PROCESS SIZING CRITERIA

5	17/11/09	Approved for Construction	OIEC	AS	MA	MMF	AAP
4	14/07/09	Approved for design	OIEC	AS	MA	MMF	AAP
3	03/12/08	Approved for design	WP	SH	AI	RPW	RC
2	22/01/08	Approved for design	WP	AIR	RPW	RPW	MS
1	04/06/07	Approved for design	WP	AIR	RPW	RPW	MS
0	30/03/07	Issued for client review	WP	AIR	RPW	RPW	MS
REV.	**DATE**	**DESCRIPTION**	**ORIG.**	**PRPD**	**CHKD**	**APP'D**	**AUTH'D**

TABULATION OF REVISED PAGES

Page	Rev. 1	Rev. 2	Rev. 3	Rev. 4	Rev. 5
1					
2					
3					
4					
5	X	X		X	
6	X			X	
7					
8		X			
9	X				
10		X			
11					
12					
13	X				
14	X			X	
15	X				
16		X			
17	X	X			
18	X	X			
19					
20	X	X		X	
21	X		X	X	
22	X			X	
23	X	X		X	
24		X			
25					
26					
27					
28		X			
29	X	X			
30					
31					

Page	Rev. 1	Rev.2	Rev. 3	Rev. 4	Rev. 5
32					
33		X		X	
34					
35					
36		X		X	
37		X		X	
38	X			X	
39				X	
40					
41				X	
42	X	X		X	
43					
44	X	X			
45					
46					
47	X	X			
48	X	X		X	
49	X				
50	X				
51		X			
52		X			
53	X	X			
54					
55					
56					
57					
58		X			
59					
60		X			
61	X			X	
62	X				

CONTENTS

1. DOCUMENT SCOPE

This document defines the main principles to be considered for the design and the implementation of the Process sizing/design criteria to be applied as a minimum in the equipment specification for both FEED and Detailed design of ONSHORE FACILITIES of SOUTH PARS GAS FIELD DEVELOPMENT (Phase 17 & 18) Project which are located at ASSALUYEH, IRAN. In the event of a conflict between this specification and any of the Licensor sizing/design criteria, the Licensors shall prevail.

The process sizing criteria shall apply to new onshore installations and to any modifications or extension to existing facilities. It is also applicable to vendor's packages.

The purpose of this document is to:

- Establish and correlate a source of basic information of typically multidiscipline interests for easy access and common reference.
- Agree and record minimum basic sizing criteria.
- Explain and clarify the options selected for equipment design during the BASIC stage.
- Define the minimum requirements for completion of missing items by the EPC contractor.
- No change will be made to the data given by the Licensor without agreement of Licensor/POGC.

2. INTRODUCTION

The IDRO/OIEC/IOEC Consortium has been contracted by the National Iranian Oil Company (NIOC) to conduct the EPCC for SOUTH PARS Phases 17 & 18 Project. This Project includes Offshore Facilities, platform and undersea pipelines, and Onshore Facilities for the processing of the reservoir fluid.

WorleyParsons Pty Ltd has in turn been contracted by the IDRO/OIEC Consortium to conduct the Basic Engineering and FEED for SOUTH PARS Phases 17 & 18 for the Onshore facilities.

The South Pars Phases 17 & 18 Project offshore work-scope is being carried out by a separate Offshore Consultant. WorleyParsons will lead, identify and coordinate deliverables required to manage the physical interfaces between the Onshore and Offshore Design Consultants.

3. PROJECT OVERVIEW

The Iranian South Pars field is the largest discovered offshore gas field in the world, located 100 km offshore in the Persian Gulf. Reservoir fluids are transported to shore via two sealines to the mainland (ASSALUYEH) at a distance of 105 km. approximately, for further treatment.

South Pars Phases 17 & 18 are a further development of the South Pars gas field consisting essentially of two 1000 MMSCFD phases.

The facilities shall be developed on the basis of supplying treated lean gas to the domestic gas network and ethane gas to the nearby petrochemical complex at the required specifications while maximising liquid recovery as propane and butane and stabilised hydrocarbon condensate for export.

A new interconnecting pipeline shall transport the lean gas produced to the domestic gas distribution network via fiscal metering facilities.

The produced condensate shall be stored in dedicated storage facilities, periodically pumped through a condensate export pipeline to a SBM via fiscal metering facilities.

The treated propane and butane shall be stored separately in refrigerated atmospheric double wall storage tanks, periodically pumped through a dedicated loading pipelines network to LPG carriers berthed at a jetty via fiscal metering facilities.

3.1 Plant Description

The main functions of the Onshore processing facilities are:

- Reception facilities for the HP fluids and separation of the raw gas and condensate/water/MEG mixture.
- Condensate stabilisation (and desulphurisation optional) producing stabilised condensate for storage and export. The light ends are recycled to the HP gas system. One condensate flashing unit, normally not operated, is provided as a back-up to the stabilisation facilities.
- Gas treatment facilities producing sales gas, gaseous ethane and NGL's consisting of:

    1. H_2S/CO2 removal from gas [deleted]

Dehydration unit, using molecular sieves technology

Mercury guard

Ethane extraction unit, producing sales gas, gaseous ethane and NGL

- Ethane treatment for CO_2 and COS removal [deleted] for further export
- NGL fractionation facilities to produce sour liquid propane, butane and condensate. This condensate is routed to the condensate stabilisation system for treatment.
- Propane and butane treatment for mercaptans removal and drying for storage and export
- Propane and butane separate atmospheric storage tanks, double wall, and full containment, refrigerated with reinforced concrete external walls.

- MEG regeneration and injection unit.
- Sulphur recovery producing liquid sulphur for solidification, granulation and export
- Utilities, offsites as required for operation.
- Gas export compression

4. CODES AND STANDARDS

Instrumentation and Systems will be designed and fabricated in accordance with engineering codes and standards listed in document:

DB-1718-999-P332-0203: List of Applicable Codes & Industry Standards

5. REFERENCE DOCUMENTS

This document is complemented by the following documents:

- RP-1718-999-6300-0002: Metallurgical Requirements for Welding of Carbon Steel in Sour Wet Service
- RP-1718-999-6300-0002: Metallurgical Requirements of Carbon Steel in Sour Service
- RP-1718-171-6300-0002: Corrosion & Material Selection Review Onshore Facilities, Process & Utilities
- DB-1718-140-P312-100: Flares & Blowdown Design Basis
- DB-1718-999-P312-203: Design Basis for P&ID Development

6. ABBREVIATIONS

AFC	Approved for Construction
AFD	Approved for Design
bbl	Barrels
BFW	Boiler Feed Water
CAD	Computer Aided Design
CAPEX	Capital Expenditure
CGR	Condensate Gas Ratio

CS	Carbon Steel
DCS	Distributed Control System
DP	Design Pressure
EPCC	Engineering, Procurement, Construction and Commissioning
FEED	Front End Engineering Design
FSLL	Flow Switch Low Low
HAZOP	Hazard and Operability
HETP	Height Equivalent to a Theoretical Plate
HIC	Hydrogen Induced Cracking
HP	High Pressure
HSE	Health, Safety & Environment
ILAH	Interface Level Alarm High
ILAL	Interface Level Alarm Low
ILALL	Interface Level Alarm Low Low
INLL	Interface Normal Liquid Level
KO	Knock-Out
LAH	Level Alarm High
LAHH	Level Alarm High High
LAL	Level Alarm Low
LALL	Level Alarm Low Low
LG	Level Gauge
LPG	Liquefied Petroleum Gas
LSHH	Level Switch High High
LSLL	Level Switch Low Low
LTCS	Low Temperature Carbon Steel
MEG	Mono Ethylene Glycol

MFD	Material Flow Diagram
MMSCFD	Million standard cubic feet per day
MSOT	Minimum Service Operating Temperature
NGL	Natural Gas Liquids
NLL	Normal Liquid Level
NPSH	Net Positive Suction Head
NSOP	Normal Service Operating Pressure
NSOT	Normal Service Operating Temperature
PAH	Pressure Alarm High
PAHH	Pressure Alarm High High
PFD	Process Flow Diagram
P&ID	Piping & Instrumentation Diagram
PSHH	Pressure Switch High High
PSV	Pressure Safety Valve
SBM	Single Buoy Mooring
SDV	Shutdown Valve
SS	Stainless Steel
SSCC	Sulphide Stress Corrosion Cracking
TEP	Total Exploration and Production
UFD	Utility Flow Diagram
WP	WorleyParsons Services Pty Ltd

7. DEFINITIONS

Within this specification the following definitions shall apply:

COMPANY: Shall mean NATIONAL IRANIAN OIL COMPANY (NIOC - POGC), or their representative.

CONTRACTOR: Shall mean the Consortium of IDRO, OIEC and IOEC in charge of EPCC realisation of SOUTH PARS Phases 17 & 18 in the CONTRACT with COMPANY, or his representative.

CONSTRUCTION CONTRACTOR: Shall mean the party sub-contracted by the CONTRACTOR for construction WORK, or his representative.

WORK: Shall mean all and any of the works and / or services and/or materials required to be provided by the CONTRACTOR or CONSTRUCTION CONTRACTOR.

PLANT: Shall mean permanent facilities designed, constructed and completed as a result of execution of the WORK.

SITE: Shall mean the premises and places on, under, in, over or through which the WORK is to be executed or carried out including CONTRACTOR's engineering office and the PLANT.

VENDOR: Shall mean any person, firm or business which manufacture or supply materials, equipment or services for the performance of any item of WORK.

PROCESS: Discipline(s) in charge to study the process units, but also associated off sites and utility fluids.

PFD (PROCESS FLOW DIAGRAM): It presents the succession of the operations in the fluid processing to reach the required products specifications set in the objective of the plant processing including associated off-sites and utility fluids. The succession of operations is pictorially represented by equipment symbols and all lines for a good understanding. The controls are indicated by simplified symbolic representation of control loops. In the upper part of the sheet, the equipment reference shall indicate the function of each piece of equipment.

