas and liquid/multiphase services is set at a liquid fraction of 5% volume in actual conditions.

11.4.1 Avoidance of erosion-corrosion problems

Reference is made to **GS EP COR 004**. Lower limits may apply for water containing specific additives such as EOR polymers.

11.4.2 Pressure drop limitation and prevention of deposition

The fluid shall be allowed to flow from one end to the other end of the pipe without excessive friction loss. For the sizing of pump suction lines, refer to section 11.1.2.

Specific services such as polymer solutions, viscous fluids, chemicals, solvents etc. may impose additional constraints on line sizing. For the particular case of rich amine solutions, TEC/COR shall be consulted.

Refer to **GS EP SAF 228** for drain connection sizing criteria.

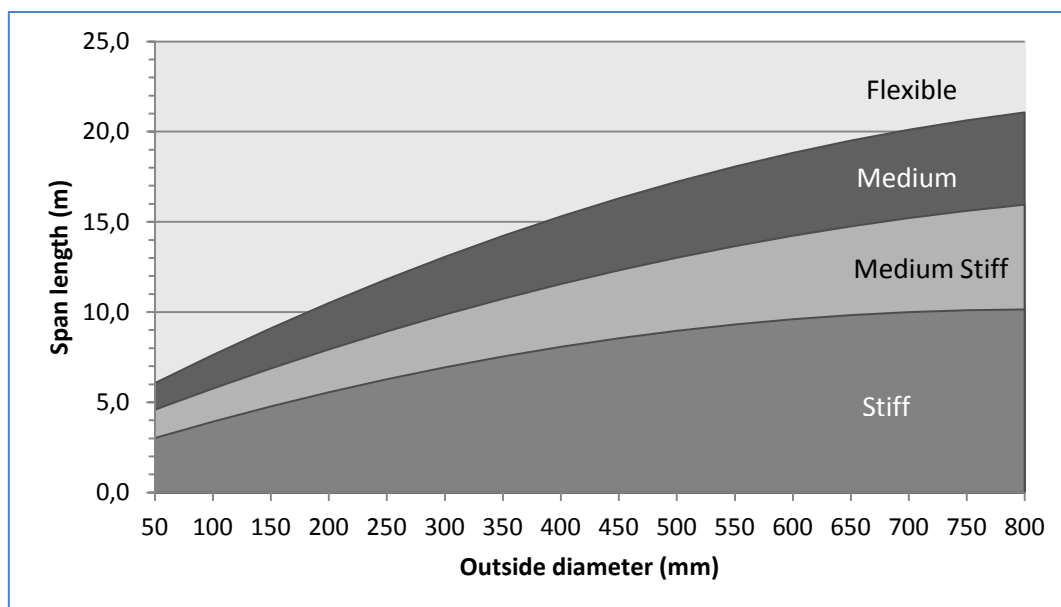
For sea water service, the velocity shall not be less than 2 m/s.

11.4.3 Avoidance of vibration induced fatigue failure

11.4.3.1 Piping support arrangement

The different types of piping support arrangement are illustrated in Figure1.

Figure 1 Pipe support arrangement map



The span length is the distance between two consecutive main line supports (pipe anchor, sliding shoe etc.; items such as spring hangers, shock arrestors etc. are not considered as main pipe supports). Refer to **GS EP PVV 119** for details.

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11.4.3.2 Low frequency vibrations in process piping

New design

The maximum allowable pv^2 to avoid vibration induced fatigue failure for a new design is given in Table 8. Medium Stiff arrangement corresponds to a standard, state-of-the-art piping support arrangement. No specific vibration analysis study is required except possibly for piping connected to rotating equipment and other services listed in **GS EP PVV 107**. These criteria apply for all fluids.

Debottlenecking

In the frame of the debottlenecking of an existing plant, with due consideration given to potential displacements (thermal stress, earthquake, swell etc.), it may be possible to stiffen the piping support arrangement and operate an existing line at higher pv^2 than originally specified.

All knowledge of the line history and its actual condition shall be used.

Piping, Mechanical and Structural specialists shall be consulted to check whether the existing piping supports and structures can actually be reinforced.

A detailed vibration analysis study shall then be conducted, and the results thereof examined, to confirm that the likelihood of fatigue failure remains remote. It shall cover not only the main line but also the small bore connections including but not limited to instrumentation tapplings, thermo-wells, injection quills etc.

Subject to the above conditions, the maximum allowable pv^2 to avoid vibration induced fatigue failure may be relaxed to the values given in Table 9 for liquids and multiphase fluids and Table 10 for gases ([EI](#)).