The operating pressure, temperature, flow rate and heat duty values are indicated in Material Balance documents and cross referenced on PFDs by means of fluid number inside a diamond. Operating pressure and temperature in the appropriate shapes are also indicated on PFD.

PROCESS DATA SHEET: It is the data sheet for equipment, packages, instruments, etc., containing the process information required for sizing and given sometime the sketch with sizes for equipment shall be issued by process.

It is different from the mechanical data sheet which defines additional information for construction and it is issued by specialists and/or vendors.

Instrumentation Process Datasheets shall also be issued to the relevant specialist group.

HEAT AND MATERIAL BALANCE: This is the document giving, for each process case, the normal operating conditions, compositions and the thermodynamic and physical properties for all streams.

MAXIMUM OPERATING CONDITIONS: These are the maximum pressure, temperature, flow rate, etc., in the equipment when the plant operates at the top of its normal operating range, conditions corresponding to the high alarm set point.

MINIMUM OPERATING CONDITIONS: These are the minimum pressure, temperature, flow rate, etc., in the equipment when the plant operates at the bottom of its normal operating range, conditions corresponding to the low alarm set point.

MINIMUM SERVICE OPERATING TEMPERATURE (MSOT): This is the minimum temperature obtained during normal operation, start-up, shutdown or process upset.

NORMAL OPERATING CONDITIONS: These are the pressure, temperature, flowrate, etc., in the equipment when the plant operates at steady state conditions corresponding to the values of the heat and material balance.

NORMAL SERVICE OPERATING PRESSURE (NSOP): This is the highest (for pressure above the atmospheric pressure) or the lowest (for pressures 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): This 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): This is the detailed drawing used for assistance of construction and operation of the processing plant including associated offsites and utility fluids. P&ID symbols are required for a good understanding and also provides symbolic representations on P&IDs.

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.

8. MECHANICAL DESIGN CONDITIONS

8.1 Design Pressures

8.1.1 General

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

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

Maximum operating pressure (barg) MOP	Minimum design pressure (barg)
< 1	2 or 3.5 (*) minimum
1 -10	MOP + 1 bar (**)
> 10	MOP + 10%

(*) 2.0 barg for PSV discharging to atmosphere, 3.5 barg for PSV discharging to flare network.

(**) 2 or 3.5 barg as minimum design pressure considering the criteria in Part (*).

For the minimum design pressure it shall be considered that:

- Unless otherwise noted, the design pressure specified by Process applies to the vapour phase at the top of the vessel.
- Minimum design pressure is not applicable for thin wall equipment such as silos and storage tanks. In that case the governing parameter is full of water or liquid/material content if specific gravity > 1. For floating roof tanks, the distributed weight of the floating roof is to be included in the design pressure of the tank.
- The design pressure shall also account for upset or transient conditions such as start-up, pressure surge, settle-out pressure at compressor suction, etc.
- Vapour pressure at design temperature should be considered as design pressure except when safety relief valves are provided.
- Equipment that operates at pressure below atmospheric pressure shall also be designed for full vacuum.

- Equipment that could face vacuum under abnormal conditions such as:
  - Vacuum conditions during start-up, shut down and/or regeneration purges
  - Normally operated full of liquid but can be blocked in and cooled down
  - Containing condensable vapour but can be blocked in and cooled down (especially equipment which is subjected to steam out).
  - Could undergo a vacuum condition through the loss of heat input
  - Loss of artificial gas blanketing on vessels containing liquids in a vessel containing liquids with a vapour pressure less than atmospheric at the minimum storage temperature
  - Product storage tanks and vessels where net output is possible (e.g. Unloading).

They shall be treated on a case by case basis and be designed for full vacuum unless fully reliable protection devices are provided (vacuum breaker, pressurisation gas, low pressure switch etc.)

- For equipment in equilibrium with (‘riding on the’) flare, the design pressure of the equipment is at least the maximum flare back pressure at any point of the flare system or the flare design pressure, whichever is higher.
- Hydraulic pressure due to the relative elevation between equipment and also the PSV’s location shall be considered.
- The design pressure of storage tanks shall be individually assessed. The vapour pressure, tank venting, purging, and relieving systems described in API 2000 shall all be considered before determining the maximum operating pressure of the tank. Tank design pressures may be selected from the recommendations in the appropriate design codes, e.g. those given in API 620 and 650, or BS 2654.
- Special equipment – it should be noted that some items of equipment, e.g. glass lined vessels, carbon block exchangers etc., may have design difference between the two sides of the unit, rather than maximum system design pressure.
- When the design pressure for an item, which is remote from the source of pressure, is being determined, it may be necessary to consider the influence of the pressure drop through the circuit, on the design pressures of all items in the circuit.
- Propane refrigeration systems shall have a minimum design pressure of 18 barg, corresponding to propane vapour pressure at 55°C.

8.1.2 PSV Setting

Refer to Figure 1- Pressure Level Relationships for Pressure Relief Valves in the 7th edition (2000) of API RP 520 Part 1 for tolerances on PSV setting.

In case of absolute necessity (for example in case of high pressure), the use of pilot operated PSV (tolerance of +/- 1% on set point) could help to reduce the design pressure.

- If two or more PSVs are in service, the set pressure shall be staggered to avoid chattering. The difference between set points shall be less than 5% of the design pressure.
- A margin of 10% is generally adequate in order to ensure protection of the equipment with one conventional pressure relief valve (PSV), one pressure alarm (PAH) and one pressure switch (PSHH).
- The following tolerances are generally admitted for conventional instrumentation:
  - PSV opening: +/- 3%
  - PSV closing: +/- 5%
  - PSV recommended leak test: 10% below set point
  - Pressure switch and transmitter: +/- 1%
- In case of PSV installed for Fire case protection only, the margin will be 21%.

8.1.3 Pipelines

For pipelines, the design pressure is a function of the Maximum Allowable Operating Pressure (MAOP) and the design factor which depends on the class location.

Process determines only the MAOP.

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

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

8.1.4 Particular Cases

The Design Pressure (DP) of the Equipment is as follows:

8.1.4.1 Compressors

- At the reciprocating compressor discharge:

  DP = MOP + 2 bar for MOP ≤ 20 bar g,

  DP = MOP + 10% for MOP > 20 bar g.

PSVs are required.

- At the discharge of the centrifugal compressor:

  DP = MOP + 1 bar for MOP ≤ 10 bar g,

  DP = MOP + 10% for MOP > 10 bar g.

Generally surge pressure is above design pressure and PSVs are required.

Consideration shall be given to 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 compressor system when the compressor is stopped or shut down. Generally the design pressure of the equipment and piping at compressor suction shall be above this settle-out pressure in order to avoid unnecessary lifting of PSVs. For variable speed compressors, the maximum discharge pressure shall be calculated from the performance curve at the maximum trip speed setting prior to arriving at design pressure considerations.

8.1.4.2 Pumps

a- Centrifugal pumps

- Generally no PSVs 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 :

$$P_d = P_{s\max} + \frac{1.2 \cdot head \cdot d_{\max}}{10.2}$$

P_d = design pressure at pump discharge (bar g)

$P_{s\ max}$. = design pressure of suction drum + static head at d_{max} and at High Liquid Alarm Level (barg)

head = head of the pump at design point (m)

d_{max} = maximum specific gravity of liquid pumped under normal operating conditions

Note:

(1) 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.

(2) Equipment elevation (from TL to grade shown on P&ID and data sheet) shall be considered as typical and shall be minimised when NPSH available / required are known.

b- Positive displacement pumps

- At discharge of positive displacement pumps:

  DP = MOP + 1 bar for MOP ≤ 10 bar g,

  DP = MOP + 10% for MOP > 10 bar g.

PSVs are required.

In case of two pumps in series, the maximum differential head shall be the sum of the maximum differential head of each pump if there is no pressure relief valve between the pumps.

8.1.5 Heat Exchangers

8.1.5.1 Heat exchangers design pressure

Maximum Operating Pressure (barg)	Minimum Design Pressure (barg)
0 < MOP ≤ 1	3.5
1 < MOP ≤ 3.5	5
3.5 < MOP ≤ 17	MOP + 2
17 < MOP ≤ 70	MOP x 1.1
70 < MOP ≤ 140	MOP + 7
140 < MOP	MOP x 1.05

This concerns TEMA and multitube heat exchangers.

According to § 5.19.2 of API 521 (5th Edition, 2007) : " Loss of containment of the low pressure side to atmosphere, is unlikely to result from a tube rupture where the pressure in the low pressure side (including upstream and downstream systems) during the tube rupture does not exceed the corrected hydrotest pressure."

It should also be noted that:

"Pressure relief for tube rupture is not required where the low pressure exchanger side (including upstream and downstream systems) does not exceed the criteria noted above."

The corrected hydrotest pressure is defined as:

"Hydrostatic test pressure multiplied by the ratio of stress value at design temperature to stress value at test temperature."

This latest edition of API 521 is based on the vessel code ASME section VIII Div 1 dated January 2004.

The recommended practice consists of oversetting, if necessary, the design pressure of the low pressure side of heat exchanger:

- In all cases up to limit of 150#
- After analysis, case by case, for higher pressure

This practice applies only to the heat exchanger itself and does not concern relevant piping and valving.

According to § 5.19.6 of API 521 (5^{th} Edition, 2007), since double pipe type heat exchangers are considered a piping item, they are excluded.

As an alternative to relief valve installation the design should consider the capacity of the shell side piping and downstream unit to accept the tube rupture case.

8.1.5.2 Heat exchangers using steam

Full vacuum conditions shall be added to design conditions since vacuum can happen during cooling of such equipment (when not connected to atmosphere) unless fully reliable protective devices are provided (vacuum breaker, pressurisation gas, low pressure switch).

8.1.6 Columns

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

The design pressure at the bottom of a column (vapour phase) is determined by adding the column pressure drop to the column overhead design pressure.