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Table 8 New Design - Gas and liquid lines - Determination of maximum allowable pv^2 to avoid excessive vibrations (no specific vibration analysis required)

Pipe support arrangement	D (inch)	D (mm)	α	β	Max. pv^2 (Pa)
Medium stiff (standard)	2"-30"	60-762	$\alpha = 283921 + 370.24 * D$	$\beta = 0.1106 * \ln(D) - 1.501$	$0.5 * \alpha * (D/\epsilon)^\beta$
Medium	2"-8"	60-219	$\alpha = 13.1129 - \frac{4.7455}{10^3} * D + \frac{1.4141}{10^5} * D^2$	$\beta = -0.132 + \frac{2.282}{10^4} * D - \frac{3.7245}{10^7} * D^2$	$0.5 * \exp(\alpha * (D/\epsilon)^\beta)$
Medium	10"-30"	273-762	$\alpha = 150412 + 208.93 * D$	$\beta = 0.0815 * \ln(D) - 1.3269$	$0.5 * \alpha * (D/\epsilon)^\beta$
Flexible	2"-8"	60-219	$\alpha = \frac{1.3175}{10^5} * D^2 - \frac{4.4213}{10^3} * D + 12.217$	$\beta = -\frac{4.622}{10^7} * D^2 + \frac{2.835}{10^4} * D - 0.164$	$0.5 * \exp(\alpha * (D/\epsilon)^\beta)$
Flexible	10"-30"	273-762	$\alpha = 41.21 * D + 49397$	$\beta = 0.0815 * \ln(D) - 1.3842$	$0.5 * \alpha * (D/\epsilon)^\beta$

Table 9 Debottlenecking – Liquid and multiphase lines - Determination of maximum allowable pv^2 to avoid excessive vibrations (specific vibration analysis required)

Pipe support arrangement	D (inch)	D (mm)	α	β	Max. pv^2 (Pa)
Stiff	2"-30"	60-762	$\alpha = 446187.1 + 645.51 * D + 9.166 / 10^4 * D^3$	$\beta = 0.1 * \ln(D) - 1.3739$	$0.5 * \alpha * (D/\epsilon)^\beta$
Medium stiff (standard)	same as for new design				
Medium					
Flexible					

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Table 10 Debottlenecking – Gas lines - Determination of maximum allowable pv^2 to avoid excessive vibrations (specific vibration analysis required)

Pipe support arrangement	D (inch)	D (mm)	α	β	Max. pv^2 (Pa)
Stiff	2" – 30"	60 - 762	$\alpha = 446187.1 + 645.51 * D + 9.166 / 10^4 * D^3$	$\beta = 0.1 * \ln(D) - 1.3739$	$0.3 * \alpha * (D/\epsilon)^\beta / FVF$
Medium stiff (standard)	2" – 30"	60 - 762	$\alpha = 283921 + 370.24 * D$	$\beta = 0.1106 * \ln(D) - 1.501$	$0.3 * \alpha * (D/\epsilon)^\beta / FVF$
Medium	2" – 8"	60 - 219	$\alpha = 13.1129 - \frac{4.7455}{10^3} * D + \frac{1.4141}{10^5} * D^2$	$\beta = -0.132 + \frac{2.282}{10^4} * D - \frac{3.7245}{10^7} * D^2$	$0.3 * \exp(\alpha * (D/\epsilon)^\beta) / FVF$
Medium	10" – 30"	273 - 762	$\alpha = 150412 + 208.93 * D$	$\beta = 0.0815 * \ln(D) - 1.3269$	$0.3 * \alpha * (D/\epsilon)^\beta / FVF$
Flexible	2" – 8"	60 - 219	$\alpha = \frac{1.3175}{10^5} * D^2 - \frac{4.4213}{10^3} * D + 12.217$	$\beta = -\frac{4.622}{10^7} * D^2 + \frac{2.835}{10^4} * D - 0.164$	$0.3 * \exp(\alpha * (D/\epsilon)^\beta) / FVF$
Flexible	10" – 30"	273 - 762	$\alpha = 41.21 * D + 49397$	$\beta = 0.0815 * \ln(D) - 1.3842$	$0.3 * \alpha * (D/\epsilon)^\beta / FVF$

where:

D: Pipe outside diameter (mm)

ϵ : pipe thickness without corrosion allowance (mm) (end-of-life conditions)

Fluid Viscosity Factor (FVF) = (gas dynamic viscosity)^{0.5} (viscosity expressed in cP)

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11.4.3.3 High frequency acoustic vibrations downstream pressure reducing systems

In gas systems, pressure let-down devices (chokes, JT valves, control valves such as compressor anti-surge or recycle valves, depressurization orifices excluding PSV) with large pressure drop can generate high frequency vibrations (500-2 000Hz). This vibration mode can be detrimental to small bore connections, welded tees etc. as it generates acoustic-induced fatigue which can lead to rapid rupture. The time until failure is related to the acoustic power level (PWL) at the location of interest. PWL can be determined as follows (E1):

$$PWL = 10 \log_{10} \left[\left(\frac{\Delta P}{P_{upstrm}} \right)^{3.6} W^2 \left(\frac{T_{upstrm}}{MW} \right)^{1.2} \right] + 126.1 - AT$$

where:

PWL (dB)	acoustic power level at the location of interest
ΔP (Pa)	pressure drop across pressure let-down device
P_{upstrm} (Pa abs.)	pressure upstream of let-down device
W (kg/s)	mass gas flow rate
T_{upstrm} (K)	absolute temperature upstream of let down-device
MW (g/mol)	gas molecular weight

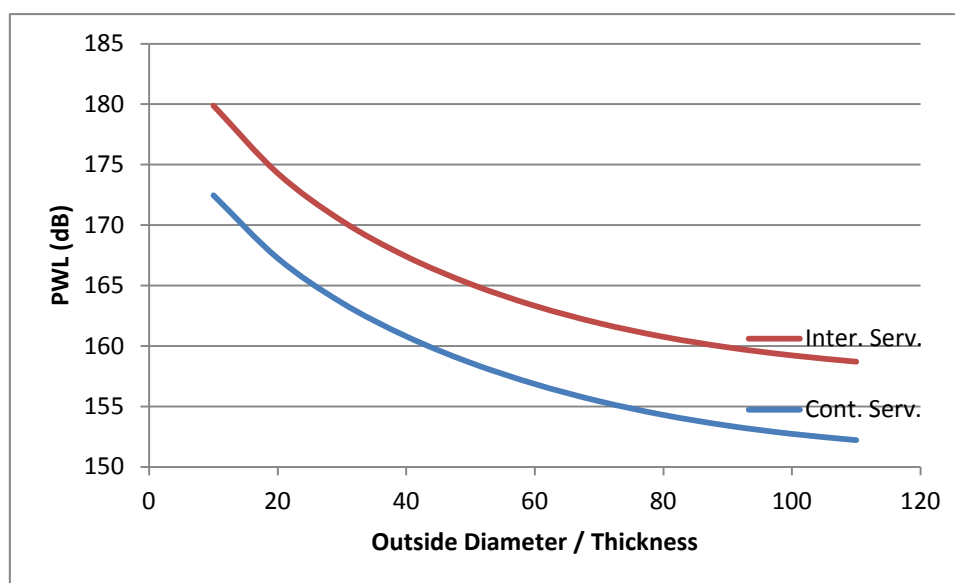
AT is the attenuation term at the location where the PWL should be evaluated (for example at the location of a branch):

$$AT = 0.06 \left(\frac{L_{downstrm}}{ID} \right), \text{ or 6 dB per 100 diameters}$$

with:

AT	dB	power attenuation term
$L_{downstrm}$	m	distance of location downstream of the pressure let-down device
ID	m	internal diameter of line downstream of the let-down device

Figure 2 gives the maximum acceptable values of PWL for intermittent and continuous services (1). Any line downstream a pressure reducing device shall comply with these upper values of PWL.

Figure 2 Acoustic Fatigue Limitation


(1): Any cumulative operation longer than 24 hours over the lifetime of the facility shall be treated as “continuous”.

11.4.3.4 Vibrations due to pressure surge (water hammer)

To calculate the effect of water hammer, Joukowsky’s method can be followed (refer also to the software recommended in the chapter 13 of this GS for more accurate calculations). In particular, it shall be checked that for any possible pressure surges inside the line, the resulting surge pressure added to the operating pressure is within the code-allowed limits.

As the result of Joukowsky’s calculation is conservative, if the predicted pressure surge does not exceed the system design pressure, then no further surge study may be required. In the frame of a debottlenecking study, it shall be checked that the higher fluid velocity does not induce a risk of pressure surge.

Note that Joukowsky’s method is valid not only for PVV piping but also for PLR piping.

12. Flare and cold vent systems

12.1 Determination of flare load

All possible causes of individual and cumulative relief shall be listed and analyzed. Reference is made to [GS EP SAF 262](#).

12.2 EDP calculations

Reference is made to [GS EP SAF 261](#). All individual reliefs from each blowdown system within a given fire zone shall be combined in an overall EDP relief to flare.

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12.2.1 Individual blowdown line sizing

For each blowdown system, the individual restriction orifice and depressurization line shall be sized on the following basis:

- Discharge of the blowdown system inventory using the system design pressure as initial pressure except for the compressor systems for which the settle out pressure shall be used, down to 7 barg or 50% of design pressure whichever is the most stringent.
- With or without fire whichever is the most severe; refer to **GS EP SAF 261** for fire case heat input calculation.
- Depressurization time as specified in **GS EP SAF 261** (typically 15 mn for fire case and cold case).
- Initial temperature equal to MinOT. For equipment which could be isolated under pressure, the initial temperature will be the minimum ambient temperature. When MinOT is higher than the minimum ambient temperature, the initial pressure will be calculated due to the temperature change considering that at the time of the isolation the pressure and the temperature are respectively the maximum design pressure and MinOT.
- For the initial liquid level in the drum, the calculations shall be conducted for both LAL and LAH. The worst case shall be retained. For pipe inventories, the normal liquid level will be used.

12.2.2 Minimum wall temperature

For each blowdown system, simulations shall be run for non-fire conditions using the orifice size as calculated in section 12.2.1 in order to determine the minimum wall temperature. The depressurization calculation shall start at the system design pressure and shall end at atmospheric pressure so as to ensure that the lowest temperature is reached.

12.2.3 Combined EDP

Once all individual restriction orifices and depressurization lines have been sized, the combined EDP calculation for a given fire zone shall be done using PSHH set points as initial pressures.

12.3 Flare tip selection

For flares and cold vents, the tip can be either open-pipe or sonic types depending on the acceptable back pressure in the flare system. The tips shall be selected on the basis of radiation and noise criteria.

When possible a sonic tip should be preferred. A sonic flare tip may not operate in sonic conditions for the entire range of specified relief flowrates. Flaring at reduced flowrate may be governing due to an increase in gas emissivity (possibly heavier gas and reduced exit velocity).

Flare tip design based on the use of Coanda effect and involving moving parts shall not be used, unless sufficient technology improvement from suppliers is found acceptable by Company. Tips with variable slots are forbidden.

The installation on a common header of flares which can run in parallel is forbidden.

A smoke suppression system shall be implemented.

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12.4 Radiation and noise levels criteria

The radiation and noise levels criteria shall follow the specifications given in the [GS EP SAF 262](#). The transmissivity factor shall be taken equal to 1, whatever the air humidity.

12.5 Emissivity coefficient

When the radiation calculations are performed by a flare tip Vendor, the Vendor shall provide the emissivity coefficient used for radiation calculations. It is necessary to check carefully the emissivity coefficient used because experience has shown that some Vendors take a very low value for this emissivity coefficient. Many times, the emissivity coefficient used by Vendors does not take into account the liquid carry over as they consider an ideal gas/liquid separation inside the flare KO drum. 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 (along with the liquid properties).

Emissivity coefficient

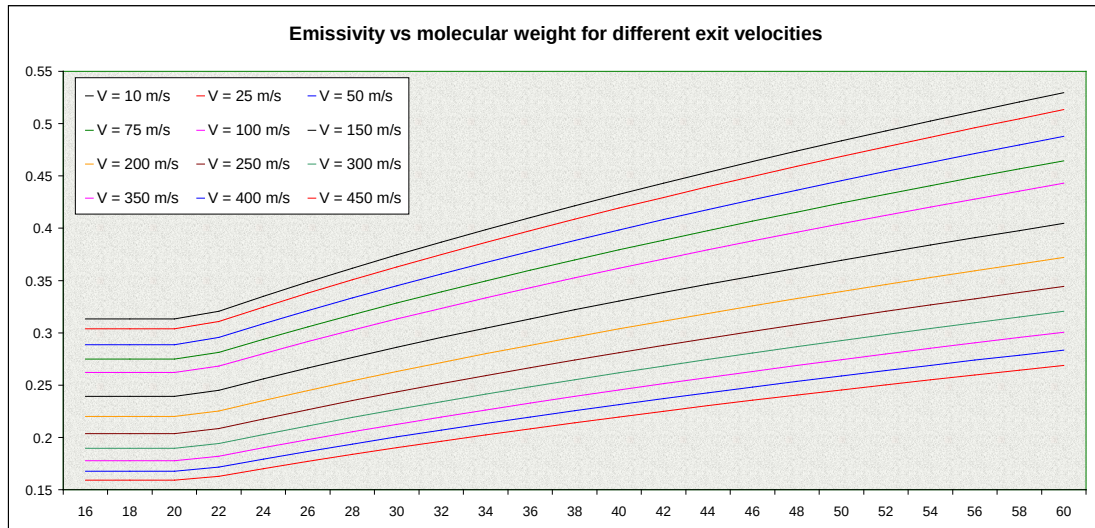
- For pipe flare:
 - Natural gas molecular weight of 18: 0.21
 - Natural gas molecular weight of 21: 0.23
 - Ethane: 0.25
 - Propane: 0.30
 - Refer also to [ISO 23251 / API STD 521](#).

- For sonic flare:

The emissivity coefficient shall be taken equal to 0.15 when the liquid carry over is assumed not to exceed 5% weight (that is the general case). For higher liquid carry over, the emissivity coefficient must be increased. For gas with no carry over, the minimum emissivity coefficient shall not be below 0.13.