Liquid density and maximum liquid height in the bottom shall be specified on the process data sheet to allow the vessel designer to calculate the bottom thickness. Special attention shall be paid to the case when the hydrostatic test is to be done in the vertical position, e.g. a field test, as the tower will be filled with water. A water column equal to column height shall be considered when calculating vessel thickness.

8.1.7 Tanks

Design Pressure of tanks (and all vertical vessels) is based on the top of the vessel.

Atmospheric tanks shall be designed full of water or full of product if 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.

8.2 Design Temperatures

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

In all cases, design temperatures of all equipment shall be quoted as Maximum design temperature / Minimum design temperature.

The design temperature shall be specified as follows; with due consideration for the special cases discussed in section 8.2.4.

8.2.1 Equipment operating above 0°C

Max. design temperature = max. operating temperature + 15°C or maximum exceptional operating temperature, whichever is the greater.

Note: exceptional operating temperature shall be considered for operations exceeding a total of 100 hours per year.

Maximum design temperature for equipment exposed to solar radiation shall be at least 85°C. This value should be examined case by case for equipment on which dilatation problems can occur (such as doublewall tank, fixed tubesheet, plate heat exchanger) and for insulated high pressure vessels (not to increase wall thickness).

For those equipment items not exposed to solar radiation the maximum design temperature for all equipment shall be at least 55°C. This is the maximum estimated temperature that can be achieved in insulated equipment and equipment shaded from the sun after prolonged shutdown.

8.2.2 Equipment operating below 0°C

- As a general rule the minimum design temperature shall be:

TD = TSM - 5°C, or minimum ambient temperature, whichever is lower.

Where:

TD: Minimum Design temperature (°C)

TSM : Minimum continuous operating temperature (°C), however see Note (1) below.

- Notes:

(1) When applicable, the exceptional temperature generated by depressurisation of equipment or interconnected items system shall be indicated as well as the related residual pressure. Depending upon the depressurisation philosophy of the plant, dynamic simulations of equipment have to be performed using commercial or in-house software so as to determine the relevant pressure and temperature of depressuring conditions (see Section 12, Software).

(2) Temperature associated with a gas blow by from one equipment item shall be considered for the buried drum (belonging to the drainage system) design temperature.

(3) When applicable an exceptional 'hot' design conditions set (e.g. temperature and related design pressure) may be added to the 'cold' design conditions set for equipment operating at low temperature. (an exceptional condition may be steam out etc.).

8.2.3 Discontinuous /Cycled Processes

- Various pressure and temperature conditions shall be specified for each phase of equipment operation.
- Mixing of extreme (and non-coincidental) conditions of pressure and temperature shall not be considered.
- A typical example is related to the molecular sieve vessel design conditions specification. The sets of conditions for each phase of the operating cycle shall be specified, e.g. Design conditions of the adsorption phase, of the regeneration phase, etc.

8.2.4 Special Cases

8.2.4.1 Emergency Depressurising

- The minimum design temperature must take into account any depressurisation and repressurisation (depending on material selection) of the equipment / piping that may occur either during an emergency or shutdown situation or gas blow-by from one equipment item to another equipment item and to the possible consequence of a change of material.
- The emergency depressurising shall impact the material selection as follows:

  - Piping material

    Piping material shall be selected taking into account the minimum temperature encountered during depressurisation. Piping repressurisation shall be considered as to be performed with the minimum depressurisation temperature.

  - Vessel material

The minimum temperature due to the blowdown conditions shall be associated with design pressure. Although depressurisation of any section of the plant cannot be performed unless the section is isolated and permission is obtained, repressurisation may take place by operator's error or a valve failure, therefore the minimum temperature shall be associated with design pressure. Credit for special devices to ensure that the plant shall remain isolated and depressurised shall not be considered.

In addition the above criteria shall ensure safe operation in case of residual piping stress being present (in particular for small diameter nozzle/piping).

8.2.4.2 Heat exchanger and air cooler

The following conditions mentioned hereafter shall be applied generally up to the next piece of process equipment:

- Consideration for design temperature definition shall 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. Where possible detailed thermal analysis of natural convection cooling under the worst design ambient conditions shall be performed to arrive at the maximum cooler outlet temperature.
- For the bypassed air cooler, the design temperature of the downstream equipment, if any, shall be the maximum upstream operating temperature of the bypassed exchanger.
- Downstream of other coolers, the design temperature shall be the upstream maximum operating temperature.
- Fixed tubesheet exchangers shall not be used and other designs shall be considered if the difference between the average shell metal temperature and the average tube metal temperature in any tube pass exceeds 28°C.

8.2.4.3 Cleaning and steam out

- Consideration to be given to conditions and fluids used for cleaning (e.g. Steam). In this case, both pressure and temperature conditions shall be provided.
- The steam-out conditions for vessels are 150°C @ atmospheric pressure. This information shall be considered by the Mechanical Department but shall be also specified on the Process Data sheet as an additional note where steam-out applies.
- Steam-out shall be not applied for Units 104/105/106/107/111/122/145/147/148/ [deleted]. For these units nitrogen purging shall be considered.

- Neither steam-out nor nitrogen purging shall be considered for Units 123/124/125/126/127/128/129/130/131/132/146.
- The requirement for steam-out for Unit 101 shall be developed in consultation with the licensor during detailed design.
- For equipment subjected to steam-out operations, full vacuum condition shall be specified at 150°C.

8.2.4.4 Exceptional cases

- Consideration shall be made for upset and transient conditions such as start-up, shutdown etc. In these cases, both pressure and temperature conditions shall to be specified.
- The exceptional temperature generated by fire shall not be considered for design temperature selection.
- A specific design temperature shall be provided with the specified vacuum design pressure.
- The following considerations shall not prevail on § 8.2.4.2 regarding cooling medium failure.

The accidental temperature which may occur in emergency situations, such as loss of utilities, valve closure, air cooler failure or any abnormal operation lasting for a short duration, is not to be taken into account as long as the temperature increase does not exceed the code's limits (investigation has to be undertaken with specialists on a case by case basis)

However equipment containing parts which can be damaged by an abnormally high temperature shall be designed for this temperature. This mainly concerns equipment internals.

8.3 Material Selection

The following section is provided as a general guideline. Final material selection is the responsibility of and shall be determined by the materials department.

8.3.1 Basis of material selection

- Material selection used for vessel

Design temperature °C (1)	Steel type
-196 ≤ T < -101	SS 304 L
-101 ≤ T < - 46	31/2 Ni / SS 304 L
- 46 ≤ T < 0	LTCS
0 ≤ T < 343	CS

(1) Design temperature corresponding to operating conditions. For temperature due to depressurisation, LTCS might be suitable to use at lower temperature (<-45°C) provided that the

vessel shall naturally reheat before repressurisation. This limit should be determined by the Mechanical Department, depending on vessel wall thickness.

- Material selection used for piping

Design temperature °C	Steel type
-196 ≤ T < -101	SS 304 L
-101 ≤ T < - 46	SS 304 L
- 46 ≤ T < - 29	LTCS
- 29 ≤ T < 343	CS

8.3.2 Corrosion due to H_2S and CO_2 services

8.3.2.1 General considerations

H_2S and CO_2 corrosion results in metal loss and in metal embrittlement (cracking). Sulphide Stress Corrosion Cracking (SSCC) can be limited by using materials listed in the NACE MR0175 / ISO15156.

Hydrogen Induced Cracking (HIC) corrosion is directly linked to the material quality, specifically the presence of elongated inclusions and microsegregations. Therefore HIC control is a matter of material specification.

8.3.2.2 Sour Service Definition

Sour service definition is based on NACE MR0175 / ISO15156 2003 including corrigendum 2005 criteria and is considered as a minimum requirement. The H_2S service definition is to base the maximum pressure criteria, mentioned in the NACE, either on equipment design pressure where no Pressure Switch High High (PSHH) protection is implemented or on PSHH level where existing.

8.3.2.3 Limits of sour service specifications

The limits of sour service specifications RP-1718-999-6300-0002 for wet sour service Sulphide Stress Corrosion Cracking (SSCC) & Hydrogen Induced Cracking (HIC) and RP-1718-999-6300-0003 (wet sour service SSCC only) are as follows:

A. When equipment is classified in sour service under the normal operating conditions of the plant as per definition of NACE / ISO standard, specification RP-1718-999-300-0002 (i.e. SSCC & HIC) shall be applied.

B. For equipment operating under transient sour service conditions only (i.e. upset, start-up,….), the use of specification RP-1718-999-6300-0002 vs. RP-1718-999-6300-0003 shall be based on the expected occurrence of the sour service conditions, as follows:

- For series 600# and above, specification RP-1718-999-6300-0003 shall be applied if transient sour service conditions are expected to represent less than 10% of the total operating time. Otherwise spec RP-1718-999-6300-0002 shall be used.
- For series below 600#, specification RP-1718-999-6300-0003 will normally be applied when the expected duration of the transient sour service conditions represents less than 20% of the operating time. Otherwise spec RP-1718-999-6300-0002 shall be used.

8.3.2.4 Application on process datasheet

The Materials department has developed two job specifications, RP-1718-999-6300-0002 and RP-1718-999-6300-0003.

On process data sheets, a note has to be added on Materials Specification to define which Specification is applicable.

8.3.2.5 Particular cases

- Flare systems:
  - FA system (high pressure flare) shall be “sour service” following spec. RP-1718-999-6300-0002.
  - FS flare (wet MP flare) is out of NACE / ISO conditions (maximum back-pressure is 3.5 bara). The maximum back pressure value is preliminary and shall be confirmed in detailed design to reconfirm whether NACE/ISO conditions apply.
  - FB network operating at very low pressure and with low H_2S partial pressure does not require sour service definition.
- Cladded carbon steel equipment (or weld overlay) in sour service conditions:

  Specifications RP-1718-999-6300-0002 or RP-1718-999-6300-0003 shall not be applied to carbon steel, used as underlining for a cladded vessel in sour service or for nozzles fitted with weld overlay.