The radiation shall be determined for both the full flow and turndown flow rates. The emissivity increase with the flow rate decrease should be obtained from the flare tip supplier. As a first approach, it can be read on Figure 3 (Modified-Chamberlain correlation):

Figure 3 Relief gas emissivity versus MW and exit velocity



12.6 Flare and cold vent line sizing criteria

12.6.1 Design flowrate

12.6.1.1 PSV inlet and outlet lines

The PSV rated capacity (maximum flowrate through the installed orifice with the lowest possible superimposed back-pressure) shall be taken as the design flowrate of PSV inlet and outlet lines, even if this figure exceeds the actual maximum flow rate expected from the process unit.

12.6.1.2 Flare and cold vent headers

The sizing of headers which collect discharges from several sources shall be conducted as per the scenarios defined in [GS EP SAF 262](#) along with the corresponding back pressure. The design basis should be the sum of the required relieving capacities of the PSV's (design rates specified in the PSV Process Data Sheet). Some relief scenarios with low back pressure may prevail over maximum relief case with high back pressure.

12.6.2 Lines upstream relieving capacities

- PSV's:

The design of these lines shall comply with the following:

- ΔP between the protected equipment and the PSV < 3% of PSV set pressure ([API RP 520 PT II](#)).
- Diameter of line $\geq \varnothing$ PSV inlet.
- $\rho V^2 \leq 25,000$ Pa for $\varnothing$ of line $\leq 2"$.
- $\rho V^2 \leq 30,000$ Pa for $P \leq 50$ bar g.
- $\rho V^2 \leq 50,000$ Pa for $P > 50$ bar g.

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- Depressurization device (BDV's + RO):
 - Minimum line size 2"
 - ρV^2 criteria are the same as for PSV's

12.6.3 Lines downstream relieving capacities, flare and cold vent headers and sub-headers

The design of these lines shall comply with the following:

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

12.6.3.1 Gas service - dynamic pressure and Mach number limitations

- Intermittent flow (peak short duration/infrequent flow)
 - Lines downstream relieving devices and sub-headers: 0.7 Mach maximum and $\rho V^2 < 150,000$ Pa
 - Headers: 0.7 Mach maximum and $\rho V^2 < 150,000$ Pa; however a velocity of 0.8 Mach may be accepted for a long straight line without elbows and connections (e.g. stack, line on bridge)

For ρV^2 values greater than 100,000 Pa, vibration analysis and line support studies shall be conducted and the results thereof examined.

- Continuous flow (flow of longer duration or more frequent)
 - Mach < 0.35 and $\rho V^2 \leq 50,000$ Pa

12.6.3.2 Multiphase fluid - dynamic pressure and Mach number limitations

For multiphase fluids (2 phase flow at the inlet of relieving device), the following line sizing criteria shall be used:

$$\text{Mach} < 0.25 \text{ and } \rho V^2 \leq 50,000 \text{ Pa for intermittent or continuous flow}$$

For ρ_m and V_m definitions refer to section 12.4.

12.7 Flare knock-out drum sizing

For flare drum and cold vent drum, the sizing criteria shall be based on [ISO 23251 / API STD 521](#) except otherwise specified.

Gas side

The drum is designed so as the gas velocity inside the drum enables to retain all the liquid droplets with a diameter greater than or equal to:

- 600 microns for remote flares or flares on not-normally manned platforms.
- 300 microns otherwise.

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The KO drum is sized to handle the peak gas rate assuming that the liquid level is at LSHH. Internals on gas side (e.g. mist eliminators) in the flare drums are forbidden.

Liquid side

The liquid side shall be based on the sum of 90 seconds of full flow production from all the sources involved (wells, trunklines in the case of a riser platform receiving remote wellhead effluent, etc.) and 15 minutes of production of the source producing the largest flow rate. The duration can be prolonged in case longer intervention time is required, but shall not be reduced even if intervention can be performed immediately. Alternative is to protect from flare overflow by installing a high high level safety instrumented system. This additional SIS shall be subject to SIL assignment according to **GS EP SAF 361**.

12.8 Purge gas

The purge gas is provided to prevent:

- 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 might induce a quicker deterioration of the tip.

In the case of cold vent, purge gas is not compulsory provided the piping can withstand the surge pressure caused by an explosion.

Since it is recognised that flare suppliers used to underestimate the amount of purge gas, flare suppliers shall be requested to guarantee the purge gas flow rate to avoid air ingress and burn back.

The purge can be eliminated if a liquid seal is installed near the base of the stack.

The purge gas rate can be greatly reduced if buoyancy or velocity seals are installed right upstream of the tip.