8.3.3 Post-weld heat treatment

For caustic soda and amine service of any concentration, a postweld treatment for stress relief shall be specified to avoid corrosion cracking. RP-1718-999-6300-0003 shall be applied for this service.

No post weld heat treatment shall be performed on Austenitic Stainless Steel materials.

8.3.4 Corrosion allowance

	Corrosion allowance on Carbon Steel and low alloy steel (mm)	Corrosion allowance on Stainless Steel, high alloy steel or cladded steel (mm)
Wet H_2S service	3 min. (1)	0
Corrosive process service (except wet H_2S service) (2)	3 min.	0
Non corrosive process services (3) (4)	3 min.	0
Utility (3) (4)	3 min.	0

Notes:

1) Minimum corrosion allowance depending on corrosion control philosophy
2) e.g. amine systems, sulphur recovery units
3) The corrosion allowance applies for pressure vessels, shell and tube type and air fin type heat exchangers
4) For piping, refer to piping classes
5) For storage tanks, corrosion allowance shall be:
   - For fixed roof tank: 1.5 mm for shell and roof except tank bottom where wall thickness is generally imposed by other constraints
   - For floating roof type tank: 0 mm

Protective painting shall be applied on roof and shell (above overflow) where condensation may occur and bottom part of shell where water may accumulate.

For tanks in licensed units, Licensor recommendations shall be followed.

8.4 Lethal service classification for pressure vessels

The pressure vessel code requires classification of the equipment in accordance with lethal service. In order to carry out this classification, the mention of “lethal service” shall be specified on equipment data sheets handling a fluid with H_2S content higher than 1000 ppm wt or where mercaptan level exceeds 100 ppm in any escaping gas.

For rotating machinery lethal service is noted on the process data sheet. The same criteria are used for pressure vessels. The purpose of the note is to assist seal selection and indicate items of equipment which may require toxic gas detection.

9. EQUIPMENT DESIGN CRITERIA

9.1 Pumps

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

- Reflux pump = +20% of normal flow
- Other process pump = +10% of normal flow
- Utility pump = +10% of normal flow
- Export pump from storage to pipeline (continuous operation) = +15% of normal flow
- Loading Pump (to road and rail tankers or marine vessels) = +0% of nominated loading rate.
- Boiler feed water pump = See applicable codes but not less than +10% of normal flow

To be noted that:

- When a non-automatically controlled minimum flow protection has been installed, the permanent recirculation flow (if required) must be added to the net process flow.
- Normal and rated flow shall be identical in such instances as:
  1) Intermittent service pumps: e.g. sump pump.
  2) When the pump has been overrated to allow for a centrifugal type and if overrating is $\geq$ 10%.
  3) Re-circulation flow such as for product loading lines or through amine filtration system.
- Pump automatic start shall generally be done through the Flow Switch Low Low (FSLL) (if flow transmitter already exists) but shall need to be examined on a case by case basis. The determination of automatic start shall be based on consideration of the following guidelines and applicability:
  1) Personnel safety: for example flare knockout drum pump shall be started in order to avoid liquid in flare tips. In that case, considering the discontinuous operation of flare drum pumps, the start of the spared pump can be performed by Level Switch High (LSH) or by Distributed Control System (DCS) logic.

2) Equipment safety: for example Boiler Feed Water (BFW) pump shall be started in order to protect the steam drum and the steam coil.

3) Severe process upset: pumps that can generate a process unit trip or that can generate an off-spec product shall have spares that can be started automatically.

4) Flaring: automatic start shall not be considered to minimise the flaring, for example reflux pumps, unless a severe process upset is faced.

9.1.1 Physical Properties

Physical characteristics of the fluid being pumped are based on the heat material balance verified during the FEED study.

9.1.2 Capacity

Design margins as set out in Section 9.1 shall be applied when setting pump design capacities.

For flows less than about 3.4 m^3/h, a larger pump is usually purchased and a fixed recycle is added to provide for recirculation of excess pumped fluid to the suction supply. An orifice plate is frequently included in the bypass line and this can be supplied by the pump vendor, who shall size the orifice for a minimum flow requirement. The pump requisition sheet shall specify a capacity equal to the desired flow (including extra capacity) plus the amount bypassed.

9.1.3 Suction Calculation

This calculation yields the system pressure available at the pump centreline of horizontal pumps or at the centreline of the suction inlet nozzle for vertical shaft pumps. It involves the summation of the feed vessel's normal operating pressure, the static head loss, the pressure drop in the suction piping resulting from friction, inlet-exit, and other losses.

The static head for vertical vessels is calculated from the bottom tangent line while for horizontal vessels, the bottom invert line is used. Usually, no credit is taken for the head contributed by liquid operating levels in a vessel. This should be reviewed on a case by case basis.

a) Suction line equivalent length (Le)

Equivalent length may be calculated in two ways for the suction lines, either the user inputs straight line length and fitting factor and the Le is calculated by multiplying the two, or the Le is estimated from the pipe diameter d (inches) as follows:

Pumping temperature < 150° C	Le = (8d+30) m
Pumping temperature >= 150° C	Le = (12d+30) m

NB: the estimation excludes an allowance for suction strainer. This shall be included as an additional loss see section 9.1.7. for DP.

b) Centreline elevations of horizontal pumps

The pump centreline elevation is selected from the table below. If the flow exceeds 4540 m^3/h the Mechanical group shall be consulted.

m^3/hr	m
Up to 45.4	0.76
45.4-227.1	0.91
227.1-2271	1.07
2271-4542	1.37

The suction pipe for fluids at or near their bubble point shall be adequately sized if the pressure drop is in the range of 0.01 to 0.06 bar/100m.

9.1.4 Net Positive Suction Head Available (NPSHA)

NPSHA is calculated by deducting the vapour pressure of the fluid at pumping conditions from the Suction Pressure and converting it to pressure head in terms of liquid column.

Process engineers shall include a Safety Margin of 1.0 m in the NPSH calculated for:-

a) All boiling point fluids either single or multi-component.

b) Fluids that contain dissolved gas.

c) Foaming fluids.

In the case of boiler feedwater pumps, a margin of 2.0 m shall be used.

The static head used in calculating the NPSH shall be taken from either the tangent line or bottom invert line in the suction vessel to one of the following:

- The centre line of a horizontal or rotary pump.
- The suction impeller on a vertical centrifugal pump.

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

Suction line sizing for reciprocating pumps shall take into account acceleration head.

9.1.5 Discharge Calculations

For pump discharge lines when fittings and valve count are not available, a reasonable estimate of the total equivalent length can be made by multiplying the approximate run of actual pipe by the multiplying factor. Details of applicable factors are given in section 10.3.6.

9.1.6 Shutoff Pressure

The shutoff pressure of a typical centrifugal pump is approximately equal to the sum of the maximum suction pressure and 120% of the net differential pressure generated by the pump, based on the maximum anticipated fluid density. Other pumps with steep Head-Flow curves such as turbine, multistage and mixed flow pumps, however, shall have higher shutoff pressures. The process engineer specifying these types of pumps shall consult with the Rotating Equipment Group to determine this value since it may influence the design pressure of downstream equipment.

The maximum discharge pressure sets the design pressure of a pump casing. This is the sum of the maximum suction pressure and maximum differential pressure, which usually occurs at zero flow. In cases where the feed vessel is protected by a safety relief valve, the maximum suction pressure shall be equal to the sum of the safety valve set pressure and the maximum static head.

9.1.7 Equipment Pressure Drops

The following typical pressure drops shall be used in line size calculation when the actual pressure drop data are not available:

TYPICAL EQUIPMENT PRESSURE DROPS (bar)	
Coalescer	0.7
Dessicant Drier	1.0
Desalter	1.7-2.7
Exchangers: S&T, Double-pipe & Air Coolers Box Coolers	 0.35-0.7 3.5
Fixed Bed Reactors	1.4-3.5
Flow Orifice	0.14
Orifice Mixer	0.35/plate
Pump Suction Strainer	0.07
Rotary & Turbine Flow Meters	0.4

For systems involving multiple heat exchangers in series, consult with the Heat Exchanger Group for pressure drop estimation.

9.1.8 Control Valve Pressure Drop

The control valve normal pressure drop is calculated in three ways:

- 33% of frictional pressure drop or,
- 10% of operating pressure or
- The value corresponding to a control valve pressure drop of 0.7 bar at maximum flow.

The maximum of these three values is inputted into the calculation for the net design discharge pressure.

For systems operating above 69 barg, the control valve may take less than 10% of the operating pressure, depending on process and control considerations.

The pressure drop of a control valve on the discharge of a pump should be a minimum of 20% of the system dynamic pressure loss at normal flowrate, or 0.7 bar, whichever is greater. (This criteria does not apply to loading pumps).

9.1.9 Pump Cooling Water Requirement

Cooling water, preferably fresh, is used to cool bearings, stuffing boxes, pedestals and glands to safe temperature levels.

The coolant rate varies with temperature and to some extent with pump size. For design purposes, the following rates shall be used:

Casing Design Temp (°C)	m^3/h
120	0.25-0.75
120-250	0.75-1.5
250+	1.5-2.5

9.1.10 Pump Efficiency

The efficiency of centrifugal pumps varies from about 20% for low capacity pumps (less than 6.3 m^3/h) to a high of almost 90% for certain large capacity pumps. Low head pumps using open type impellers are less efficient than closed impellers. For pump efficiency estimation see Appendix, Figure 1.

9.2 Compressors, Fans & Blowers

International Standards are utilised to identify tolerances for rotating equipment, such as the API 600 series. In addition the following criteria shall be applied.

- Normally no margin is taken if the flow is constant, a 10% margin can be used if the flow is directly coming from a production separator to take into account slugging regime.
- The variations of gas compositions, molecular weight, specific heat ratio etc., and the operating conditions (mainly suction pressure and temperature) shall be taken into account to determine the sizing case, and shall be listed on the Process Data Sheet.

9.2.1 Compressor Process Specifications

9.2.1.1 Operating Case

If more than one case exists, all these alternative cases shall be included in the specification so that the compressor vendor is able to evaluate the most stringent case for design.

9.2.1.2 Capacity

The volumetric flowrate capacity shall be determined by the compressor manufacturer from the data sheet provided. The process engineer shall determine the mass flowrate based on minimum, normal and maximum flow conditions. Design margins are as set out in Section 9.2.

9.2.1.3 Suction Temperature

Suction temperature is to be accurately specified since it is directly related to the volume of gas at suction conditions, the discharge temperature, and the horsepower requirements. It is important for the vendor to know the minimum and maximum temperatures for proper compressor design and selection of correct driver rating.

9.2.1.4 Suction Pressure

Suction pressure is the pressure at the suction flange of the compressor and not before filters, pulsation dampers, etc. The suction pressure shall be accurately specified.

9.2.1.5 Molecular Weight

Molecular weight is an important consideration in the design of a centrifugal compressor. When this or any type of compressor is to be used in multiple services, the vendor is to be supplied with data on the molecular weight of the gases in each of these services.

9.2.1.6 Specific Heat Ratio

The specific heat ratio is also an important consideration in the design of centrifugal and reciprocating compressor as it affects both power and efficiency of the machines. It shall be clearly documented what the basis for the stated specific heat ratio e.g. ideal or polytropic etc.

9.2.1.7 Compressor Power Estimation

Compressor power estimates shall include gear losses. When a compressor is to be used in vacuum or refrigeration service, peak driver load may be required during start-up and a footnote to this effect is to be added to the specification form. The final determination of compressor power requirements and discharge gas temperatures is part of the vendor's responsibility.

9.2.1.8 Gas Composition

This is to be supplied by the process engineer and is to be expressed on a wet basis if the gas contains moisture.

9.2.1.9 Discharge Temperature (maximum allowable)

This is to be supplied by the process engineer when a known process limitation exists. Discharge temperatures are limited by gas reactions, eg. polymerisation or in the case of air compressors with the lube oil, safe lubrication temperatures. Some compressors are limited by mechanical considerations and these shall be defined by the Mechanical Equipment Group and the compressor vendor.

9.2.1.10 Corrosive Compounds

Corrosive compounds in the gas (such as sulphur oxides, hydrogen sulphides, acidic compounds, chlorides, etc.), shall be specified by the process engineer as these may determine the selection of materials by materials group or the compressor manufacturer.

9.2.1.11 Start-up considerations

Start-up methods shall be considered by the process engineer since items such as anti-surge control systems, bypass lines, valve lifters and pockets on reciprocators, etc., are involved. In addition, compressors generally require a running-in period during which time an alternative feed gas may be used. If air is to be used for running-in, then suitable vents, etc. may be an additional requirement.

9.2.1.12 Compressor Selection and Comparison

Centrifugal compressors are the preferred type for the majority of applications.

Reciprocating compressors shall be considered for conditions of low flow, high differential pressures, intermittent loads, varying gas densities, and varying discharge pressures, combined with moderate temperatures.

Screw compressors shall be employed for applications involving relatively low flows and differential pressures. Their selection shall be referred to the rotating equipment specialists.

9.2.1.13 Safety Considerations

The following potential hazards shall be considered for compressor installations.

a) At high pressures, many reactions proceed at higher rates, e.g. the reaction between a hydrocarbon lube oil and oxygen or air. The discharge temperature of air from reciprocating compressors is generally limited to about 149-166°C. Compressor circuits frequently have automatic shutdown instrumentation, which operates on high gas discharge temperature.

b) Excessive discharge pressures from positive displacement machines can be attained if a discharge valve is inadvertently closed. Therefore, safety valves are mandatory for this class of compressors.

c) Adequate ventilation of the compressor house shall be provided when compressing toxic or flammable gases. This is frequently accomplished by omitting the siding from a portion of the compressor house.

d) Adequate inlet KO drums shall be provided where necessary to prevent liquid slugs from damaging compressors. Providing demisters in the KO drum can reduce entrainment.

e) Rotating compressors and their drivers have speed limitations. Trip-outs are indicated and these are usually supplied by the vendor and specified by the Mechanical Equipment Section.

9.2.1.14 Bearing and Seal Losses

The polytropic horsepower absorbed by the gas compression phase does not include additional power, which is required for bearing and seal losses.

The combined losses shall be estimated from the table below and shall be added to the polytropic power requirement.

Polytropic Power, kW	Power Loss, kW
Up to 4500	19
Above 4500	38

9.2.1.15 Gear Losses

The mechanical efficiency of gears used to transmit power from a driver to a compressor varies as follows:

Type of Gear	Mech. Efficiency %	Gear Loss
Single Reduction	98-98.5	2 – 1.5 %
Double Reduction	97-97.5	3 – 2.5 %
Triple Reduction	96-96.5	4 – 3.5 %

9.3 Heater and Boiler

The design margins to be applied are as follows:

- Fired heaters and furnaces : 10% on design duty
- Boilers : 10% on design flow rate

9.4 Vessel

9.4.1 Overdesign Factor

- First separation equipment (plant inlet) : 10% on inlet gas flow rate
- Other drums : 0% unless specific requirements
- Fractionation column : 0% unless specific requirements

9.4.2 Vapour Area Sizing

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

- If internals are installed, the common vapour internal shall be a wire mesh but for some services a vane pack can be used with COMPANY approval.
- The use of other vapour internals such as cyclones, etc. Requires COMPANY approval.

- The basis of sizing is the critical velocity V_c (m/s)

$$V_C = 0.048\left(\frac{\rho_l - \rho_g}{\rho_g}\right)^{0.5}$$

ρ_l = liquid density in kg/m^3

ρ_g = vapour density in kg/m^3

V_c = critical velocity in m/sec

The maximum gas velocity is K^*V_c

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

- 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 shall be confirmed with the vendor.
- For horizontal vessels without vapour internal (wire mesh, vane pack,…..), the minimum distance between the top of the vessel and the LSHH (level switch high high alarm) set point is the largest of 300 mm or 0.2 internal diameter.
- Vessels handling paraffinic oil shall not be equipped with gas internals
- The above separation criteria do not apply to slug-catchers which are not vessels and are indeed a coarse Vapour-Liquid Separator. (Refer to section 9.4.3).

A high efficiency inlet distributor can be considered to improve gas/liquid separation provided that EPCC Contractor verify pressure drop through distributor and dimensions between inlet distributor/mesh and inlet distributor/High Liquid (HLA-high liquid alarm point).

9.4.3 Hold-up and Residence Time of Liquids

- If the vessel is sized to receive a slug, that slug volume shall be taken between Normal Liquid Level and High Liquid Level.
- The residence time corresponds to half of the hold up time, the Normal Liquid Level (NLL) being set at 50% of the High and Low Liquid Level range. Exceptions shall be specified on data sheet.
- The minimum liquid hold up time between Low Level Alarm and High Level Alarm are as follows:

SERVICES	TIME (MINUTES)
Feed Surge Drum	
A. to heater	5
B. to others	3 without pump
	5 with pump
Reflux Drum	5
Fractionation tower bottom : the largest of	
A. product to next process	5
B. product to other column	5
C. product to storage tank	3 without pump
	5 with pump
Steam flash drum (process units)	5
Steam drum (utility generation and sulphur recovery unit)	10
Desalter	15
Deaerator (note 1)	15
Atmospheric degassing drum	15
Others Drums	3 without pump
	5 with pump

Note 1: Liquid hold up time is based on one deaerator shutdown associated with the normal liquid flowrate.

Note 2: The above criteria apply generally to all vessels (horizontal or vertical) where the liquid volume is one of the controlling cases for sizing.

- Two phase separators

ISA SYMBOL	CORRESPONDING DATA SHEET SYMBOL	VERTICAL DRUM	HORIZONTAL DRUM
LAHH/LSHH	HHLA/HHLS (HHL)		
		At least 1 to 2 min. with 150 mm minimum To verify : minimum 10% of control range (1) If only HHL:HLA-HHL: 10% of control range.	At least 1 to 2 min with 100 mm minimum To verify : minimum 10% of control range (1) If only HHL:HLA-HHL: 10% of control range.
LAH	HLA		
		Liquid hold up time to be considered with 300 mm minimum	Liquid hold up time to be considered with 300 mm minimum
LAL	LLA		
		At least 1 to 2 min. with 200 mm minimum To verify: minimum 10% of control range (1) If only LLL:LLA-LLL: 10% of control range.	At least 1 to 2 min. with 100 mm minimum To verify : minimum 10% of control range (1) If only LLL:LLA-LLL: 10% of control range.
LALL/LSLL	LLLA/LLLS (LLL)		
		300 mm minimum, but to be compatible with time required to close a Shutdown valve (SDV)	150 mm minimum, but to be compatible with time required to close a Shutdown valve (SDV)
Tangent line (Vertical Drum) Bottom (Horizontal Drum)			

(1) Control range is the vertical distance between the high level alarm (HLA) and the low level alarm (LLA) with the normal control liquid level set point (NLL) set usually at 50% of the HHL-LLL distance.