Without seals case

To avoid air ingress into the flare stack, the purge flow rate shall be calculated as follows:

$$Q = 31.25 D^{3.46} K$$

where:

Q is the purge flow rate in normal cubic meter per hour,

D is the flare stack diameter in meters,

K is a parameter depending on the gas composition to be calculated as follows:

$$K = \sum_{i=1}^n C_i^{0.65} k_i$$

where

C_i is the volume fraction of component i,

k_i is a constant depending on the component i. The following give typical values:

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- hydrogen : $k_i = 5.783$
- helium : $k_i = 5.078$
- methane : $k_i = 2.328$
- nitrogen (no wind) : $k_i = 1.067$
- nitrogen (with a wind speed about 7 m/s) : $k_i = 1.707$
- ethane : $k_i = -1.067$
- propane : $k_i = -2.651$
- CO_2 : $k_i = -2.651$
- C_4^+ : $k_i = -6.586$

The purge gas flow rate to avoid air ingress into the stack is usually too low to avoid back burning of the tip which is known to reduce the service life of the tip and as such should to be avoided as much as possible. The purge gas rate to avoid burn-back is indeed much greater and depends somewhat upon the tip itself: it shall therefore be requested from the tip supplier.

As a first approach, the corresponding burn-back velocity U_b (m/s) can be estimated using the following formula:

$$U_b = \frac{\left(\frac{D}{MW}\right)^{0.75} \sqrt{LHV}}{15.101}$$

where:

D (m) is the flare stack diameter,

MW (g/mol) is the molecular weight of the purge gas,

LHV (MJ/kg) is the purge gas Lower Heating Value.

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 on a case by case basis with the flare tip supplier).

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With seals case

Buoyancy seals can help reduce the purge rate needed to avoid air ingress into the stack to values as low as 0.003 m/s.

Velocity seals can help reduce the purge rate needed to avoid air ingress into the stack to values in the range (0.006 m/s and 0.012 m/s).

12.9 Installation

12.9.1 Headers and sub-headers

The flare knock-out drum shall be the only low point of the flare system. 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 shall not be installed on flare headers/sub-headers. Block valves shall be avoided. Block valves downstream relieving devices (if required) shall be full bore ball valves and shall be safely blocked in open position. Downstream valves should be installed flange to flange with PSV unless built up back pressure is excessive. Other types of valves shall not be used.

Flow orifice or flowmeter, which can be blocked by foreign matters, shall not be installed on flare headers. For flare flow metering, only the following types of instrument can be used:

- Averaging pitot tube type flowmeter (annubar)
- Thermal Mass flowmeter
- Ultrasonic flowmeter

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 including 4" nominal diameter.

12.9.2 Relief valve arrangement

Refer to:

- Appendix 1: PSV's installation with one upstream isolating valve
- Appendix 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 upstream and downstream block valves should be installed to avoid purging and inerting the protected equipment and the flare lines for the relief valve maintenance. The spare relief valve shall be stored in the workshop (note on P&ID to be added).

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12.9.3 Depressurization valve arrangement

Refer to Appendix 3.

12.9.4 Drainage arrangement

See general specification **GS EP SAF 228**.

12.9.5 Bursting disk

Bursting disks shall be used any time a fast response time is required. For the protection of equipment some components of which shall not be in contact with hydrocarbon fluids (e.g. shell side of heat exchangers), two bursting disks in series shall be used. The arrangement shall provide an indication of bursting disk rupture. This indication will cause valves shutdown on the low pressure side so as to provide containment of the high pressure fluid.

12.9.6 Pilot

Where the flare tip is equipped with several pilots, each pilot shall have its own dedicated feeding line.

Flame front generators are forbidden for the ignition of the tip. Piezoelectric-based system shall be used.

13. Software

The parties in charge of the Process design shall produce a list of the Process software they intend to use for their study, including the software version. In case the selected software is not listed in this section, a description of its capabilities and a relevant reference list shall be provided.

This list shall be approved by Company.

The following softwares are accepted by Company for process related calculations. Contractor shall verify with Company which software version is validated by Company at the time of the study.

13.1 Plant thermodynamic simulations

- PRO II (Invensys) which is Company's preferred option
- Unisim (Honeywell), Hysys (AspenTech)

13.2 Fluid property table generation

- PVTsim (Calsep)

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13.3 Multiphase pipelines calculations

- Steady state:
 - PIPEPHASE (Invensys) (option TACITE or OLGA compulsory)
 - OLGA (SPT)
 - PIPESIM (Schlumberger) (option OLGA or LedaFlowPM compulsory)
 - LedaFlow 1D (Kongsberg)
- Transient conditions
 - OLGA
 - LedaFlow 1D for gas-condensates pipelines; for oil pipelines, the use of LedaFlow 1D shall be first approved by Company.
- Slug flow simulation (for oil pipelines only, deemed inappropriate for gas-condensates pipelines):
 - OLGA Slug-Tracking

13.4 Surge protection on liquid systems and water hammer calculations

- PIPENET software from Sunrise (UK)
- FLOWMASTER software from Flowmaster (France)

13.5 Heat exchangers thermal calculations

- Shell and tube:
 - HTRI (HTRI)
- Air coolers
 - HTRI
 - HTFS (AspenTech)

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13.6 Depressurization calculations (excluding pipelines)

One of the following softwares shall be used for simulating the depressurization of any capacity.

Table 11 Conditions of use of depressurization simulation softwares

Software	One vessel	A group of vessels (1)	Margin on calculated flowrate	Margin on calculated wall temperature
BLOWDOWN (Imperial College)	Yes	Yes	0%	3°C
LNGDyn (Technip)	Yes	Yes		
DynSim (Invensys)	Yes (2)	Yes (2)	10%	15°C for gas-condensates fluids 10°C for oil fluids
Hysys steady-state (AspenTech)	Yes (3)	No		
ProII steady-state (Invensys)	Yes (4)	No		

Notes:

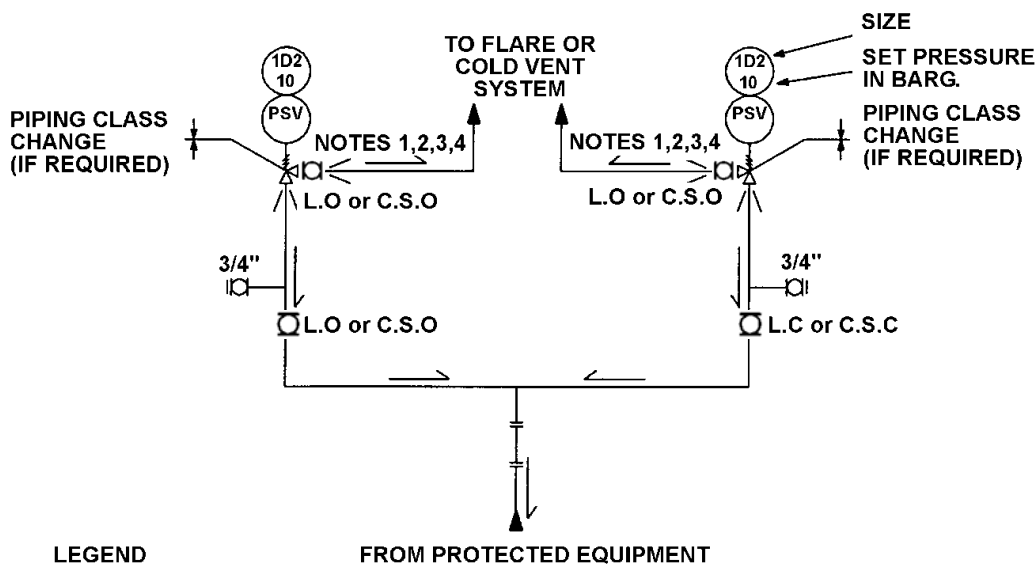
- (1) For instance, a fractioning system including the distillation column, the reboiler, the condenser, the reflux drum etc.
- (2) DynSim can be used with the following parameters:
 - a. Gas Blowdown as heat transfer model
 - b. Internal phase option = vapour-liquid
 - c. External phases option = two phases
- (3) Hysys latest version validated by Company can be used with the following parameters:
 - a. SRK method for EOS and enthalpy computations
 - b. recycle efficiency = 100% for all phases
 - c. Solver option 'Close Component and Mass Balance' active
 - d. PV work term contribution = 100%
 - e. Masolenian equation for the valve characteristics
 - f. Time step = 0.01 second
- (4) ProII latest version validated by Company can be used with the following parameters:
 - a. Masolenian / User model for the valve characteristics
 - b. Rigorous Blowdown as heat flow model for the cold case (non-fire case)
 - c. Isentropic efficiency of expansion = 1 (default value for Rigorous Blowdown)

For cases where the predicted wall temperature is less than the minimum mechanical design temperature of the selected material plus 5°C, and if BLOWDOWN or LNGDYN softwares have not been chosen in the first place, a second simulation shall be run with BLOWDOWN or LNGDYN.

13.7 Flare system

- Radiation:
 - FLARESIM (Softbits)(mixed method with 25 points minimum shall be used, wind chill effect to be ignored)
- flare network design
 - FLARENET (AspenTech)
 - VISUALFLOW (Invensys)