- Three phase separators

ISA SYMBOL	DATA SHEET SYMBOL	VERTICAL DRUM	HORIZONTAL DRUM
LAHH/LSHH	HHLA/HHLS (HHL)		
		At least 1 to 2 min. with 150 mm minimum To verify : minimum 10% of control range (1) If only HHL:HLA-HHL: 10% of control range.	At least 1 to 2 min with 100 mm minimum To verify : minimum 10% of control range (1) If only HHL:HLA-HHL: 10% of control range.
LAH	HLA		
		Lightest density liquid hold up time to be considered with 200 mm minimum	Lightest density liquid hold up time to be considered with 200 mm minimum
LAL	LLA		
		At least 1 to 2 min. with 200 mm minimum To verify : minimum 10% of control range (1) If only LLL:LLA-LLL: 10% of control range.	At least 1 to 2 min. with 100 mm minimum To verify : minimum 10% of control range (1) If only LLL:LLA-LLL: 10% of control range.
LALL/LSLL	LLLA/LLLS (LLL)	450 mm minimum	450 mm minimum
LDAHH	HHIA (HIL)		
		At least 1 to 2 min. with 150 mm minimum To verify : minimum 10% of control range (1) If only HHL:HLA-HHL: 10% of control range.	At least 1 to 2 min. with 100 mm minimum To verify : minimum 10% of control range (1) If only HHL:HLA-HHL: 10% of control range.
LDAH	HIA		
		Highest density liquid hold up time to be considered with 200 mm minimum	Highest density liquid hold up time to be considered with 200 mm minimum
LDAL	LIA		
		At least 1 to 2 min. with 200 mm minimum To verify : minimum 10% of control range (1) If only LLL:LLA-LLL: 10% of control range.	At least 1 to 2 min. with 100 mm minimum To verify : minimum 10% of control range (1) If only LLL:LLA-LLL: 10% of control range.

LDALL	LLIA(LLIS)		
		300 mm minimum, but to be compatible with time required to close a Shutdown Valve (SDV)	150 mm minimum, but to be compatible with time required to close a Shutdown Valve (SDV)
Tangent line (Vertical Drum) / Bottom of Vessel (Horizontal Drum) (2)			

1) Control range is the vertical distance between the high level alarm (HLA) and the low level alarm (LLA) with the normal control liquid level set point (NLL) set usually at 50% of the HHL-LLL distance. This definition also applies to each discrete and separated liquid phase for three phase separators.

2) Exception is made for vertical vessel with negligible liquid on clean service with manual or on/off liquid outlet valve; in that case volume of the hemi-spherical head can be used : Low Low Liquid Level (LLL) (or LLLA/LLLS) location to be still compatible with Shutdown Valve (SDV) or control valve closing time.

3) When applicable, the hold up time below the low low liquid level (LLLA or ILLLA) has to be compatible with the time required to close a Shutdown Valve (SDV) or to trip the pump(s) taking suction from the vessel.

4) Stand pipe shall be installed on clean service when at least 3 level instruments are required to be installed (independently from level instrument required for safety actions) e.g.: one level transmitter with two level gauges.

5) Minimum size for stand pipe: 3”

6) Particular case: slug catcher: stand pipe shall be installed.

7) Gauge glasses and level controller shall cover the full range of level transmitters and alarm switches.

8) Connections for level instruments generating a trip function shall be independent from control function.

General notes for Three Phase Separator:

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

1. The effective retention volume in a vessel is that 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.

2. The highest density liquid retention volume is taken between the bottom for horizontal vessels and bottom tangent line (for vertical vessels) and the normal interface liquid level (INLL).

3. The lightest density liquid retention volume is taken between the INLL and the normal liquid level (NLL)

9.4.4 Diameter

- As a general rule, inside diameter shall be specified on process data sheets (in mm)
- If the required inside diameter for a vessel is lower than 800 mm, a note shall be added specifying that a piping element is acceptable.
- For vessels less than 1000 mm ID, flanged heads ***may*** be specified.
- Recommended L/D ratio for horizontal vessel:

PRESSURE (barg)	**L/D**
Lower than 17	3
17 up to 34	4
Higher than 34	5

9.4.5 Manholes

Size of manholes

- For vessel diameter < 1000 mm
- Flanged vessel shall be considered if equipment contains internals
- Otherwise, size of manholes = 18”
- For vessel diameter ≥ 1000 mm
- Toxic service size of manholes = 24”
- Non-toxic service size of manholes = 20” (or up to 24” if internals need to be removable through manhole.)

Location of manholes

- At the opposite side of the utility connection for horizontal vessel

Number of manholes

- Vessel
- For vessel length/height less than 6 m a single manhole shall be provided. For other vessel (length/height > 6m), two manholes to be provided at least ; one manhole each 6 m for longer / higher vessel. If vessel is equipped with internals (baffles etc.) , one manhole to be provided on each compartment.
- Trayed column
- Manhole shall be provided at the top, below the bottom tray, at the feed tray, at any other tray at which removable internals are located, and at intermediate points so that the maximum spacing of manholes does not exceed 15 trays. Tray spacing with manholes in the internal shall be at least 900 mm.

9.4.6 Handhole

Handhole size = 8”. Handhole to be installed on vessel with diameter less than 800 mm or on vessel where severe fouling of internals is expected.

9.4.7 Vortex Breaker

Vortex breaker to be installed for the following services:

- Pump suction
- Outlet to thermosiphon or kettle reboilers
- Letdown to a low pressure system

A vortex breaker in fouled/dirty service shall have a standoff of 150 mm from vessel wall /bottom.

9.4.8 Drains, Vents and Overflow Connections

- Location

  The drain of the vessel shall always be at the lowest point of a vessel. For vertical vessels they shall be connected to the bottom outlet line at the low point. For horizontal vessels the drain point shall be directly on the bottom of the drum at the lowest point ensured through vessel slope (1:100).

- Vent and drain diameter shall be defined as follows :

Volume or diameter of vessel (m^3 or mm)	Vent diameter (minimum)	Drain diameter (minimum)
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”

- Drain number :

  For horizontal drums having a length greater than 6 m TL to TL, additional drain connections shall be considered. An additional drain is also required on each compartment of a vessel. On toxic service, an open drain connection (washing out) is to be provided with a blind flange. Size of open drain connection shall be of the same diameter as drain connection.

- Overflow :

  For vessels equipped with overflow connections, the overflow nozzle line size shall be at least one size greater than the inlet/outlet nozzle, whichever is greater.

9.4.9 Utilities Connections (steam out, purging)

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

- Drums and heat exchangers (when applicable): 2”
- For large vertical drums, two 2” connections shall be provided for diameter >= 4.5m
- For horizontal vessel with a length >= 6m and operating in toxic service, two 2” connections shall be provided.
- If vessel is equipped with internals (baffle), one 2” connection shall be provided on each compartment
- Columns: as follows with regard to column diameter, D (m)

  1. D ≤ 4 : 2”
  2. 4 < D ≤ 5.5 : 3”
  3. D > 5.5 : 4”

Utility connections, when specifically required, are not necessarily located on vessels (advantage may be taken to use connection on drain to steam out / nitrogen purge the vessel) but should remain operational even when the vessel is isolated.

No hard piping connection for steam out / nitrogen purge shall be provided.

9.4.10 Elevation of Equipment

As a general rule for a vessel containing a liquid at its boiling point, a minimum elevation of 5000 mm shall be specified when supplying a centrifugal pump. The elevation shall be updated when NPSH requirements are defined with rotating equipment specialist.

If there is no process requirement regarding the elevation, a note on PID shall be indicated "minimum for piping".

9.4.11 Nozzle Sizing

The following criteria for vessel and column nozzles design shall be used:

Inlet line:

- $\rho * v^2$ max = 1500 if no inlet device is foreseen
- $\rho * v^2$ max = 3000 if half pipe or baffle inlet device is foreseen
- $\rho * v^2$ max = 6000 if Schoepentoeter or other vane pack inlet device is foreseen

Outlet line:

The same criteria which are used for line sizing (see below paragraph 10.3) shall be used.

9.5 Heat Exchangers and Air Coolers

9.5.1 Oversizing

- Shell and tube heat exchangers and air coolers: 10% on surface based on design duty.

9.5.2 Fouling Factors

The following gives some indicative fouling factors for process and utility fluids which can be reviewed case by case. They can be applied to items such as electric motor cooling and used to check vendor's data.

▪ <u>Process fluids</u>	$m^2.°C/W$
Acid gas	0.00020
Sour natural gas from slug catcher	0.00035

Sour natural gas downstream unit 100	0.00017
Sweet gas	0.00017
Liquid LPG	0.00020
Raw Feed Condensate from slug catcher	0.00052
Raw Feed Condensate	0.00035
Stabilised Condensate	0.00020
Process water	0.00035
Stripped water	0.00030
Glycol	0.00040
Refrigerant (propane)	0.00015

- Licensed units fluids — by Licensor

- Utilities

Sea Water	0.00050
Chilled water	0.00020
Potable water	0.00020
Saturated steam/LP condensate	0.00017
BFW/Demineralised water	0.00017
Nitrogen	0.00017
Instruments air	0.00017
Fuel gas	0.00017
Diesel	Light:0.00030 Heavy 0.00035

9.5.3 Fouling Factors for Plate Exchangers

For Plate Frame Heat Exchangers, a general fouling factor of 0.00005 m^2 °C/W shall be taken for all fluids (or Process Licensor recommendation).

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

9.5.4 Temperature Approach for Heat Exchangers

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

- 5°C for TEMA type heat exchangers (shell and tube)
- 10°C for air coolers
- For plate type heat exchangers (PHE) and printed circuit heat exchangers (PCHE) the vendor shall confirm and specify the requirement on a case by case basis.
- 3°C for kettle type

9.5.5 Specific Requirements for Heat Exchangers

TEMA R shall be generally used for all shell and tubes and air fin type heat exchangers. Fixed tube sheet exchangers are acceptable for non fouling service on the shell side. In this case Licensor or EPCC Contractor shall define all exceptional operating conditions (start-up, shutdown) to check the necessity to provide an expansion bellow on the shell.

9.5.6 Air Cooler Type

The air cooler shall be of an induced type when the air cooler is installed on pipe rack. Forced draft type shall be considered:

- If air cooler is installed at grade
- If inlet process temperature is above 175°C or calculated air outlet temperature is over 93°C
- Several sections stacked (in the case of multiple service exchangers).

When control of process side temperature is required then the control method shall be defined on a case by case basis.

In all cases design margins for air cooler fans shall be specified.

9.5.7 Allowable Pressure Drop- Shell and Tubes

Typical allowable pressure drops are given below:

9.5.7.1 Liquids

Total Pressure Drop (bar)

-Shells in series-

Viscosity (cp). (at average temperature)	Total Pressure Drop (bar) -Shells in series- One	Two	Three
Less than 1.0	0.35 to 0.7	0.35 to 0.7	0.7 to 1.0
1.0 –5.0	0.7	1.0	1.0 to 1.4
5 – 10	1.0	1.0 to 1.4	1.4
Above 10	1.4	1.4 to 2.0	2.0

Notes:

1) Under the following circumstances, ∆P's approaching the higher recommended values shall be employed: when the ∆T is small (28°C or less) or when the temperature range is large, (above 111°C).
2) Calculated tube side pressure drop values are subject to greater variation than shell side values, because of the nature of tube bundle construction.
3) It must be realised that little can be gained by specifying increased pressure drop for one fluid in an exchanger when the other fluid has a significantly lower film coefficient.
4) For gravity flow, the pressure drop is usually limited from 0.07 to 0.14 bar.

9.5.7.2 Gases

Operating Pressure (barg)	Pressure Drop (bar)
0 – 0.7	0.035 – 0.07
Above 0.7	0.14 – 0.35

9.5.7.3 Condensers

Types	Pressure Drop (bar)
Partial	0.14 – 0.35
Total	Negligible

9.5.7.4 Reboilers

Types	Pressure Drop (bar)
Kettle	Negligible (shell side)
Thermosyphon:	
- Horizontal	0.02 – 0.035
- Vertical	Equivalent to approx. 3 – 5m tube length

9.5.8 Allowable Pressure Drop- Air Fin Exchangers

Suggested pressure drop for various services are given below. However, care should be taken to ensure that the selected pressure drop results in the most economic overall installation.

The allowable pressure drop for product cooling and non-critical services shall not control the size of the exchanger, as this may result in an uneconomic design which could be avoided by reconsidering the hydraulics of the process circuit. Special consideration is required for wide temperature range cooling of viscous liquids, low pressure gases or condensation of vapours at very low pressures. In these services, pressure drop is a critical requirement, which greatly influences the size of the heat transfer surface.

ALLOWABLE PRESSURE DROP		
Service	**Allowable pressure drop (bar)**	
Liquid Cooling	0.7	Note 1
Gas Cooling:		
Operating pressure 1.0 to 3.5 barg	0.2	
Operating pressure 3.5 to 17.5 barg	0.35 to 0.7	
Condensing: (atmospheric pressure and above)		Note 2
Total condensation	0.035 min	Note 3
Partial condensation	0.14 to 0.35	Note 3

Notes:

1) Not valid for viscous fluids.

2) For vacuum service the selection of an allowable pressure drop should be from the results of an economic study. Pressure drops are usually in the range of 3-5 mmHg.

3) For multi-pass air coolers high pressure drops assure proper flow distribution. The higher pressure drop shall also assure proper distribution at lower than design throughput.

9.6 Columns and Trays

Towers shall be sized based on flows that are 110% of the respective material balance figures to allow for any vagaries in the equations of state, operational control around the material balance duty point and effects of fouling etc. Tray loadings used for sizing should be the vapour to the respective tray and the liquid leaving it.

10. PIPING

10.1 General Design and Hydraulic

The guidelines and process sizing criteria detailed below shall be implemented in conjunction with the latest COMPANY approved Piping Specification, e.g. use of non standard pipe sizes.

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

Line size of each line shall be firstly selected based on the mass flow rate and in accordance with the velocity range criteria and then be checked in accordance with the friction loss range criteria as given in paragraph 10.3 Line Sizing Criteria.

10.1.2 Minimum piping sizes

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

- 3/4” for pipe when located above ground
- 2” for process line on pipe rack
- 2” for utility lines on pipe rack
- 2” for pipe on pipe sleeper
- 2” for underground steel pipe
- 2” for underground non metallic piping

10.1.3 Pump suction calculations and NPSHA

For the NPSHA calculation method refer to Section 9.1.4. For the value of NPSHA specified on the process data sheet, the referenced elevation shall be indicated (e.g. grade, pump, centreline, etc.)

With pumps in parallel and/or spare pumps, consideration shall be given to ensure that the common suction line leading to the individual pump suctions are sized adequately to cater for the additional flow imposed whether it is for a short or extended duration.

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

Pump suction line shall not be smaller than suction nozzle and shall be at least the same diameter as the line. If ball valves are appropriate for the service then 'full bore' (FB) ball valves shall be used for all valves on the suction and marked on the P&ID.

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

10.1.4 Control valve

- Control Valve pressure drops are defined in section 9.1.8.
- In case of mal-operation, the gas blow-by calculation shall consider the flow rate through the control valve when fully open and also through its by-pass when fully open, where installed. If the calculated flowrate oversizes the flare, the manual by-pass could be removed or a mechanical interlock between the associated manual block valves could be installed.
- For the control valve arrangement, refer to Design Basis for P&ID development DB-1718-999-P312-0203.

10.1.5 Piping, Vents and Drains

Pipe Size (ins)	Vent Size (ins)	Drain Size (ins)
¾ to 8"	¾"	¾"
10" and 12"	1½"	1½"
14" and over	2"	2"

Air Coolers

On air coolers one 2" vent shall be located at the highest point in the inlet header and one 2" drain at the lowest point in the outlet header. The exact location of these vents and drains is dependent on the actual cooler design. Connections shall be valved and blanked off.

Pump Casings

For non-operating vents and drains, provide ¾" valved and blanked off vents and drain (not shown on P&ID). EPCC Contractor shall provide drain connections for pump casings, with the drain piped to the edge of the baseplate by the pump vendor.

For non-volatile services, casing vents and pump drains shall be valved and piped to pump baseplate or into a sewer

For volatile services, casing vents and drains shall be piped to relief header and sewers via a sample cooler.

For details of pump vents and drains see PID-1718-999-0030-0006.

Additional Notes

- Valved / blanked off vent and drain connection shall be furnished on all equipment that is not self-venting or self-draining. Connection shall be located on equipment, if practical, but may be located on connected piping when there are no valves or blocks between the vent or drain connections and the equipment.
- Hydrostatic vents and drains shall be provided by EPCC Contractor, and shall not be shown on P&ID.
- For total condensing service in shell and tube exchangers, EPCC Contractor to provide a flanged 2" vent nozzle on shell at opposite end to shell inlet.
- At relief valves, a ¾" valved blanked off bleed shall be shown between safety valve and any block valve on inlet and discharge side.
- Vents from vessels that may chill and freeze during depressuring shall have double block valves separated by at least 900mm.

10.2 Insulation and Heat Tracing

Thermal insulation for hot or cold services may be required 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 thermal insulation if it is not required for process reasons,
- To avoid external water condensation or ice,
- Steam or electrical heat tracing.

In all cases, insulation shall be minimised in order to limit CUI (Corrosion Under Insulation).

10.3 Line Sizing Criteria

This paragraph shall not be applied to the flare lines. The pressure drop and velocity guidelines provided may be used for the preliminary sizing of lines. However, the final sizing shall also take into account other factors, such as pump NPSH requirements, pressure drops available, and specific process requirements. Where specific maximum velocity limits are given these shall not be exceeded.

10.3.1 Line sizing criteria for gases and steam

Vapour and steam lines	ρv^2 max. ($kg.m^{-1}.s^{-2}$)	Max. Velocity (m/s)	DP (bar/km)	
			Normal	Maxi
- Continuous operation				
P <= 20 bar g	6000		)	
20 < P <= 50 bar g	7500		)	
50 < P < =80 bar g	10000		) Pressure drop must be	
80 < P <= 120 bar g	15000		) considered compatible	
P > 120 bar g	20000		) with corresponding service	
- Compressor suction	Compatible with above		0.2	0.7
- Compressor discharge	Compatible with above		0.45	1.15
- Discontinuous operation				
P <= 50 bar g	10000		)	
50 < P <= 80 bar g	15000		) Pressure drop must be	
P > 80 bar g	25000		) considered compatible	
- Column overhead	15000 (high pressure columns)		) with corresponding service	
- Stripper vapor return			0.2	0.45
- Kettle vapor return			0.2	0.4
Steam lines				
- P <= 10 bar g				
Short line (L <= 200 m)	15000		0.5	1.0
Long line (L > 200 m)	15000		0.15	0.25
- 10 < P <= 30 bar g				
Short line (L <= 200 m)	15000	42	1.2	2.3
Long line (L > 200 m)	15000	42	0.25	1.0
- P > 30 bar g				
Short line (L <= 200 m)	15000	30	1.2	2.3
Long line (L > 200 m)	15000	30	0.35	1.0
Vacuum (<0.2 bara)			0.001	0.002

10.3.2 Line sizing criteria for liquids

Liquid line type	ΔP (bar/km)		Max. Velocity. (m/s) (2)			
	Norm.	**Max.**	**To 2”**	**3” to 6”**	**8” to 18”**	**from 20”**
Pump suction						
- Liquid at bubble point 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 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 (1)						
- Disch. pres. <= 50 bar g	3.5	4.5	1.5 to 4.5 m/s			6.0
- Disch. pres. > 50 bar g	7.0	9.0	1.5 to 4.5 m/s			6.0
Column outlet	0.6	0.9	0.6	0.9	0.9	0.9
Gravity flow	0.25	0.45	0.6	0.6	0.6	0.6
Water lines (CS)(3)						
- Cooling water & service water (4)						
Large feeders between pumps and units	1.5		1.5 to 3.0 m/s			
Unit lines (long)		1.5	1.5	2.5	3.0	3.0
Unit lines (short)		3.5	1.5	2.5	3.0	3.0
- Boiler feed						
Pres. <= 50 bar g	3.5	4.5	1.5 to 4.5 m/s			6.0
Pres. > 50 bar g	7.0	9.0	1.5 to 4.5 m/s			6.0
-Sea water lines			2.5 to 3.5 m/s (2 m/s mini)			
- Steam cond. return			1 to 1.5 m/s			
- Reboiler feed (for indication)	0.2	0.4				

Notes:

1) 3.0 m/s maximum (2 m/s average) at storage tank inlet or in loading.
2) Vendor and/or Licensor requirements could supersede maximum velocity values upon COMPANY approval.
3) Special considerations can be applied for copper-nickel or glass reinforced plastic piping upon COMPANY approval.
4) Velocities below 1 m/s shall not be used for cooling water service to avoid solids deposition.
5) For amine service velocity should not exceed 1 m/s to avoid corrosion/erosion.
6) For lines containing mixtures of hydrocarbon and water, velocity should be limited to 1 m/s to avoid generation of static.
7) 60 to 98% sulphuric acid lines velocity should not exceed 1.2 m/s to avoid corrosion.