Appendix 1 PSV's installation with one upstream isolating valve



LEGEND

	FULL BORE 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

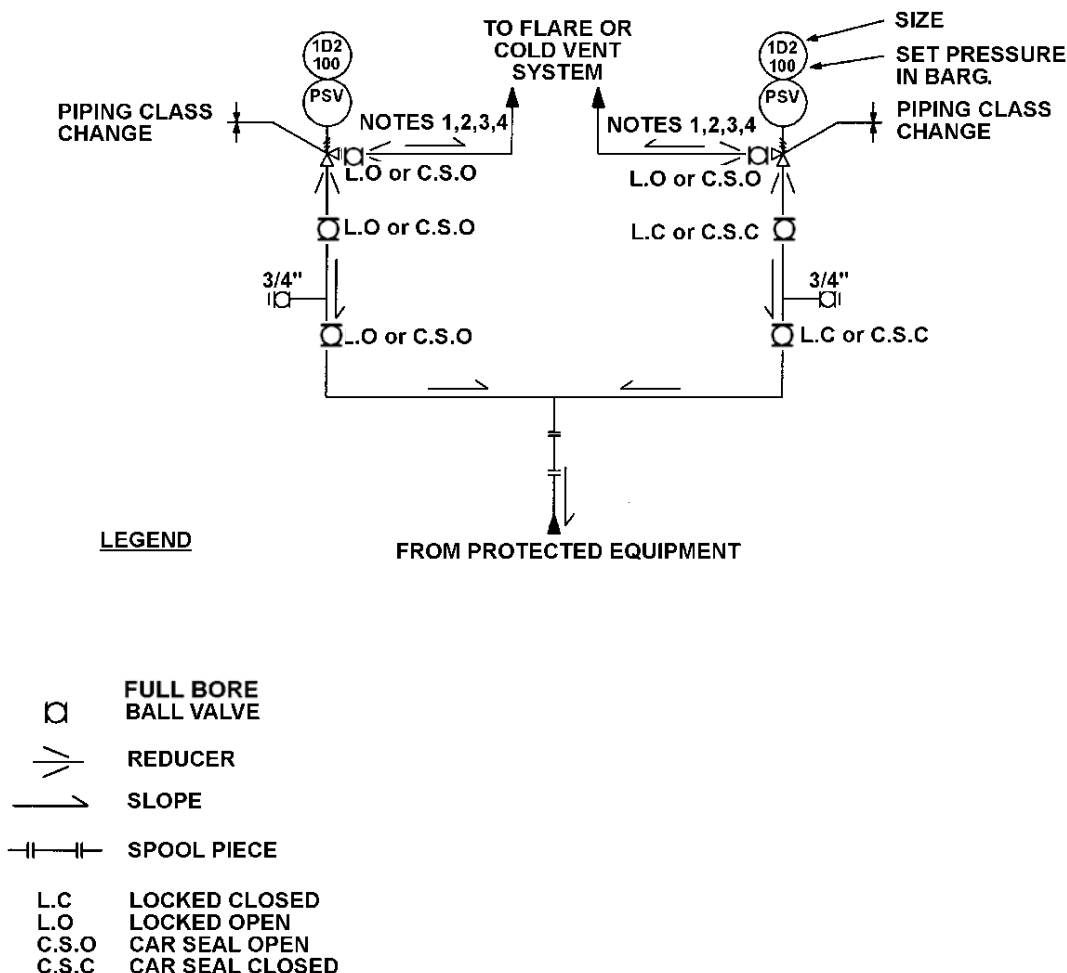
Notes:

- Distance between PSV and reducer is minimum.
- 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.
- Installation of elbow between PSV and reducer is forbidden.
- Piping class change could be required downstream PSV, depending of relief pressure and PSV size.

Appendix 2 PSV's installation with two upstream isolating valves

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

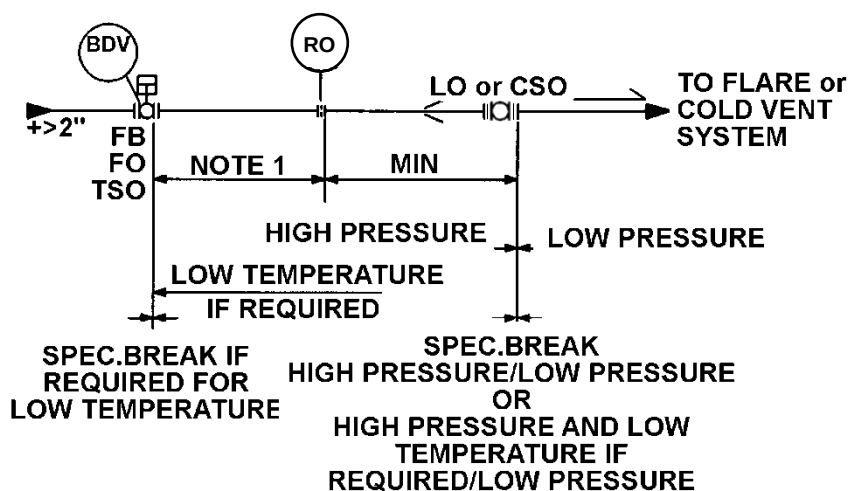
- Operating pressure above 70 barg.
- Operating pressure above 35 barg, only for LNG.
- H₂S partial pressure > 1 bar.
- Fluid very corrosive and abrasive.



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 3 Depressurization valve arrangement



LEGEND

	FULL BORE BALL VALVE
	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