10.3.3 Line sizing criteria for two phase flow

For preliminary mixed phase fluid line size calculations, the average density method shall be used while considering the following criteria:

- V_m : 10 to 23 m/s
- $\rho_m V_m^2$: 5000 to 10000 Pa
- $\rho_m V_m^3$: 100 000 to 150 000 kg/s^3

Where:

- $\rho_m = W / ((W_L / \rho_L) + (W_V / \rho_V))$ in kg/m^3

  $W = W_L + W_V$ = Total rate in kg/h ρ_L = liquid density in kg/m^3

  W_L = liquid flow rate in kg/h ρ_V = vapour density in kg/m3

  W_V = vapour flow rate in kg/h

And the apparent fluid velocity V_m expressed as:

- $V_m = 4W / (3600)\ \rho_m\ \pi.D^2)$ in m/s

  D = internal diameter of the line in m

In general, continuous flow patterns should be ensured such as:

- Stratified, annular, bubble, wavy flow patterns, etc. For horizontal lines or slightly sloped.
- Annular or bubble flow, etc. For the vertical lines
- For horizontal lines in slug and plug flow regimes and for vertical line in slug flow regimes reinforced anchoring shall be specified.

10.3.4 Line sizing criteria for offsite line

The following criteria are typical and shall have to be supported by economic appraisal.

LINE TYPE	DP (bar/100m)		Max. velocity (m/s)
	Normal	Maximum	
Long Carbon steel water line	0.058	0.116	-
Bonna concrete pipe	-	-	2.5 to 3
Steam condensate (mixture)	0.02 to 0.03	-	-

10.3.5 Corrosion/Erosion Criteria

Corrosion

For corrosion resistant material (SS, Special alloys…), no limitation of flowing velocity up to 100 m/s and no requirement for corrosion allowance.

For non corrosion resistant material, in corrosive fluid service, a corrosion allowance for the design service life and corrosion inhibitor injection are required. The flowing velocity is limited by the inhibitor film integrity. The process designer shall consult the project material and corrosion specialist who shall be responsible for implementing COMPANY approved guidelines. Refer to the latest revision of the "Corrosion and Material Selection Review Onshore Facilities Process and Utilities" (Doc. No. RP-1718-171-6300-0002).

Erosion

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

- 100 m/s in 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 ensures a lubricating effect which drifts the erosion velocity limit. The corrosion inhibitor erosion velocity limit shall be calculated taking into account the inhibitor film wall shear stress.
- In case of uninhibited fluid, the API RP 14 E recommendation shall apply : the flowing velocity must be maintained below the erosional limit :

$$V_e = C / (\rho_m)^{0.5}$$

With : V_e erosional velocity in m/s

ρ_m gas / liquid mixture density at flowing conditions in kg/m^3

The empirical constant 'C' is equal to 183 to 207. C values of up to 245 can be considered on peak flow rates only in case of absence of abrasive (solid) particles such as sand. When solid and/or corrosive contaminants are present C values shall not be higher than 122 in SI units.

10.3.6 Equivalent Line Length Calculation

The total equivalent length (Le) can be calculated using a factor multiplied by the straight length of pipe or by adding up the equivalent length of pipe fittings and the straight length of pipe. This method shall be used when pipe routing has not been finalised / defined.

TABLE OF FITTING FACTORS

Line sizes, diameter	Approximate line length , ft		
	100	200	500
	Multiplying Factor		
3in or less	1.9	1.6	1.2
4in	2.2	1.8	1.3
6in	2.7	2.1	1.4
8in or over	3.4	2.4	1.6

10.4 Absolute Roughness

Values for absolute roughness for commonly used materials, which are used in liquid and vapour line sizing calculations, are as follows:

ABSOLUTE ROUGHNESS, mm

Material	Roughness mm (Inches)
Carbon Steel	0.05 (0.0018)
Corroded Carbon Steel (For Flare Lines)	0.5 (0.018)
Stainless Steel, Duplex Steel, (New, Seamless, Cold Drawn)	0.03 (0.0012)
Stainless Steel (Hot Rolled, Longitudinally Welded)	0.05 - 0.1 (0.0019 - 0.0039)
Titanium, (New, Seamless, Cold Drawn)	0.03 (0.0012)
Titanium (Cold Rolled, Longitudinally Welded)	0.05 - 0.1 (0.0019 - 0.0039)
Galvanised Carbon Steel	0.15 (0.0059)
GRP	0.02 (0.0008)
Epoxy Lined Pipe	0.15 (0.0059)

11. FLARE AND COLD VENT SYSTEMS

11.1 Type of Flare Tip

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

When possible a sonic tip shall be preferred. Sonic tip with Coanda effect and/or with variables slots are prohibited.

The flares shall be smokeless. Suitable media (Steam, Air, Fuel Gas) shall be considered for smokeless operation of Flares.

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.

11.2 Flaring Flowrates

For detail, refer to Flares & Blowdown Design Basis Doc. No. DB-1718-140-P312-100.

11.3 Radiation Levels Criteria

The radiation levels criteria shall follow the Basic Engineering Design Data. The minimum relative humidity stated on the basis of design shall be applied.

11.4 Emissivity Coefficient

When the radiation calculations are performed by a flare vendor it is necessary to check carefully the emissivity coefficient used. 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 minimum emissivity coefficient = 0.13 for all gases without liquid carry over, and 0.15 with liquid carry over not exceeding 5% weight.

11.5 Flare and cold vent lines sizing criteria

11.5.1 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)

Inlet line diameter ≥ PSV inlet diameter

- $\rho V^2 \leq$ 25 000 kg/m/s^2 for ∅ of line ≤ 2”
- $\rho V^2 \leq$ 30 000 kg/m/s^2 for P ≤ 50 bar g
- $\rho V^2 \leq$ 50 000 kg/m/s^2 for P > 50 bar g

DEPRESSURISATION DEVICE

- Minimum line size 2”
- ρV^2 criteria are the same as for PSV’s

11.5.2 Line downstream relieving devices

Flare and cold vent headers and sub-headers:

- Minimum line size 2”
- Back pressure to be compatible with the protected equipment
- Velocity and ρV^2 :

SINGLE PHASE (GAS AT THE RELIEF DEVICE INLET) :

- Intermittent flow:
- Lines downstream relieving devices and sub-headers : 0.7 Mach maximum and ρV^2 < 150000 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, lines 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
- MULTIPHASES (2 phase flow at the inlet of relieving device) :
- Velocity ≤ 0.25 Mach and $\rho_m v_m^2 \leq$ 50 000 kg/m/s^2
- For ρ_m and V_m definition see § 10.3.3

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 header and sub-headers.

11.6 Flare Drum Sizing

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

- Remote flare or cold vent offshore : 600 μm
- Vertical flare or cold vent onshore : 600 μm

11.7 Purge Gas

The purge gas is provided to avoid:

- An explosive mixture in the stack or header due to air intake into the flare or cold vent stack.
- The risk of burn back which induces the quickest deterioration of the flare tip.

The purge gas flowrate shall not be lower than the value given by the following equation:

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

Where : Purge gas flow in Sm^3/h

D is the tip internal diameter (1) in m

MW is the purge gas molecular weight in kg/kmol

(1) For sonic flare, the tip internal diameter is taken as the equivalent diameter corresponding to exit gas area

The above formulae shall be applied only to the flare tip, assuming COMPANY/Project procedures do not allow the application of flare tip vendor's guaranteed purge rates.

For purging of sub headers and headers, the purge velocity shall be a minimum of 0.03m/sec as per API RP521 (Section 4.4.3.4.2).

For flare, if 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 preferably be 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.

In some cases, nitrogen could be used as purge gas. In these situations, pilots able to run in a predominantly inert gas environment shall be installed after detailed case by case evaluations with the flare tip vendor.

12. SOFTWARE

Process engineering software that is available for use during the FEED stage:

Software	Description	Publisher
SmartPlant P&ID	An intelligent P&ID design system developed by Intergraph. This system is also integrated with the line list and Instrument data base system.	Intergraph
HYSYS/HYSIM	Process simulation package with extensive component library and a wide selection of property packages/Windows version.	AspenTech
ZyQad	Process engineering database with direct interface to simulation packages. Can be used for the production of PFD's and data sheets.	AspenTech
KYPipe	Water network analysis	University of Kentucky
Sulpack	Column internal design	Sulzer
AFT Fathom	Piping network analysis	AFT
AFT Impulse	Piping network analysis	AFT
FlareNet	Flare piping calculations	AspenTech
PROII/Provision	Process simulation package	Invensys
PipeSim	Piping and network analysis	Baker Jardine
Aspen Plus/Dynamics	Process simulation package	AspenTech
Koch Tower Design	Tower design	Koch Engineering
Korf	Pipe pressure drop calculations	Korf
ESI (PESP)	Equipment /Piping design	Engineering Software
SCond2	Sulphur condenser design	WorleyParsons
SpiraCalc	Steam system calculations	Spirax-sarco Limited
SulSim	Sulphur plant simulation	Bovar-Western Research

Trapselect	Steam system calculations	Spirax-sarco Limited
VIB	Vibration analysis	WorleyParsons
HTFS	Heat exchanger thermal calculations	Hyprotech
Pipephase	Multiphase calculation for steady state or transient conditions	
HTRI	Heat exchanger thermal calculation	

13. APPENDIX

Figure 1- Pump Efficiency vs Flowrate (GPM)